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Design Manual

Municipal Wastewater Disinfection

U.S. Environmental Protection Agency
Office of Research and Development
Water Engineering Research Laboratory
Center for Environmental Research Information
Cincinnati, OH 45268

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Chapter 1
Introduction

1.1 General

Chlorination has long been the accepted and preferred method of disinfection for both water and wastewater in the United States. Major factors in the implementation of chlorination were efficiency and low cost when compared to other means of disinfection. New considerations that are now being examined very closely are the environmental and biota impacts of chlorination versus attainment of public health. The U.S. Environmental Protection Agency recognized the adverse effects of chlorination of wastewaters and reported the following conclusions in its Task Force Report (1):

- disinfection requirements (i.e., the need) should be evaluated on a case-by-case basis with consideration of beneficial uses and criteria.
- chlorine and subsequent residuals are extremely toxic to aquatic wildlife.
- chlorine and organic compounds form chloro-organics and are potentially toxic to man.
  In effect, the policy and summary conclusions have not changed since 1976. The agency policy essentially states:

- disinfection should not be required in those instances where significant benefits are not demonstrated.
- prospective disinfection benefits must be weighed against the environmental risks and costs.
- chlorine should be considered only when there are public health hazards to control.
- alternative disinfection methods and/or dechlorination must be considered when and where public and aquatic health and/or life impacts co-exist.

With the enactment of the Clean Water Act, its amendments and new assertions on site specific criteria, there is no longer a "generic" disinfection procedure. The realization that chlorine is toxic to aquatic life and reacts with precursors of trihalomethanes (THMs) and other chlorine oxidizable and substitutable compounds in wastewater effluents has caused considerable concern. As a result, the existing practices of disinfection have fallen under close scrutiny. Disinfection studies, chemical characterizations, and other evaluations are ongoing even now to assess more accurately the impact of present-day disinfection practices.

That a specific disinfection process has been reported as troublesome at a particular facility could very well be the result of many factors, some or all of which may not have been considered prior to engineering and installation. Critical factors such as disinfection type, tank configuration, contact time, conditioning and pre-conditioning criteria, power costs, maintenance requirements, and others may not have been fully understood or evaluated. Recent information released by the EPA and the General Accounting Office (GAO) has indicated that many of the municipal wastewater treatment plants in the United States have not been meeting their effluent discharge limitations (2). An estimated 18,000 municipal wastewater treatment plants were in operation or under construction by the end of 1979. More than half of these plants were not functioning as designed with respect to BOD, TSS, and fecal coliform removal. Reasons for plant problems were site-specific and complicated with long-term violations determined to be due to a combination of problems, as follows:

- operation and maintenance (O&M) deficiencies.
- equipment problems.
- infiltration and inflow.
- industrial waste overloads.
- design deficiencies.

With respect to design deficiencies, four basic types of design problems that have been implicated include the following:

- limited state-of-the-art during the design phase.
- lack of expertise.
- sampling errors prior to plant design.
- time and funding constraints.

All these problems have been responsible for design deficiencies associated with wastewater disinfection facilities. Limited state-of-the-art during the design phase has been a real problem associated with the design of alternative disinfection processes, such as ozone and ultraviolet light. With the increasing need for wastewater treatment plant upgrading modifications, including more stringent disinfection criteria, elimination of design deficiencies and expanding the state-of-the-art for process design become very important considerations.

The importance and need for a comprehensive design document for municipal wastewater disinfection is apparent. A single source information document does not exist today that will assist and direct the design engineer, and other concerned parties, in an educated and informed decision, selection, and design procedure for the best disinfection alternative process suited to site specific constraints. A major tool that is lacking in the area of disinfection is a concise and unbiased design manual dealing with the disinfection alternatives, and the design of the most effective alternative for the application desired. The purpose of this manual is to bridge this gap and present the engineer with comprehensive design guidance for the implementation of the most efficient and cost effective disinfection method needed for a particular site.

1.2 Purpose and Objectives

Over the past 14 years the EPA has actively pursued an extensive program to investigate alternatives to chlorination of wastewaters through internal and externally funded research and development. This research has produced a wealth of information on disinfection alternatives from literature surveys, laboratory studies, pilot plant studies, and full-scale investigations.

The EPA contractors, grantees, and project officers involved in this extensive effort have developed results and experience in design and operation of the most applicable disinfection alternatives. The EPA disinfection program has advanced to the point where the wealth of information and expertise developed on disinfection alternatives can be compiled into a single document for evaluation of process alternatives and design of the selected disinfection process. The primary objective of this document is to provide a comprehensive process design manual for wastewater disinfection to be used by design engineers, regulatory and review agencies, and owners and operators of disinfection processes. The development of this document has been based on both the theoretical and practical application of process design criteria for the implementation of disinfection technology in municipal treatment facilities. The information in this process design manual for disinfection has been developed from the following sources:

- Literature
- Extensive experience of the individuals preparing the manual from participation in disinfection research and development.
- Active involvement and participation by the EPA project officers involved in the disinfection program.
- Information and results developed from the EPA internal efforts and externally funded research programs.
- Communications with investigators and equipment manufacturers.
- Plant site visits and discussions with operating personnel.

The disinfection alternatives that are the subject of this manual and of which sufficient information exists for design purposes are halogenation/dehalogenation (including chlorine, bromine chloride, and chlorine dioxide), ozonation, and ultraviolet irradiation. In the future the acceptable risks associated with the various disinfection alternatives and the levels of disinfection required may be refined and thus create disinfection criteria other than those in present use. The design approaches presented in this manual are flexible and applicable to the various levels of disinfection presently required as well as those that may be required in the future by covering the range of no detectable coliforms up to any desired level in wastewater effluents.

1.3 Scope

The first part of this manual presents an overview of the disinfection process, the types of disinfecting agents, and the advantages and disadvantages of each. The manual then discusses in separate chapters how to design each of the primary disinfection alternatives. Although many more alternative disinfection methods have been identified, only those considered in this manual are cost effective and presently being implemented. The following criteria were used to select the disinfection processes discussed: technical feasibility, flexibility, reliability, complexity, safety, costs, environmental impacts, and hazardous material impact or formation.

The primary thrust of this manual is to present thorough design guidelines on ozone, ultraviolet light, chlorination, and chlorination/dechlorination facilities. Bromine chloride and chlorine dioxide are discussed in the manual but not to the level of detail as the previous alternatives. Bromine chloride appears to be as flexible and effective as chlorine; however, some questions still remain regarding the use of bromine chloride, such as equipment reliability and future chemical cost. Full-scale operating data and a proven track record for bromine chloride disinfection are still lacking. Chlorine dioxide is a proven bactericide and virucide and has certain features making it attractive for drinking water treatment. However, for wastewater disinfection, chlorine dioxide is not so attractive, principally because of its high cost. Also, its persisting residual is toxic and may require removal through chemical reduction prior to discharge of the disinfected effluent.

Each of the respective chapters presents a brief historical background of the development and use of the particular disinfectant of interest. The chapters then review process chemistry and disinfection kinetics. Analytical measurement methodology is discussed relative to both wet chemistry and instrumentation analysis, including disinfection process control concepts. Then the actual process design factors, considerations, and experience are presented along with case histories, operation and maintenance, and safety considerations.

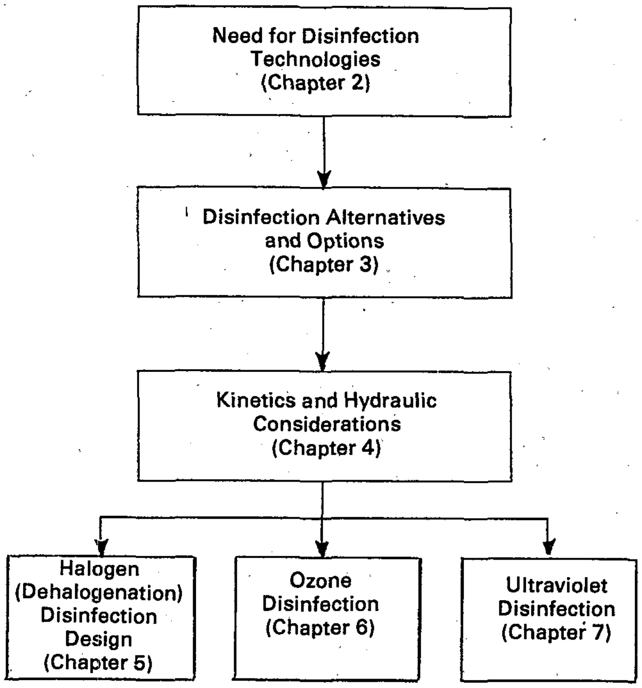
1.4 How to Use this Manual

After discussing the need for wastewater disinfection, along with disinfection criteria in Chapter 2, the manual in Chapter 3 discusses the disinfection alternatives and options considered feasible for municipal wastewater disinfection. A qualitative screening procedure is presented in Chapter 3 for evaluating and selecting an appropriate disinfection technology for a specific application. Examples are presented where each of the primary disinfection technologies is selected. In Chapter 3, the predominant advantages and disadvantages of each of these disinfection alternatives are also discussed.

After a disinfection alternative has been selected for a specific application, the manual user can proceed directly to the appropriate chapter on design of that technology. Halogen disinfection is discussed in Chapter 5, ozone disinfection is discussed in Chapter 6, and UV disinfection is discussed in Chapter 7. However, before the user proceeds to the appropriate design chapter, he may want to review Chapter 4 on disinfection kinetics and hydraulic considerations. Chapter 4 presents a general overview of kinetic considerations, mixing requirements, and contacting requirements that apply to each of the disinfection alternatives. The suggested sequence to follow for use of this disinfection design manual is presented in Figure 1.1.

After the appropriate disinfection technology has been selected for design, the user can proceed through the specific design chapter indicated in Figure 1.1. Each design chapter includes a brief history, overview, and application discussion of the respective disinfection technology. More detail on the fundamental chemical and kinetic aspects of each disinfectant is presented, compared to the general overall discussion in Chapter 4. Specific aspects of mixing and contactor hydraulic considerations are also presented as they relate to the specific disinfection technology under discussion. Following this, each chapter presents an in depth discussion of equipment design considerations, including materials of construction, similarities and differences among available equipment suppliers, and factors to be aware of in specifying equipment. Process design procedures are then presented, detailing the mechanics of how to size and specify equipment based on dose requirements, disinfectant demand, NPDES permit limitations, and cost minimization. The latter section of each chapter is the heart of the manual, but the previous sections provide essential technical back-up for making the proper design decisions.

Figure 1-1. Sequence of Manual's use.



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Chapter 2
Need for Disinfection Technologies

2.1 Need for Disinfection

Population increases and much greater demands for water supply and water recreational uses within the past 15 to 20 years have significantly increased the opportunity for human exposure to wastewaters being discharged into the environment. Natural safeguards, such as dilution and distance or time before contact or use, have been reduced due to the large volumes of wastewater being discharged and the number of discharge locations. Domestic wastewaters carry human pathogens excreted in the fecal discharges of infected individuals. Even treated effluents can affect sources of domestic water supply, recreational waters, and shellfish growing areas. Disinfection is necessary to reduce transmission of infectious diseases when human contact is probable. Chlorine in the past has been used almost universally as the disinfectant for wastewaters. However, studies have shown that chlorine and its by-products can be toxic to aquatic life, repel and deny spawning grounds to anadromous fish, and decimate fish larvae and other forms of life. Out of these concerns, questions have arisen relative to both disinfection needs in general and disinfection with chlorine. Although the need for disinfection is site specific, in general, disinfection is considered to be a beneficial unit process and required for most discharge applications.

The organisms of greatest concern in human exposure to wastewater-contaminated environments are the enteric bacteria and viruses and the intestinal parasites. Diseases that are spread via water consumption and/or contact can be severe and sometimes crippling. Bacterial diseases such as salmonellosis (including typhoid and paratyphoid fevers), cholera, gastroenteritis from enteropathogenic *Escherichia coli*, shigellosis (bacillary dysentery) and viral diseases caused by infectious hepatitis virus, poliovirus, coxsackieviruses A and B, echoviruses, reoviruses, and adenoviruses may be contracted by contact with or by consumption of wastewater contaminated water supplies (i.e., potable and/or recreational).

The alternative of discharging wastewater that has not been disinfected allows discharging of pathogenic organisms and other resultant hazards posed to humans. Pathogenic organisms, by definition, cause disease in human beings. Waterborne transmission of these disease-causing organisms can occur via four pathways(1):

- direct ingestion of untreated water.
- direct ingestion of treated drinking water.
- ingestion of aquatic food species infected with pathogens absorbed from contaminated waters.
- invasion resulting from skin contact with contaminated water.

The first three pathways are sometimes classified as the *fecal-oral* route. The second pathway described above occurs when a drinking water treatment system fails or the integrity of the water distribution system is violated. The fourth pathway is likely to result in skin, mucous membrane, or urinary tract infections but is seldom implicated in gastrointestinal illness in the United States. The risk of disease by exposure to wastewater effluent in recreational water, especially non-disinfected effluent, is not well established on epidemiological grounds; however, recent work has demonstrated a cause-effect relationship via this pathway (2).

Wastewater treatment plants historically discharge their effluents to natural receiving streams that are often tributaries of larger recreational bodies of water or that are used as water supply sources by downstream communities. City potable water supplies are often extracted from these tributaries or lakes, physically and chemically treated and distributed to customers. The only protection that the recreational users receive is the hope that the wastewater was adequately disinfected prior to discharge. So long as disinfection guidelines and standards are being met at the sources, public safety and water quality will be protected. Factors that influence and potentially bias this type of rational thinking are equipment design, operator training, equipment dependability, operator attention, and others.

It is significant that infectious hepatitis has maintained a level of 50,000 to 60,000 cases per year in the United States (3), while typhoid fever dropped from 2000 cases in 1955 to 300 in 1968 (4). There are more than 100 viruses excreted in human feces that have been reportedly found in contaminated water, of which any one could cause a waterborne disease.

A wide variety of enteric pathogens including viruses, bacteria, and parasites is known to occur in all community-derived wastewaters. Infective dose studies with a variety of enteric organisms have been conducted over the past 30 years in human volunteers (5). The widest dose range required to produce a response was found with the bacterial agents. *Salmonella* spp. required the largest dose with the ingestion of 10^5 to 10^8 cells needed to produce a 50 percent illness rate. In contrast, three species of *Shigella* produced illness in a significant percent of subjects dosed with 10 to 100 cells. Protozoan infections have been produced with *Entamoeba coli* and *Giardia lamblia* dosed in gelatin capsules at the level of 1 to 10 cysts. Enteric viruses have produced infection at low dosage levels via oral ingestion, inhalation, and conjunctival exposure. These studies have clearly shown that specific enteric organisms of all three classifications, i.e., bacteria, viruses, and animal parasites, can produce infections at relatively low exposure levels. Available data are insufficient to evaluate the actual health hazards that exist for individuals exposed to wastewater subjected to various degrees of treatment and dilution; however, the data indicate that enteric pathogens can cause infections at exposure levels found in undisinfected wastewater.

Sobsey reported the viral content of wastewater (United States origin) ranged from 2 to 3 to more than 1000 infectious units/100 ml sample (6). Peak periods were observed to occur in late summer and early fall. Most enteric virus concentrations have been isolated in heavily polluted surface waters, but Berg and coworkers were able to detect enteric viruses in the Missouri River having fecal coliform concentrations as low as 60/100 ml (3). Although very little quantitative information is available with respect to the concentration levels of enteric viruses in the United States surface and groundwaters, much evidence leads to the indication that wastewaters are a primary source.

It is difficult to accurately identify general municipal wastewater characteristics due to differences in locations, water uses, seasonal variations, diurnal variations, etc. However, it is important that the design engineer have a knowledge of the wastewater characteristics for which he is designing a disinfection system. He must have a knowledge of the composition of the conventional parameters as well as the concentrations of the pathogenic agents or indicator organisms for which he is designing the disinfection system. The influent pathogen or indicator organism concentration is a critical parameter for design of any of the disinfection technologies. Typical compositions of raw wastewater through various levels of treatment are summarized in Table 2-1. The total or fecal coliform (indicator organism) level is a critical design parameter that should be determined on a site specific basis where possible.

The total bacterial population of human feces has been estimated to reach a density of 10^{12} organisms per gram (3). The density range of fecal coliforms in human feces has been estimated at 10^6 to 10^9 per gram with total coliforms estimated at 10^7 to 10^9 per gram. Hubley et al. (1) presented an assessment of the concentration ranges of certain organisms in domestic wastewater, reduction through primary and secondary treatment, and estimated secondary treated effluent concentrations, as summarized in Tables 2-2, 2-3, and 2-4. If one assumes that the pathogenic organisms are removed in proportion to

Table 2-2. Typical Influent Concentration Ranges for Pathogenic and Indicator Organisms (7)

Organism	Number/100 ml Minimum	Maximum
Total Coliforms	1,000,000	—
Fecal Coliforms	340,000	49,000,000
Fecal Streptococci	64,000	4,500,000
Virus	0.5	10,000

Table 2-3. Microorganism Reductions by Conventional Treatment Processes (8)(9)

Microorganisms	Primary Treatment Removal (%)	Secondary Treatment Removal (%)
Total coliforms	<10	90–99
Fecal coliforms	35	90–99
Shigella sp.	15	91–99
Salmonella sp.	15	96–99
Escherichia coli	15	90–99
Virus	<10	76–99
Entamoeba histolytica	10–50	10

Table 2-1. Typical Composition of Domestic Wastewater

Water Quality	BOD mg/l	TSS mg/l	Total N mg/l	Total Coliforms #/100 ml	Fecal Coliforms #/100 ml
Raw Wastewater	200	200	35	10^7–10^8	10^6–10^7
Primary Effluent	130	100	30	10^7–10^8	10^6–10^7
Secondary	20	20	20	10^5–10^6	10^4–10^5
Filtered Secondary	12	5	18	10^4–10^5	10^3–10^5
Nitrified	7	10	18	10^4–10^5	10^3–10^5
Filtered Nitrified	5	5	18	10^4–10^5	10^3–10^5

Table 2-4. Secondary Effluent Ranges for Pathogenic and Indicator Organisms Prior to Disinfection

Organism	Number/100 ml Minimum	Number/100 ml Maximum
Total Coliforms	45,000	2,020,000
Fecal Coliforms	11,000	1,590,000
Fecal Streptococci*	2,000	146,000
Viruses	0.05	1,000
Salmonella sp.	12	570

*Assuming removal efficiencies for fecal streptococci similar to the fecal coliform removal efficiencies

the indicator organisms (total and/or fecal coliforms), conventional treatment of domestic wastewaters without disinfection cannot be considered sufficient for removal and control of human pathogens where beneficial uses and body contact occur.

2.2 Disinfection Criteria

In 1972 Congress enacted the Federal Water Pollution Control Act (PL 92-500) (amended in 1977, Clean Water Act, PL 95-217), to "restore and maintain the chemical, physical and biological integrity of the nation's waters." Effluent standards were to be imposed on industrial and municipal wastewater dischargers based on the limits of current technology. Today, secondary treatment is the minimum level required for municipal wastewater treatment plants. The Secondary Treatment Information Regulation was promulgated by the EPA in 1973 as Part 133, Title 40 of the Code of Federal Regulations (40 CFR 133). Included in the definition of secondary treatment was a fecal coliform limitation, but this was omitted from the CFR in July 1976, thereby delegating to the individual states, the responsibility to establish site specific water quality criteria.

Municipal wastewater treatment plants often discharge into potable water sources and recreational bodies of water. The disinfection practices for potable water have helped control waterborne disease outbreak from such sources; however, wastewater disinfection through the destruction of pathogenic agents, provides a barrier to possible waterborne disease before the wastewater is released into the environment. Public health and protection of the aquatic and human environments should be the overriding considerations affecting disinfection requirements. With the site specific constraints and considerations, such as water quality, seasonal versus year round disinfection requirements, etc., each individual state has established its own wastewater disinfection policies. A review of the types of disinfection criteria of various states follows.

As a result of the 1924-1925 typhoid epidemic resulting from consumption of contaminated shellfish, and because of several other less severe incidents related to shellfish contamination, a national shellfish sanitation program was established by the United States Public Health Service to protect the quality of coastal shellfish beds. The 1964 National Shellfish Sanitation Workshop defined the bacteriological water quality for approved shellfish areas (those areas where shellfish may be taken for direct marketing) as having a total coliform mean Most Probable Number (MPN) not exceeding 70 per 100 ml. In 1977 a fecal coliform median concentration of 14 MPN per 100 ml with no more than 10 percent of the samples exceeding 43 MPN per 100 ml was recommended for approved shellfish waters by the EPA. These fecal coliform levels were based on an extensive total coliform to fecal coliform correlation study and were subsequently accepted for wastewater effluent discharges into shellfish waters.

The two effluent guidelines or standards most commonly used by the various states and territories in the United States have been a total coliform value of 1000 per 100 ml and a fecal coliform limit of 200 per 100 ml. The microbial guideline for primary contact recreational waters as adopted by many states requires that the fecal coliform content shall not exceed a geometric mean of 200 per 100 ml and that no more than 10 percent of the total number of samples taken during any 30 day period shall exceed 400 fecal coliforms per 100 ml.

Some states have adopted even more stringent disinfection requirements than 14 MPN per 100 ml for mean effluent fecal coliform levels. Since November 1976, the Maryland Department of Health and Mental Hygiene has required all facilities plans to provide for disinfection to total coliform levels of 3 MPN per 100 ml unless a special exemption is granted by the department (10). This high level disinfection is applicable to both coastal and inland waters but is geared toward protection of the state's intensive shellfish industry. The California State Department of Health "Uniform Guidelines for Sewage Disinfection" incorporates consideration of dilution, receiving water quality, and beneficial uses in disinfection requirements, which results in varying coliform standards for different discharge situations (11). The standard for nonrestricted recreational uses of wastewater and shallow ocean discharges in close proximity to shellfish areas specifies a 7-day median total coliform value of 2.2 per 100 ml or less at some point in the treatment process. The 2.2 total coliform per 100 ml standard applies to those discharge situations where reclaimed wastewater (i.e., 100 percent wastewater effluent) is impounded for body contact recreation activities, where wastewater is discharged to ephemeral streams with little or no dilution, and where body contact recreation has been designated as a beneficial use of the stream.

The various states have fecal coliform standards ranging from less than 2.2/100 ml up to 5,000/100 ml and total coliform standards from less than 2.2/100 ml up to 10,000/100 ml. Some states have seasonal disinfection requirements, and Illinois, at the time of writing this manual, has no bacterial standards for certain, designated non-contact waters. These disinfection criteria have been established relative to discharge stream water quality with the most common standards being fecal coliform limits of 200/100 ml to 1,000/100 ml. Over 45 states have multi-level standards for disinfection relative to the discharge stream water quality criteria, with 200 fecal coliforms per 100 ml as the most common standard. At least 15 states have both fecal and total coliform disinfection criteria. At least 15 states also require effluent disinfection levels of 14 fecal coliforms per 100 ml for discharge into shellfish waters, and 9 states have more stringent disinfection standards than those imposed on shellfish water discharges.

As indicated by Cabelli (2), the development of water quality guidelines, standards, and associated health effects for recreational waters, such as disinfection requirements, has historically followed a pattern characteristic of many such efforts to control pollution; the establishment of health and ecological effects. The first step has been the development of guidelines and standards dictated largely by attainment with the Best Control Technology Available (BAT). The guidelines have been based upon limited epidemiological and ecological evidence in many cases, and little, if any, data quantifying the risk in relation to the level of the pollutant in the environment. The second step has been the modification of the guidelines and standards on the basis of detectable risk with a limited quantity of data. The last step is then the development of guidelines based upon acceptable risk, which requires an epidemiological or ecological data base broad enough to mathematically model the relationship of some measure of water quality to the risk. As more studies are conducted and more data become available, the effluent disinfection guidelines may change. This design manual has been prepared with the realization that different disinfection criteria are presently required and still others may be required in the future. Therefore, the design of the various disinfection alternatives has been presented with this flexibility in mind, as well as the considerations of different water qualities. For example, this manual can be used to design disinfection facilities for primary treatment through advanced wastewater treatment. It can be used for effluents from lagoons, activated sludge, trickling filters, oxidation ditches, and rotating biological contactors, as well as other unit processes.

Attainment of the disinfection guidelines can only be achieved by the disinfection process, which, from a disease prevention standpoint, is the most important unit process in the wastewater treatment system. Disease transmission via the aquatic route including recreational water, drinking water, and seafood from polluted water has been and continues to be a problem. To the extent that the wastewater disinfection process mitigates that problem asserts for the continued use of that practice in this country.

2.3 Projected Applications of Disinfection Technologies

A summary of the 1984 EPA wastewater treatment plant needs survey has been prepared (12). A breakdown of the number of municipal wastewater treatment plants by flow capacity in 1984 and those projected for the year 2000 is presented in Table 2-5.

Table 2-5. Number of Wastewater Treatment Plants by Flow Capacity (12)

Flow Capacity m^3/d (mgd)	Number of Facilities 1984	2000	Percent of Total Wastewater Flow 1984	2000
380 (0 – 0.1)	5,032	8,416	0.7	0.9
380 – 3,800 (0.1 – 1.0)	6,962	8,313	7.4	6.9
3,800 – 38,000 (1.0 – 10.0)	2,833	3,255	25.7	24.6
>38,000 (>10.0)	551	687	66.2	67.6
Total	15,378	20,671		

The total number of facilities broken down by type of treatment system in 1984 and projected for the year 2000 is shown in Table 2-6. An evaluation of the number of facilities and quantity of flow treated by municipal wastewater treatment facilities with primary treatment through advanced wastewater treatment shows that approximately 41 percent of the total wastewater flow presently receives secondary treatment, with similar requirements in the year 2000. Approximately 39 percent of the total flow presently treated receives greater than secondary

Table 2-6. Summary of Wastewater Treatment Processes in the United States (12)

Type of Process	Number of Processes 1984	2000
Lagoons	7,500	12,210
Activated Sludge	5,690	8,275
Trickling Filter	2,463	2,570
Land Treatment	926	1,454
Oxidation Ditch	741	1,215
Rotating Biological Contactor	347	291
Total Design Flow, m^3/d (mgd)	1.3×10^8 (35.9×10^9)	1.6×10^8 (43.2×10^9)

treatment, with around 51 percent expected to require this level of treatment in the year 2000.

In 1984 at least 80 percent of the total wastewater flow treated required secondary or more stringent levels of treatment. This value is expected to increase to over 90 percent of the total flow treated in the year 2000. Most of this increase will be in requirements for advanced secondary treatment with the number of facilities increasing from around 3,000 in 1982 to around 6,400 facilities in the year 2000. The percentage of flow treated that requires secondary treatment will remain about the same; however, an estimated increase of around 5,000 new facilities has been projected. These more stringent requirements for treated effluent quality are related to concerns for protection of public health and the aquatic environment. These concerns will most certainly place greater emphasis on disinfection requirements and alternatives to chlorine disinfection. Advanced wastewater treatment facilities are excellent candidates for alternative disinfection with ozone and ultraviolet irradiation. Due to the health hazards and environmental concerns with chlorination, there appears to be an excellent market potential for alternative disinfectants, especially ozone and ultraviolet irradiation, over the next 15 years.

Many of the existing treatment facilities are required to disinfect their effluents, although the degree of bacterial reduction varies from plant to plant. However, there are many facilities that presently do not disinfect their effluents prior to discharge. Due to the present concerns and future requirements for more stringent effluent qualities, more stringent disinfection criteria relative to both microorganism control, disinfectant residual, and biotoxicity or bioassay testing requirements can be expected. Many of the existing facilities may be required to upgrade their present disinfection systems to include dechlorination or install alternative disinfection processes. New facilities being built can be expected to require proper disinfectant residual controls or alternative disinfection processes.

Over 400 facilities have been designed for chlorination/dechlorination, and approximately 50 percent of these facilities are presently in use. At the time of writing this manual, there were 24 wastewater treatment plants in the United States confirmed to be using ozone disinfection, 20 water treatment plants using ozone either for disinfection or chemical oxidation, and 30 drinking water plants in Canada using ozone (R.G. Rice, Personal Communication, 1985). An estimated 125 wastewater treatment facilities were reported to be using ultraviolet light disinfection in 1984. As of March, 1985, 33 municipalities with ultraviolet light disinfection had been granted funds by the EPA under the Innovative and Alternative (I/A) Technology Program.

2.4 References

When an NTIS number is cited in a reference, that reference is available from:

National Technical Information Service
5285 Port Royal Road
Springfield, VA 22161
(703) 487-4650

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Chapter 3
Disinfection Alternatives and Options

3.1 General Considerations

Since municipal effluents are an identifiable source of pathogenic contamination and disinfection processes themselves can create hazards to human health and the aquatic environment, the decision to disinfect or not disinfect is a very complicated one that must be made on a site-specific basis. It is therefore, impossible to establish universal policies on wastewater disinfection requirements. Resolution of the need for municipal wastewater disinfection at a particular site involves the investigation of receiving water uses and the associated risks to human health, an assessment of the options that are available for control of fecally-contaminated discharges, and an evaluation of the environmental effects that control measures may create. Figure 3-1 presents an approach for the type of rationalization that can be involved in assessing the need for, and consequences of, disinfecting municipal wastewaters (1).

In general, Figure 3-1 indicates that human health risks are the initial concern, and upon establishing the level of risk involved and the potential for reducing or eliminating the risk, the environmental considerations determine the applicability of the proposed control measures. Development of an option, chlorination or an alternative disinfectant, that satisfies both the human health and environmental concerns at a specific site is the next step.

3.2 Selecting a Disinfection Alternative

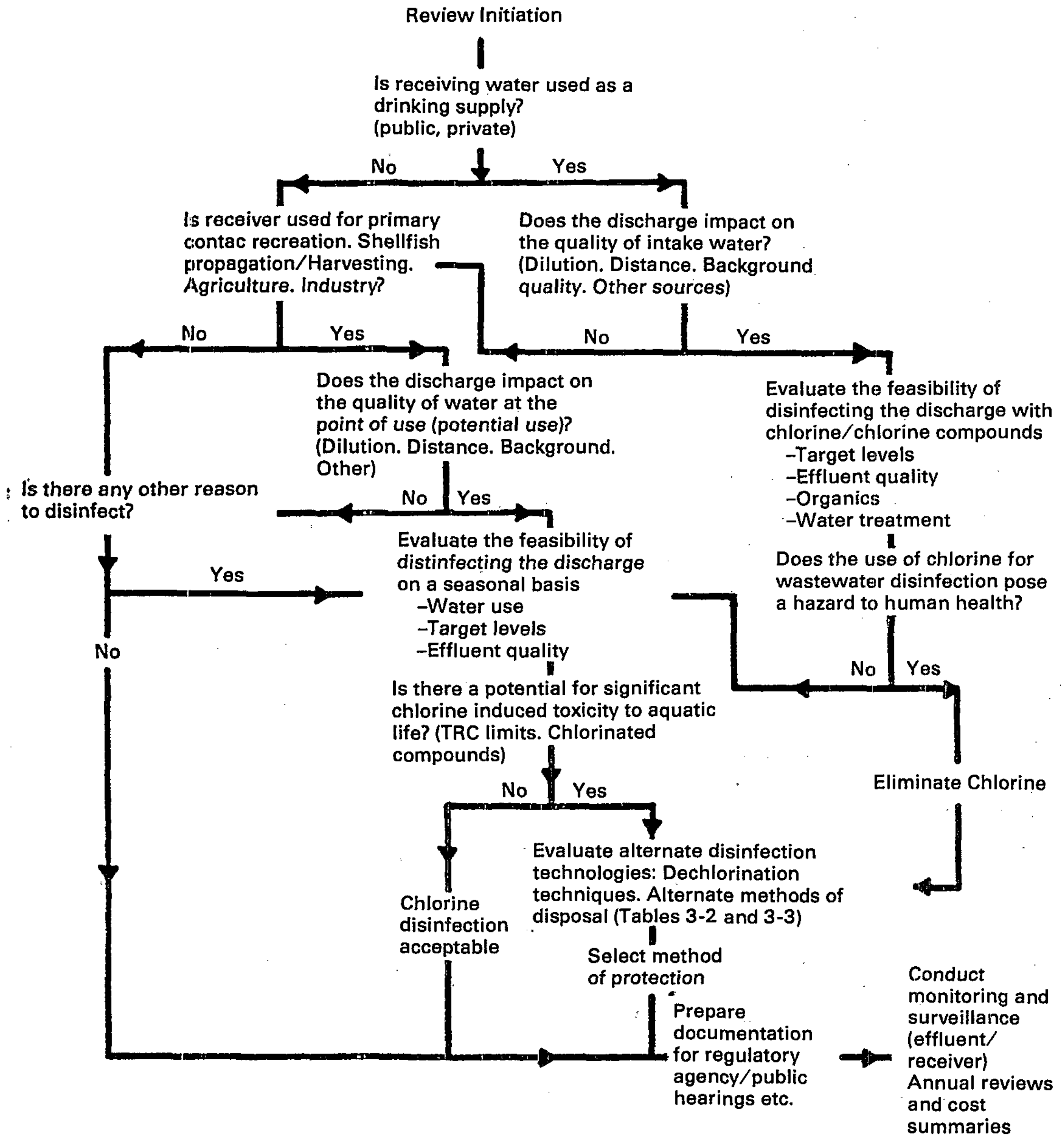
As many as 25 disinfection alternatives could be considered and have been previously identified from the literature without regard for physical or operational constraints (2). The major factors that must be considered when evaluating disinfection alternatives are summarized in Table 3-1 (3). The first four factors, effectiveness, use cost, practicality, and pilot study requirements relate to the disinfection process itself. The fifth factor, potential adverse effects, relates to effects of the disinfectant on the receiving water and other environmental concerns and considerations. Evaluation and thorough consideration of all the criteria listed in Table 3-1 relative to practical, physical, and operational constraints of municipal wastewater disinfection reduces the available alternatives to chlorination, hypochlorination, chlorination/dechlorination with sulfur dioxide, chlorine dioxide, bromine chloride, ozone, and ultraviolet light.

To properly evaluate and select alternative disinfection systems two levels of review are required. In the first level of review, a number of non-monetary factors are considered. This qualitative assessment is comprised of three primary components, including the previously described technical factors, environmental impacts, and safety. In order to assess the disinfection alternatives with respect to their non-monetary factors, a qualitative matrix approach, as shown on Table 3-2, can be used. A relative ranking of the alternatives based on this qualitative assessment can also be made, as shown in Table 3-3. The ranking scale is based on a scale of one to five, with one indicating the least impact or best degree of confidence. From these types of analyses, the number of appropriate alternatives can be narrowed, and some alternatives may be completely eliminated.

The remaining acceptable alternatives can then be evaluated in a second more detailed level of review. In this second level of review, a preliminary design can be developed, cost estimates performed, and an economic analysis comparing the alternatives on an equitable basis can then be evaluated. Detailed capital and operation and maintenance costs can be developed for each alternative disinfection system. Capital costs include structures, process equipment, major auxillary equipment, special foundation requirements, electrical and instrumentation, site work, miscellaneous process and piping, construction contingencies, engineering, project administration, and interest during the estimated period of construction. The operation and maintenance costs are annual costs and include labor, electrical power, chemicals, routine equipment maintenance, and materials and supplies. The specific details required for performing the second level of review are addressed in the respective design chapters for each of the disinfection alternatives.

The predominant advantages and disadvantages of these disinfection alternatives are well known and commonly cited in the literature. Some of the more obscure elements have not been emphasized or have been considered secondary or insignificant. A brief

Figure 3-1. Framework for evaluating site-specific wastewater disinfection requirements.



review of the pertinent factors associated with each of the alternatives is presented in the following sections of this chapter.

From Tables 3-2 and 3-3, it may be possible to select an alternative disinfection system, as demonstrated in the following situations. Examples follow where ozone, ultraviolet light, and chlorination/dechlorination alternatives were selected for different specific applications.

The cost of installing and operating an ozone disinfection process is dependent on many variables. Major factors include size, flexibility, local construction costs, energy costs, energy efficiency and power consumption, ozone dose requirements, and site specific constraints. The cost for producing ozone is normally higher than the alternative disinfection methods; however, the other advantages associated with ozone disinfection (Tables 3-2 and 3-3), may outweigh cost considerations in some cases. The following example is cited where ozone would be selected as the disinfectant of choice based on site specific constraints; however, this situation may be unique since ozone would be cost competitive.

Table 3-1. Major Factors in Evaluating Disinfectant Alternatives

Effectiveness	- Ability to achieve target levels of selected indicator organisms - Broad spectrum disinfecting ability - Reliability
Use-Cost	- Capital cost - Amortization cost - Operating and maintenance cost - Cost of special wastewater pretreatment
Practicality	- Ease of transport and storage, or ease of on-site generation - Ease of application and control - Flexibility - Complexity - Ability to predict results - Safety considerations
Pilot Studies Required	- Dose requirements - Refine design details
Potential Adverse Effects	- Toxicity to aquatic life - Formation and transmission of undersirable bio-accumulating substances - Formation and transmission of toxic, mutagenic, or carcinogenic substances

A large tourist community has selected the oxygen activated sludge process as the secondary treatment process of choice. The treatment plant is to be located next to the municipal utility district power plant with electrical costs of $0.025/kWh. The following design conditions were established due to the stringent effluent discharge criteria:

- Average Flow = 378,000 m^3/d (100 mgd)
- Peak Daily Flow = 568,000 m^3/d (150 mgd)
- Effluent BOD = 5 mg/l
- Effluent TSS = 5 mg/l
- Effluent NH_3-N = 1 mg/l
- 30-Day Geometric Mean Fecal Coliforms = 200/100 ml
- Maximum Daily Fecal Coliforms = 400/100 ml
- Maximum Chlorine Residual = 0.02 mg/l
- Dissolved Oxygen = 6.0 mg/l

Due to the recreational nature of the community and the use of the downstream water, major considerations included effective virus destruction, *Giardia* cyst destruction, and color removal. Maintaining the pristine image of the service area and protection of the quality of the downstream water supplies were the primary reasons for the choice of ozone disinfection. The total cost for ozone disinfection represented a small part of the total plant costs, and in this unique situation, at the design ozone transferred dose of 5.0 mg/l, ozone disinfection was cost competitive with chlorination/dechlorination and ultraviolet disinfection. The safety concerns for handling and transporting chlorine through the recreational community were also resolved with the selection of an ozone disinfection process.

In a second community, ultraviolet disinfection was chosen over chlorination/dechlorination and ozonation when the following factors were given high priority:

- effectiveness in destroying bacterial and viral pathogens.
- safety precautions for transport, storage, and application.
- environmental impacts, especially as relates to halogenated organics formation and chlorine residual toxicity.
- equipment reliability and ease of maintenance.
- process reliability and simplicity.
- process control flexibility.
- overall construction, operation, and maintenance costs.

The treatment process consists of activated sludge, final clarification, chemical addition for phosphorus removal, and filtration prior to disinfection. The following design conditions applied to this specific case:

- Average Flow = 7,600 m^3/d (2 mgd)
- Peak Daily Flow = 13,200 m^3/d (3.5 mgd)
- Effluent BOD = 10 mg/l
- Effluent TSS = 10 mg/l
- Effluent NH_3-N = 1.0 mg/l
- Effluent PO_4-P = 1.0 mg/l
- 30-Day Geometric Mean Fecal Coliform = 200/100 ml
- Maximum Daily Fecal Coliforms = 400/100 ml
- Maximum Chlorine Residual = 0.1 mg/l

Ultraviolet light was considered to be the most desirable of the three disinfection alternatives considered with respect to non-monetary factors. Ultraviolet light systems were considered to be simple to operate and maintain relative to ozone systems, yet provide the same advantages for overall disinfection capacity relative to bacteria, spores, and virus inactivation. Ultraviolet light was the safest alternative since it is a physical disinfection process, and its use eliminated the handling, transportation, and storing of toxic, hazardous, and corrosive chemicals. Due to the size of the plant and assurance of good effluent quality especially with respect to the suspended solids levels, ultraviolet light was chosen as the disinfectant of choice over both ozone and chlorination/dechlorination. Ultraviolet light disinfection was cheaper than ozonation and cost competitive with chlorination/dechlorination in this particular application.

The third specific case cited here involved a community in the northern part of the United States

Table 3-2. ***Applicability of Alternative Disinfection Techniques***

Consideration	Cl_2	$Cl/deCl_2$	BrCl	ClO_2	O_3	UV
Size of plant	*all sizes*	*all sizes*	*all sizes*	*small to medium*	medium to large	small to medium
Applicable level of treatment prior to disinfection	all levels	all levels	secondary	secondary	secondary	secondary
Equipment Reliability	good	fair to good	?	?	fair to good	fair to good
Process Control	well developed	fairly well developed	problematic	no experience	developing	developing
Relative Complexity of Technology	simple to moderate	moderate	moderate	moderate	complex	simple to moderate
Safety Concerns Transportation on site	yes substantial	yes substantial	yes substantial	yes substantial	no moderate	no minimal
Bactericidal	good	good	good	good	good	good
Virucidal	poor	poor	fair to good	good	good	good
Fish Toxicity	toxic	non-toxic	slight to moderate	toxic	none expected	non-toxic
Hazardous By-products	yes	yes	yes	yes	none expected	no
Persistent Residual	long	none	short	moderate	none	none
Contact Time	long	long	moderate	moderate *to long*	moderate	short
Contributes Dissolved Oxygen	no	no	no	no	yes	no
Reacts with Ammonia	yes	yes	yes	no	yes (high pH only)	no
Color Removal	moderate	moderate	?	yes	yes	no
Increased Dissolved Solids	yes	yes	yes	yes	no	no
pH Dependent	yes	yes	yes	no	slight (high pH)	no
O&M Sensitive	minimal	moderate	moderate	?	high	moderate
Corrosive	yes	yes	yes	yes	yes	no

Table 3-3. Technical Factors and Feasibility Considerations

Considerations	Cl_2	$Cl/deCl_2$	BrCl	ClO_2	O_3	UV
Flexibility	2	2	2	2	2	2
Reliability	1	2	3	2	3	2
Complexity	2	2	2	3	3	2
Effectiveness	2	2	1	1	1	2
Pilot Studies	1	1	4	4	3	3

Rating based on scale of 1 to 5, with 1 indicating best degree of confidence.

where seasonal disinfection was required during the early spring and summer months. In this particular application there was a concern for chlorine residuals and associated fish toxicity; however, the discharge stream was not used as a water supply downstream. The following design conditions apply to this plant:

- Average Flow = 3,800 m^3/d (10 mgd)
- Peak Daily Flow = 5,700 m^3/d (15 mgd)
- Effluent BOD = 30 mg/l
- Effluent TSS = 30 mg/l
- 30-Day Geometric Mean Fecal Coliforms = 200/100 ml
- Maximum Daily Fecal Coliforms = 400/100 ml
- Chlorine Residual = 0.1 mg/l

Due to the seasonal disinfection requirements, capital costs, and operation and maintenance costs, chlorination/dechlorination was chosen as the design alternative. The non-monetary advantages typically associated with ozone and ultraviolet light over chlorination/dechlorination did not outweigh the advantage of chlorine disinfection in this particular application.

As indicated in the previous examples, site specific constraints and criteria will set guidelines for selection of appropriate disinfection alternatives. Important considerations include factors such as the size of the plant and effluent discharge quality requirements. The more stringent the effluent requirements, the more feasible alternative disinfectants such as ozone and ultraviolet light become. Capital and operation costs are obviously important factors; however, the cost of any of the disinfection options is a very small part of the total system costs.

3.3 Chlorination

Chlorine has not been required by EPA for inclusion in States' nonconventional pollutant standards, but approximately 15 states have taken steps to develop specific criteria for determining the impact and adverse effects of chlorine on aquatic life. Most states have not adopted site-specific criteria to determine the need(s) and adverse effects of disinfection using chlorine. As a result, there has been limited success in ensuring that public health and aquatic wildlife are adequately protected.

Current developments consist of revised water quality standards regulation and guidance documents to ascertain the appropriateness of existing water quality standards and assess use attainability. EPA's Office of Research and Development has completed a draft criteria document for chlorine with directions on how to apply the criteria and the chemistry and fate of chlorine in natural receiving bodies of water. It is anticipated that guidance for establishing chlorine effluent limitations for NPDES permits will be published as well as other documents to help encourage states to consider public and wildlife health issues. The Advanced Technology (AT) review policy requires the evaluation of chlorine toxicity and the construction and operation of dechlorination facilities where chlorine will exceed EPA criteria.

However, due to the toxicity of chlorine residuals at extremely low concentrations (11 to 19 μg/l) and the relatively high limit of detection of chlorine residual test procedures (50 to 100 μg/l), it is difficult to control chlorine-induced toxicity in the receiving stream. Therefore, use of alternative disinfection processes should be considered where aquatic toxicity is the overriding concern.

Today, chlorination is the most used disinfectant at water and wastewater treatment plants in the United States. Chlorine reacts very rapidly when mixed with water, and both hydrolysis and ionization occur. Environmental factors such as temperature, pH, alkalinity, suspended solids, chemical oxygen demand (COD), and nitrogen containing compounds influence the effectiveness of chlorine disinfection. Chlorine reacts rapidly with ammonia and certain organic compounds to form chloramines and chlorinated organic compounds. The combined chloramines are lower in germicidal value compared to the free chlorine residuals, with organic chloramines in wastewater offering significantly lower germicidal value than inorganic chloramines.

The use of chlorine disinfection of wastewater can result in several adverse environmental impacts, especially due to toxic levels of total residual chlorine in the receiving water and formation of potentially toxic halogenated organic compounds. Chlorine residuals have been found to be acutely toxic to some species of fish at very low levels. The chlorine residuals are stable and can persist for many hours at toxic levels. Other toxic or carcinogenic chlorinated compounds can bioaccumulate in aquatic life and contaminate public drinking water supplies.

Chlorine is normally handled in steel containers of 68 kg (150 lb) cylinders up to 82 metric ton (90 ton) railroad cars. Chlorine is an extremely volatile and hazardous chemical, and proper safety precautions must be exercised during all phases of chlorine shipment, storage, and use.

Hubley et al. conducted an extensive analysis of the transportation risks of chlorine (4). A summary of some of their findings is presented. A compilation of the Department of Transportation (DOT) accident reports for chlorine transportation from 1971 to 1980 is summarized in Table 3-4. These reports have included information on number of deaths or injuries and amount of chlorine released. The data in Table 3-4 are summarized by railroad, truck, and barge shipment. The major accident at Youngstown, Ohio in February 1978, has been broken out separately. Chlorine movements for 1972, obtained from the U.S. Bureau of Census, in metric ton-kilometers by transportation mode and by shipment weight are presented in Table 3-5 (5).

A breakdown of the information in Table 3-5 for chlorine shipments by transportation mode and container is presented in Table 3-6 (4). The information in Table 3-4 was averaged to obtain estimates of deaths, injuries, and property damage per year.

The percentage breakdown by transportation mode and container in Table 3-6 was then used to calculate

Table 3-4. Compilation of Department of Transportation Accident Data (January 1971 to December 1980) (4)

	Number of Accident Reports	Deaths	Injuries	Property Damage ($)	Amount Released (kg)
Railroad	72	8	247	1.1×10^6	1.4×10^5
Railroad Excl. Youngstown	71	0	87	22,500	99,300
Truck:					
Cylinders to 114 kg	14	0	60	8,000	574
0.911 metric ton containers	4	0	15	23,500	245
Tanker trucks	2	0	71	15,000	23
Barge	2	0	3	0	*

*Information not available prior to 1976.

Table 3-5. Percent Distribution of Chlorine Shipped by Transportation Mode and by Shipment Weight (metric ton-kilometers) (4)

	Percent
A. Transportation Mode	
Rail	84.9
Truck (combines DOT motor carrier and private truck data)	15.0
Water	0.3
	100.2
B. Shipment Weight	
Under 454 kg	0.1
454 to 4,500 kg	5.0
4,500 to 13,600 kg	6.3
13,600 to 27,200 kg	6.1
27,200 to 40,900 kg	2.5
40,900 kg and over	80.2
	100.2

Table 3-6. Breakdown of Chlorine Shipments by Transportation Mode and Container (4)

	Percentage (for 1972)	Annual Average metric ton-km (for 1971 to 1980)
Rail	84.9	1858
Truck		
Cylinders to 114 kg	0.1	2.2
0.91 metric ton cylinders	5.0	109
Tank Truck	9.9	219
Barge	0.3	6.4
	100.2	2194.6

a corresponding breakdown by ton-km, assuming a yearly total of 2191 million ton-km.

The results of this analysis, shown in Table 3-6, were then used to normalize the accident data per ton-km as shown in Table 3-7. It can be observed in Table 3-7 that the accident rate for truck-transported cylinders was consistently higher than the other categories listed.

Chlorination systems are reliable and flexible, and the equipment is not complex. It is relatively easy to apply and control in wastewater treatment, and low use cost is often a great advantage for chlorine. Even when dechlorination is required, it is still normally the lowest cost alternative in most cases.

Hypochlorination refers to the use of solid (calcium) or liquid (sodium) hypochlorite compounds as the disinfecting agent. The active compounds are the same as with chlorination, primarily monochloramine in wastewater effluents. The mechanism of bacterial kill is the same as with liquid or gaseous chlorine. Adverse environmental impacts associated with the use of chlorine are also applicable with sodium hypochlorite. Concentrated solutions from 10 to 15 percent chlorine must be stored in rubber-lined steel or fiberglass storage tanks. Sodium hypochlorite is a hazardous and corrosive material, but it will not volatilize to a toxic gas as liquid chlorine does. The required equipment is relatively simple and easy to operate and maintain. The primary advantage of hypochlorination over chlorination is the increased safety in transporting, storing, and handling of chemicals; however, chemical costs per unit of free chlorine are generally much higher.

Dechlorination, normally with sulfur dioxide, has been used to reduce the environmental impacts and concerns associated with chlorination or hypochlorination. Other dechlorinating agents such as other sulfite reducing compounds or activated carbon have been used; however, costs normally make these options prohibitive. The levels of total chlorine residuals can normally be reduced to below levels that are toxic to aquatic life. The potential for formation of halogenated organics may be reduced; however, it appears that many halogenated organics are formed rapidly upon chlorine addition, and the application of sulfur dioxide will probably not affect these compounds.

Dechlorination facilities are similar in most respects to chlorination systems. Sulfonators meter gaseous sulfur dioxide and are similar to chlorinators. The amount of sulfur dioxide applied is normally based on

Table 3-7. Accident Rates Per Metric Ton-KM (4)

	Deaths	Injuries	Property Damage ($)	Chlorine Released (kg)
Railroad	4.3×10^{-10}	1.4×10^{-8}	6.0×10^{-5}	3.3×10^{-5}
Railroad excluding Youngstown	0	4.7×10^{-9}	1.2×10^{-6}	2.3×10^{-5}
Truck:				
Cylinders to 114 kg	0	2.7×10^{-6}	3.6×10^{-4}	1.2×10^{-4}
0.91 metric ton containers	0	1.4×10^{-8}	2.15×10^{-5}	1.0×10^{-6}
Tanker Truck	0	3.2×10^{-8}	0.7×10^{-5}	4.6×10^{-8}
Barge	0	4.7×10^{-8}	0	*

*Information not available.

flow and/or chlorine residual signals because continuous sulfur dioxide monitoring equipment is not presently available. Reaction with chlorine residual is very rapid and separate contact chambers are normally not used. Chlorination/dechlorination is more complex to operate and maintain than chlorination or hypochlorination alone. The total cost of chlorination will be increased by approximately 30 to 50 percent with the addition of the dechlorination step.

3.4 Chlorine Dioxide

Chlorine dioxide has been used to disinfect potable waters, especially waters containing phenols or other taste and odor producing compounds. It is a proven bactericide equal to or greater than chlorine in disinfecting power, and has a higher oxidation potential. Chlorine dioxide has been indicated to be a much more effective virucide than chlorine. The environmental impacts associated with chlorine dioxide use as a wastewater disinfectant have not been well established. Advantages of chlorine dioxide include its lack of reaction with ammonia and reduced halogenated organic compound formation.

Chlorine dioxide is an extremely unstable and explosive gas and any means of transport is potentially very hazardous. Therefore, it must be generated onsite. Gaseous chlorine dioxide is normally generated using a process that involves a reaction between sodium chlorite and chlorine. Sodium chlorite is much more expensive than gaseous chlorine per unit weight expressed as chlorine. Chlorine dioxide has not received a great deal of attention as a wastewater disinfectant due to its on-site generation requirements and high chemical costs. The overall system is relatively complex to operate and maintain compared to chlorination. Safety hazards include the handling of two dangerous chemicals. Chlorine dioxide has been shown to be an effective wastewater disinfectant; however, few pilot studies have been done and experience is quite limited in this country. No applications of chlorine dioxide disinfection of municipal wastewaters are known at the present time.

3.5 Bromine Chloride

Bromine chloride hydrolyzes to hyprobromous acid very rapidly and forms monobromamines and dibromamines when added to water containing ammonia-nitrogen. Bromamines have been shown to be very effective as a disinfectant with shorter lived residuals compared to chloramines. Because bromamines are more unstable than chloramines, shorter contact times are needed. Environmental impacts associated with bromine chloride disinfection are less adverse than those associated with chlorine. Contrasting results relative to the toxic effects of bromine residuals have been cited in the literature, possibly due to the stability of the bromine residual and their rapid decay to chloride and bromide. Other brominated organic compounds can also be formed during the disinfection process. These compounds may not be produced in appreciable quantities; however, they may be toxic at high enough concentrations and may bioaccumulate.

Bromine chloride is manufactured from bromine and chlorine. It is a hazardous and corrosive chemical and requires special transportation, handling, storage, and use precautions. Bromine chloride is shipped in modified chlorine cylinders, or tank cars, and essentially the same safety and handling procedures are required as required for chlorine.

Bromine chloride disinfection facilities are very similar to those used for chlorination. It is applied in much the same manner as chlorine, using a vacuum chlorobrominator to prepare a concentrated solution, which is then mixed with the wastewater. Shorter contact times are possible, resulting in a construction cost saving due to a smaller contactor. The use cost is heavily dependent on the cost of the bromine chloride itself. Secondary treated wastewaters require no special pretreatment prior to disinfection; however, bromine chloride use for wastewater disinfection is relatively new and a good data base and track record are presently not available. Bromine chloride disinfection has been proven effective in field-scale applications for wastewater, but problems with the liquid feed equipment have occurred. Brominated organics such as bromoform and mixtures of chlorinated and brominated organics can be expected to be formed.

3.6 Ozone

Ozone is an unstable gas that is produced when oxygen molecules are dissociated into atomic oxygen and subsequently collide with another oxygen molecule to produce ozone. The ozone molecule can coexist as a gas with other gases such as air or oxygen, or it can dissolve in a liquid such as water. Ozone is an extremely reactive oxidant and a very effective bactericide and virucide. Over 40 full-scale facilities have been constructed in the United States, and an extensive amount of research and development has been expended to develop ozone disinfection of wastewaters.

Unlike some of the other chemical disinfecting agents previously discussed, ozone can exert beneficial impacts on the environment. Since ozone decomposes rapidly to oxygen after application, the dissolved oxygen levels in the treated effluent can be elevated significantly, often to saturation levels. In most, if not all cases, the need for effluent reaeration to meet required dissolved oxygen water quality standards can be eliminated. Ozone residuals can be acutely toxic to aquatic life; however, since ozone dissipates rapidly, residuals are normally not found in the effluent by the time it reaches the receiving water. Ozonation has been shown in some instances to produce toxic mutagenic and/or carcinogenic compounds, but little is presently known about these organic byproducts (6). Ozone is believed to present fewer potential environmental and health hazards than chlorine.

Due to the instability of ozone, it must be generated on-site from air or oxygen carrier gas. The most efficient method of producing ozone today is by the electric discharge technique, which involves passing the air or oxygen carrier gas across the gap of narrowly spaced electrodes under a high voltage. Due to this expensive method of producing ozone, it is extremely important that the ozone is efficiently transferred from the gas phase to the liquid phase. The two most often used contacting devices are bubble diffusers and turbine contactors. With the bubble diffusers, deep contact tanks are required. Ozone transfer efficiencies of 85 percent and greater can be obtained in most applications when the contactor is designed properly. The contactors must be covered to control the off-gas discharges. Since any remaining ozone would be extremely irritating and possibly toxic, the off-gases from the contactor must be treated to destroy the remaining ozone. Ozone destruction is normally accomplished by thermal or thermal-catalytic means.

An ozonation system can be considered to be relatively complex to operate and maintain compared to chlorination. The process becomes still more complex if pure oxygen is generated on-site for ozone production. Ozonation system process control can be accomplished by setting an applied dose responsive to wastewater flow rate (flow proportional control), by residual control, or by off-gas control strategies. Ozone disinfection is relatively expensive, with the cost of the ozone generation equipment being the primary capital cost item, especially since the equipment should be sized for the peak hourly flow rate as with all disinfectant technologies. Operating costs can also be very high depending on power costs, since ozonation is a power intensive system. The important criteria for design include maximum transfer efficiency in the contactor to maximize ozone utilization and minimize applied dose and power consumption requirements along with efficient ozone generation equipment design. Equipment design is presently provided by the equipment manufacturers, and they are improving and updating their equipment to improve production efficiencies and reduce the associated operating costs.

3.7 Ultraviolet Light

The effectiveness of ultraviolet light as a bactericide and virucide has been well established. It is a physical disinfecting agent compared to the previously discussed chemical agents. Radiation at a wavelength of 254 nm penetrates the cell wall and is absorbed by the cellular nucleic acids. This can prevent replication and cause death of the cell. Since ultraviolet light is not a chemical agent, no toxic residuals are produced. Although certain chemical compounds may be altered by the radiation, the energy levels used for disinfection are too low for this to be a significant cause for concern.

Major advantages of ultraviolet light are its simplicity, lack of impact on the environment and aquatic life, and minimal space requirements. There is a negligible likelihood of producing harmful chemicals in the wastewater. Required contact times are very short, on the order of seconds rather than minutes. The equipment is simple to operate and maintain, but fouling of the quartz sleeves or Teflon tubes must be dealt with on a regular basis. Fouling is normally handled by mechanical, sonic, or chemical cleaning. High suspended solids concentrations, color, turbidity, and soluble organic matter in the water can react with or absorb the ultraviolet radiation reducing the disinfection performance. High levels of wastewater disinfection (e.g., 2.2 total coliforms per 100 ml) will be difficult to achieve with ultraviolet disinfection.

Relationships between effluent quality, effectiveness, and use costs have become better defined in recent months. Total costs appear to be competitive with chlorination. The major operating costs are power consumption and annual replacement of the ultraviolet lamps. The process has been proven to be applicable with over 120 installations in the United States in small to medium size treatment plants. This process is considered to be an effective alternative to chlorination. Increased popularity and lowered costs have occurred due to improvements in modern lamp and system designs, increased competition, and improved reliability and simplicity of operation.

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When an NTIS number is cited in a reference, that reference is available from:

National Technical Information Service
5285 Port Royal Road
Springfield, VA 22161
(703) 487-4650

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Chapter 4
Kinetics and Hydraulic Considerations

4.1 Disinfection Kinetics

4.1.1 Natural Die-Off

The natural processes of dilution, physical removal, and die-off or inactivation can reduce pathogen levels. The impacts of dilution on the concentration of organisms in a wastewater discharged into a stream, river, or lake can be estimated. Several mathematical models have been proposed for estimating bacterial die-off, and these models are mostly based on first order kinetics. For example, the first order model for die-off in streams is (1):

$$N = N_o e^{(-kt)} \qquad (4\text{-}1)$$

where:

N_o = the initial concentration of microbes discharged into the stream
N = the concentration of the microbes t time units after discharge into the stream
k = rate constant

The first order model for die-off in standing water bodies (i.e., lagoons, lakes, etc.) is as follows:

$$N = N_o/(1+kt_d) \qquad (4\text{-}2)$$

where:

N_o = the concentration of microbes in the water body's inflow
N = the concentration of microbes in the water body's discharge
t_d = the hydraulic detention time in the water body based on the water body's discharge
k = *rate constant*

The rate constant k can be determined from a die-away study of typical lakes or streams in a planning area. Rate constants may also be found in the literature (2).

Johnson, et al. (3) found that coliform die-away or removal rates can be as much as 16 times higher in summer months when compared to winter months. Results of a study on the Logan City, Utah lagoon system indicated that the summer coliform decay rate coefficient was 0.5 per day, and that the winter coliform decay rate coefficient was 0.03 per day. Based on these results and using Equation 4-1, a hydraulic residence time of at least 23 days in the summer months would be required to reduce an influent coliform level of 10^7 organisms/100 ml to an effluent coliform level of 10^2 organisms/100 ml. Redesign of lagoon systems to meet stringent effluent coliform disinfection criteria appears to be economically impractical, especially because of the low die-away rates during winter months. The only remaining alternative consists of adding disinfection processes to the final lagoon effluents.

4.1.2 Die-Off In the Presence of a Disinfectant

In cases where natural die-off is not sufficient to prevent the potential for humans to ingest pathogenic organisms, disinfection should be required prior to discharge. Disinfection is a time-dependent process. The ultimate outcome of bacteria and virus destruction is the result of a series of physical, chemical, and biochemical actions that can be approximated by simple kinetic expressions. It must be pointed out that although the kinetic descriptions are simple, application of these kinetics cannot be used universally. Site specific conditions at one site may create problems with precision and accuracy in the use of an empirical relationship that was found effective at a different site.

The information needed for the design of a disinfection system includes knowledge of the rate of inactivation of the target organism(s) by the disinfectant. In particular, the effect of disinfectant concentration on the rate of the process will determine the most efficient combination of contact time and disinfectant dose to use. The major precepts of disinfection kinetics were first enunciated by Chick (4), who recognized the close similarity of microbial inactivation by chemical disinfectants to chemical reactions.

Chick stated that "disinfection is a gradual process, without any sudden effects, and if the disinfectant is sufficiently dilute to admit a reasonable time being taken for the process, the reaction velocity can be studied by enumerating the surviving organisms at successive intervals of time." Therefore, for a given number of organisms and chemical disinfectants, the rate of disinfection (number of organisms inactivated/ volume-time) can be described by:

$$-\frac{dN}{dt} = kN \qquad (4\text{-}3)$$

where:

$-\frac{dN}{dt}$ = rate of change in organism population

k = organism die-off rate constant

N = number of surviving organisms per unit volume at any given time

Equation 4-3 expresses the rate of die-off of microorganisms as an empirical first order kinetic model and is commonly referred to as Chick's Law. Chick's Law is presented graphically in Figure 4-1. Divergences from exponential decay are commonly observed, and it is recognized that many factors can cause these deviations, such as changes in disinfectant concentration with time, differences in resistance between individual organsims of various ages in the same culture, existence of clumps of organisms, occlusion of organisms by suspended solids, etc.

Watson analyzed data with varying concentrations of disinfectant and demonstrated a definite logarithmic relationship between concentration of disinfectant and the mean reaction velocity (5). He proposed the following equation to relate the rate constant of inactivation to the disinfectant concentration:

$$k = k'C^n \qquad (4\text{-}4)$$

where:

C = disinfectant concentration

n = coefficient of dilution

k' = corrected die-off rate constant, presumably independent of disinfectant concentration and organism concentration

The process of disinfection is influenced by temperature, and the Arrhenius equation can be used to predict temperature effects when direct heat-kill is not a significant factor:

$$k'_T = k'_{20}\, B^{(T-20)} \qquad (4\text{-}5)$$

where:

k'_T = rate constant at some temperature T (°C)

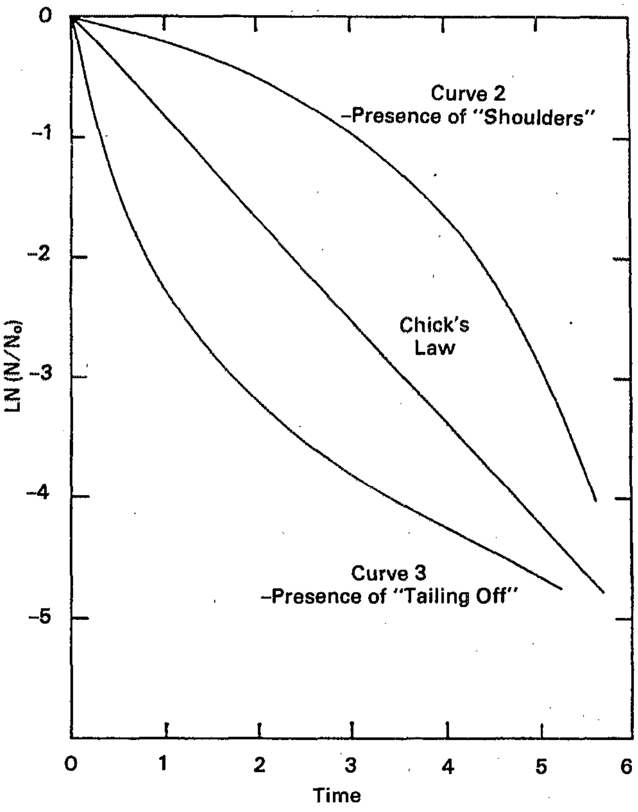
k'_{20} = rate constant at 20°C

B = empirical constant

Little is known about disinfection efficiency at elevated temperatures, but for agents such as ozone, a significant reduction would occur due to the lower efficiency of ozone mass transfer and greater ozone decay.

The observation has frequently been made that inactivation of organisms in batch experiments, even when disinfectant concentration is kept constant, does not follow the exponential decay pattern predicted by Equation 4-3. Two common types of deviation are noted, as shown in Figure 4-1. In addition to the linear Chick's Law decay, the presence of "shoulders" or time lags until the onset of disinfection, is often observed (Curve 2 in Figure 4-1). Also, some organisms and disinfectants exhibit a "tailing" wherein the rate of inactivation progressively decreases (Curve 3 in Figure 4-1). In some cases, a combination of both of these behaviors can even be observed.

Figure 4-1. **Chick's Law and deviations.**



The presence of "shoulders" can be accounted for if it is assumed that multiple targets within each organism must each be damaged by independent disinfectant molecules prior to kill (6). An alternative explanation is that diffusion through microbial outer layers, or binding of disinfectant to the cell is a necessary precursor to inactivation (7). The "tailing" survivor curve may be explained (in the absence of changes in disinfectant concentration with time) by either inherent differences in sensitivities of organisms present, by nonuniformity in a spatial sense (e.g., some organisms protected by enmeshment in solids), or by the induction (chemical or biological) of resistance in survivors over the time course of exposure to disinfectant (8). The tailing phenomenon has been explained by Poduska and Hershey (9), who assumed different subpopulations with differing sensitivities, and by Hom (10), who developed a flexible, but highly empirical kinetic formulation based on modifying Equations 4-3 and 4-4 to the following form:

$$\frac{dN}{dt} = -kNt^mC^n \qquad (4\text{-}6)$$

Depending upon the value of m, both "shoulders" and "tailing" may be depicted by Equation 4-6.

4.1.3. Summary of Kinetic Considerations

In reality, the rate of bacterial kill generally increases or decreases with time. This can be attributed to several factors such as oxidation, complexation, poor mixing, cell resistance, dispersion, clumping, and others. Deviations from Chick's Law can generally be described by modifying the first-order expressions. It must be emphasized that all the reported disinfection kinetic equations are empirical in nature. Unfortunately, Chick's Law does not accurately predict coliform numbers as a function of dose in real world, continuous flow systems, and therefore, the kinetics of disinfection, as with any process, must be determined experimentally. As a result, pilot studies are required for effective design of any disinfection system. But pilot studies are expensive and many times not conducted. Therefore, the need exists for compiling all the available data and information on disinfection to address this problem.

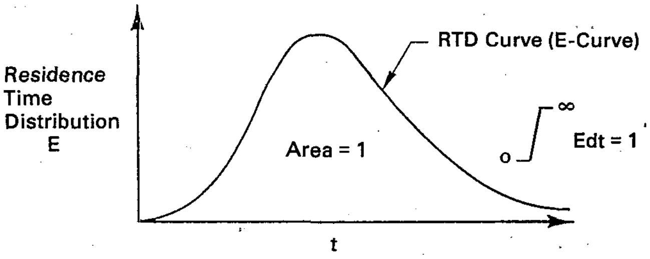
4.2 Mixing and Contactor Hydraulics

In dealing with the hydraulic behavior of any continuous flow reactor, the ideal situation is one in which the velocity profile is known at any point within the reactor. This most often is not possible; the flow characteristics are instead partially described by the residence time distribution (RTD). If one considers the time each element resides in the reactor, the frequency distribution of these times when plotted against the time forms the residence time distribution curve. This is presented graphically in Figure 4-2. By normalizing the distribution, the RTD can be represented in such a way that the area under the curve is equal to unity:

$$\int_0^\infty Edt = 1 \qquad (4\text{-}7)$$

where E = the residence time distribution.

Figure 4-2. Presentation of the residence time distribution (RTD).



One should note that in these discussions, it is assumed that the system is at steady-state, that there are no reactions taking place (conservative tracer), and that the density of the liquid is uniform throughout the reactor.

Given an E curve, the system mean residence time may be determined using:

$$\theta = \int_0^\infty tEdt \qquad (4\text{-}8$$

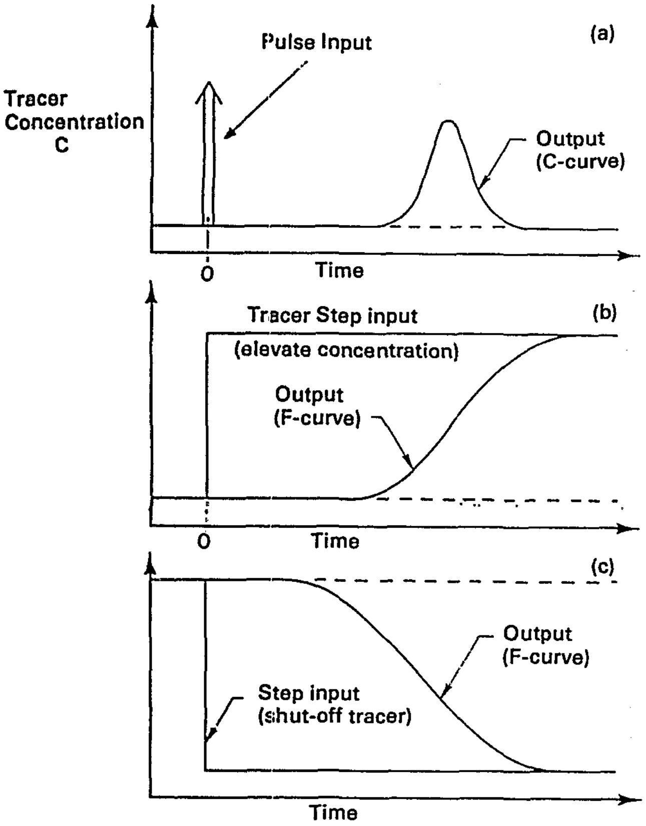
It is noted that this definition is similar to that for a moment of inertia. However, knowing the value for θ, an infinite number of E curves may still exist, differing in the spread of actual residence times. It can readily be shown that θ is identical to V/Q in systems such as those considered here (11). The determination of the system RTD provides information on the numerical spread of actual residence times, otherwise referred to as the dispersion.

Thus, the evaluation of a specific reactor relies on the experimental determination of RTD for that reactor. This can be accomplished by a number of procedures; subsequent analysis of the RTD curves may be used to characterize mixing behavior of the unit.

4.2.1 Experimental Determination of the Residence Time Distribution Curve

The means to develop the RTD of a reactor is generally referred to as a stimulus-response technique. The state of the reactor is perturbed in some fashion and the resulting response is observed. In this case, the stimulus is the input of a conservative (non-reactive) tracer to the fluid entering the vessel and the response is the time record of the tracer leaving the vessel. The method by which the tracer is introduced to the fluid will influence the type of response seen at the outlet of the vessel. Typically, direct experimental tracer analyses entail two types of inputs. The first is the pulse input in which the tracer is injected in a relatively short period of time to the influent. This is shown on Figure 4-3(a); the response is referred to as a C-curve.

Figure 4-3. **Representation of pulse and step inputs and resulting outputs.**



The second method is the step input. In this case, the tracer is introduced as a constant input, i.e., the fluid is adjusted from one steady-state concentration level to a new steady-state concentration level. The response is the time record of the concentration in the effluent reaching this adjusted level, as shown on Figure 4-3(b) and (c). This response is referred to as an F-curve. The step input can be applied in either direction: on Figure 4-3(b), it is shown as an input to increase the concentration; the tracer can also be applied as a steady input and then shut-off, with the response as illustrated by Figure 4-3(c). The E-curve (Figure 4-2) and the C-curve are equivalent:

$$E = C \qquad (4\text{-}9)$$

The F-curve can be transformed to the C-curve by taking its derivative:

$$dF/dt = C \qquad (4\text{-}10)$$

The utility of this equation will be demonstrated in a later discussion.

Experimentally, the development of the RTD curve is fairly straight-forward. A substance that is conservative in the carrying fluid is injected into the influent of the reactor, either as a pulse or as a step input. Suitable tracers are typically a salt (NaCl or LiCl) or a fluorescent dye. These are then measured directly in the effluent by rapidly taking aliquots for subsequent analysis or by direct instream measurements (e.g., conductivity). For RTD determinations in chlorine or ozone systems, batch sampling techniques suffice because of the long system mean residence times.

The nature of typical UV system designs presents logistical problems to the experimental procedures for the tracer studies. These arise from the very short average residence times typical of most UV systems, and in some cases to the difficulty of injecting the tracer at an appropriate point in the approach to the reactor. The average residence times generally encountered in these systems are between 1 and 50 seconds; at the shorter times it is difficult to collect samples manually at frequent enough intervals to construct a good time record of the effluent concentration. High frequency automatic samplers and a very rapid injection of the tracer at the upstream portion are generally required. Johnson and Qualls (12,13) described such a system for a pulse injection and sampling method. The method is generally appropriate where there are distinct inlet and outlet pipes on the UV reactor (see Figure 4-4(a)). The tracer is injected quickly into the inlet pipe: the time of injection must be small with respect to the average residence time of the reactor and there must be adequate mixing of the tracer and wastewater prior to entry into the reactor. Samples should be drawn immediately after exiting the reactor; alternatively, conductivity can be measured directly if salt is used as the tracer.

A step injection method was demonstrated in the Port Richmond study (14). The reactors in this evaluation were open vessels, in effect simulating open channel flow. Thus, there were no distinct inlet or outlet constrictions in which to inject or measure the tracer. Additionally, the times were very short, such that direct sampling of the effluent would be difficult. The technique that was developed (Figure 4-4(b)) incorporated the use of the F-curve; the step input was generated by first injecting a steady-state stream of tracer into the unit and then shutting off the injection. A conductivity probe was used to search for the point of maximum concentration on the exit side of the lamp battery. Note that the probe can also be used to scan the entire exit plane in order to define the conductivity profile and location of the plume as it exits the lamp battery.

Once the probe is situated and fixed at the center of the plume, a steady-state condition is allowed to develop at fixed wastewater and salt solution flow rates. The high frequency output from the probe is

Figure 4-4(a). **Techniques for experimentally determining the RTD curves of UV reactors. (closed vessel analysis)**

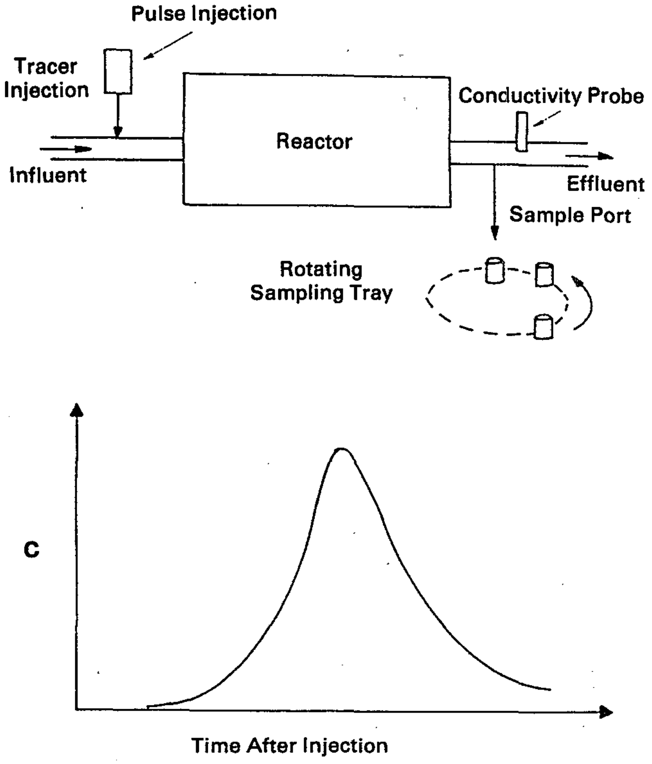
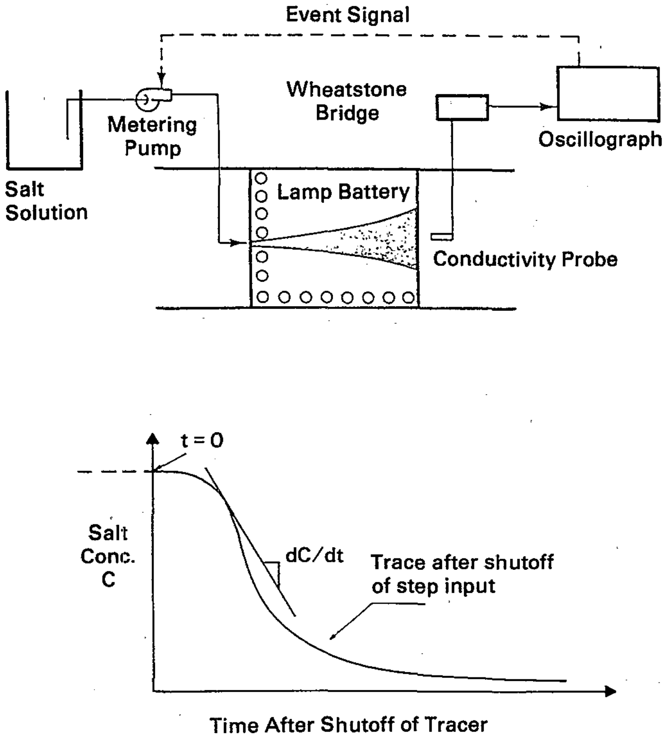


Figure 4-4(b). **Techniques for experimentally determining the RTD curves of UV reactors. (open vessel tracer analysis)**



amplified and recorded by an oscillograph in this case; it is necessary to have a high speed recorder in order to obtain a readable and undistorted trace in the short time frames. Once steady-state is indicated by the recorder, the salt solution pump is shut-off. This event should be automatically signalled and recorded. The die-away of the salt tracer is then monitored by the fixed probe and continuously recorded. Readings (conductivity) can then be taken of the tracer, converted to salt concentration (mg/l), and transposed to a plot of concentration against time. For a particular type of system this can also be repeated at several points across the plane of the lamp battery.

The step input method is generally applicable to any type of reactor system. It is particularly useful in situations where there is a short average residence time and where direct in-line monitoring of the tracer is possible.

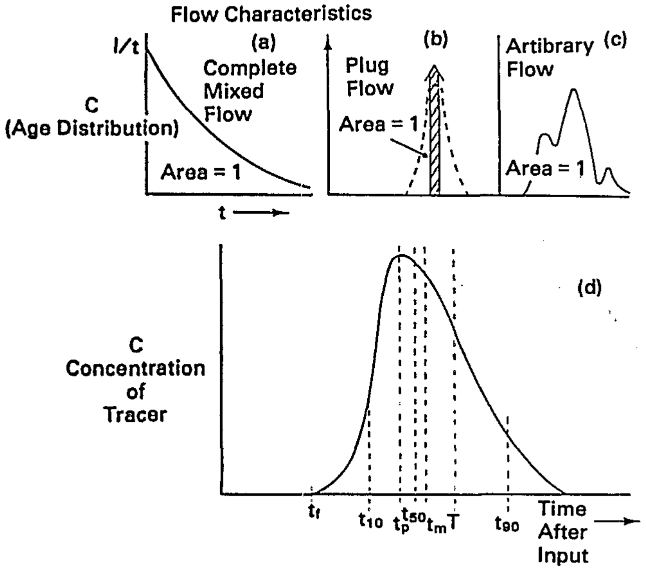
4.2.2 Analysis of Residence Time Distribution Curves

There are a number of uses for the RTD as a tool for design and as a diagnostic to determine the effect of the hydraulics on the system's performance. The shape of the curve and the distribution of the area under the curve will describe much of the hydraulic characteristics of a system and indicate if it conforms with proper design for a disinfection process.

Consider the RTD curves on Figure 4-5, for three different flow regimes. The responses shown are all to a pulse input. The worst case for a disinfection reactor is shown as a complete mixed flow reactor (Figure 4-5(a)). A pulse input enters the reactor and is completely mixed in the total volume of the reactor instantaneously. This implies that a fraction of the input will be immediately discharged without any real time of contact with the disinfectant. This is a wholly unacceptable condition for a disinfection process. Recall that the inactivation of microorganisms requires the time element; thus, in a completely mixed reactor, a significant fraction of the microorganisms in the wastewater will exit with little chance of being inactivated.

The ideal case is the pure plug flow (Figure 4-5(b)). The response to the pulse input is a spike of essentially zero width. This implies that every element of the pulse input resides in the reactor for an equal amount of time. This ideal situation is not achieved in actual applications; there will be some degree of dispersion such that the width of the plug flow C-curve will expand. The objective of the hydraulic design aspect of a system will be to minimize the degree of this spread in the RTD curve.

Figure 4-5. Examples of RTD curves for various flow characteristics.



A third flow condition is displayed (Figure 4-5(c)). This is in a sense an arbitrary response, having little similarity to a uniform flow condition. Such a curve is indicative of different volume elements of the reactor yielding different flow responses; examples are channeling, eddies, or quiescent zones. Such a response from a contactor would suggest potential problems.

A number of numerical characteristics can be used to describe the RTD curve and particularly its spread for a particular system. These can then be used to determine the adequacy of its design. The C-curve on Figure 4-5(d) is a hypothetical response to a pulse input for a given reactor. First, define the theoretical detention time of the reactor as the void volume V_v divided by the rate of flow, Q:

$$T = V_v/Q \tag{4-11}$$

The mean residence time, θ, is the mean value, or the centroid of the distribution:

$$\theta = \int_0^\infty tC\,dt \Big/ \int_0^\infty C\,dt \tag{4-12}$$

This can be approximated by summing the areas of discrete time intervals:

$$t_m = \frac{\Sigma\, t_i C_i \Delta t_i}{\Sigma\, C_i \Delta t_i} \tag{4-13}$$

The mean of the RTD curve is the first moment; the second moment of the curve about its mean is the variance. Whereas the mean locates the RTD, the variance will represent the spread of the distribution:

$$\sigma^2 = \frac{\Sigma\, t_i^2 C_i \Delta t_i}{\Sigma\, C_i \Delta t_i} - t_m^2 \tag{4-14}$$

The parameters shown on Figure 4-5(d) are defined as follows:

t_f = time at which the tracer first appears
t_p = time at which the peak concentration of the tracer is observed
t_{10} = time at which 10 percent of the tracer had passed through the reactor
t_{50} = time at which 50 percent of the tracer had passed through the reactor
θ = the meantime (centroid) of the curve
T = the theoretical mean time (V_v/Q)
t_{90} = the time at which 90 percent of the tracer had passed through the reactor

These parameters describing the RTD curve can be combined to a series of indices that in turn describe the hydraulic behavior of the reactor (15). In the evaluation of reactors specifically designed for disinfection, the critical areas of the RTD are the initial portions and their relationship to the theoretical and the mean residence times. Referring to Figure 4-5(d):

t_f/T ... Measures the most severe short-circuiting. In an ideal plug flow reactor, the ratio is one, and approaches zero with increased mixing.

t_p/T ... Measures the average degree of short-circuiting and will indicate the presence of significant dead space areas. This in turn provides an estimate of the effective volume of the reactor. This ratio will approach one in a plug flow reactor and zero with

t_{90}/t_{10} ... This is the Morrill Dispersion Index, and is the ratio of the time for 90 percent of the tracer to pass to the time for 10 percent of the tracer to pass. It is a measure of the spread of the curve; a value of 1.0 would indicate the ideal plug flow case, and 21.9 for ideal mixing. Reactor designs should allow for this index to be less than 2.0 for an effective design.

θ/T ... For any reactor, this should equal 1.0; this would imply that full use is being made of the volume. When this value is significantly less than 1.0, it indicates that the effective volume is much less than the actual volume. By definition, the ratio of θ/T cannot be greater than 1.0.

t_{50}/θ ... In an effective plug flow reactor, the RTD curve is very similar to a normal or Gaussian distribution. The ratio of the median (t_{50}) to the centroid (θ), will be a measure of the skew of the curve. A skew to the left, in which t_{50}/θ is less than 1.0, would be detrimental to an effective reactor design.

4.2.2.1 Dispersion Models

The above discussion centered on indices that could be readily obtained from the RTD to evaluate or diagnose the flow regime of a specific reactor. However, recalling that the primary design approach incorporates the use of a disinfection model, it is necessary to also evaluate the RTD as a means of estimating the dispersive characteristics of the system. To accomplish this, one can write a mass balance over the reactor for a conservative substance (at steady-state), where E is defined as the dispersion coefficient (length^2/time), u is the linear axial velocity, and x is the axial distance from the inlet.

$$\theta = \frac{E}{uX}\frac{(d^2C)}{dx^2} - \frac{dC}{dx} \qquad (4\text{-}15)$$

The dimensionless group, E/uX, is called the dispersion number (d), and is a measure of the axial dispersion in a reactor:

as $\frac{E}{uX} \rightarrow 0$; plug flow

as $\frac{E}{uX} \rightarrow \infty$; complete mix

Levenspiel (11) gives the solution to Equation 4-15 for pulse inputs in relatively low levels of dispersion. First define the dimensionless parameter, θ';

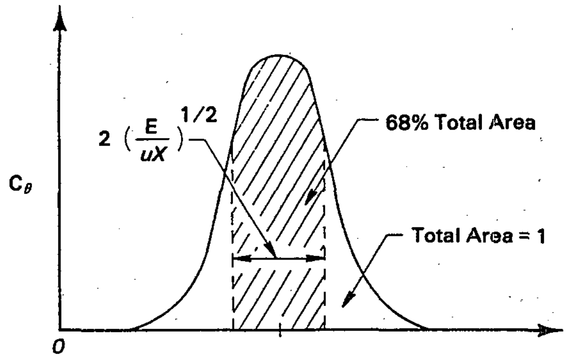
$$\theta' = \frac{t}{\theta}$$

which describes time in units of the mean residence time. In case of small values of d the solution of Equation 4-15 gives the symmetrical C-curve:

$$C_\theta = \frac{1}{2(\pi d)^{1/2}} \exp \frac{-(1-\theta')^2}{4(d)} \qquad (4\text{-}17)$$

As shown on Figure 4-6, E/uX (or d), is effectively the single parameter for the resulting C-curve. The mean and variance of this curve, C, are:

Figure 4-6. Relationship of C-curve to E/uX for small degrees of dispersion.



$$\theta'_{m,c} = \frac{t_{m,c}}{\theta} = 1 \qquad (4\text{-}18)$$

and

$$\theta_m^2 = \frac{\sigma^2}{\theta_m^2} = 2\left(\frac{E}{uX}\right) = 2d \qquad (4\text{-}19)$$

Thus, if the RTD closely approximates a normal distribution, the parameter d = (E/uX) can be obtained from the C-curve. It can be shown that minimal error (<5 percent) will be incurred by the above estimate when:

$$d < 0.01$$

In cases where a significant degree of dispersion exists, (i.e., d > 0.01), causing the C-curve to skew, Levenspiel gives the mean and variance for both closed and open vessels. Note that "closed" vessels are those in which fluid enters and exits by plug flow. In open vessels, flow is undisturbed at the boundaries of the vessel. For closed vessels the mean and variance are:

$$\theta'_{m,c} = \frac{t_{m,c}}{\theta} = 1 \qquad (4\text{-}20)$$

$$\sigma_\theta^2 = 2\frac{E}{uX} - 2\left(\frac{E}{uX}\right)^2 (1 - e^{-uX/E}) \qquad (4\text{-}21)$$

In the case of open vessels with a significant degree of axial dispersion, the mean and variance are:

$$\theta'_{m,c} = \frac{t_{m,c}}{\theta} = 1 + 2\frac{E}{uX} \qquad (4\text{-}22)$$

$$\sigma_\theta^2 = 2\frac{E}{uX} - 8\left(\frac{E}{uX}\right)^2 \qquad (4\text{-}23)$$

Note that in situations of small E/uX, the RTD is not affected by boundary conditions. From Equation 4-22 it can be implied that at values of E/uX << 0.5, "open" and "closed" vessels are equivalent.

It is suggested that for closed vessels, i.e., plug flow entrance and exit reactors, different values of E/uX can denote the following degrees of dispersion: (11)

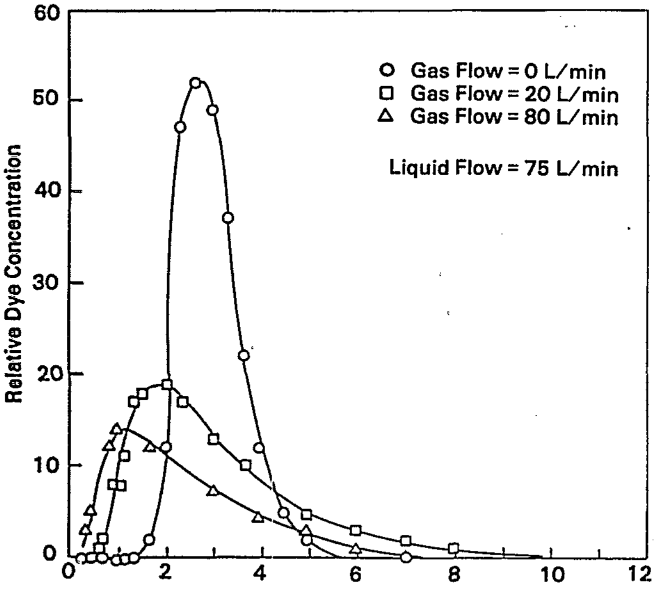
No Dispersion	E/uX = 0
Low Dispersion	E/uX = <0.01
Moderate Dispersion	E/uX = 0.01 to 0.10
High Dispersion	E/uX = >0.1

A distinct advantage of the use of the dispersion number for the characterization of mixing behavior is that several predictive models are available. In the chapter on chlorination, correlation for dispersion in pipes and baffled contact chambers will be presented. There are experimental measurements reported for UV and ozone systems; however, these have yet to be reduced to simple correlations.

The characterization of mixing in an ozone contact basin is more complex and is dependent on the type of contactor used. For example, with turbine contactors a high degree of mixing is a necessary part of the process of dissolving ozone into the wastewater. At the same time, the violent mixing helps bring the organisms into contact with the ozone. On the other hand, back-mixing can increase d, and care must be taken' to ensure that good baffling is provided to prevent short-circuiting.

With bubble diffuser ozone contact basins the mixing is much less violent than with turbine contactors. Even here, however, basin back-mixing in each stage of the contactor due to the flow of gas in that stage is sufficient to change the liquid flow characteristics from a plug flow toward a complete mix pattern. The tendency toward complete mix operation with increasing gas flow for bubble diffuser ozone contact basin was demonstrated by Venosa and Opatken (16) by performing dye tests with and without gas flow (Figure 4-7).

Figure 4-7. **Dye test shows effects of gas flow rate on plug flow characteristics through a bubble diffuser ozone contact basin.**



Each stage of a bubble diffuser contactor would exhibit similar characteristics; therefore, a similar potential for short-circuiting would exist. Also, each stage of other high mixing type reactors, such as the turbine mixer reactor, would exhibit characteristics of a complete mix reactor. To minimize the effect of short-circuiting, multiple stages that are positively isolated from each other should be provided.

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When an NTIS number is cited in a reference, that reference is available from:

National Technical Information Service
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Springfield, VA 22161
(703) 487-4650

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Chapter 5
Halogen Disinfection and Dechlorination

5.1 Coverage

This chapter covers the use of chlorine, chlorine dioxide, bromine chloride and related compounds for wastewater disinfection. In addition, design of systems for the dechlorination of wastewaters will be discussed.

The material in this chapter may be used in several ways. Sections 5.2 to 5.6 present an overview of the history of chlorination and the other halogens, and the fundamental chemical and kinetic aspects of these materials as disinfectants. The chemical aspects of dechlorination are also reviewed.

Section 5.7 covers in detail the specification and design of the various components comprising a halogen disinfection system, and a dechlorination system. The engineer may proceed directly to this section, referring to material in Sections 5.2 to 5.6 as background. For application to a particular case under consideration, the engineer may also use the design example presented at the end of Section 5.7 as a checklist in the construction of other calculations.

Section 5.8 treats safety considerations in the use of halogens, to which the design engineer must pay particular attention. Section 5.9 and the following material summarize important O&M considerations that should be addressed in the design stage, and relate performance experience at several full-scale treatment plants.

5.2 History of Halogen Disinfection

5.2.1 Chlorine and Hypochlorites

While chlorides in the form of salts were known by the ancients, the first preparation of chlorine gas, then known as "dephlogisticated marine acid air" is credited to Scheele who achieved this result from the reaction of manganese dioxide with hydrochloric acid in 1774. However, it was not until 1808 that this material was regarded by Davy as a chemical element, and general acceptance of Davy's hypothesis did not occur until 1815 (1).

Medical germicidal applications of the new compound soon followed, with hospital disinfection equipment installed as early as 1823, and with the use of chlorine in surgical applications by Semmelweis in 1826 (1). The first public health or environmental application of chlorine appears to have been its use as a prophylactic agent during the European cholera epidemic of 1831 (1), although Baker (2) asserted that Javelle water (chlorine gas dissolved in an alkaline potassium solution) was used in France for waste treatment as early as 1825.

The earliest U.S. reference to chlorine as a disinfectant occurs in 1832, when the known germicidal properties of chlorine for control of disease epidemics were summarized by Averill (3). It is particularly noteworthy that these results were obtained approximately 50 years before the advent of the germ theory of disease.

The first formal recognition of chlorine or chlorine compounds for wastewater treatment occurred in 1854, by the English Water Commission (2), although not until 1884 was wastewater in England actually chlorinated (4).

During the last years of the nineteenth century, electrolytic generation of chlorine and hypochlorites became sufficiently competitive with the chemical oxidation synthesis routes to spur the use of chlorine compounds for disinfection. The growth of chlorine disinfection for both water and wastewater applications occurred simultaneously.

The three earliest applications of chlorine as a wastewater disinfectant used patented on-site electrolytic generation. In 1892, operations commenced at both Hamburg, Germany and Brewster, New York (4).

The electrolytic generation of chlorine for wastewater treatment was covered by the patenting of the Woolf Process in 1893 (5), which superseded the earlier Powers process for the chemical production of chlorine by a method similar to Scheele's original procedure. Many variants on the electrolytic process were reported up to and including the work of Rideal in 1908.

Systematic investigations of the efficiency of chlorine disinfection of wastewater were carried out in the first decade of the twentieth century. Three research groups reported on studies conducted almost contemporaneously.

During 1907 and 1908, Phelps conducted laboratory and field investigations of chlorination using chloride of lime. Field chlorination at Boston, Baltimore and Red Bank, NJ led to the conclusion that a dose of several milligrams per liter chlorine and 15 minutes contact time resulted in effective microbial reductions. Furthermore, Phelps asserted that it was infeasible to attain complete microbial reduction using chlorination. In his work, Phelps used the coliform organism as an indicator of disinfection efficiency. As a justification for this, he demonstrated that these organisms had equivalent sensitivity to the typhoid bacillus in wastewater chlorination (6).

Kellerman et al. (7) conducted two field studies on the chlorination of (slow) sand filter effluents in Ohio and came to conclusions similar to Phelps. In addition, this latter work estimated costs for chlorination at two particular plants studied as $7.50 and $16/million gallons ($2.00 and $4.25/1,000 m^3) using chlorinated lime. Futhermore, it was concluded that "the quantity of chlorine immediately absorbed cannot be estimated from the determination of the oxygen consumed factor of the sewage effluent," thus contradicting the hypothesis advanced by Rideal (8).

The third set of initial investigations was conducted by Clark and Gage (9) at the Lawrence, MA, experiment station. In addition to corroborating the studies at Ohio and by Phelps, these workers appear to have been the first to investigate the phenomenon of post-disinfection regrowth. Their cost estimates for the use of calcium hypochlorite for disinfection were $1.75 to $3.75/million gallons ($0.45 to $1.00/1,000 m^3).

Following these early studies, a rapid introduction of chloride of lime disinfection into wastewater treatment plants occurred. While a commission of the City of Milwaukee, WI recommended, in 1909, that a primary treatment/chlorination plant be built, the first major city to install chlorination appears to have been Philadelphia, PA in October 1910 (10). In addition, Baltimore, MD commenced chlorination in February of 1912 and Providence, RI commenced chlorination of a 20 MGD (0.9 m^3/s) stream that same year (10,11). As of 1913, chlorination facilities using chloride of lime were installed at Red Bank, NJ; Rahway, NJ; Shore Harbor, NJ; Ventnor, NJ; Atlantic City, NJ; Bridgeton, NJ; Keyport, NJ; Margate City, NJ (10); and at Pleasantville, NY (9) and College Park, CA (12).

The rise in chlorine treatment of wastewater was paralleled by the use of chloride of lime in water disinfection. The first applications of this technology occurred in 1908 at Boonton, NJ and Bubbly Creek (Chicago), and in 1909 at Poughkeepsie, NY (13).

While Phelps, Kellerman, and the Lawrence, MA group contributed to knowledge concerning the application of chlorinated lime to wastewater disinfection, perhaps the first serious study of the kinetics of wastewater disinfection, per se, was that of Avery in 1913 (14). In contrast to prior work, and particularly, that of Phelps, Avery concluded that it was important to achieve some minimum amount of contact time between the points of chlorine addition and discharge. He suggested that this time be at least 20 minutes. Furthermore, Avery's studies were significant in noting the relative difficulty of disinfecting septic, as opposed to fresh, wastewater.

The use of solid hypochlorite compounds as disinfectants was superseded following the development of practical and economic means of generating and dispensing liquid chlorine. In 1909, production of liquid chlorine commenced at Niagara Falls, NY, by Electro Bleaching Gas Co. In 1912, Dr. Georg Ornstein of this company developed and patented a device for the application and feeding of chlorine in solution form. Simultaneous to this, C.F. Wallace and M.F. Tiernan developed and installed a direct feed gas chlorinator at the Boonton, NJ water works, and went on to develop solution feed chlorinators. The acquisition and abandonment of the Ornstein patent rights by W&T in 1917 finalized the adoption of modern solution feed liquid chlorination treatment for both water and wastewater (13).

In wastewater treatment, the liquid chlorine direct feed process was used as early as 1918 in Millville and Camp Merritt, NJ (14). The W&T solution feed liquid chlorination process was used at New Haven, CT commencing in 1919, where costs were estimated at $4/million gallons ($1.00/1,000 m^3) (15).

In light of the apparently rapid adoption of chlorination for wastewater disinfection it is also of interest to note that reservations were expressed then, some of which are still relevant to practice today. For example, as noted above, the issue of regrowth of bacteria subsequent to chlorination and release was noted by Clark and Gage at Lawrence (9). In addition, as early as 1922 (16) reservations were expressed about the relative resistance of coliforms to chlorine vis a vis pathogenic bacteria, and the resulting adequacy of the coliform test as an indicator of disinfection efficiency.

The genesis of modern chlorination practice is seen in the experiences at the Easterly and Westerly plants of the city of Cleveland, OH (17). At these plants, with average flows of 90 and 30 mgd (3.9 and 1.3 m^3/s), respectively, disinfection using liquid chlorine was commenced in 1923, using heated water evaporators. In 1929, at Easterly, one-ton container handling systems were installed, while at Westerly, the 15-ton unit tank car system was placed on line. At these two plants, over the period 1923 to 1929, costs averaged $4.09 and $5.50/million gallons ($1.08 and $1.45/1000 m^3), respectively. The final stage in development of modern wastewater chlorination concepts was the demonstration of the validity of operational control based on measurement of the chlorine residual by Tiedeman (18).

The geographic and temporal spread of wastewater disinfection by chlorine is illustrated in Tables 5-1 and 5-2, which indicate the geographic distribution of chlorination plants in 1910 and 1916, and the growth in chlorination from 1910 to 1957.

Table 5-2. Development of Chlorination Installations for Wastewater Treatment after Laubusch (5, 20)

	Wastewater Treatment Plants		% of Population with Wastewater Chlorination	
Year	Total	With Chlorination	Of Those with Waste Treatment	Of Total U.S. Population
1910	619	22	2.4	0.12
1916	846	55	4.6	0.28
1934	3697	655	N/A	N/A
1940	5580	1127	35.3	10.9
1945	5786	1262	34.2	12.1
1948	6058	1307	37.0	12.3
1957	7518	2216	49.5	N/A

N/A not available.

5.2.2 Chlorine Dioxide

The discovery of chlorine dioxide, produced from the reaction of potassium chlorate and hydrochloric acid, is attributed to Davy in 1811 (21). However, it was not until the industrial scale preparation of sodium chlorite, from which chlorine dioxide may more readily be generated, that its widespread use occurred (22).

Chlorine dioxide has supplemented and supplanted chlorine as a bleaching agent in pulp and paper manufacture (22); however, despite early investigations on the use of chlorine dioxide as an oxidant and disinfectant (23), its ascendancy in both water and wastewater treatment has been slow.

As of 1977, 84 potable water treatment plants in the United States used chlorine dioxide treatment, although only one of these relied upon it as a primary disinfectant (21). In Europe, chlorine dioxide is used as either an oxidant or disinfectant in almost 500 potable water treatment plants (21).

While there have been numerous laboratory and pilot plant investigations of chlorine dioxide disinfection of wastewater, there does not appear to have been any full-scale operating experience with this disinfectant in wastewater treatment. The situation in Europe is not known in detail; however, White cites one unpublished study of chlorine dioxide at a full scale wastewater treatment plant in France (24).

5.2.3 Bromine Chloride and Bromine

Due to the relative economics of bromine and chlorine, little interest in the former compound as a disinfectant was shown until the Second World War, when shortages of chlorine and increased supplies of bromine as a byproduct occurred (24). McCarthy demonstrated that, as a potable water disinfectant, elemental bromine was of similar effectiveness as chlorine (25). In a more detailed study, Wyss and Stockton demonstrated the relative insensitivity of bromine disinfection to high ammonia nitrogen concentrations (26). These results were confirmed by Johannesson (27), who subsequently attributed this finding to chemical properties of bromamines (28).

While the existence of bromine chloride has been known since the work of Balard in 1826, final equilibrium coefficients were not well defined until the work of Mattraw et al. (29). However, little interest in the industrial applications of bromine chloride were evidenced until work by Mills and associates at Dow Chemical (30-32).

In the mid 1970s, several investigations of the effectiveness of bromine chloride in laboratory and pilot plant disinfection of wastewaters were reported.

Table 5-1. Early Geographic Distribution of Chlorination Facilities (19)

	1910		1916	
State	No. of Plants	% of Serviced Population with Chlorination	No. of Plants	% of Serviced Population with Chlorination
California	1	0.7	1	0.7
Connecticut	0	0	1	5.9
Maryland	2	0.7	2	0.4
New Jersey	6	0.7	16	38.8
New York	1	0.4	1	0.2
North Carolina	0	0	1	8.2
Ohio	1	0.4	2	1.7
Pennsylvania	4	7.0	5	9.9
Texas	6	3.8	25	15.0

These included pilot or demonstration studies at Hampton Roads, VA (33,34), Freedom District (Sykesville), MD (35,36), Hatfield, PA (37), Hawaii (38-41), and Michigan (42,43).

The number of full scale operating wastewater disinfection systems using bromine chloride is unknown, although a number of bromine chloride installations for power plant condenser biofouling control exist (24).

5.3 Chemistry and Physical Characteristics of Disinfectants

The fundamental mechanisms of action of disinfectants and the problems raised in the design of wastewater disinfection systems can be related, in many cases, to the physical properties of the disinfectants themselves, and their chemical reactions with other constituents that may be present. In this section, these properties will be reviewed.

5.3.1 Properties of Disinfectants

5.3.1.1 Chlorine and Hypochlorites

Elemental chlorine Cl_2 is a gas of density greater than air at room temperature and pressure. When compressed to pressures in excess of its vapor pressure, chlorine condenses into a liquid with the release of heat and with a reduction in specific volume of approximately 450 fold. Hence, commercial shipments of chlorine are made in pressurized tanks to reduce shipment volume. When chlorine is to be dispensed as a gas, it is necessary to supply thermal energy to vaporize the compressed liquid chlorine.

Table 5-3 summarizes the major physical properties of chlorine (44). Figure 5-1 summarizes the vapor pressure of chlorine as a function of temperature. These data are obtained from a publication of the Chlorine Institute (45), from which more extensive tabulations of the physical properties of gaseous and liquid chlorine may be obtained.

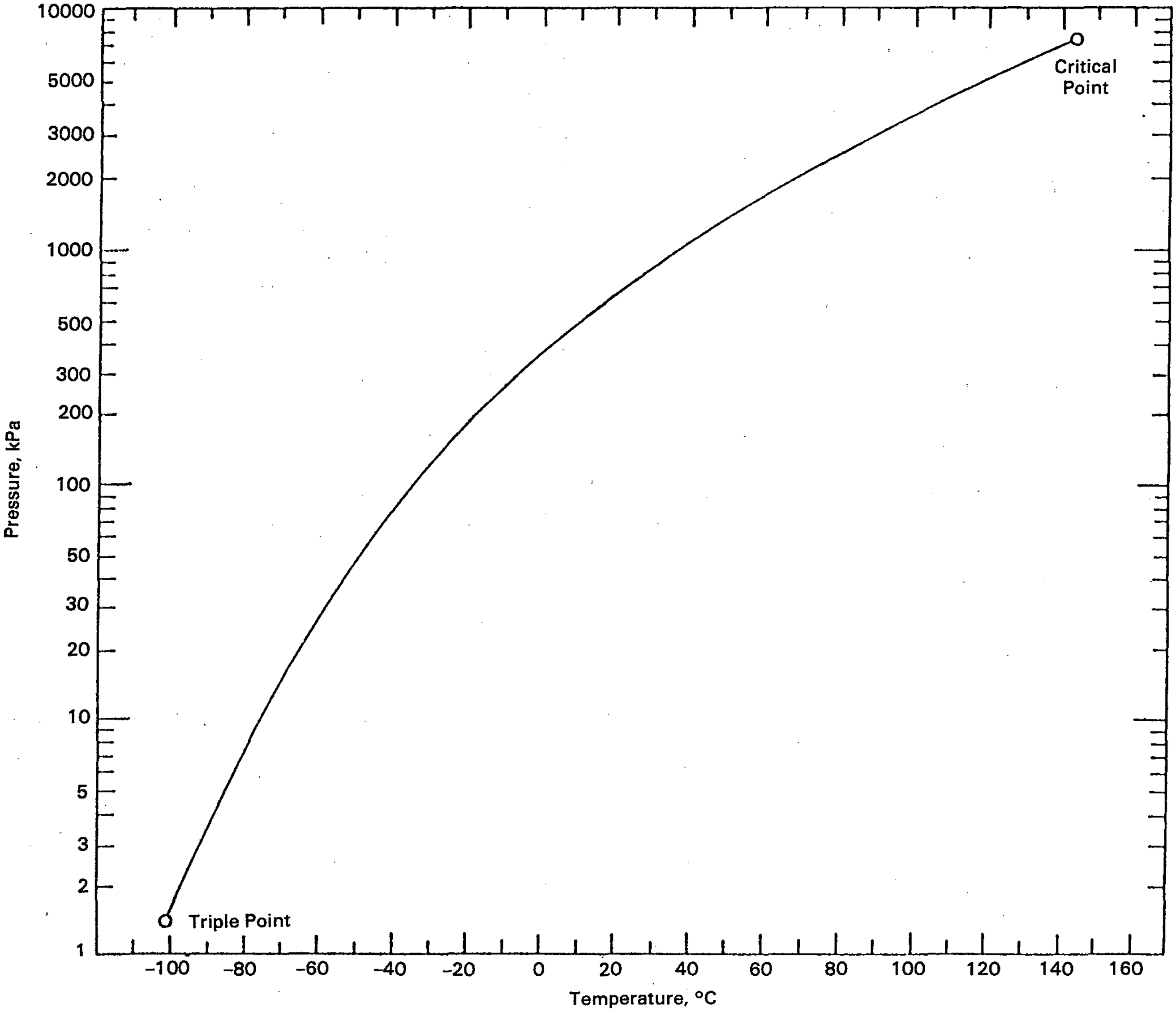
Commercially, sodium hypochlorite and calcium hypochlorite are also used as sources of chlorine compounds for wastewater disinfection. The relative amount of chlorine present in these alternative sources of chlorine is expressed in terms of "available chlorine."

The concentration of hypochlorite (or any other oxidizing disinfectant) may be expressed as available

Table 5-3. Physical Properties of Chlorine (44)

	Liquid	Gas
Affinity for Water	Slight	Slight
Boiling Point (@ 1 atm)	−34.05°C (−29.3°F)	
Color	Clear amber	Greenish-yellow
Corrosivity	Extremely corrosive to steel in presence of small amounts of moisture. See discussion.	Same as liquid
Density, lb/cu ft	88.79 @ 60°F (85.61 psia)	0.2003 @ 32°F and 1 atm
Explosive limits (in air)	Non-explosive	Non-explosive
Flammability	Non-flammable	Non-flammable
Melting (freezing) Point (@ 1 atm)	−100.98°C (−149.76°F)	Non-flammable
Odor	Penetrating and irritating	Same as liquid
Solubility	---	Below 9.6°C (49.3°F)
Chlorine hydrate		Cl_2:$8H_2O$, may crystallize
Specific gravity (compared to 4° water)	1.468 @ 32°F and 3.617 atm	---
Relative Vapor Density (Air = 1)	---	2.482 @ 32°F and 1 atm.
Vapor pressure	See Figure 5-1	---
Viscosity	0.385 centipoise @ 0°C 0.729 centipoise @ −76.5°C	167.9 micropoise @ 100°

Figure 5-1. Vapor pressure of liquid saturated chlorine gas. (45) (Reprinted by permission of The Chlorine Institute.)



chlorine by determining the electrochemical equivalent amount of Cl_2 to that compound. By Equation 5-1, it can be seen that one mole of elemental chlorine is capable of reacting with two electrons to form inert chloride:

$$Cl_2 + 2e^- = 2\ Cl^- \qquad (5\text{-}1)$$

From Equation 5-2, it can also be noted that one mole of hypochlorite (OCl^-) may react with two electrons to form chloride:

$$OCl^- + 2e^- + 2H^+ = Cl^- + H_2O \qquad (5\text{-}2)$$

Hence, one mole of hypochlorite is equivalent (electrochemically) to one mole of elemental chlorine, and may be said to contain 70.91 grams of available chlorine (identical to the molecular weight of Cl_2).

Since calcium hypochlorite ($Ca(OCl)_2$) and sodium hypochlorite ($NaOCl$) contain two and one moles of hypochlorite per mole of chemical, respectively, they also contain 141.8 g and 70.91 g available chlorine per mole. The molecular weights of $Ca(OCl)_2$ and $NaOCl$ are, respectively 143 and 74.5, so that pure preparations of the two compounds contain 99.2 and 95.8 weight percent available chlorine; hence they are effective means of supplying chlorine for disinfection purposes.

Commercially, calcium hypochlorite is available under a variety of trade names (HTH, Pittchlor, Perchloron) as a dry solid. It is relatively stable in the dry form, subject to a loss in strength of approximately 0.013 percent/d (46).

Sodium hypochlorite is available in solution form commercially in strengths of 1 to 16 percent by weight. It is not practical to provide higher solutions since chemical stability rapidly diminishes with increasing strength. At ambient temperatures, the half-life of sodium hypochlorite solutions varies between 60 and 1700 days, respectively, for solutions of 18 and 3 percent available chlorine (46,47).

When either chlorine gas or one of the hypochlorite compounds is added to a water containing insignificant quantities of Kjeldahl nitrogen, organic material, and other chlorine demanding substances, a rapid equilibrium is established among the various chemical species in solution. The term "free available chlorine" is used to refer to the total of the concentrations of molecular chlorine (Cl_2), hypochlorous acid (HOCl) and hypochlorite ion (OCl^-), each expressed as "available chlorine" as described previously.

The dissolution of gaseous chlorine to form dissolved molecular chlorine is expressible as a phase equilibrium, and may be described by Henry's Law:

$$Cl_{2(g)} = Cl_{2(aq)} : H\,(mol/L\text{-}atm) = [Cl_{2(aq)}]/P_{Cl_2} \qquad (5\text{-}3)$$

In Equation 5-3, quantities within square brackets represent molar concentrations, P_{Cl_2} is the gas phase partial pressure of chlorine in atmospheres, and H is the Henry's Law constant.

The dissolved aqueous chlorine is capable of reacting with water to form hypochlorous acid, chloride ions, and protons as indicated by:

$$\begin{aligned} Cl_{2(aq)} + H_2O &= H^+ + HOCl + Cl^- \\ K_H &= [H^+]\,[HOCl]\,[Cl^-]/[Cl_{2(aq)}] \end{aligned} \qquad (5\text{-}4)$$

The kinetics of this reaction have been discussed in detail by Morris and have been found to be extremely rapid (47). The reaction mechanism appears to involve elementary reactions between dissolved molecular chlorine and hydroxyl ions.

The extent of chlorine hydrolysis, or disproportionation, as described by Equation 5-4, is dependent upon the pH and the salinity of the solution. The extent of reaction decreases with decreasing pH and increasing salinity; hence, the solubility of gaseous chlorine may be increased by the addition of alkali or by the use of fresh, rather than brackish water.

Hypochlorous acid is a weak acid and may dissociate according to:

$$\begin{aligned} HOCl &= OCl^- + H^- \\ K_A &= [OCl^-]\,[H^+]/[HOCl] \end{aligned} \qquad (5\text{-}5)$$

The pK_A of hypochlorous acid at room temperature is approximately 7.6 (48).

The three equilibrium constants describing the free chlorine system are each a function of system temperature. Morris has provided a correlating equation for K_A as a function of temperature (48):

$$\ln(K_A) = 23.184 - 0.0583\,T - 6908/T \qquad (5\text{-}6)$$

In Equation 5-6, T is specified in degrees Kelvin (°K = °C + 273).

For engineering purposes the chlorine equilibria in water may be adequately described using the Gibbs free energy and enthalpy at 298°K with the assumption of constancy of enthalpy over the practical range of interest. For each of the important species, Table 5-4 indicates the relevant thermodynamic quantities.

Using the standard thermodynamic relationships from the data in Table 5-4, the equilibrium constants for each of the chemical reactions governing free available chlorine may be calculated. These constants are summarized in Table 5-5.

Table 5-4. Thermodynamic Functions of Free Chlorine Species-298°K (49)

Species	Standard Gibbs Free Energy of Formation (G) (kJ/mol)	Standard Enthalpy of Formation (H) (kJ/mol)
Cl_2 (g)	0	0
Cl_2 (aq)	+6.90	−23.43
HOCl	−79.9	−120.92
OCl^-	−36.82	−107.11
Cl^-	−131.30	−166.94

Table 5-5. Equilibrium Constants for Free Chlorine (Applicability Approximately 283 – 308°K)

$K_A = 7.349 \times 10^{-6} \exp(-1660.89/T)$	(mol/l)
$K_H = 2.581 \exp(-2581.93/T)$	(mol^2/l^2)
$H = 4.805 \times 10^{-6} \exp(2818.48/T)$	(mol/l – atm)

Note: T is in degrees Kelvin.

From the use of these equilibrium constants it is possible to determine the relationship between gas phase partial pressure of chlorine (P_{Cl_2}) and solubility under two particular conditions of importance. If gaseous chlorine is dispersed into pure water with negligible amounts of demand, buffering potential, and chloride ion, then the following equation can be used to determine solubility:

$$S = P_{Cl_2}H + (K_H P_{Cl_2} H)^{1/3} \qquad (5\text{-}7)$$

where S is solubility in mol/l as Cl_2.

To determine the solubility as free available chlorine in mg/l multiply the latter quantity by 71,000.

However, if a solution of chlorine is prepared using water with significant buffering potential and/or chloride, or if a solution is prepared from calcium or sodium hypochlorite, both of which may contain chloride, then Equation 5-8 is appropriate. This would also be applicable to determine the partial pressure of chlorine above a solution containing the free chlorine species.

$$(5\text{-}8)$$

$$S = P_{Cl_2}H\,[1 + (K_H/[H^+][Cl^-]) + (K_A K_H/[H^+]^2[Cl^-])]$$

The use of these equations is illustrated by way of two examples.

Example 1:

A solution of chlorine is prepared by bubbling chlorine gas at 0.5 atm pressure into pure water until equilibrium is reached. The temperature is 54°F (285°K). Determine the solution concentration of free chlorine.

Step 1—Determine constants at the temperature of concern (285°K)

From Table 5-5:

$$K_H = 2.581 \exp(-2581.93/285) = 3 \times 10^{-4}\ \text{mol}^2/\text{l}^2$$

$$\begin{aligned} H &= 4.805 \times 10^{-6} \exp(2818.48/285) \\ &= 0.0948\ \text{mol/l-atm} \end{aligned}$$

Step 2—Apply Equation 5-7 to determine solubility:

$$\begin{aligned} S &= (0.50 \times 0.0948) + (0.50 \times 3 \times 10^{-4} \times 0.0948)^{1/3} \\ &= 0.0716\ \text{mol/l} \\ &= 5{,}084\ \text{mg/l free available chlorine} \end{aligned}$$

Example 2:

An effluent is used to make up a chlorine stock solution. The effluent is chlorinated until it contains 1,000 mg/l free available chlorine. At this point, it has a pH of 6.0 and a chloride ion concentration of 500 g/m^3 (this includes the initial chloride that was present). What would be the equilibrium pressure of chlorine in the gas phase? The temperature is 285° K.

Step 1—We may use the same values for K_H and H as the prior problem, since the temperature is the same. A value for K_A is needed, however.

From Table 5-5:

$$\begin{aligned} K_A &= 7.349 \times 10^{-6} \exp(-1660.89/285) \\ &= 2.16 \times 10^{-8}\ \text{mol/l} \end{aligned}$$

Also, since the pH = 6, $[H^+] = 10^{-pH} = 10^{-6}$ mol/l

Step 2—Apply Equation 5-8:

(1,000 mg/l free chlorine/71,000 = 0.0141 mol/l)

$$\begin{aligned} 0.0141 = P_{Cl_2} &\times 0.0948 \\ &\times [1 + (3 \times 10^{-4})/(10^{-6} \times 0.0141) + (3 \times 10^{-4}) \\ &\times (2.16 \times 10^{-8})/[(10^{-6})^2 \times 0.0141]] \end{aligned}$$

$$0.0141 = P_{Cl_2} \times 2061$$

$$P_{Cl_2} = 6.84 \times 10^{-6}\ \text{atm}$$

The second example illustrates an important aspect to chlorine chemistry, namely that adjustment of pH upwards (in this case to pH 6) results in a substantial diminution of the equilibrium gas pressure. This is of practical importance in the proper responses to spills of hypochlorite, or solutions of chlorine. Adjustment of pH, by the addition of an alkaline material such as lime or sodium bicarbonate will reduce the volatility of chlorine from such spills and thus minimize danger to exposed personnel.

5.3.1.2 Chlorine Dioxide

Chlorine dioxide (ClO_2) is a neutral compound of chlorine in the +IV oxidation state. It has a boiling point of 11°C at atmospheric pressure. The liquid is denser than water and the gas is denser than air (50). Relevant physical properties of chlorine dioxide are summarized in Table 5-6 (51).

Table 5-6. Physical Properties of Chlorine Dioxide (50)

Property	Value
Melting point	−59°C
Boiling point	11°C
Density (0°C)	1.640 g/ml (liquid) 2.4 g/l (vapor)
Heat of Vaporization	27.28 kJ/mol
Critical temperature	153°C
Vapor Pressure (0°C)	0.626 or 0.673 atm
Heat of Solution (0°C)	27.61 kJ/mol
Color	
- Solid	Red
- Liquid	Orange
- Gas	Orange

Chemically, chlorine dioxide is considered to be a stable free radical. Hence, at high concentrations it may react violently with reducing agents. It is explosive, with the lower explosive limit in the vapor phase variously reported as 10 percent (49,51) or 39 percent (50). As a result, virtually all applications of chlorine dioxide require the synthesis of the gaseous compound in a dilute stream (either gaseous or liquid) on location as needed.

The solubility of gaseous chlorine dioxide in water may be described by Henry's Law, and a fit of the available solubility data (52) results in the following relationship for the Henry's Law constant:

$$\begin{aligned} H(\text{mol/l-atm}) &= [ClO_{2(aq)}]/P_{ClO_2} \\ &= 3.819 \times 10^{-6} \exp(3680/T) \end{aligned} \qquad (5\text{-}9)$$

In dilute aqueous solution, the spontaneous disproportionation of chlorine dioxide into chlorate (ClO_3^-) and chlorite (ClO_2^-) leads to a continuous loss of measurable chlorine dioxide from solution. Under alkaline conditions, the following disproportionation occurs (53):

$$2\,ClO_2 + 2\,OH^- = H_2O + ClO_3^- + ClO_2^- \qquad (5\text{-}10)$$

This reaction has been reported to be catalyzed by carbonate ions; in the absence of carbonate, the reaction is governed by parallel first and second order kinetics according to the following rate law (53,54):

$$\begin{aligned} r_{ClO_2} &= \text{rate of } ClO_2 \text{ loss (moles/liter-second)} \\ &= k_1[ClO_2] + k_2[ClO_2]^2 \end{aligned} \qquad (5\text{-}11)$$

$$k_1 = (5 \times 10^5 + 1.7 \times 10^8[OH^-]) \exp(-7348/T) \qquad (5\text{-}12)$$

$$\begin{aligned} k_2 &= 2.7 \times 10^9 \exp(-7348/T) \\ &+ 7.3 \times 10^6[OH^-] \exp(-3875/T) \end{aligned} \qquad (5\text{-}13)$$

These kinetics imply that the half-life of aqueous chlorine dioxide solutions decrease substantially with increasing concentration, and that above pH 9, the further increase in pH decreases the half-life. Even at neutral pH values, however, in the absence of carbonate at room temperature, the half-life of chlorine dioxide solutions of 0.01, .001, and 0.0001 mol/liter is respectively, 0.5, 4, and 14 hours. Hence, the storage of stock solutions of chlorine dioxide for even a few hours is impractical.

Recent evidence also suggests that the simple disproportionation reaction to chlorate and chlorite is insufficient to explain the decay of chlorine dioxide in water free of extraneous reductants. Equation 5-10 predicts that the molar ratio of chlorate to chlorite formed should be 1:1; however Medir and Giralt found that the molar ratio of chlorate:chlorite:chloride:oxygen produced was 5:3:1:.75, and that the addition of chloride enhanced the rate of decomposition and resulted in the predicted 1:1 molar ratio of chlorite:chlorate (55). This suggests that the oxidation of chloride by chlorate, and the possible formation of intermediate free chlorine, may be of significance in the decay of chlorine dioxide in demand free systems (53).

The concentration of chlorine dioxide in solution may be expressed in terms of "g/l as chlorine" by multiplying the molarity of chlorine dioxide by the number of electrons transferred per mole of chlorine dioxide reacted and then multiplying this by 35.5 grams Cl_2 per electron mole. Conventionally, the five electron reduction (Equation 5-14) is used to carry out this conversion.

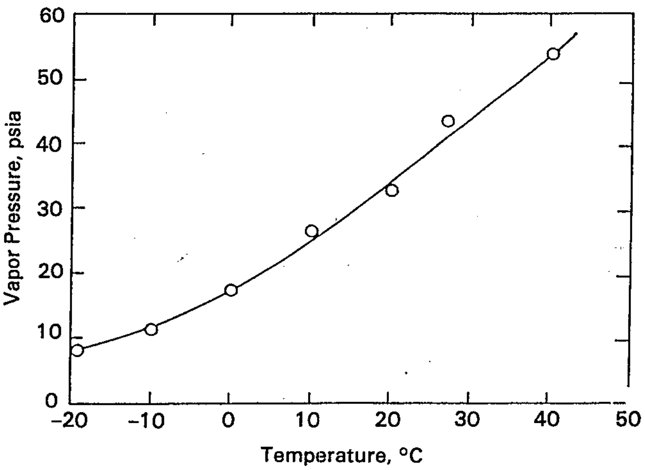
$$ClO_2 + 5e^- + 4H^+ = Cl^- + 2\,H_2O \qquad (5\text{-}14)$$

Hence, one mole of chlorine dioxide contains 67.5 grams of mass, and is equivalent to 177.5 (= 5 x 35.5) grams Cl_2. In other words, one gram of chlorine dioxide contains 2.63 grams "as chlorine." In examination of any studies on chlorine dioxide, due care with regard to units of expression of disinfection concentration is warranted.

5.3.1.3 Bromine Chloride

Bromine chloride (BrCl) is a chemical compound formed by the combination of an atom of chlorine and an atom of bromine. Below -5°C, BrCl exists as a dark red liquid at atmospheric pressure. The vapor pressure above bromine chloride liquid (Figure 5-2) is intermediate between the vapor pressures for the molecular halogens chlorine and bromine (Table 5-7).

Figure 5-2. **Vapor pressure over liquid bromide chloride. (56)**



Physical properties of BrCl are summarized in Table 5-7 in comparison to properties of molecular bromine

Table 5-7. Comparison of Properties of BrCl and Br_2 and Cl_2 (57)

Property	Br_2	BrCl	Cl_2
Freezing Point	−7.3°C	−66°C	−101°C
Boiling Point	−58.8°C	5°C	−34°C
Density (gas, 20°C, 1 atm)	3.12 g/l	2.34 g/l	1.40 g/l
Vapor Pressure atm, 20°C	0.98	3.03	7.81
Solubility (g/100 g H_2O)	3.3	8.5	0.75

and chlorine. In general, the properties of BrCl are intermediate between the two halogens, with the exception of the solubility of BrCl, which exceeds that of either of the two molecular halogens.

Commercially, bromine chloride is supplied as a containerized liquid under pressure. Since the vapor pressure over liquid bromine chloride at ambient temperatures is substantially in excess of atmospheric (Figure 5-2), the chemical may be withdrawn under its own pressure.

In both liquid and gas phases, BrCl may dissociate into its constituent elements according to the following:

$$2\ BrCl_{(g)} = Br_{2(g)} + Cl_{2(g)} \qquad (5\text{-}15)$$

$$2\ BrCl_{(l)} = Br_{2(l)} + Cl_{2(l)} \qquad (5\text{-}16)$$

The equilibrium constant for the gas phase reaction (Equation 5-15) is given by:

$$K_g = P_{Br_2}P_{Cl_2}/P^2_{BrCl} \qquad (5\text{-}17)$$

This equilibrium has been well studied, and the value for K_g appears to vary only slightly with temperature and assumes the value of approximately 0.12 at 25°C (29,30,58,59). No equilibrium constant for Equation 5-16 has been reported; however, it has been asserted on the basis of unpublished results that liquid bromine chloride is less than 20 percent dissociated (30).

Due to the unavoidable presence of molecular bromine and molecular chlorine in both the gas and liquid phases of a bromine chloride system, the thermodynamic significance of the vapor pressure, boiling and melting point data for BrCl is ambiguous. In a closed vessel containing BrCl in both liquid and vapor phases, the vapor phase will be enriched in Cl_2 and the liquid phase will be enriched in Br_2 under normal conditions. To ensure withdrawal of bromine chloride containing a 1:1 stoichiometric ratio of the halogens it is necessary that all withdrawals be made from dip tubes below the liquid level.

Aqueous solutions of bromine chloride in which the concentrations of reducing agents are negligible undergo a series of reactions similar to solutions of chlorine gas. Bromine chloride may hydrolyze to form hypobromous acid (HOBr) and chloride according to the following equation, which is analogous to Equation 5-4 for chlorine:

$$BrCl + H_2O = HOBr + H^+ + Cl^- \qquad (5\text{-}18)$$

$$K_H = [H^+][HOBr][Cl^-]/[BrCl]$$

The value of K_H at 25°C has been reported as $2.94 \times 10^{-5}\ M^2/l^2$, which is approximately fivefold less than that for molecular chlorine (30,60).

Hypobromous acid (HOBr) is formed on hydrolysis in preference to hypochlorous acid, since the latter is able and will readily act to oxidize bromide ion to HOBr. Hypobromous acid is a weak acid and will dissociate in accordance with:

$$HOBr = H^+ + OBr^- \qquad (5\text{-}19)$$

$$K_A = [H^+][OBr^-]/[HOBr]$$

Using thermodynamic data on hypobromous acid and hypobromite (51), the following relationship for K_A may be derived for temperatures in the range of normal operation:

$$K_A = 5.24 \times 10^{-6} \exp(-2265/T) \qquad (5\text{-}20)$$

In general, the equilibrium constant for HOBr dissociation is lower than that for HOCl at a given temperature (i.e. the pK for HOBr is higher than that for HOCl). Hence, the percentage of bromine present as hypobromous acid at a given pH is greater than would be the relative abundance of hypochlorous acid in a system containing chlorine at identical concentrations.

Concentrations of bromine chloride can be expressed in an analogous manner to those for free chlorine. From Equation 5-21, in comparison with equation (1), it may be seen that one mole of BrCl is equivalent to one mole of chlorine:

$$BrCl + 2e^- = Br^- + Cl^- \qquad (5\text{-}21)$$

Since one mole of BrCl contains 115.4 grams of material, there are 0.62 (71/115.4) grams of available halogen as chlorine per gram of bromine chloride.

5.3.2 Onsite Production Chemistry

In the case of chlorine and chlorine dioxide, technologies exist that enable a wastewater treatment plant to produce disinfecting chemical at the plant location itself, rather than relying upon shipment of externally generated material. In the case of chlorine, these technologies are alternative to external production, while in the case of chlorine dioxide the onsite synthesis of disinfectant is mandatory due to decomposition discussed above. The chemistry associated with the production of these chemicals will now be discussed.

5.3.2.1 Chlorine and Hypochlorites

Chlorine and hypochlorites have been produced from the electrolysis of brines and saline solutions since the early days of the 20th century (8). The basic principle is the use of a direct current electrical field to effect the oxidation of chloride ion with the simultaneous and physically separated reduction usually of water to gaseous hydrogen.

At the anode of such an electrolytic cell, one or both of the following reactions takes place:

$$Cl^- = 1/2\ Cl_{2(g)} + e^- \qquad E^\circ = 1.36 \text{ volts} \qquad (5\text{-}22)$$

$$Cl^- + 2OH^- = OCl^- + H_2O + 2e^- \quad E^\circ = 0.90 \text{ volts} \qquad (5\text{-}23)$$

In Equations 5-22 and 5-23, E° is the standard electromotive potential at unit activity of all products and reactants at 298K, and the actual half-cell voltage changes may depend slightly on concentration. This dependence may be determined by standard relationships.

At the cathode of an electrolytic cell, reduction of water occurs producing molecular hydrogen gas according to Equation 5-24, written for alkaline conditions.

$$H_2O + e^- = OH^- + 1/2\ H_{2(g)} \quad E^\circ = 0.83 \text{ volts} \qquad (5\text{-}24)$$

Combination of Equations 5-22 and 5-24 or 5-23 and 5-24 yield the two overall possibilities, depending upon whether the chlorine is withdrawn as gaseous chlorine evolving at the anode or as hypochlorite in spent brine.

$$Cl^- + H_2O = 1/2\ Cl_{2(g)} + 1/2\ H_{2(g)} + OH^- \qquad (5\text{-}25)$$
$$E^\circ = 2.19 \text{ volts}$$

$$Cl^- + H_2O = OCl^- + H_2O + 1/2\ H_{2(g)} \qquad (5\text{-}26)$$
$$E^\circ = 1.73 \text{ volts}$$

From these thermodynamic relationships, it may be concluded that for each mole of Cl_2 (or the equivalent one mole of OCl^-) produced, two moles of electrons are required. Since there are 96,493 coulombs (= ampere seconds) per equivalent, theoretically 1.93×10^5 coulombs are required per mole of available chlorine, or 2.72 C/mg available chlorine.

Additionally, from Equations 5-25 and 5-26, the minimum voltage difference between the plates of an electrolytic cell ranges from 1.73-2.19 volts. The minimum theoretical energy required is the product of this and the quantity (2.72 C/mg available chlorine). Thus, 4.7-5.96 Watt-seconds are required to produce 1 mg available chlorine. This may also be expressed as 1.31-1.66 kWh/kg available chlorine.

In actual practice, it is necessary to operate electrolytic chlorine generating units at voltages as high as 3.85 volts in order to increase the rates of the generation reaction. At these overvoltages, however, additional oxidations such as formation of chlorate, ohmic heating, and incomplete separation of hydrogen from oxidized products with subsequent dissipative reaction, combine to produce system inefficiencies. For typical electrolytic generating units, current efficiencies (based on actual to theoretical C/mg) of 97 percent may be obtained along with energy efficiencies (based on kw-hr/kg) of 58 percent (49). These efficiencies are related to the physical configuration of the electrolysis cells, brine concentration, and desired degree of conversion to available chlorine (61,62).

5.3.2.2 Chlorine Dioxide

Theoretically, chlorine dioxide may be produced by either the oxidation of a lower valence compound or reduction of a more oxidized compound of chlorine. Chlorites (ClO_2^-) or chlorous acid ($HClO_2$) may be oxidized by chlorine or persulfate to chlorine dioxide, or may undergo autooxidation (disproportionation) to chlorine dioxide in solutions acidified with either mineral or organic acids. Chlorates (ClO_3^-) may be reduced by use of chlorides, sulfuric acid, sulfur dioxide, or oxalic acid, or electrochemically to form chlorine dioxide (51).

For practical purposes in wastewater treatment, chlorine dioxide is generated exclusively from chlorite inasmuch as the reductive processes using chlorate as a starting material are capital intensive and competitive only at larger capacities (51).

In the acid-chlorite process, sodium chlorite and hydrochloric acid react according to:

$$5\ NaClO_2 + 4\ HCl = 4\ ClO_2 + 5\ NaCl + 2\ H_2O \qquad (5\text{-}27)$$

The resulting chlorine dioxide may be evolved either as a gas, or removed in solution. Mechanistically, this process occurs by a series of coupled reactions, some of which may involve the in situ formation of chlorine, the catalysis by chloride, and the oxidation of chlorite by chlorine (50,51,53). In addition, the yield of the reaction as well as the rate of the process are improved by low pH values in which both gaseous chlorine and chlorous acid formation are favored. Under these favorable conditions, the reaction proceeds in the order of minutes; however, to achieve these conditions, excess hydrochloric acid is required.

Alternatively, chlorine dioxide may be produced by the oxidation of chlorite with chlorine gas according to:

$$2\ NaClO_2 + Cl_2 = NaCl + 2\ ClO_2 \qquad (5\text{-}28)$$

As in the previous case, low pH accelerates the rate of this process, as does excess amounts of chlorine gas. However, if chlorine gas is used in stoichiometric excess, the resultant product may contain a mixture of unconsumed chlorine as well as chlorine dioxide.

5.3.3 Disinfection Demand Reactions

In the presence of such dissolved impurities as exist in wastewater effluents, each of the halogen disinfectants may undergo reactions in which they decompose or transform to less effective chemical forms. In the case of chlorine and bromine chloride, these principally involve reactions with ammonia and amino-nitrogen compounds, while chlorine dioxide demand may result from reactions with other organic materials. The nature of these processes will now be discussed.

5.3.3.1 Chlorine and Hypochlorites

Available chlorine, whether in the form of chlorine gas or hypochlorite, when added to wastewater, can undergo a series of dissipative reactions that result in a loss of disinfectant from the system, or a change in disinfectant form to a less active chemical species. The following reactions of chlorine or hypochlorite may occur (63):

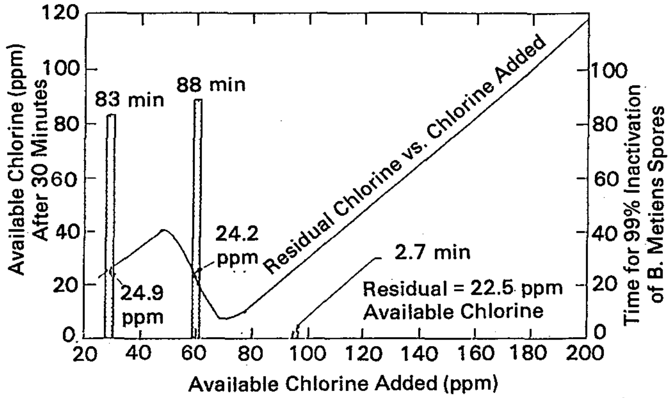
- cyanide may react with chlorine to form chlorides, bicarbonate, and a variety of oxidized nitrogen forms;
- ammonia or amino-nitrogen groups may react with chlorine to form chloramine compounds by the replacement of a proton by a chlorine atom;
- organic molecules containing unsaturated (double or triple) bonds may react to form chlorinated organic molecules;
- a variety of inorganic reducing agents, including hydrogen peroxide and reduced iron and sulfur compounds, may consume chlorine by redox reactions.

Of these processes, probably the most significant process in determining the fate of chlorine added to wastewater is the reaction with ammonia or amino-nitrogen groups. However, the reactions with organic materials are also of importance in the production of chlorinated organic byproducts during wastewater chlorination.

To understand chlorine demand reactions, a further definition is necessary. As noted above, the term "free available chlorine" is used to denote the concentrations of hypochlorous acid plus hypochlorite ion expressed on a mass equivalent Cl_2 basis. The term "total available chlorine" is defined as the mass equivalent Cl_2 contained in all materials that contain chlorine in an oxidized state. The difference between total available chlorine and free available chlorine is defined as "combined available chlorine" and represents the amount of chlorine that is in chemical association with various compounds (usually amino- or ammonia-nitrogen) but that is also capable of carrying out oxidation. The significance of this distinction, which will be further discussed in the section on kinetics, is that the free chlorine forms are generally more effective disinfectants than the combined chlorine forms.

One example of this behavior, along with a graphical demonstration of typical chlorine behavior in wastewater, is given in Figure 5-3 (64). In this depiction, as the dose of chlorine is increased, the total available chlorine residual (i.e. remaining in the system after 30 minutes) increases until a dose of approximately 50 mg/l, whereupon residual chlorine decreases to a very low value, and subsequently increases linearly with dose indefinitely. The "hump and dip" behavior is paralleled by the sensitivity of microorganisms to the available chlorine residual (indicated by the time required for 99 percent inactivation of *Bacillus metiens* spores). At the three points indicated, the total available chlorine is approximately identical at 22 to 24 mg/l yet there is a 32-fold difference in microbial sensitivity.

Figure 5-3. **Effect of increased chlorine dosage on residual chlorine and germicidal efficiency: pH 7.0, 20°C NH_3 10 mg/l. (64) (Reprinted by permission of the American Water Works Association.)**



The explanation for this behavior is founded in the "breakpoint" reaction between free chlorine and ammonia, which is present in non-nitrified effluents. At doses below the "hump" in the chlorine residual curve, only combined available chlorine is detectable in the wastewater. At doses between the "hump" and the "dip" in the curve, there is an oxidative destruction of combined residual chlorine accompanied by the loss of ammonia nitrogen from the wastewater (65). In fact, this particular reaction may be used as a means to remove ammonia nitrogen from wastewaters. Finally, after the ammonia nitrogen has been completely oxidized, the residual remaining consists almost exclusively of free chlorine. The minimum in the chlorine residual vs dose curve (in this case 70 mg/l) is called the "breakpoint" and denotes the amount of chlorine that must be added to a wastewater before a stable free residual can be obtained. The relative inefficiency of combined chlorine forms as disinfectants vis a vis free chlorine forms has been known at least since the work of Enslow (66).

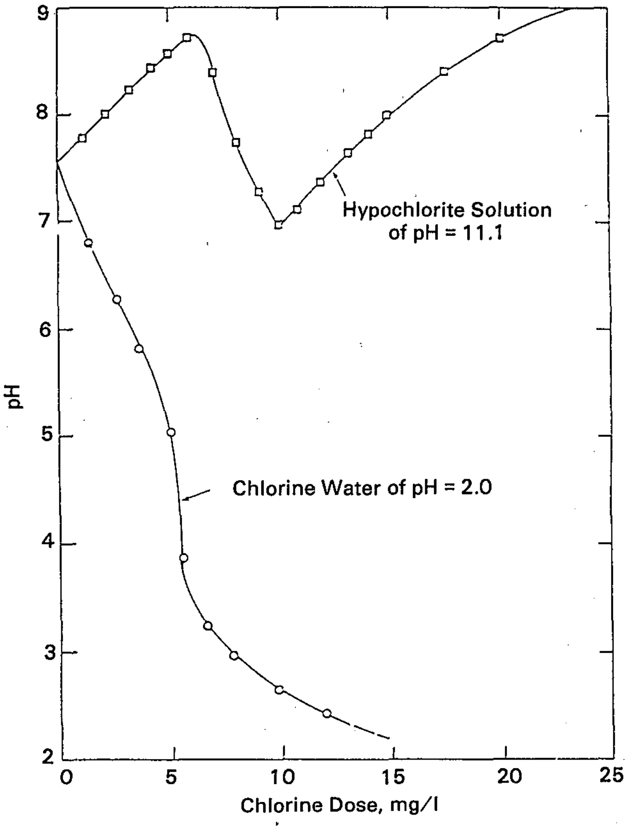
Much of the knowledge of breakpoint reactions during chlorination arises from studies on potable water chlorination. In investigating chlorination of drinking water, Griffin and Chamberlain (67,68) observed that:

- the classical "hump and dip" curve is only seen at water pH's between 6.5 and 8.5;
- the molar ratio between ammonia nitrogen and chlorine dose at the breakpoint under ideal conditions is 1:2 corresponding to a mass dose ratio (ammonia N:chlorine) of 1:10;
- in practice, mass dose ratios of 1:15 may be needed to reach breakpoint.

In later work with breakpoint chlorination of wastewaters, Griffin and Chamberlin found that measurement of the 30 minute and 2 hour residuals failed to produce a "hump and dip" curve, and that only at 18 hours could a classical breakpoint pattern be observed (69). Furthermore, it was found that even above the breakpoint, the concentrations of organic nitrogen were unaffected. However, this work did confirm that a mass dose ratio of 10 mg chlorine to 1 mg ammonia nitrogen results in breakpoint at 18 hours in raw, primary, and secondary (activated sludge) wastewaters.

The breakpoint reaction may also affect the pH of a wastewater. Figure 5-4 indicates that, if sodium hypochlorite is used as the source of active chlorine, as breakpoint occurs, the pH decreases due to an apparent release of protons during the breakpoint process (70). If gaseous chlorine is used, this effect is obscured by the release of protons by hydrolysis of gaseous chlorine according to Equations 5-4 and 5-5.

Figure 5-4. **Effect of chlorine or hypochlorite dose on pH of settled wastewater.(70) Reproduced from Journal of the Water Pollution Control Federation, by permission.**



In wastewaters, the chlorine demand, representing the difference between the applied chlorine dose and the total chlorine residual at a given contact time, is due to ammonia nitrogen, as well as organic and inorganic compounds. In early work, Symons et al. determined that domestic wastewater contains an average chlorine demand of 45 g/capita-d, including all sources (this is demand in the raw wastewater and fails to account for possible changes in quality through prior unit processes) (71,72). In addition, superimposed upon this average were significant fluctuations of various frequencies.

It was found that chlorine demand fluctuated seasonally approximately with temperature. Weekly and diurnal fluctuations also existed, and the latter were found to be more variable than flow. During storms, the chlorine demand was found to increase shortly after the onset of the storm flow, and then to rapidly return to baseline levels.

The rate of exertion of chlorine demand in both potable water and wastewater chlorination has been the subject of numerous studies. The most systematic work has been that of Taras (73), who chlorinated pure solutions of various organic compounds and found that chlorine demand exertion follows:

$$D = k(t)^n \qquad (5\text{-}29)$$

where t is the time in hours, D is the chlorine demand, and k and n are empirical constants. It was found that n for various substances ranged from 0.02 to 0.29.

In subsequent work, Feben and Taras chlorinated potable water, and potable water blended with up to 1.5 percent wastewater to a free residual, and found that the results could be correlated to Equation 5-29 with the value of n determined by the one hour chlorine demand as follows (74):

$$n = 0.18 - 0.017 \log (D1) \qquad (5\text{-}30)$$

where D1 is the one hour chlorine demand.

Lin and Evans extended the significance of Taras' work by chlorinating secondary effluents to combined residuals (75). They found that Equation 5-29 could be used to describe the kinetics of chlorine demand exertion except that two sets of k and n values were required. At times from 1 to 12 minutes, the rates of chlorine demand development were significantly greater than between 12 and 60 minutes. It was also found that when calcium hypochlorite was used as a chlorinating agent, a lower chlorine demand was obtained as compared with chlorination using dissolved chlorine gas.

A more recent kinetic model for chlorine demand exertion has been developed by Haas and Karra (76):

$$D = C_o (1 - X \exp(-k_1 t) - (1 - X) \exp(-k_2 t)) \qquad (5\text{-}31)$$

In Equation 5-31, X is an empirical constant typically 0.4-0.6, and k_1 and k_2 are rate constants, typically 1.0 min^{-1} and 0.003 min^{-1}, respectively, and C_o is the chlorine dose in mg/l.

While it has been generally observed that nitrified effluents that are low in ammonia-nitrogen have lower chlorine demands in comparison to nonnitrified effluents that have higher chlorine demands, recent work suggests some anomalies. White et al. (77) observed that the addition of low levels of ammonia to nitrified effluents may reduce chlorine demand. This phenomenon may be due to competitive formation of inorganic versus organic chloramine compounds, and will be discussed in greater detail below.

In the presence of ammonium ion, free chlorine may react in a step-wise manner to form chloramines. This process is depicted by the following equations:

$$NH_3 + HOCl = NH_2Cl + H_2O + H^+ \qquad (5\text{-}32)$$

$$NH_2Cl + HOCl = NHCl_2 + H_2O \qquad (5\text{-}33)$$

$$NHCl_2 + HOCl = NCl_3 + H_2O \qquad (5\text{-}34)$$

These three compounds, monochloramine (NH_2Cl), dichloramine ($NHCl_2$), and trichloramine (NCl_3), each contribute to the total (or combined) available chlorine residual in a wastewater. Each chlorine atom associated with a chloramine molecule is capable of undergoing reduction to chloride, and in the process accepting 2 electrons; hence, each mole of monochloramine contains 71 grams available chlorine; each mole of dichloramine contains 2 x 71 or 142 grams; and each mole of trichloramine contains 3 x 71 or 213 grams of available chlorine. Inasmuch as the molecular weights of mono-, di-, and trichloramine are 51.6, 86, and 110.5, this means that the chloramines contain respectively 1.38, 1.65, and 2.02 grams available chlorine per gram. However, the efficiency of the various combined chlorine forms as disinfectants differs, and thus the concentration of available chlorine also is insufficient to characterize process performance.

As Equation 5-32 indicates, the formation of monochloramine is accompanied by the loss of a proton, which corresponds to the experimental findings of McKee et al. (70). The loss of a proton is due to the fact that chlorination reduces the affinity of the nitrogen moiety for protons; this was verified experimentally for ammonium as well as a variety of amines by Weil and Morris (78).

Under conditions where ammonia nitrogen is in excess of the chlorine dose (i.e. below the "breakpoint") and where the pH is below 9.0 (so that the dissociation of ammonium ion is negligible), the amount of combined chlorine in dichloramine relative to monochloramine after Equations 5-32 and 5-33 have attained equilibrium may be given by the relationship described by McKee et al. (70):

$$A = BZ/(1 - (1 - Z(2 - Z)B)) \;\; -1 \qquad (5\text{-}35)$$

In Equation 5-35, A is the ratio of available chlorine in the form of dichloramine to available chlorine in the form of monochloramine, Z is the ratio of moles of chlorine (as Cl_2) added per mole of ammonia nitrogen present, and B is defined by:

$$B = 1 - 4 K_{eq} [H^+] \qquad (5\text{-}36)$$

The equilibrium constant in Equation 5-36 refers to the direct inter-conversion between dichloramine and monochloramine as follows:

$$\begin{aligned} H^+ + 2\,NH_2Cl &= NH_4^+ + NHCl_2 \\ K_{eq} &= [NH_4^+][NHCl_2]/[H^+][NH_2Cl]^2 \end{aligned} \qquad (5\text{-}37)$$

At 25°C, K_{eq} has a value of 6.7×10^5 liters/mole (70,79).

Using these relationships, it is possible to determine the equilibrium ratio of dichloramine to monochloramine as a function of pH and applied chlorine dose ratio (assuming no dissipative reactions other than those involving the inorganic chloramines). Table 5-8 summarizes these calculations. As pH decreases and the Cl:N dose ratio increases, the relative amount of dichloramine also increases.

Table 5-8. Ratio of Dichloramine Combined Chlorine to Monochloramine Combined Chlorine as a Function of pH and Applied Molar Dose Ratio (Equilibrium Assumed)

Molar Cl_2:N Ratio	pH 6	7	8	9
.1	.13	.014	1E-03	0
.3	.389	.053	5E-03	0
.5	.668	.114	.013	1E-03
.7	.992	.213	.029	3E-03
.9	1.392	.386	.082	.011
1.1	1.924	.694	.323	.236
1.3	2.7	1.254	.911	.862
1.5	4.006	2.343	2.039	2.004
1.7	6.875	4.972	4.698	4.669
1.9	20.485	18.278	18.028	18.002

As the Cl:N molar dose ratio increases beyond unity, the amount of dichloramine relative to monochloramine rapidly increases as well. For the conversion from dichloramine to trichloramine, the equilibrium constant given at 0.5 M ionic strength and 25°C indicates that the amount of trichloramine to be found in equilibrium with di- and mono-chloramine at molar dose ratios of up to 2.0 is negligible (79). This is in line with experimental measurement of the individual combined chlorine species as a function of approach to breakpoint, which indicates that detectable trichloramine is not formed until shortly before, or after the break point itself, and then only in relatively small amounts (80).

These findings, coupled with the routine observation of the breakpoint at molar doses below 2:1 (weight ratios Cl_2:N below 10:1) indicate that trichloramine is not an important species in the breakpoint reaction. Rather, the breakpoint reaction leading to oxidation of ammonia nitrogen and reduction of combined chlorine is initiated with the formation of di-chloramine.

The kinetics of formation of chloramine species have been investigated by various researchers. It has been found that the formation of monochloramine is a first order process in each of hypochlorous acid and uncharged ammonia. Solely through kinetic arguments, however, it is not possible to determine whether this, or a process involving hypochlorite ions reacting with ammonium cations, is the actual mechanism of reaction. If the neutral species are selected as the reactants, then the rate of formation of monochloramine may be given by (81):

$$r\,(\text{mol/l-s}) = 6.6 \times 10^{8} \exp(-1510/T)\,[HOCl]\,[NH_3] \qquad (5\text{-}38)$$

As noted above, hypochlorous acid dissociates into hypochlorite with a pK of approximately 7.5; analogously, ammonia is able to associate with a proton to the ammonium cation, with the pK for the latter of approximately 9.3. For a constant chlorine:nitrogen dose ratio, the maximum rate of monochloramine formation occurs at a pH where the product HOCl x NH_3 is maximized, which is at the midpoint of the two pK values or 8.3. At this optimum pH and the usual temperatures encountered in practice, the formation of monochloramine attains equilibrium in less than a second; however, at either a higher or lower pH, the speed of the reaction slows.

The formation of dichloramine from reactions between hypochlorous acid and monochloramine obeys the following rate law (81):

$$r\,(\text{mol/l-s}) = 3.0 \times 10^{7} \exp(2010/T)\,[NH_2Cl]\,[HOCl] \qquad (5\text{-}39)$$

While this reaction has been reported to be *catalyzed* by protons and acetic acid, the effectiveness of these catalysts is such that they would not be of importance at the usual concentrations encountered.

Given the situation where free chlorine is contacted with ammonia, the initial velocity of the monochloramine formation process given by Equation 5-38 is substantially greater than the velocity of the subsequent formation of dichloramine given by Equation 5-39. Hence, relative to equilibrium levels, there will be an initial accumulation of monochloramine if large dose ratios are used until the dichloramine formation process can be driven. Thus in practice, the ratio of dichloramine to monochloramine found may be less than that given by equilibrium analysis as in Table 5-8.

Additional formation reactions of the chloramines are disproportionation processes, such as Equation 5-37. However, both Morris (82), and Gray et al. (79) provide data to indicate that the formation of dichloramine via monochloramine disproportionation is relatively slow compared to Equation 5-38 under the usual conditions encountered in wastewater disinfection.

Morris has also studied the reaction of organic amines to form organic monochloramines (82). The rate laws for these reactions follow a similar pattern as those for the inorganic monochloramine formation process, except that the rate constants are generally less. In addition, the rate constants for this process correlate to the relative basicity of the amine reactant.

The mechanism of the breakpoint reaction has been extensively studied by Morris, and more recently by Saunier and Selleck (83). The oxidation of ammonia nitrogen by chlorine to gaseous nitrogen would theoretically yield a stoichiometric ratio of 1.5 moles of chlorine (Cl_2) consumed per mole of nitrogen oxidized according to:

$$NH_3 + 1.5HOCl = N_2 + H^+ + Cl^- + H_2O \quad (5\text{-}40)$$

As noted previously, the observed minimum stoichiometric ratio between chlorine added and ammonia nitrogen consumed at breakpoint is 2:1, suggesting that more oxidized nitrogen compounds are produced at breakpoint. Evidence (83) suggests that the principal additional oxidized product may be nitrate formed via:

$$NH_4^+ + 4HOCl = NO_3^- + 4\,Cl^- + 6\,H^+ + H_2O \quad (5\text{-}41)$$

As indicated, the production of nitrate from ammonia results in the consumption of four moles of Cl_2 (or four moles of HOCl) per mole nitrate formed. Hence, depending upon the relative amount of nitrate formed in comparison to nitrogen at breakpoint, between 1.5 and 4.0 moles of available chlorine may be required, which is consistent with the available data.

The breakpoint reaction consists of a complex series of elementary reactions, of which Equations 5-40 and 5-41 are the net results. On the basis of extensive kinetic investigations, Saunier and Selleck proposed that hydroxylamine (NH_2OH) and NOH may be intermediates in this reaction (83). A complete kinetic scheme for the breakpoint process as proposed by these authors is given in Figure 5-5; however, there does not yet appear to be sufficient experience with the use of this reaction mechanism to justify its use in wastewater applications.

At chlorine doses below breakpoint, the inorganic chloramines can decompose by direct reactions with several compounds. For example, Trofe et al. determined that monochloramine may react with bromide ions to form monobromamine, the chemistry of which will be discussed below (84). If trichloramine is formed, as would be the case for applied chlorine doses in excess of that required for breakpoint, it may decompose either directly to form nitrogen gas and hypochlorous acid or by reaction with ammonia to form monochloramine and dichloramine. Saguinsin and Morris determined the kinetics of these processes, the relative importance of which are dependent upon solution pH, trichloramine, and ammonia concentrations (85).

Organic chloramines may be formed by processes similar to inorganic chloramine formation, although usually at lower rates. Pure solutions of amino acids and some proteins may display breakpoint curves of identical shape to those of ammonia solutions (86,87). Organic chloramines may also be formed by the direct reaction between monochloramine and the organic amine, and this is apparently the most significant mechanism of organic N-chloramine formation at higher concentrations such as might exist at the point of application of chlorine to a wastewater (88).

Free chlorine residuals may react with other inorganic compounds likely to be present in wastewaters. Table 5-9 summarizes available data on rates of these processes (63). These reactions are generally first order in each of the oxidizing agent (hypochlorous acid or hypochlorite anion) and the reducing agent. Notable among these possible reactions are the reductions in the presence of nitrites, and sulfites. Nitrites might be present in incompletely nitrified effluents and react via a complex, pH-dependent mechanism (89). The reaction with sulfite is used in dechlorination and will be discussed subsequently.

Figure 5-5. **Proposed kinetic mechanism for the breakpoint reaction. (83) (Reproduced from American Water Works Association by permission.)**

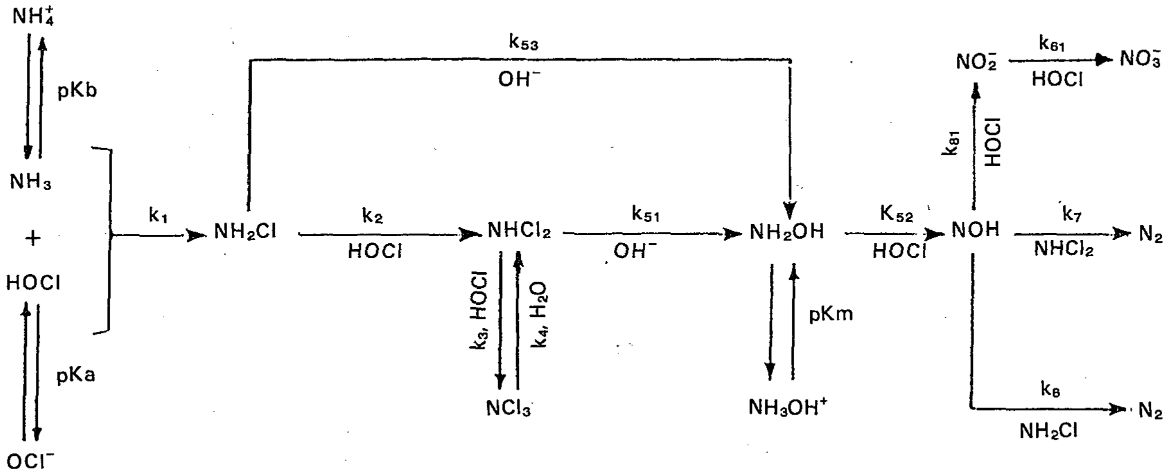


Table 5-9. Summary of Kinetics of HOCl and OCl⁻ Reduction by Miscellaneous Reducing Agents (63)

Oxidizing Agent	Reducing Agent	Oxidation Product	Log k ($M^{-1} s^{1}$) 25°C
OCl^-	IO_3^-	IO_4^-	−5.04
OCl^-	OCl^-	ClO_2-	−7.63
OCl^-	ClO_2-	ClO_3-	−5.48
OCl^-	SO_3^{2-}	SO_4^{2-}	3.93
HOCl	NO_2-	NO_3-	0.82
HOCl	$HCOO^-$	H_2CO_3	−1.38
HOCl	Br^-	BrO^-	3.47
HOCl	OCN^-	HCO_3-, N_2	−0.55
HOCl	HC_2O_4-	CO_2	1.20
HOCl	I^-	IO^-	8.52

Free chlorine can react with organic constituents to produce chlorinated organic by-products. Murphy et al. surveyed the reactivity of many classes of organic materials, and indicated that phenols, amines, aldehydes, ketones and pyrrole groups are all readily susceptible to chlorination (90). The first rigorous study of this point was that of Granstrom and Lee, who noted that phenol could be chlorinated by free chlorine to form chlorophenols of various degrees of substitution (54). The kinetics of this process were dependent upon both phenolate ions and hypochlorous acid. However, if excess ammonia was present the formation of chlorophenols was substantially inhibited.

In more recent work, De Laat et al. determined the rate of reaction of a variety of organic compounds with hypochlorous acid to form chloroform (91). This reaction, the prototype of which is the reaction of acetone with hypochlorous acid to form chloroform as in the following, is of major interest in potable water chlorination.

$$H_3CCOCH_3 + 3\,HOCl = CHCl_3 + H_3CCOOH + 2\,H_2O \tag{5-42}$$

Polyhydric phenols are substantially more reactive than simple ketones in the production of chloroform, and that the rates of these processes are typically first order in each of the phenol concentration and the free chlorine concentration (91). More significantly, it was observed that the reactivity of these compounds, many of which are analogous to materials such as tannins and humic acids that could be present in effluents, is greater than the reactivity of ammonia with hypochlorous acid. Therefore, even if sub-breakpoint chlorination is practiced, some chloroform may be formed rapidly prior to the conversion of free to combined chlorine.

Experimentally, this has been verified by Jolley who noted that more than 1 percent of the applied chlorine dose may be converted to chlorinated organic materials during wastewater chlorination to residuals on the order of 12 mg/l (92). However the kinetics of this process were sufficiently slow in that the formation of organic chlorine compounds was found to increase with time up to at least 4 hours following chlorination. Similar findings were noted by Chow and Roberts, who determined that the production of both organic halogen compounds, and trihalomethanes (chloroform and its analogs) specifically was greater in nitrified than in non-nitrified effluents (93). In breakpoint chlorination of effluents, a large spectrum of individual organic compounds may be isolated from wastewater effluents at levels of 10 to 500 micrograms/l (94).

5.3.3.2 Bromine Demand

The reaction of free bromine (hypobromous acid and hypobromite ion) follows many of the same pathways as the reactions involving hypochlorous acid and hypochlorites. The key differences in chemistry between free chlorine and free bromine, as outlined by LaPointe et al are as follows: (95)

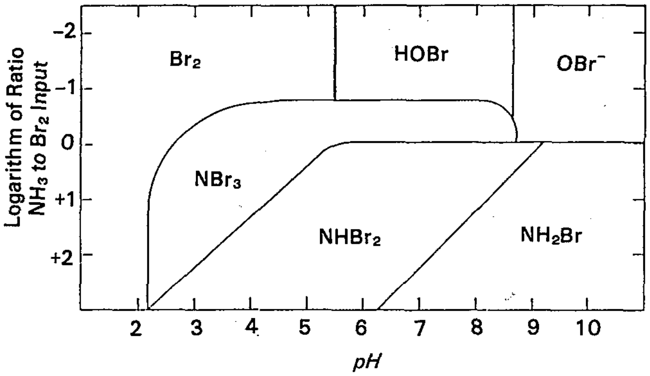
- formation of bromamines is rapid in comparison to chloramines;
- except for the irreversible oxidation of ammonia in the breakpoint reaction, the bromamine system is adequately described purely by equilibrium relationships;
- at normal pH and Br:N dose ratios, the major form of bromamine is dibromamine ($NHBr_2$) versus monochloramine; and
- at low pH values and relatively high Br:N dose ratios, tribromamine may be of significance, whereas trichloramine almost always is insignificant.

The comparative rate of formation of monobromamine versus monochloramine has been studied by Wajon and Morris (96). These results indicate that the monobromamine is formed 14 to 20 times more rapidly than monochloramine at pH 7 and equivalent molar concentrations of free halogen and ammonia.

On this basis, it is satisfactory to describe the predominance of various bromamine compounds by the use of an equilibrium diagram in which pH and N:Br molar dose ratio are the two master variables. Such a diagram is reproduced in Figure 5-6 (97).

The oxidation of ammonia nitrogen by free bromine residuals during the break point process follows the analogous reaction as in the case of chlorine. As described by Equation 5-43, 1.5 moles of hypobromous acid are consumed by the reaction with one mole of ammonia. This corresponds to the equivalent of 1.5

Figure 5-6. **Distribution of bromamine species as a function of pH and N:Br molar dose ratio (97). (Reproduced by permission, American Society of Civil Engineers.)**



moles of molecular bromine per mole ammonia nitrogen (97).

$$2NH_4^+ + 3HOBr = N_2 + 3Br^- + 3H_2O + 5H^+ \quad (5\text{-}43)$$

There do not appear to have been any detailed studies on the mechanism of this reaction in real wastewaters, or the production of more oxidized nitrogen forms, such as nitrate, which the reports noted above indicate may be produced during breakpoint chlorination. It is reasonable to suspect that the amount of bromine consumed in the breakpoint reaction would be somewhat greater than that indicated by Equation 5-43 due to the possible existence of these other processes.

One factor that must be mentioned is that combined bromine residuals (inorganic bromamines) possess disinfection efficiencies comparable to free bromine residuals. Thus, while the ammonia-bromine reactions do exist and do occur in wastewater disinfection, their significance in governing process performance is somewhat less than the comparable ammonia-chlorine reactions.

The kinetics of decomposition of bromamine residuals have been studied in some detail in pure laboratory systems (98,99). Both of these studies suggest a mechanism of the bromine breakpoint process that may involve the elementary reaction between dibromamine and tribromamine acting via a series of complex intermediates. One consequence of this process appears to be that high ammonia concentrations may reduce the rate of the breakpoint reaction due to a reduction in the amount of tribromamine present (98). In addition, the decomposition of dibromamine may lead to the reappearance of free bromine (which could, in turn, form additional amounts of combined bromine) via the following (98):

$$NHBR_2 + NBr_3 + 2H_2O = N_2 + 3H^+ + 3Br^- + 2HOBr \quad (5\text{-}44)$$

Comparatively little information exists on the formation of organobromine compounds during wastewater bromination. However, it is now well known from water treatment practice that transient free bromine may react to form bromoform and mixtures of chlorinated and brominated compounds, and it would be prudent to anticipate that these reactions would also occur during wastewater bromination.

5.3.3.3 Chlorine Dioxide Demand

The reaction of chlorine dioxide with material present in effluents resulting in chlorine dioxide demand appears to be less significant than in the case of bromine and chlorine. Rather, the dominant causes of loss of chlorine dioxide from wastewaters may be the direct reactions with water and interconversions to chlorite and chloride, as outlined above.

In one of the only studies directly applicable to wastewater, Roberts et al. (100) and Chow and Roberts (93) determined the chlorine dioxide demand of wastewater effluents and the formation of halogenated organics and trihalomethanes. It was found that at equal mass doses (i.e. 10 mg/l chlorine versus 10 mg/l chlorine dioxide), the chlorine dioxide demand of a wastewater (expressed as electron equivalents) was less than the chlorine demand of that wastewater. In addition, no trihalomethanes were formed from the application of chlorine-free chlorine dioxide (prepared via the acid-chlorite process, rather than the chlorine-chlorite process), and the amount of halogenated organics produced was 10 to 20 times less than those produced by equal mass doses of chlorine.

It has been established that, at usual concentrations, ammonia nitrogen, peptone, urea, and glucose have insignificant chlorine dioxide demand in 1 hour (101,102).

Masschelein reviewed the reactions of various classes of organic materials with chlorine dioxide, and only the following would appear to be of significance to wastewater applications (51):

1. ClO_2 may oxidize tertiary amines to secondary amines and aldehydes, with the formation of chlorite. For trimethylamine, this reaction is first order in chlorine dioxide and the amine, and has a rate constant of approximately 100,000 liters/mol-s.
2. Ketones, aldehydes, and alcohols may be oxidized to acids. While Masschelein cites this as a significant reaction, Somsen indicates that the rate of ethanol reaction with chlorine dioxide is insignificant at pH 7, while carbonyl compounds do exhibit reactivity (103).

3. Phenols and phenol derivatives may react with chlorine dioxide to form oxidized and chlorinated products.

4. Sulfhydryl amino acids, particularly cystine, may be oxidized to cysteic acid.

Wajon et al. have recently studied the reaction of phenols with chlorine dioxide under dilute conditions typical of water and wastewater disinfection applications (104). The reaction's stoichiometry is 2 moles of chlorine dioxide consumed per mole of phenol (or hydroquinone) consumed. Products formed included chlorophenols, aliphatic organic acids, benzonquinone, and (in the case of phenol) hydroquinone. The mechanism appeared to include the possible formation of hypochlorous acid as an intermediate which would thus chlorinate, and the rate of this process was found to obey the following law:

$$r = 2(k_b + k_a K_a/[H^+])\,[\text{Phenol}]\,[ClO_2]\ (\text{mol/l-s}) \tag{5-45}$$

Where K_a is the acidity constant for the phenol, and k_b and k_a are the rate constants. For phenol, these were 0.24 and 2.4 x 10^7 l/mol-s, respectively, while for hydroquinone, these rate constants were respectively 39,000 and 6.5 x 10^9 l/mol-s.

5.3.4 Dechlorination Chemistry

In certain circumstances, it is desirable to reduce the chlorine residual in a disinfected wastewater prior to discharge. In order to do this, it is necessary to contact the chlorinated wastewater with a substance that reacts with, or accelerates the rate of decomposition of, the residual chlorine. While many compounds may perform this function, including thiosulfate, hydrogen peroxide, ammonia, sulfite/bisulfite/sulfur dioxide, and activated carbon, only the latter two materials have been widely used for this purpose in either water or wastewater treatment (105). In this section, the chemistry of the reaction between residual chlorine and the latter two substances will be discussed.

5.3.4.1 Sulfur (IV) Compounds

Sulfur dioxide, and its aqueous dissolution products, sulfite and bisulfite ions, are reduced sulfur compounds with an oxidation state of +4. They may react with free or combined chlorine residuals to form oxidized sulfur products, principally sulfate, and hence convert available chlorine to chloride.

Commercially, S(IV) used for dechlorination is supplied principally as sulfur dioxide under pressure as a gas in equilibrium with its liquid. Sulfur dioxide is an irritant gas, with a water solubility 20 times greater than chlorine gas, and with a gas density relative to air of 2.26 (106). Sulfur dioxide gas is supplied in cylinders similar to those used for the supply of gaseous chlorine.

Table 5-10 summarizes physical properties of sulfur dioxide at various temperatures.

Table 5-10. Physical Properties of Sulfur Dioxide (106)

Temperature (°C)	Liquid Density (g/ml)	Vapor Pressure (atm)	Solubility (g/l)
−40	1.530	—	—
−29	1.510	—	—
−18	1.480	—	—
−6	1.450	0.15	—
4	1.420	0.81	199.8
16	1.390	1.74	130.5
27	1.360	3.83	87.6

Upon addition to water, sulfur dioxide forms sulfurous acid (H_2SO_3), which may lose, successively, two protons, according to the following reactions:

$$H_2SO_3 = H^+ + HSO_3^- \tag{5-46}$$

$$HSO_3^- = H^+ + SO_3^{-2} \tag{5-47}$$

The pK for the second dissociation (Equation 5-47) is 7.2 at room temperature (107), and hence sulfur (IV) is present in most wastewater effluents as a mixture of both sulfite and bisulfite ions. In some circumstances, sulfur (IV) may be applied as sodium sulfite or bisulfite, and the resulting equilibrium mixture is established via the above equations; however, the form of application of sulfur (IV) does not, in and of itself, affect the nature of the subsequent dechlorination reactions.

The reaction between sulfur dioxide and free chlorine is such that one mole of sulfur dioxide reacts with one mole of either free chlorine or monochloramine via the following stoichiometric equations (105):

$$SO_2 + H_2O + HOCl = 3H^+ + Cl^- + SO_4^{-2} \tag{5-48}$$

$$SO_2 + 2H_2O + NH_2Cl = NH_4^+ + 2H^+ + Cl^- + SO_4^{-2} \tag{5-49}$$

The rate of reaction between free chlorine and sulfur (IV) compounds appears to be quite rapid. This rate has been found to be first order in each of sulfite and free chlorine concentration. However, Lister and Rosenblum (108) have indicated that the kinetics are dependent on OCl^- concentration, while Srivastava et al. (109) proposed that HOCl is the reactive species.

The rate laws given by these investigations are, respectively:

$$r = 2.7 \times 10^{9} \exp(-3773/T) \times [OCl^{-}][SO_3^{-2}] \text{ (M/l-s)} \quad (5\text{-}50)$$

$$r = 1.2 \times 10^{15} \exp(-7851/T) \times [HOCl][SO_3^{-2}] \text{ (M/l-s)} \quad (5\text{-}51)$$

At 25°C and 0.2 mM sulfite concentrations (16 mg/l as sulfite), these rate laws give half-lives of 0.4 and 0.8 s, respectively. Thus, it appears, for free chlorine, that the rate of reaction is sufficiently rapid so as to be regarded as instantaneous (80).

Surprisingly, there does not appear to have been any direct measurements made of the rate of reaction of sulfur (IV) species with inorganic combined chlorine. However, based on the results reported for organic chloramines, it may be expected that the dechlorination of combined chlorine residuals is slightly slower than free chlorine. Stanbro and Lenkevich reported that the sulfite dechlorination of organic chloramine residuals follow kinetics first order in both total reduced sulfur and chloramine concentrations (110). For the monochloramine derivatives or methylamine, N-alpha-acetylysine, alanine, leucine, and alanylalanylalanine (N-chloro derivative), at pH 7 and 25°C, the measured second order rate constants were 4,333, 2,167, 1,167, 2,333, and 83 liters/mol-s. For the peptide chloro derivative, in particular, this is sufficiently slow to suggest that small amounts of some organochloramines may not be dechlorinated completely in some sulfur (IV) processes, particularly where extremely short contact times are used.

One side reaction that may occur during sulfur dioxide dechlorination, and that may be of concern for the receiving water quality, is the deoxygenation of the effluent. This occurs by the following reaction stoichiometry (105):

$$O_2 + 2SO_2 + 2H_2O = 4H^{+} + 2SO_4^{-2} \quad (5\text{-}52)$$

This reaction, which is used during the standard oxygen transfer tests (111) is catalyzed by a variety of trace metals, including copper, cobalt, iron, cerium, and manganese, and is inhibited by ethanol, glycerol and mannitol. In uncatalyzed systems, at sulfur (IV) concentrations below 0.02 M it has been found to be first order in reduced sulfur and zero order in oxygen, with a first order rate constant of 0.005 s^{-1} at pH 7.35, and increasing rates as pH increases (105). Thus, unless catalyzed, the rate of deoxygenation is much less than the rate of sulfur dioxide dechlorination. In pilot studies, it has been noted that no significant oxygen depletion occurred until sulfur dioxide overdoses exceeded 50 mg/l (112). However, the degree to which catalysis of Equation 5-52 might occur in dechlorinated effluents has not been thoroughly explored.

5.3.4.2 Activated Carbon Dechlorination

Activated carbon is a material prepared by controlled combustion and oxidation of an organic material. This section will describe the chemical properties of activated carbon peculiar to dechlorination applications. However, it should be noted that the use of activated carbon in wastewater treatment may simultaneously achieve other objectives, such as the reduction of total organic carbon to low levels, or the specific removal of biologically refractory trace pollutants.

The reduction of free chlorine residuals by activated carbon at low loadings has been found to be due to an actual reaction of the chlorine with the carbon surface, rather than merely by a catalytic decomposition (113). Products of this reaction include chloride and surface-bound oxides of carbon that evolve as carbon monoxide and carbon dioxide, as well as water, when the spent carbon is thermally regenerated. Hence, the dechlorination process using activated carbon represents a consumptive use of the activated carbon itself.

At higher loadings, once the surface of the carbon has accumulated surface oxides (denoted by CO*), reduction of free chlorine residuals may continue to occur. This has been postulated to be due to the catalysis of free chlorine decomposition to chloride and chloric acid according to (114):

$$3HOCl = 2HCl + HClO_3 \quad (5\text{-}53)$$

The reaction of combined chlorine residuals with the carbon surface is more complex, and differs between mono and dichloramine. For fresh carbon, before a large amount of surface oxides have formed, monochloramine is reduced by the carbon surface (denoted by C*) to form surface oxides (which are converted to carbon monoxide and dioxide when thermally regenerated) according to (115):

$$NH_2Cl + H_2O + C^{*} = NH_4^{+} + Cl^{-} + CO^{*} \quad (5\text{-}54)$$

After some surface oxides have formed, they may react with monochloramine by the additional reaction (115):

$$2NH_2Cl + CO^{*} = N_2 + 2H^{+} + 2Cl^{-} + H_2O + C^{*} \quad (5\text{-}55)$$

Reaction 5-55 serves to reduce the concentration of ammonia nitrogen present in a wastewater, and this process has been suggested as a means for the removal of ammonia nitrogen from effluents.

Dichloramine residuals appear to react primarily by the formation of nitrogen gas and surface oxides by (115):

$$2NHCl_2 + H_2O + C^* = N_2 + 4H^+ + 4Cl^- + CO^* \qquad (5\text{-}56)$$

The kinetics of monochloramine-carbon dechlorination occurring by Equations 5-54 and 5-55 have been studied in detail (116). Since the rate of monochloramine decomposition is accelerated by the formation of surface oxides, the overall process is autocatalytic, and for low effluent monochloramine residuals, prolonged contact times are necessary—for example, for a 2 mg/l monochloramine residual, 16 minutes contact in a packed bed carbon contactor are necessary to reduce the residual below 0.3 mg/l.

The reaction between dichloramine and carbon is faster than that between free chlorine and carbon, and much faster than that between monochloramine and carbon. In kinetic investigations, Kim et al., (117,118) determined that the reduction of dichloramine occurred by Equation 5-56 in parallel with a catalytic breakpoint reaction, given by:

$$NH_4^+ + 3NHCl_2 = 2N_2 + 6Cl^- + 7H^+ \qquad (5\text{-}57)$$

Hence, dechlorination of predominantly dichloramine residuals will also produce significant degrees of ammonia nitrogen removal.

Since, as noted above, in sub-breakpoint chlorination, monochloramine predominates, it would appear that the design of activated carbon beds for dechlorination would be limited by the reduction of monochloramine residuals, which are the more slowly reacting form, particularly when final chlorine residuals below 0.1 mg/l were desired.

Only a small number of wastewater treatment plants practice GAC dechlorination. Other than the models of Kim et al. (118) and Kim and Snoeyink (116), little guidance for design is available.

5.4 Analysis of Disinfectant Residuals

The monitoring and control of halogenation processes requires that a means be available for the analysis of residual disinfectant, which, as previously noted, is one major variable governing microbial inactivation. Ideally, the analytical method should be simple, rapidly performed, and not subject to analytical interferences by other constituents likely to be present in the wastewater.

5.4.1 Chlorine

Chlorine may be analyzed as total residual, including all chemical forms capable of functioning as oxidizing agents (free chlorine plus inorganic chloramines), or it may be differentiated as free chlorine and combined chlorine. Each of the latter two fractions may be further subdivided; free chlorine into HOCl and OCl⁻, and combined chlorine into mono-, di-, and tri-chloramine (as well as organic chloramine compounds). In most cases, however, the concentration of total residual chlorine is sufficient to allow adequate process monitoring.

All currently available methods for the analysis of total residual chlorine rely upon the oxidation of iodide to form molecular iodine, which then can be analyzed. The chemistry of this process, which is favored at low pH, is as follows:

$$HOCl + H^+ + 2I^- = I_2 + Cl^- + H_2O \qquad (5\text{-}58)$$

$$OCl^- + 2H^+ + 2I^- = I_2 + Cl^- + H_2O \qquad (5\text{-}59)$$

$$NH_2Cl + 2H^+ + 2I^- = NH_4^+ + I_2 + Cl^- \qquad (5\text{-}60)$$

$$NHCl_2 + 3H^+ + 4I^- = NH_4 + 2I_2 + 2Cl^- \qquad (5\text{-}61)$$

$$NCl_3 + 4H^+ + 6I^- = NH_4^+ + 3I_2 + 3Cl^- \qquad (5\text{-}62)$$

At a pH of 4.0, potassium iodide will react with all of the above forms of chlorine residual to produce iodine.

At this point, the analysis can be conducted in two general manners. First, in the so-called forward methods, the released iodine may be directly measured by titration using either a starch colorimetric indicator or amperometric measurements for endpoint determination. In the reverse methods, simultaneously with the production of iodine in Equations 5-58 to 5-62, reductant is added. By measuring the amount of this reductant that has been oxidized during the analysis, the amount of total residual chlorine may be determined.

The two most common forward titration methods involve the use of starch indicator in conjunction with sodium thiosulfate ($Na_2S_2O_3$) titrant and the use of amperometric titration equipment in conjunction with phenylarsine oxide titrant.

In the first case, sodium thiosulfate is added to the sample until the reddish iodine color is almost discharged. At this point, starch indicator is added, which forms a strong bluish complex with iodine, and sufficient additional titrant is added until the color disappears. The amount of sodium thiosulfate that has been added is equivalent to the amount of iodine that had been released, and this in turn is equivalent to the total residual chlorine.

The second procedure substitutes the amperometric titrator for the starch indicator solution (119). This instrument measures the current developed across electrodes maintained at a constant potential sufficient to reduce free halogen residuals. In the forward titration, phenylarsine oxide (PAO) titrant is added until there is no further change in the indicated current; this is the point at which all halogen residual has been reduced, and the amount of PAO added is equivalent to the total residual chlorine.

In the reverse titration, the amount of consumed reductant (either thiosulfate or PAO) is measured by titration with standard iodine solution using either the starch indicator or amperometric end-point detection methods. Alternatively, the reductant used may be a compound that changes color with oxidation, and thus the amount of residual chlorine may be determined either colorimetrically or by titrating the color with reductant until disappearance. This latter procedure is exemplified by the use of DPD (N,N-diethyl-p-phenylenediamine) or Leuco Crystal Violet (LCV).

In the DPD procedure, DPD reagent is added prior to iodine addition in the reverse titration. Oxidized DPD is reddish in color, whereas DPD itself is colorless; thus the amount of residual chlorine may be determined by a spectrophotometric determination of the amount of oxidized DPD produced. This method is widely available in a field kit form. Alternatively, the oxidized DPD may be titrated with ferrous ammonium sulfate reductant.

The LCV method is a colorimetric assay based upon the vivid bluish color exhibited by oxidized LCV.

For each of the above procedures, detailed methods may be found in Standard Methods (111). All colorimetric and visual titrimetric methods are subject to interferences in highly colored or turbid wastewaters. Oxidized manganese and chromium, as well as other materials capable of converting iodide to iodine, interfere with all of the above methods. Silver or Cu(I) ion interfere with the amperometric methods.

In previous years, orthotolidine reagent had also been used for total chlorine determinations in wastewater. However, a number of studies have indicated that the OT method is substantially less satisfactory than either the amperometric or starch iodide titration methods (120-126). In addition, orthotolidine is a suspect carcinogen and therefore, on the basis of these two reasons, this method is no longer approved (111).

The amperometric and starch-iodide forward titration methods for total chlorine residuals have been found to produce comparable results (122). For wastewaters, it is recommended that the total chlorine reverse titrations be used. In the forward methods, free iodine residuals are in contact with the wastewater for a period of time during the completion of the titration. During this time, there may be a loss of halogen residual due to exertion of demand on the part of the wastewater. In the reverse titrations, any iodine produced is rapidly removed by reaction with the reducing reagent (thiosulfate, PAO, LCV or DPD), and thus there is little opportunity for demand exertion.

Experimental studies comparing these methods, however, have tended to yield disparate results, which appear to vary with the quality of the wastewater being chlorinated. Browning and McLaren (127) developed a modified starch-iodide forward titration (MSI), which was found to yield results closely comparable to the reverse amperometric method and greater than the forward amperometric method, and reverse starch-iodide methods. In contrast, Collins and Deaner (128) found that the MSI procedure consistently produced lower chlorine residual readings than did the reverse starch-iodide method, and that the extent of this difference increased with the degree of pollution of the effluent (e.g. BOD).

Lin et al. compared the reverse starch iodide titration (SIB), the DPD method, the LCV method and the MSI titration using three wastewaters. These authors concluded that the MSI procedure generally, but not always, yielded results less than SIB and LCV methods, but that the results were indistinguishable from the DPD procedure. Furthermore, following chlorine addition, in some cases, the total chlorine residual measured by the DPD and LCV procedures increased with time; the MSI procedure did not produce such anomalous results. On this basis, Lin et al. suggested that the MSI procedure was preferable to either DPD or LCV methods (129).

Recent data suggest that all of the methods used for total chlorine residual may produce false negative readings if sulfite or similar reducing agents are present (110). This situation may occur when dechlorination using S(IV) compounds is practiced, particularly if organic chloramines are also present. In this case, it is possible that total chlorine residuals actually exist in the effluent, but that when the pH is lowered during the initial addition of iodide, the reaction between S(IV) and chlorine residuals may be more rapid than the reaction with iodide. Thus, chlorine may be deemed to be absent when it is in fact present.

Two additional methods for the analysis of total residual chlorine in wastewater have been reported; however, neither of them has yet appeared in Standard Methods. Jenkins and Baird (130) utilized a polarographic iodine electrode to measure the iodine released following acid-iodide addition to wastewater. This method is essentially a forward titration procedure using a direct electrode analysis of iodine production. It was found that this method was equal to the SIB method in precision and accuracy at total chlorine residuals above 1 mg/l; below 1 mg/l, the iodine electrode procedure was more precise than the SIB procedure, which is not recommended for residuals below 1 mg/l (111). The principal interference with the electrode iodometric assay occurred in effluents with more than 50 mg/l BOD—in these cases, the iodine progressively declined to zero prior to a stable electrode reading, presumably due to exertion of iodine demand.

Liebermann et al. (131) investigated the use of the syringaldizine reagent for total chlorine determination following acid-KI addition, with a colorimetric assay. This test yielded a linear response to 10 mg/l total chlorine, and was found to be comparable to the forward amperometric method. Syringaldizine forms the basis of the FACTS test for free chlorine which has been accorded tentative status in the current edition of Standard Methods (111).

In certain cases, particularly in breakpoint chlorination for nitrogen removal, or the chlorination of nitrified effluents, it may be desirable to measure free chlorine apart from total chlorine. The iodometric starch-iodide, amperometric, DPD, and LCV methods may be modified to be relatively selective for free chlorine. Further details on these methods may be obtained in Standard Methods (111). It should be noted that the determination of small concentrations of free chlorine in the presence of large amounts of total chlorine (as might be found immediately prior to the breakpoint itself) is an extremely difficult analytical problem (132).

5.4.2 Bromine

The analysis of bromine residuals, as might be produced from BrCl treatment, proceeds directly analogous to chlorine residuals. Free or combined bromine can oxidize iodide to iodine, and any of the methods for total chlorine can be used to measure total bromine. If mixtures of chlorine and bromine residuals are present, it becomes impossible to distinguish between the two by iodometric methods.

5.4.3 Chlorine Dioxide

For the determination of chlorine dioxide in wastewaters, it is particularly necessary to differentiate this compound from chlorite (which is produced as a reaction product) and free and combined chlorine (which may arise if the chlorine-chlorite process is used to produce chlorine dioxide). Both the DPD and the amperometric methods may be modified to determine chlorine dioxide apart from total chlorine (111); if the total chlorine methods are used, they will tend to indicate the sum of total chlorine, chlorine dioxide, and a portion of the chlorite. Roberts et al. (100) indicated that, as for total chlorine, use of the reverse titration procedures produces results in excess of the forward titration methods.

In addition to the above procedures, Knechtel et al. (133) determined that acid chrome violet K (ACVK) reagent is suitable for the direct colorimetric determination of chlorine dioxide in wastewaters. ACVK is not subject to interference by free or combined chlorine, chlorite, or nitrite, and results in wastewater were found to be comparable to determinations using electron spin resonance. However, if large amounts of turbidity are present, a pretreatment via centrifugation is necessary to prevent optical interference.

Similarly, Wheeler et al. (134) proposed chlorophenol red (CPR) as a selective indicator for chlorine dioxide; this compound is suitable for use in a direct colorimetric or a forward titrimetric assay. No interferences have been reported by chlorite, chlorate, free chlorine, Cr(VI), Fe(III), or Mn(V).

5.5 Kinetics of Microbial Inactivation

The information needed for the design of a disinfection system includes knowledge of the rate of inactivation of the target, or indicator, organism(s) by the disinfectant. In particular, the effect of disinfectant concentration on the rate of this process will determine the most efficient combination of contact time (i.e. basin volume at a given design flow rate) and dose to use.

Chick's Law and deviations therefrom are discussed in Section 4-1, and the reader is referred to that section for the historical development of disinfection theory. The discussion that follows here is a practical and logical outgrowth of that theory.

In the disinfection of a wastewater by halogens the concentration of disinfectant changes with time, and, particularly during the initial moments of contact with either chlorine or bromine, the chemical form of the halogen undergoes a rapid transformation from the free to the combined forms. Since C in Equation 4-6 would thus not be a constant, typically wastewater disinfection results obtained in batch systems exhibit "tailing," the degree of which may depend on the chlorine demand and the ammonia concentration of the system (135). It is more critical to determine what the chlorine residual is, rather than the chlorine dose in these systems. According to Heukelekian and Smith: "...the control of chlorination cannot be based on the dosage of chlorine, because neither for sewage from different sources, nor for a sewage from a given source, but at different times, does a constant dosage of chlorine produce a constant number of coliform organisms per ml in effluent." (136)

As a result, the analysis of the kinetics of microbial inactivation in wastewater effluents by disinfectants has often been accomplished by empirical models deviating substantially from Chick's Law. For example, McKee et al., chlorinated primary effluent, and found that his data could be described by (70):

$$\log(N) = 1/(a + bC_o) \qquad (5\text{-}63)$$

where N is the viable microorganism concentration and C_o is the chlorine dose; in this equation, the contact time is held constant. McKee et al. also proposed the following equation as a means to extrapolate inactivation results from one contact time to another (70):

$$N_t/N_a = (t/a)^{n'} \qquad (5\text{-}64)$$

In Equation 5-64, N_t and N_a are the microorganisms remaining at time t and time a, respectively (chlorine dose fixed), and n′ is a constant.

The following equation, similar in concept to McKee's model, was developed to describe inactivation in various chlorinated effluents (137,138):

$$N/N_o = (bCt)^{-a} \qquad (5\text{-}65)$$

In Equation 5-65, a and b are constants, and C is the chlorine residual remaining at time t. At values of the product $(Ct) < 1/b$, by definition, N/N_o equals unity. The use of this equation will be further discussed below.

While it is recognized that the presence of solids in wastewater effluents may harbor and protect enmeshed microorganisms from the action of disinfectants, few methods are available for the quantitative consideration of these phenomena.

However, it has been shown that organic solid materials present in actual wastewaters can render enmeshed microorganisms some protection from a measurable chlorine residual (139,140).

Analysis of data on the inactivation of a variety of microorganisms by free and combined chlorine suggests that the combined Chick-Watson law (Equation 4-4 substituted into Equation 4-3) provides a satisfactory description of the disinfection process. Values for k′ and n are a function of the microorganism, the pH, and the temperature of the system, and also differ depending upon whether free or combined chlorine species are used as disinfectants. Table 5-11 summarizes a number of values of the Chick-Watson parameters for microbial inactivation. In general, microbial inactivation increases with a decrease in pH for either free or combined chlorine residuals. An increase in temperature also increases inactivation. At a given temperature and pH, free chlorine residuals are more effective than combined chlorine residuals. These data may underestimate the resistance of microorganisms in wastewater. However, the degree of underestimation is unknown.

Table 5–11. Chick–Watson Parameters for Microbial Inactivation by Chlorine

Reference	Organism	pH	Temp. (°C)	k′ $l^n mg^{-n} min^{-1}$	n
			FREE CHLORINE		
141	*E. coli*	8.5	2–5	10.9	1.2
		9.8	2–5	1.18	1.29
		10.7	2–5	0.279	1.11
		8.5	20–25	30.6	1.46
		9.8	20–25	5.91	1.34
		10.7	20–20	1.30	0.79
	Aerobacter aerogenes	7	20–25	1.39×10^4	3.78
	Pseudomonas Pyocyanea	8.5	20–25	312	2.74
		9.8	20–25	2.13	1.26
		10.7	20–25	0.738	0.711
	Salmonella typhi	7	2–5	7.87×10^{12}	7.26
		9.8	2–5	0.962	0.76
		7	20–25	8.15×10^6	4.07
		8.5	20–25	2.45×10^4	1.78
	Shigella dysenteriae	7	20–25	9.07×10^7	4.92
142	*E. coli*	7	25	4.02	.801
	Micrococcus pyogenes var. aureus	7	25	3.32	1.10
143	*Bacillus metiens*	10	20	.00577	.483
		6	20	.029	1.24
		7	20	.0219	1.18
		8	20	.0209	1.12
		9	20	.008	0.99

Table 5-11. Continued

Reference	Organism	pH	Temp. (°C)	k′ $l^n mg^{-n} min^{-1}$	n
		9.35	20	.0086	1.04
		12.86	20	.0015	.58
		10	30	.00324	.868
		10	35	.0044	1.0
		10	50	.0075	1.26
144	Poliovirus	6	2	11.28	.766
	Type I	6	10	12.78	.818
	(Mahoney)	6	20	30.12	.615
		6	30	75.12	.608
145	Poliovirus	6	20	12.20	0.69
	Type I	10	20	6.5	3.23
	(Brunhilde)				
146	*Naegleria*	7.2	25	.171	.786
			COMBINED CHLORINE		
147					
	Mycobacterium	7	20	.226	.545
	phlei	7	20	6.21×10^{-4}	3.51
	M. fortuitum	7	5	5.68×10^{-4}	2.68
	Candida	7	20	6.85×10^{-3}	2.38
	parapsilosis	7	5	.656	1.16
	E. coli	7	20	1.01	0
148	*E. coli*	7	35	.084	1.39
		8.5	35	.0109	1.52
		9.5	35	2.48×10^{-5}	13.3
		6.5	20–25	.483	1.07
		7.0	20–25	.316	1.04
		7.8	20–25	.193	1.18
		8.5	20–25	.0854	1.125
		9.5	20–25	.049	1.37
		10.5	20–25	.0125	2.27
	Aerobacter	6.5	20–25	.363	1.19
	Aerogenes	7	20–25	.241	1.35
		7.8	20–25	.095	1.18
		8.5	20–25	.0715	.917
		9.5	20–25	.0358	1.16
		10.5	20–25	.00809	1.7
	Shigella	6.5	20–25	.821	1.3
	dysenteriae	7.0	20–25	.55	1.15
		7.8	20–25	.341	1.32
		8.5	20–25	.151	1.02
		9.5	20–25	.064	.995
		10.5	20–25	.0301	1.52
	Salmonella	7	2–6	.0902	1.32
	typhi	8.5	2–6	.0182	1.67
		9.5	2–6	6.8×10^{-4}	6.26
		6.5	20–25	.491	1.13
		7.0	20–25	.290	1.84
		7.8	20–25	.211	1.07
		8.5	20–25	.113	1.16
		9.5	20–25	.0417	.878
	Pseudomonas	6.5	20–25	.44	1.27
	pyocyanus	7.0	20–25	.301	1.44
		7.8	20–25	.174	1.55
		8.5	20–25	.102	1.01
		9.5	20–25	.0483	1.05

The use of the parameters in Table 5-11 for the description of wastewater microbial inactivation entails the assumption that the microorganisms for which k′ and n have been measured are identical (in terms of sensitivity towards disinfectant) to those likely to be present in wastewater to be disinfected. Unfortunately, all of the data summarized in Table 5-11 are based upon the use of pure, laboratory strains of microorganisms. These may have different sensitivities as compared with indigenous wastewater organisms.

As an alternative approach, the model of Collins et al. (Equation 5-65) has been used to describe inactivation of coliforms indigenous to wastewater. Table 5-12 summarizes values for the a and b parameters in Equation 5-65 which have been determined for the inactivation of coliforms indigenous to wastewaters.

Table 5-12. Parameters in the Collins et al. Model Describing Wastewater Coliform Inactivation by Chlorine (177)

System	b l/mg-min	a
Palo Alto—1978	.193	3.15
Palo Alto—1979—Unfiltered	.452	2.22
Palo Alto—1979—Lab Filtered	1.190	2.10
Dublin—San Ramon Lab	1.54	1.62
Dublin—San Ramon Field	0.598	1.79
San Jose	0.246	2.82

Note - Palo Alto and San Jose are non-nitrified secondary effluents. Dublin is nitrified activated sludge.

Use of the Chick-Watson and the Collins et al. models for the estimation of microbial inactivation in wastewater chlorination are described in Section 5.7.3.

5.6 Process Options

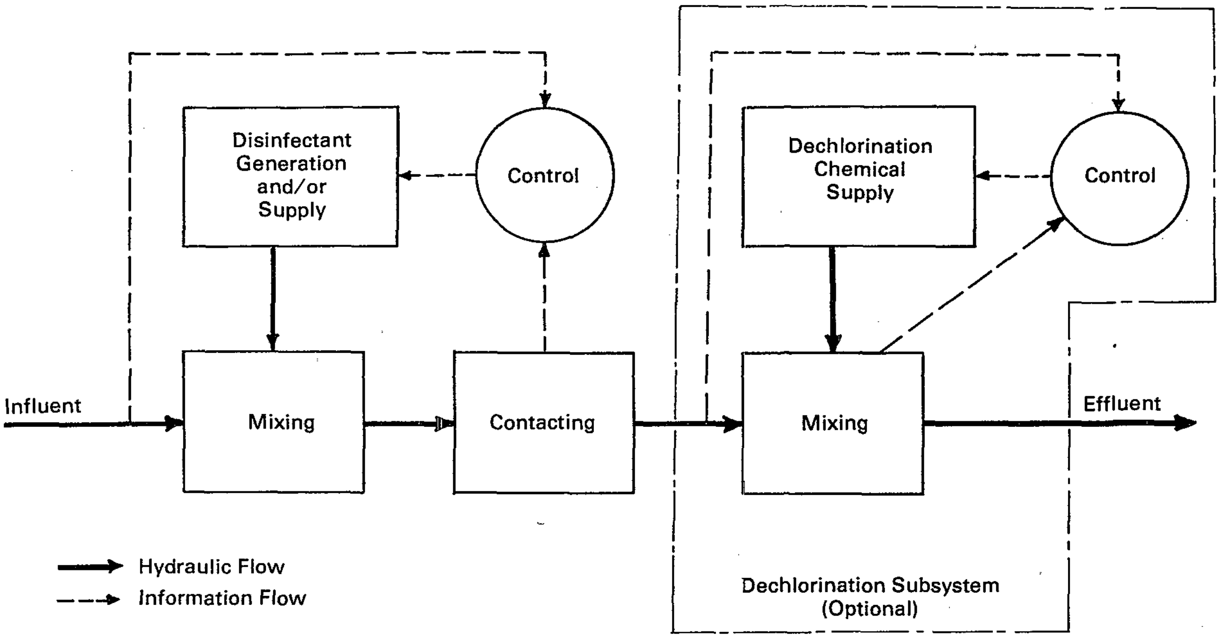
Disinfection systems using chlorine, chlorine dioxide, or bromine chloride, with or without dechlorination share certain common elements. For each of these elements, several possible options exist for the system designer to consider. By specifying the potential options, the many available halogenation disinfection systems may be enumerated.

In this section, these elements and options will be summarized. In the following section, the methodology for detailed sizing and specification of each option element will be discussed.

The general block diagram of a halogen disinfection system, with optional dechlorination, is shown in Figure 5-7. In general, the disinfection system contains four elements and the dechlorination system contains three elements. Flows of information are depicted by dashed lines and flows of material are depicted by solid lines.

The disinfection subsystem elements are chemical generation and/or supply, mixing, contacting and control. In the generation/supply element the disinfecting agent is maintained and fed continuously. In the mixing element, the incoming wastewater is blended with the disinfectant. In the contacting element, the blended wastewater is held for a sufficient period until the desired microbial inactivation has been attained. The control element uses

Figure 5-7. Elements of halogen disinfection systems, with optional dechlorination.



information about incoming wastewater flow and residual disinfectant to adjust the operating parameters of the generation and supply element so as to maintain consistent disinfectant performance.

In the dechlorination subsystem, the dechlorination chemical supply element functions to provide a feed. The mixing elements and the control elements have identical functions as in the case of the disinfection subsystem. In dechlorination, there is generally no substantial contacting element (other than, perhaps, the final outfall pipe).

For the halogen generation/supply subsystem, a number of choices exist as to the option to be selected. These may be enumerated as follows:

- I. Choice of chemical
  - A. Chlorine Gas
    - 1. Method of Withdrawal
      - (a) Gas
      - (b) Liquid/external evaporation
    - 2. Method of feeding
      - (a) Solution
      - (b) Direct Gas
  - B. Sodium Hypochlorite
    - 1. Commercial supply
    - 2. Onsite generation
  - C. Calcium Hypochlorite
    - 1. Solid Feed
    - 2. Solution Feed
  - D. Chlorine Dioxide
    - 1. Method of Generation
      - (a) Chlorine-chlorite process
      - (b) Acid-chlorite process
  - E. Bromine Chloride

For chlorine gas, the choice of gas or liquid withdrawal is dictated primarily by the size of the disinfection system to be designed. It is impractical to use gas withdrawal in very large systems due to limitations on safe withdrawal of gas from ton containers of chlorine. This decision between solution and direct gas feed processes is also based upon the size of the plant; in all but very small wastewater treatment plants (less than 1 MGD), or if chlorine is only to be used on an occasional basis, the solution feed process is generally used.

When sodium hypochlorite is to be used as the disinfecting agent, the chemical may be purchased in bulk as a liquid, or may be produced continuously onsite by electrolysis of brine. The latter process becomes feasible at remote locations, or at wastewater treatment plants where inexpensive sources of brine are available (i.e., plants on ocean coast lines).

The use of calcium hypochlorite is confined primarily to smaller wastewater treatment plants due to economics of chemical supply. In these systems, the chemical may be fed directly as a solid (proprietary tablet feed chlorinators) or may be prepared as a slurry/solution and fed in a similar manner as sodium hypochlorite. Solution feed of calcium hypochlorite has inherent disadvantages in that the addition of water to calcium hypochlorite will produce a calcium carbonate sludge that will foul surfaces and storage tanks.

The choice among gaseous chlorine/sodium hypochlorite/calcium hypochlorite is governed primarily by economic considerations in conjunction with inherent safety and handling hazards with the use of gaseous chlorine. In wastewater treatment plants serving large municipalities, it may be advantageous to use sodium hypochlorite, despite higher unit costs, to minimize risks associated with the transport of liquid chlorine through populated urban areas.

In chlorine dioxide systems, the primary design choice in the generation/ supply subsystem is the type of the generation process. As discussed above, the utilization of chlorite is greater in the chlorine/chlorite process than in the acid-chlorite process. However, in the former process, the possibility for the production of chlorine dioxide containing chlorine at low levels exists, and thus some of the deleterious aspects of chlorine residuals and byproduct reactions may exist.

For BrCl systems, liquid withdrawal with external evaporation is necessary to prevent chemical dissociation and enrichment of gaseous chlorine, which would occur in gas phase withdrawal. Solution feed systems for BrCl plants are universally used.

In the mixing subsystem, the primary design options are the use of devices (static mixers or hydraulic jumps) that dissipate hydraulic head and the use of mechanical mixers (jet mixers, impeller mixers, etc.) that require external motive power such as pumps or motors. In plants where there is little available head between the influent and effluent (such as may exist in wastewater treatment plants serving municipalities with relatively flat topography), the use of the former processes may be infeasible. In general, however, the former processes offer an advantage in that their performance is insensitive to any power fluctuations or outages that may occur, and thus they are inherently more reliable.

The contacting subsystem may consist of a separate contact basin or the outfall pipe itself. The latter option reduces capital costs; however, it necessitates the existence of a sufficiently long outfall pipe, which may not exist at all plants.

The control subsystem serves to produce an effluent in which a consistent effluent quality is attained at a minimum chemical dose. Three options exist: manual, or "open-loop" control (relying upon operator intervention), simple flow-proportional feed-forward control, and true closed-loop feedback control (relying upon continuous sensing of halogen residual, usually in conjunction with feed-forward flow proportional control). The last option provides minimum chemical utilization, but is more capital intensive. In very small plants, where dechlorination is not practiced, manual, or feed-forward control is sufficient. However, in plants subject to effluent constraints on chlorine residual, or in very large plants where small decreases in chemical dosage result in large economic savings, the last option may be most desirable. Also, in dechlorination plants, the consumption of dechlorinating chemical will be a function of the chlorine residual leaving the chlorine contact system, and thus the use of true feedback control will help to minimize costs associated with the dechlorination process.

For the dechlorination sub-system, the options for chemical supply and generation are as follows:

I. Choice of chemicals
  A. Sulfur Dioxide
    1. Method of Withdrawal
      (a) Gas
      (b) Liquid/External evaporation
  B. Sulfites or Bisulfites

The choice of sulfur dioxide versus sulfites/ bisulfites is dictated by considerations similar to those that govern the selection of gaseous chlorine versus calcium or sodium hypochlorites. If sulfur dioxide is used, selection of gas or liquid withdrawal is governed by considerations identical to those that pertain to the selection between these options for the use of gaseous chlorine. It is virtually universal to use solution feed processes for sulfur dioxide dechlorination processes.

The types of mixing processes that may be used for dechlorination are identical to those used for chlorination. Due to the rapidity of the dechlorination reaction, and the absence of deleterious side reactions, it is considered unnecessary to use separate contacting systems for dechlorination, since the short residence time in the mixing and outfall structures generally suffices for the reduction of halogen residuals to desired values.

The choice of process control options for dechlorination is identical to that for halogenation processes, with one exception. In wastewater treatment plants that discharge effluents into low dilution receiving streams, the use of manual control will result in overdosing of dechlorination chemicals into the wastewater. This may produce deoxygenation of the effluent and the receiving stream, with undesirable results. Hence, in some plants, even where simpler control systems for chlorination are used, flow-proportional, or feedback control for dechlorination may be necessary to minimize the possibility for deoxygenation in the effluent and the receiving stream.

5.7 Design Coordination

As previously indicated, an overall halogen disinfection system may be subdivided into several components. In this section, the procedures for sizing and specifying each of these components are discussed.

5.7.1 Disinfectant Generation and Handling

In the case of chlorine or bromine chloride this subelement includes possible on-site generation (in the case of hypochlorites), storage, metering, and mixing of the disinfectant for delivery to the point of application. In the case of chlorine dioxide, the items included are identical to the above, except on-site generation is mandatory.

5.7.1.1 Chlorine and Hypochlorites

The first decision to be made by the designer is whether gaseous chlorine or liquid sodium hypochlorite is to be the chemical used for disinfection. For small wastewater treatment plants, generally up to 100,000 gpd (378 m^3/d), proprietary solid hypochlorite systems are available. However, these are generally not used in larger plants.

The basis for deciding between chlorine gas and sodium hypochlorite as a disinfecting chemical is primarily one focusing upon risk of chlorine transport versus additional costs of sodium hypochlorite. Only in rare instances will the delivered cost of NaOCl solutions per unit available chlorine be less than the delivered cost of gaseous chlorine.

If sodium hypochlorite is used as the chlorinating agent, it may be purchased in strengths of 5 to 15 percent, or it may be generated onsite. If purchased, storage tanks for sodium hypochlorite solution are necessary. It is recommended that the size of these storage tanks should equal the amount of disinfectant required for a time equal to the shipping time from the vendor plus a 15 day emergency reserve against strikes and transportation problems (149). At smaller wastewater treatment plants, the reserve will be higher due to the necessity of purchasing minimum loads of solution to achieve reasonable chemical costs. The concentration of available chlorine in NaOCl solutions diminishes with time, and allowance must be made for this in computation of reserve requirements. A loss of 0.031 percent/d from 10 percent NaOCl solutions, and 0.075 percent/d from 15 percent NaOCl solutions have been reported (46).

Example: A wastewater treatment plant has an average daily flow of 0.116 m^3/s (2.6 MGD) and an average chlorine dose of 5 mg/l is necessary. Chlorine is to be supplied in the form of 10 percent NaOCl solution. The shipping time for the vendor is 3 days. Determine the required storage capacity assuming a 15-day emergency reserve.

Step 1. Hourly chlorine requirement = (0.116 m^3/s) (5 g/m^3) (3600 s/hr) (24 hr/d) = 50 kg/d

Step 2. Daily NaOCl volume = (50 kg /d)/ (100/ kg/m^3) = 0.5 m^3/d

Step 3. Storage volume = (3 d + 15 d) (0.5 m^3/d) = 9 m^3

Step 4. Correct for hypochlorite decay. (18 d)(0.031 percent/d) = 0.56 percent loss in 18 d.
(9 m^3) (10 percent)/(10-0.56 percent) = 9.5 m^3

Therefore a storage tank of at least 9.5 m^3 (2,500 gal) would be necessary.

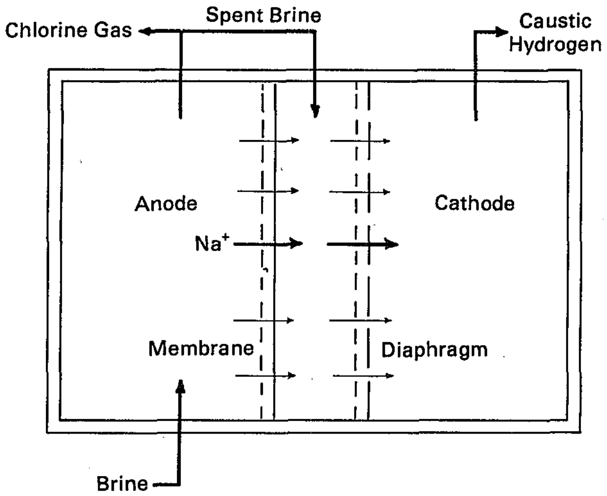
Onsite Generation. For onsite generation of sodium hypochlorite, several systems are available. All require a source of DC electricity (rectification facilities may be included as part of the generation package). In addition, a source of brine is necessary, and particularly if seawater or rocksalt is used, pretreatment of brine for the removal of calcium carbonate and iron (which precipitates in the alkaline side of the electrolysis cell) may be required.

In one such system (Chloromat-TM, Ionics Inc.), electrodes are separated by two membranes. One membrane is a cation exchanger permitting only the passage of cations (such as sodium), and in this manner, chlorine gas is evolved, which is then immediately dissolved to form hypochlorite solution. In addition, caustic (NaOH) and hydrogen gas are produced as by-products. The hydrogen gas must be flared off to minimize fire hazards. A schematic of this electrolytic system is shown in Figure 5-8.

Individual cells are manifolded together to produce hypochlorite generation capacities of 9-150 kg/hr (500-8,000 lb NaOCl/d). Capital costs have been estimated for the electrolytic units, exclusive of power rectification and brine pretreatment, and fit the following equation (150):

$$\text{Capital Cost (\$, 1974)} = 10386\,(\text{kg NaOCl/hr})^{0.66} \qquad (5\text{-}66)$$

Figure 5-8. **Schematic of Chloromat^{TM} (Ionics, Inc.) electrolytic hypochlorite cell (Reproduced by permission of the Water Pollution Control Federation).**



Replacement of the cell anode and membranes has been estimated as a contribution to operating costs to be $4.40-$13.20/1000 kg NaOCl (150). Power requirements vary with the desired product strength of hypochlorite and are 3.5-5.5 kWh/kg (1.6-2.5 kWh/ lb) available chlorine, increasing with increasing strength (Ionics, Inc.); sodium chloride requirements are 2 kg/kg available chlorine without brine recycle, and 1.75 kg/kg with brine recycle. Doan and Haimes presented a case study of the application of this system to a 24 MGD (1 m^3/s) wastewater treatment plant at Amherst, NY (151). In that case, brine pretreat ment with sodium carbonate and caustic (some of which may be recovered from the electrolysis operation) to less than 10 mg/l (as calcium carbonate) hardness and less than 1 mg/l iron was needed since brine was produced from rocksalt. Capital costs for the entire system (electrolytic apparatus, power rectification, and brine pretreatment) were estimated at $1,400,000 (1978 dollars).

Electrolytic hypochlorite generating equipment suitable for on-site use is also manufactured by Engelhard Corporation (152) and Diamond Shamrock (61).

If onsite hypochlorite generation is used, short term NaOCl storage (several days) may be desirable to permit maintenance of the electrolytic units, and to eliminate the need for continuous operation. In wastewater treatment plants serviced by electric utilities with off-peak billing schedules, it may also be desirable to operate the electrolytic generation equipment only during offpeak hours and to store NaOCl solution for use during peak periods.

Gaseous Chlorine Supply. When chlorine gas supplied in cylinders or containers is to be used as the disinfecting chemical, it is important to decide upon the necessary inventory in the same manner as outlined for hypochlorite. In addition, the working inventory should be increased by an amount equal to the number of chlorine containers simultaneously under service. However, an inventory above that minimum calculated may be desirable since the shipping cost of bulk chlorine in the form of cylinders or containers is a function of the number of cylinders or containers shipped, and hence a larger working inventory may enable the treatment plant to take advantage of quantity price discounts.

Chlorine is generally supplied either in 150 lb cylinders, ton containers, or larger rail or truck tankloads. In such containers, liquid chlorine exists in a pool at the bottom of the container and is in equilibrium with gaseous chlorine in the container vapor space.

The chlorine may be withdrawn for use from ton containers either as a gas or as a liquid. If it is withdrawn as a gas, it may be directly piped to the ejector, which produces a chlorine solution in water. If it is withdrawn as a liquid, it is necessary to allow the liquid chlorine to vaporize in an external evaporator prior to its dispersal in water.

There is a *maximum rate of safe withdrawal* of gaseous chlorine from cylinders or containers. This limit exists because, as gaseous chlorine is withdrawn, liquid chlorine must vaporize to maintain equilibrium within the container. As this vaporization occurs, heat must be withdrawn from the surroundings. If the rate of gas withdrawal is too high, the rate of heat abstraction from the surroundings will not be sufficient to prevent cooling of the chlorine to low temperatures. At a sufficiently low temperature, the vapor pressure of chlorine will be below the exit pressure of chlorine gas from the cylinder, and thus no further flow will occur. To prevent this, a maximum gas withdrawal rate is permissible. This is dependent upon surrounding temperature and exit gas pressure. For a 21°C (70°F) external temperature and 2.4 atm exit pressure, the maximum gas withdrawal rate from a 68 kg (150 lb) cylinder is 0.8 kg (1.75 lb)/hr, and from a 910 kg (1 ton) container is 6.8 kg (15 lb)/hr (44). For liquid withdrawal, no such cooling occurs (since the heat of vaporization is supplied by an external evaporator), and the maximum withdrawal rate is set primarily by the hydraulics of chlorine flow through the fittings; it is 90 kg (200 lb)/hr and 180 kg (400 lb)/hr, respectively, for the two types of containers.

In general, a liquid withdrawal system will be used when the number of parallel containers necessary to maintain the average daily chlorine flow while obeying the maximum rate of safe withdrawal per container becomes unwieldy. For example, if 90 kg (200 lb)/hr of chlorine are to be supplied from ton containers, at least 14 ton containers must be simultaneously manifolded (90 kg/hr÷6.8 kg/hr/container). In addition, treatment plants using bulk rail or truck shipments generally practice liquid withdrawal. Benas presents a detailed case study of bulk chlorine storage utilized at a 2.8 m^3/s (65 mgd) and a 1.3 m^3/s (30 mgd) treatment plant in San Francisco, CA (153).

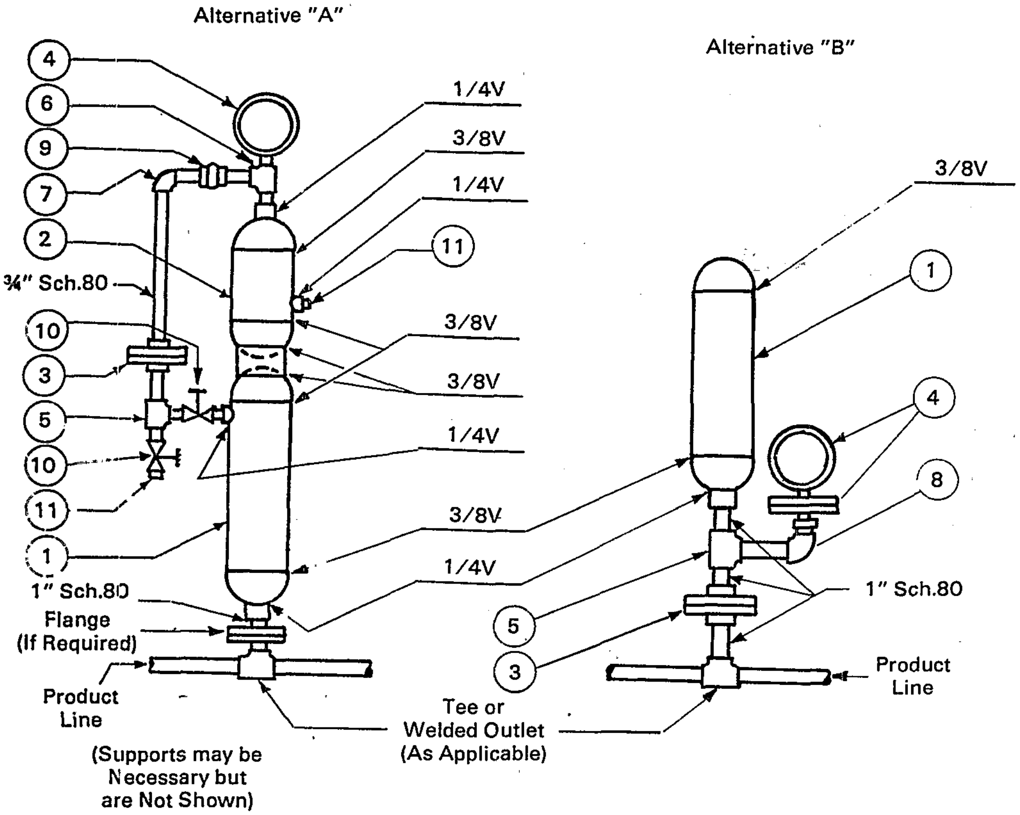
If liquid withdrawal is practiced, the contents of a single container are piped to an evaporator. Multiple evaporators and containers may be used; however, it is undesirable to manifold several cylinders or containers each with liquid withdrawal due to the possibility of liquid transfer between containers.

In piping systems where liquid chlorine is transported, particular attention must be given to preventing the possibility of heating of the pipeline. If this occurs (such as by exposure to sunlight), the liquid flowing full in the pipeline will expand and could hydrostatically rupture the line. To guard against this possibility, an emergency expansion chamber (or chambers) equivalent to at least 20 percent of the volume contained in the liquid chlorine pipeline must be provided, and should be protected by a rupture disk rated at less than the maximum bursting pressure of the supply line. Figure 5-9 details this provision. In addition, pipelines should be insulated to minimize heating. Further information on the design of chlorine pipelines may be found in Chlorine Institute Pamphlet #60 (154).

While it is desirable to minimize the length of liquid chlorine pipe, to reduce the risk of leakage in the event of pipeline breaks, instances exist wherein this may be difficult. Cameron reported on the use of a 490 m (1,600 ft) liquid chlorine pipe handling a flow of 1,365 kg (3,000 lb)/d of chlorine supplied to the City Island Plant at Atlantic City, NJ (155).

Chlorine evaporators are usually supplied as integral units by various suppliers and consist of a pressure vessel surrounded by a heating bath (hot water or thermostated electrical) that serves to convert the incoming liquid chlorine into a gaseous chlorine product stream. Evaporators are inherently high in energy requirements, and manufacturers' literature should be consulted for such details. Design and safety information for evaporators are available from the Chlorine Institute (156). In particular, pressure relief devices leading to gas absorption tanks and super heating of effluent gas are necessary. It is recommended that an absorption tank (of caustic or lime solution) capable of neutralizing all liquid chlorine in the supply line plus 70 kg (150 lb)/evaporator be provided, and also that the exit gas be 11°C (20°F) higher in temperature than the boiling point at the exit pressure (157). This latter requirement will prevent reliquification in the subsequent gas piping as the pressure of the gas is reduced.

Figure 5-9. Chlorine expansion chambers (reproduced from Chlorine Institute Pamphlet #60, 1982 by permission).



Item	Name of Part
1	Primary Expansion Chamber (Note 1)
2	Secondary Expansion Chamber (Note 2)
3	Rupture Disc (400 psi) (Note 3)
4	Pressure Indicator or Alarm Switch, 1/2" Conn. (Note 4)
5	Tee (Note 5)
6	Reducing Tee (Note 5)
7	Elbow (Note 5)
8	Reducing Elbow (Note 5)
9	Union (Note 5)
10	Valve (Note 5)
11	Plug (Note 5)

Notes:

1. Capacity—20% of line volume
2. Capacity—10% of line volume
3. 400 psi setting suitable for many systems, setting must not exceed system design pressure
4. Liquid-filled protective diaphragm optional
5. Fittings shall be forged carbon steel, 3000 lb cwp

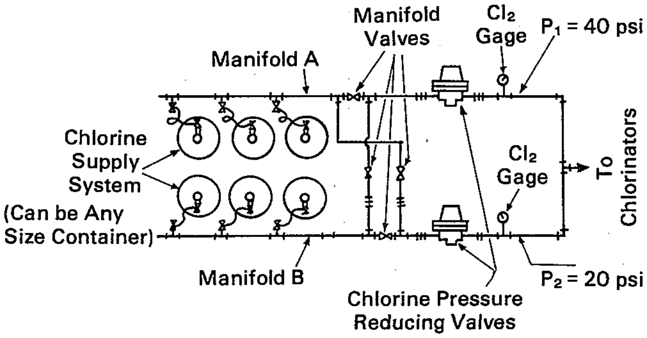
For gas withdrawal systems, two major design options exist. These have been referred to as pressure/vacuum/pressure (PVP) and all vacuum systems (158). In the former system, chlorine gas is withdrawn from the supply containers into a pipe at greater than atmospheric pressure, whereupon it flows to a chlorinator at which point the pressure is reduced below atmospheric until the chlorine flows to the ejectors. In the all vacuum system, the chlorine gas is immediately expanded to a pressure below atmospheric, eliminating the initial pressure line. In a PVP system, the initial pressure line introduces some potential risk of gas loss in the event of line rupture; however, it is easier to manifold multiple cylinders than in the all vacuum case.

In gas withdrawal systems, two major concerns are temperature in cylinders and supply lines, and impurities in the gas. Filters should be placed inline as close as possible to the chlorine cylinder or manifold to remove liquid droplets and solid impurities (80). It is desirable that the chlorine gas lines and subsequent connections be warmer than the chlorine cylinders or containers to prevent liquification of chlorine in the downstream lines. In an all vacuum system (as opposed to a PVP system) this possibility is minimized since the gas pressure is reduced considerably below the vapor pressure of saturated chlorine gas at the existing temperature. In PVP systems, it is also desirable to have duplicate piping and manifold systems to facilitate cleaning.

Gas cylinders manifolded together should be at similar temperatures. This may deserve special consideration in wastewater treatment plants located in areas of extreme temperature variability. Otherwise, there will be a transfer of contents between containers, which could lead to overfilling and container rupture. Nevertheless, multiple ton containers (or gas cylinders) may be readily manifolded to provide easy switchover of cylinders upon emptying. One such configuration is indicated in Figure 5-10 (80). Pressure reducing valve 1 (PRV 1) is initially set at a higher pressure, e.g. 375 kPa (40 psig), than PRV 2, e.g. 238 kPa (20 psig). Initially, gas will flow only from the cylinders in header A. When the pressure in these cylinders approaches 20 psig, flow will commence from header B. The valves connecting the individual cylinders to header A can now be shut, and the cylinders replaced (since 20 psig is below the vapor pressure of chlorine at usual ambient temperatures, relatively little chlorine will be remaining in these cylinders). Upon replacement, valves a, b, c, and d can now be switched such that the header containing the "old" cylinders is connected to the 375-kPa (40-psig) pressure reducing valve.

In both liquid and gas withdrawal systems, it is desirable to have weighing scales for each gas cylinder under service. Since chlorine liquid and gas are at equilibrium at a constant pressure given a fixed temperature, pressure will only drop from a cylinder upon exhaustion of all of the contained liquid and will thus provide little warning to the operator of imminent emptying. Thus, the only reliable means of inventory control, as well as verification of chlorine consumption, is weight. A variety of weighing devices exist; however, these should permit the weighing of cylinders or containers without lifting.

Figure 5-10. Chlorine manifold and switchover system (after White, 1972 reproduced by permission of Van Nostrand Reinhold).



The chlorine gas from the cylinder manifold, or in the case of a liquid withdrawal system, from the evaporator, flows to a chlorinator. The function of this chlorinator is to regulate the flow of chlorine gas, and to couple with the control elements so that this chlorine gas flow rate (and ultimately the chlorine dosage to the wastewater) varies with the chlorine demand.

Chlorinator Design and Hydraulics. Chlorinators are sized at a variety of gas flow rates. Typical upper limit gas flow rates are 100, 200, 500, 1,000, 2,000, 4,000, 6,000 and 8,000 lb chlorine/d (1.9, 3.8, 9.4, 18.8, 37.5, 112.5 and 150 kg/hr). The lower limit to flow is 1/20 of this upper limit. If there are frequent conditions necessitating a chlorine dose below 1/20th of the maximum, it may be necessary to have two chlorinators in parallel, with a second used to regulate gas flow under low flow conditions.

In PVP systems, the lines between gas supply and the chlorinator should be on an uphill grade, to help minimize liquid chlorine carryover. Typically the exit pressure reducing valve at the gas source is set at 172 kPa (25 psia). In ton containers, it is customary to install a blind drip leg prior to the chlorinator as a final trap against liquid chlorine and debris passage, which could damage chlorinator components.

The output from the chlorinator in a PVP system is a regulated chlorine gas flow stream to the ejector where an aqueous solution is produced (however, often the ejector may be physically housed close to or in the same mounting as the chlorinator). The chlorine ejector (also variously called the eductor or injector) is a device that conveys the chlorine gas into the converging region of a jet of high velocity water. The diverging water downstream of the point of gas injection produces a vacuum, which draws the chlorine gas into solution and serves to convey the gas stream from the pressure reducing valve at either the gas manifold or the evaporator. Figure 5-11 illustrates the typical configuration of ejectors with fixed water flow rates, and with variable water flow rates set by a variable orifice. Additional configurations are available wherein the gas flow rate may be controlled by a variable orifice on the gas inlet pipe.

The inlet water for the ejector may be either city water, or, more commonly, chlorinated secondary effluent. As will be indicated below, the inlet water supply to the ejector must be at relatively high pressure, and therefore if secondary effluent is used, a booster pump is required. In all but the smallest wastewater treatment facilities it will be less expensive to provide this booster pump for secondary effluent ejector water supply rather than to rely upon purchased city water.

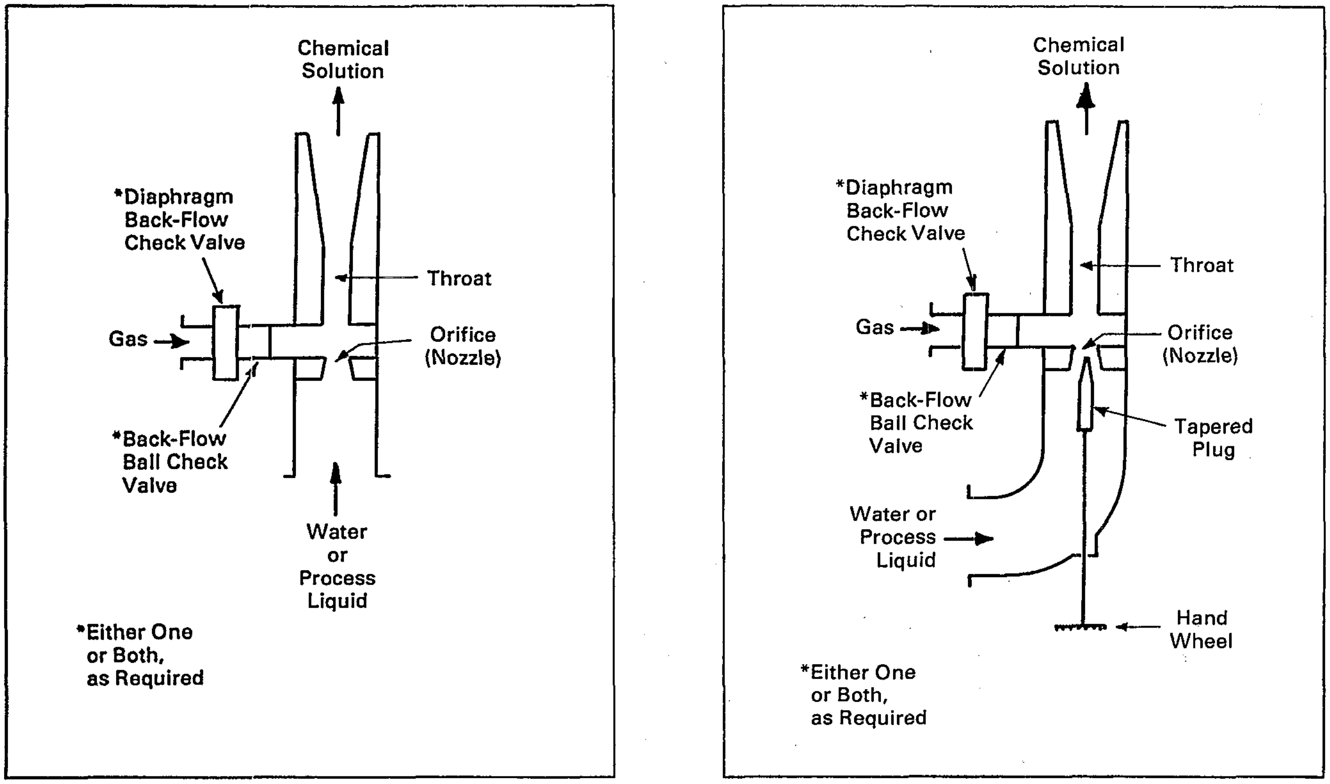
Important considerations for ejector sizing include (157):

1. The ejector water supply should be no less than that required to produce a chlorine solution of less than 3,500 mg/l to minimize the possibility of chlorine gas volatilizing from the feed solution.
2. The ejector back pressure (sum of static and dynamic pressure losses between the outlet side of the ejector and the mixing device) should be at least 1.2 m (4 ft) of water (1.7 psi) to prevent chlorine release in the solution line.
3. The ejector should be installed as close as possible to the mixing point to minimize the time lag in the solution line and the back pressure at the ejector.
4. The vacuum line carrying gas between the chlorinator and the ejector should have a pressure drop of less than 3.8 cm (1.5 in) of mercury (0.7 psi).

Specific information needed to size and specify the ejector is the following (Fischer & Porter):

- maximum capacity of the chlorinator;
- inlet water supply pressure (this can be assumed, with several values, for the purpose of determining ejector water booster pump capacity);
- back pressure at the ejector outlet;
- elevations of the ejector and the diffuser; and
- distance between ejector and diffuser, and any piping bends or constrictions.

Figure 5-11. Schematic of fixed and variable orifice ejectors. (Reproduced courtesy of Fischer and Porter.)



Fixed Orifice Ejector

Variable Orifice Ejector

Ejector sizing procedures will be illustrated by example. Assume a maximum chlorine application rate of 18.8 kg/hr (1,000 lb/d) has been calculated. It will also be assumed that the booster pump supplying feedwater to the ejector can provide a pressure of 515 kPa (60 psig)). At the point of mixing (the outlet side of the diffuser nozzles) the pressure is 1.8 m (6 ft) of water or 17.9 kPa (2.6 psi). The diffuser is located 1.5 m (5 ft) higher than the ejector. There are 100 ft (30.5 m) of equivalent pipe (pipe length plus length equivalents of fittings obtained from standard hydraulic handbooks) between the ejector outlet and the diffuser. It is assumed that the headloss through the diffuser itself is 13.8 kPa (2 psi) or 1.4 m (4.6 ft).

For one possible ejector (Fischer & Porter, Figure 5-12), manufacturer's data are available for acceptable combinations of inlet and back pressure and water flow rate. Since the inlet pressure is assumed at 515 kPa (60 psig), from Figure 5-11, the acceptable back pressure is 273 kPa (25 psig) and the acceptable water flowrate is 2.8 l/s (45 gpm). Now it is necessary to verify that the system backpressure is below the maximum acceptable (273 kPa). The backpressure is the sum of the following elements:

- elevation of diffuser with respect to ejector (A);
- back pressure on diffuser outlet (B);
- friction loss through diffuser (C); and
- friction loss in piping (D).

From the assumptions, A + B + C = 5 ft + 6 ft + 4.6 ft = 15.6 ft or 6.8 psig. In order to calculate D, the pipe diameter needs to be specified. Assume that 1-¼ in ID Schedule 80 PVC pipe is used. Frictional losses may be estimated using the graph in Figure 5-13. At a water flow of 45 gpm in 1-¼ in PVC Schedule 80 pipe, the frictional loss is 37 ft/100 ft. Therefore the total

Figure 5-12. **Ejector sizing curve (Reproduced by permission of Fischer and Porter).**

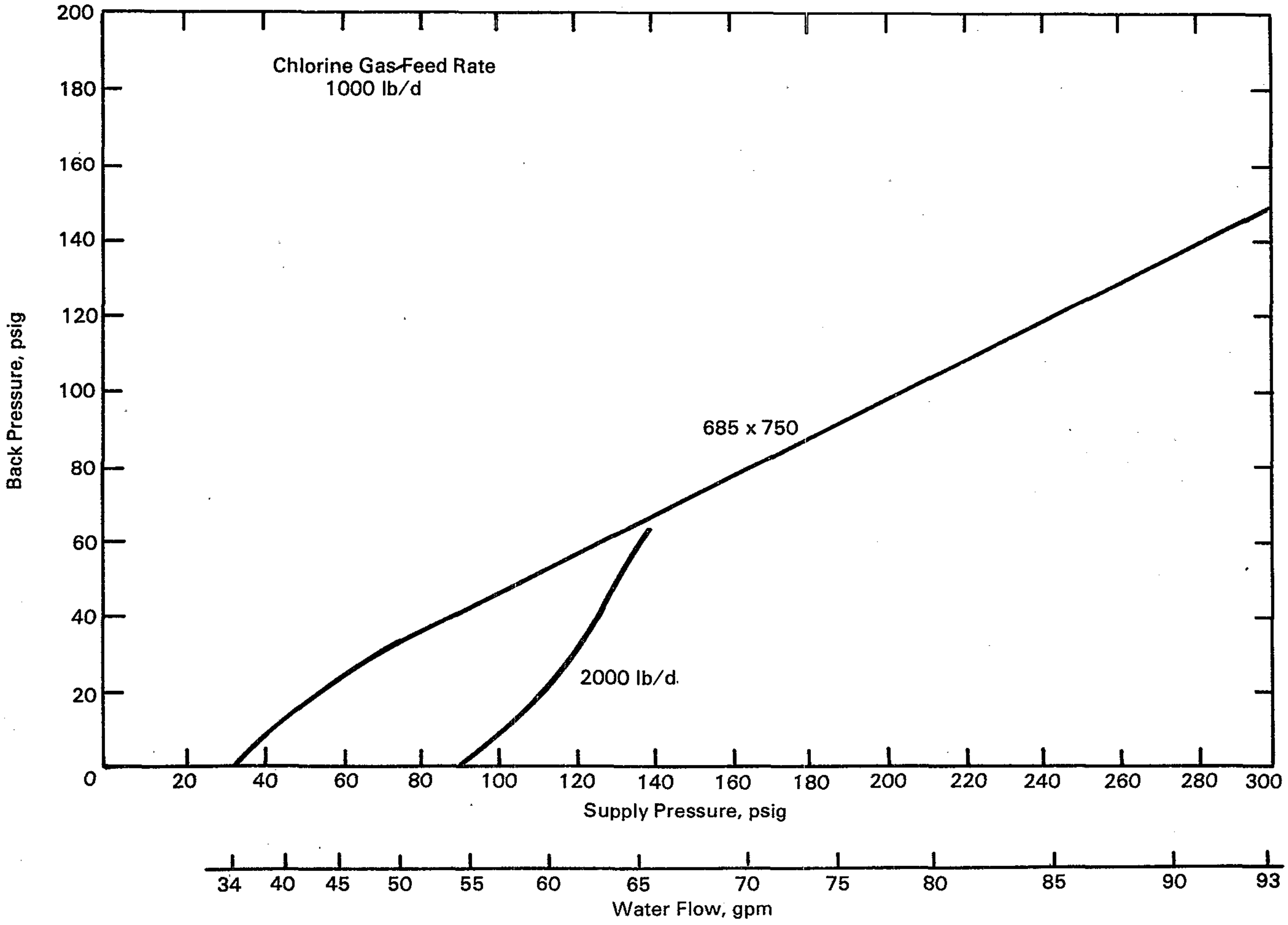
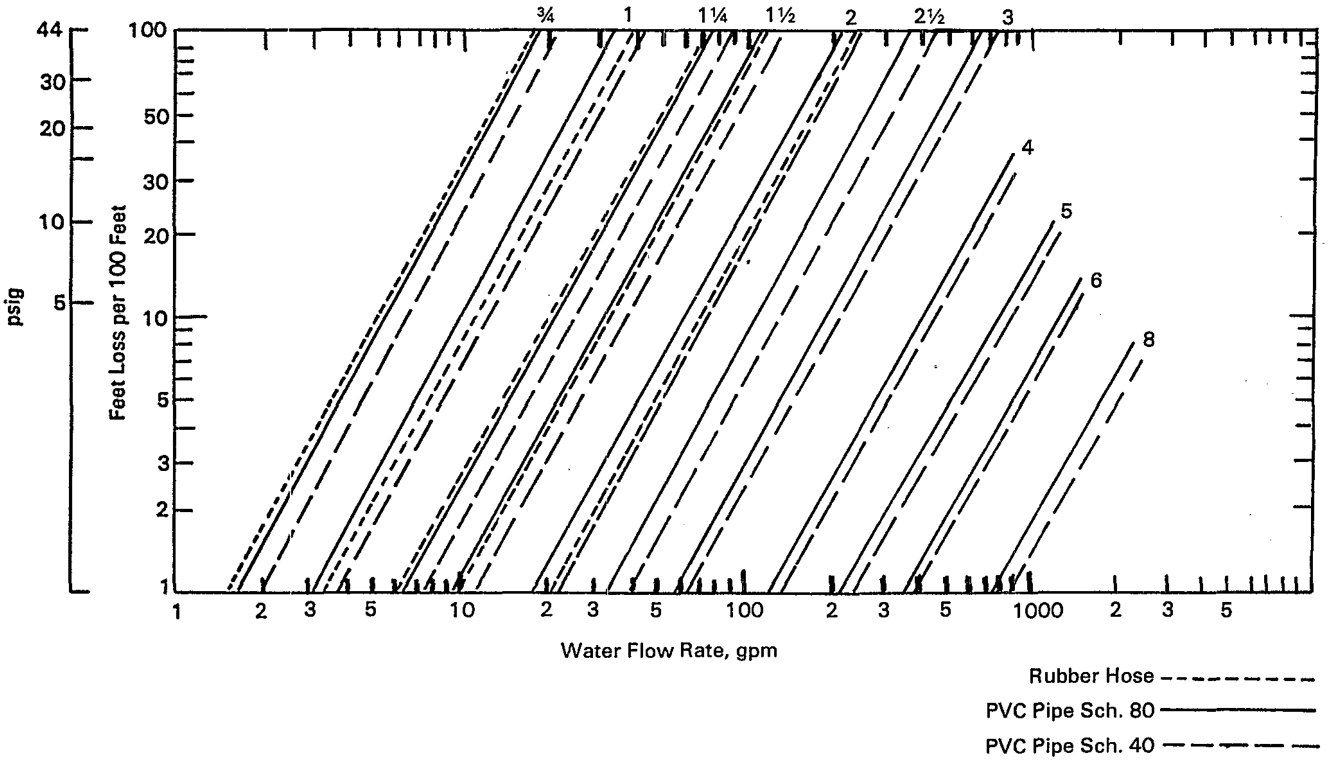


Figure 5-13. Frictional losses in solution piping (Reproduced courtesy of Fischer & Porter Co.).



system backpressure (A+B+C+D) is 53 ft or 260 kPa (23 psig). Since this is below the maximum acceptable back pressure, the design is satisfactory. At this point, a different inlet assumed pressure may be tried to investigate the trade-offs inherent in various booster pumps. It is also necessary to determine the transit time in the solution piping, which is necessary, as will be noted later, to ensure that unreasonably long control lags are not present in the system.

The flow rate in the solution pipe is 2.8 l/s (45 gpm). Assuming that the entire pipe is straight, the volume of a full 100 ft pipe of 1-¼ in ID is 24.5 l (0.85 cu ft). Therefore, the residence time in the line is 0.14 minutes (8.5 sec) which is reasonably short. A larger diameter line, while serving to reduce back pressure, would also have increased this system delay time. As will be noted, the total of the lag time in the solution line, the time for analysis, and the residence time between mixing and sample collection for analysis in the contactor should be under several minutes for best performance when residual control or compound loop control is practiced.

As a more complex example, a second manufacturer (Wallace and Tiernan Division of Pennwalt, Figure 5-14) produces ejector specifications in which the relationship between water flow rate and injector pressure is more variable than in Figure 5-12. For the above example, at an ejector inlet pressure of 515 kPa (60 psig) and 273 kPa (25 psig) backpressure, a water flow of 15.5 l/s (245 gpm) is required for this particular ejector. Losses A, B and C will remain as above (15.6 ft). If schedule 80 PVC pipe is used, from Figure 5-12, to keep the total backpressure below 25 psig (273 kPa) (and, in other words, to keep the frictional losses in the solution line below 231 kPa (19 psig), 3 inch (76 mm) ID pipe must be used. In this case, the residence time in the solution line is 9 sec, or nearly identical to the lag obtained with the preceding ejector. In this case, however, the necessary ejector water flow rate is substantially higher, which may necessitate the use of more costly pumping equipment and larger operational pumping costs. It should be noted that for both chlorine solution piping itself and the ejector, maximum pressure limits exist above which material failure becomes possible. Generally these pressures are in excess of 100 psig (790 kPa); however, they decrease at above-ambient temperature. Prudent practice dictates that the inlet water supply pressure be maintained below that specified by the ejector or pipe manufacturer.

At this point, knowing the flow rate of chlorine solution and the desired maximum headloss through

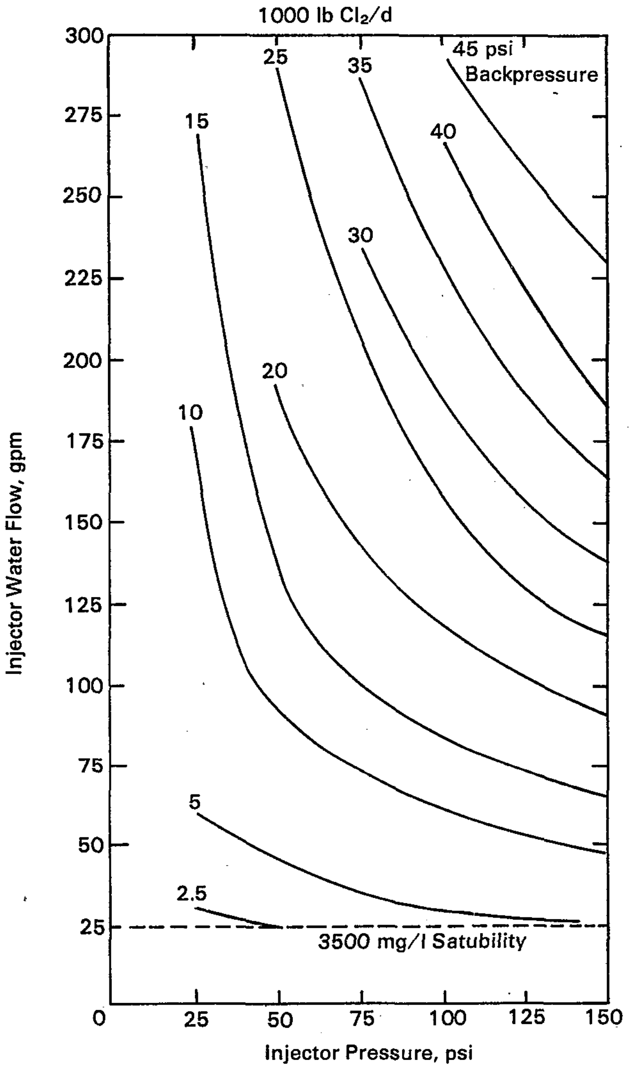


Figure 5-14. Ejector sizing curve (Reproduced by permission, Wallace & Tiernan Division, Pennwalt Corp.).

Note: Extrapolations and interpolations should not be used.

the diffuser structure, one can design the diffuser itself. For small solution flows, a single spray nozzle on the end of the solution pipe or hose may be adequate. Figure 5-15 can be used to determine the headloss for this design as a function of hose diameter.

For the solution flow rates used in the example case, however, the use of a diffuser with multiple perforations is more likely. Such a system consists of a pipe lateral of diameter equal to the solution pipe itself, with multiple outlet holes. This diffuser pipe may be suspended horizontally or vertically in a pipe or in a rapid mixing chamber or device. Figure 5-16 provides a nomograph for the estimation of headlosses through perforated pipe diffusers. In the present example, given a solution flow of 2.8 l/s (45 gpm), a diffuser with a headloss of less than 13.7 kPa (4.6 ft) is desired.

A vertical line at 1.4 m (4.6 ft) headloss is constructed to the point of intersection with the loss of head curve. At this point of intersection, a horizontal is constructed to the left. Each intersection of this horizontal with an orifice diameter represents suitable combination of chlorine solution flow and orifice diameter necessary to achieve the design headloss. The number of orifices needed is obtained by dividing 2.8 l/s (45 gpm) (the total solution flow) by the flow per orifice and rounding up to the nearest whole number. Thus, 6 ½-in orifices would be needed; 5 9/16-in orifices, etc.

Figures 5-15 and 5-16 could also be used to design the diffuser system for a sodium hypochlorite feed unit in which the flow rate of sodium hypochlorite is known. For example, to feed 18.8 kg/hr (1,000 lb/d) of chlorine using a NaOCl solution containing 10 percent available chlorine:

$$\frac{1000 \text{ lb/d}}{100 \text{ g/l}} \times 454 \text{ g/lb} = 4{,}540 \text{ l/d}$$
$$= 1{,}200 \text{ gpd}$$

would be needed. This flow rate is equal to 0.05 l/s (0.83 gpm). Hence, from Figure 5-15, if a ½-in hose spray nozzle is used, a headloss of approximately 8.1 kPa (2.7 ft) would result.

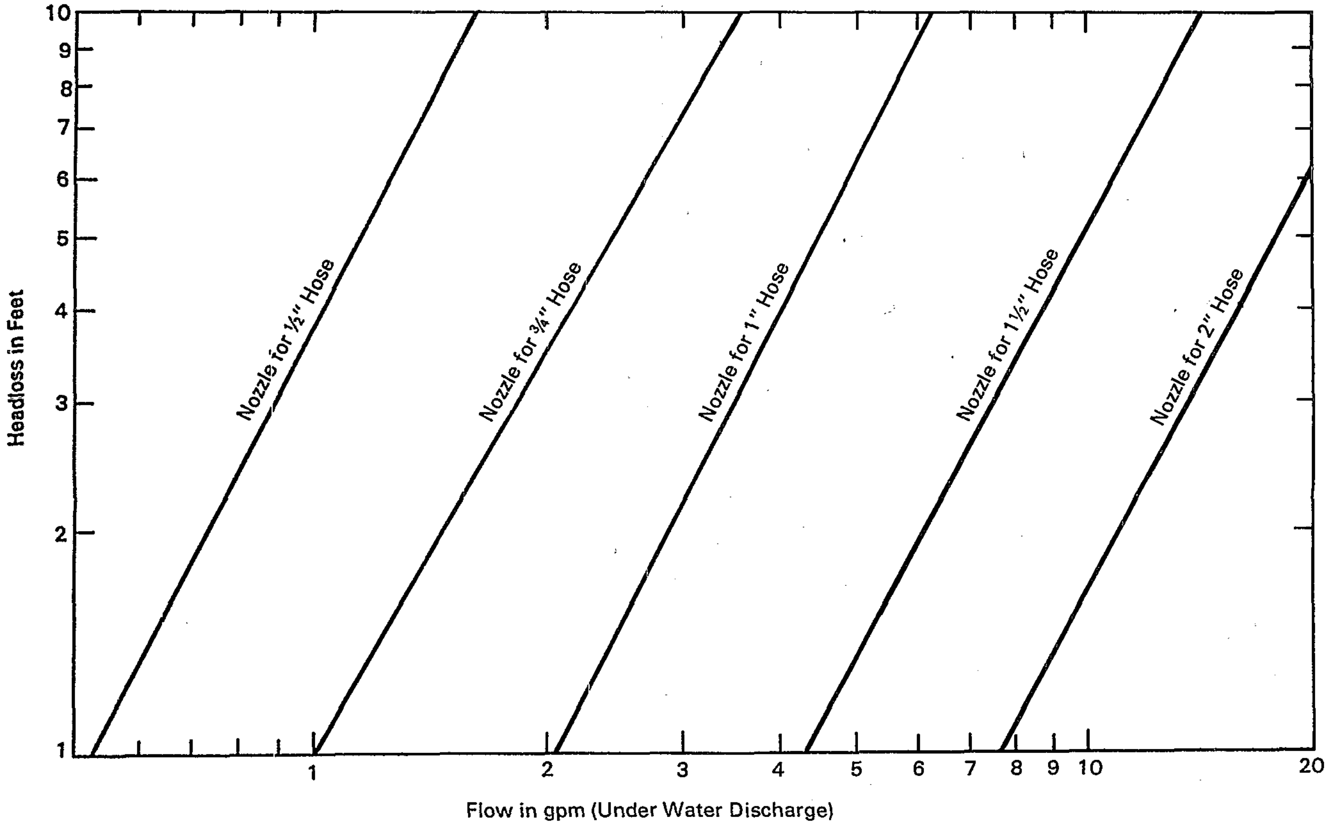
Materials Compatibility. In the design of piping and handling systems for both chlorine and hypochlorite, attention must be given to the corrosive nature of both of these materials. Factors involved in the handling of liquid chlorine have already been discussed. For dry chlorine gas, copper, iron, and steel pipe are satisfactory (44). For hypochlorite solutions (including the solution piping from the chlorine ejector), rubber, ceramic, glass, Tyril, saran, PVC, vinyl and Hypalon are all suitable materials (159).

5.7.1.2 Bromine Chloride Supply

Systems for the supply and handling of bromine chloride are similar to those for the supply and handling of gaseous chlorine. BrCl is supplied in cylinders of 68 kg (150 lb) capacity and containers of 1,365 kg (3,000 lb) capacity. Since the vapor phase above liquid BrCl is relatively enriched in molecular Cl_2 due to gas phase dissociation, it is necessary to withdraw BrCl from either type of container from the liquid phase.

The vapor pressure of BrCl is less than that of gaseous chlorine at identical temperatures. This means that,

Figure 5-15. Headloss thru spray nozzle diffusers. (Reproduced courtesy of Wallace & Tiernan Division, Pennwalt Corp.)



while similar liquid evaporators may be used for BrCl service, it is necessary to attain a higher degree of superheating of the BrCl product gas to prevent reliquifaction. White recommends that 30°F superheat be provided to the product gas (24).

Bromochlorinators (analogous to chlorinators) and solution ejectors may be designed on a similar basis to those used for chlorine service. The only substantial differences are that materials specified for chlorine service may not necessarily be suited for BrCl service. It is recommended that piping should be black iron or carbon steel, but not cast iron (56). Other suitable materials for BrCl service are tantalum, nickel-molybdenum alloy, lead, silver, platinum, glass ceramic and some plastics (56). Early pilot plant studies noted frequent clogging of BrCl evaporators (42). This problem persists in BrCl evaporators, due, apparently, to substantially greater impurities existing in BrCl as compared to Cl_2. As a result, at least one manufacturer has withdrawn its designs for BrCl evaporation systems, and is currently testing direct liquid BrCl feeding equipment (T. Zeh, Capital Controls Co., personal communication).

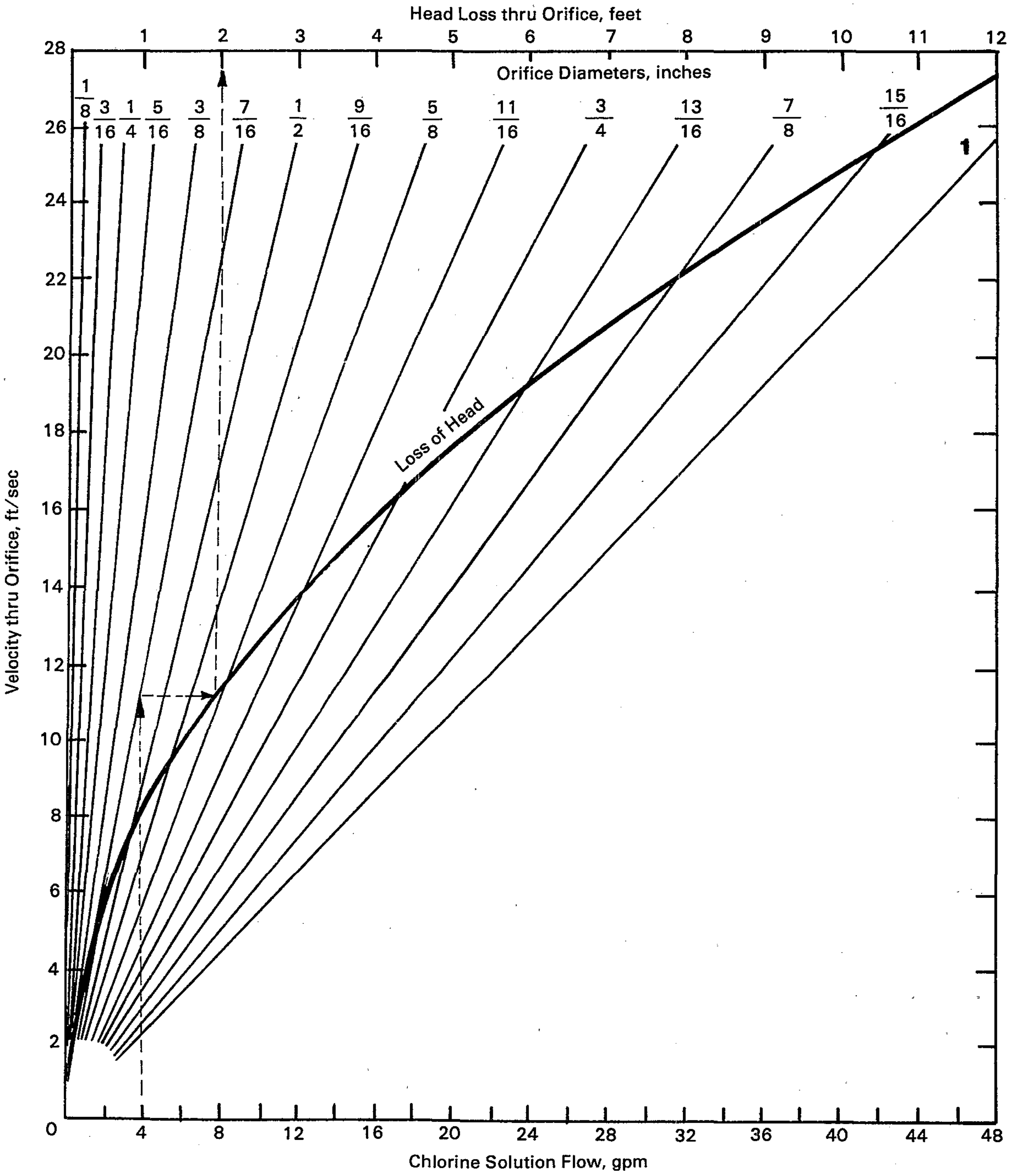
Similarly, design of ejectors and diffuser structures for the resulting bromine chloride solution follows principles noted above for gaseous chlorination systems.

5.7.1.3 Chlorine Dioxide Supply

It is necessary to generate chlorine dioxide on a continuous basis for use as a disinfectant. Although a few European potable water treatment plants use the acid-chlorite generation process (21), the most common synthesis route for disinfectant ClO_2 generation is the chlorine-chlorite process.

In the chlorine-chlorite process, sodium chlorite is supplied as either a solid powder or a concentrated solution. A solution of chlorine gas in water is produced by a chlorinator-ejector system, of design similar to that used in chlorination. The chlorine-water solution and a solution of sodium chlorite are simultaneously fed into a reactor vessel, typically PVC, 36-42 inches (91-107 cm) in height and 8 inches (20 cm) in diameter, packed with Raschig rings to promote mixing (21). From Equation 5-28, one mole of chlorine is required for two moles of sodium

Figure 5-16. **Nomograph for design of multiple perforated diffusers (Reproduced courtest of Wallace & Tiernan Division, Pennwalt Corp.)**



chlorite—or 0.78 part Cl_2/part $NaClO_2$ by weight. It has been found, however, that for this reaction to proceed to completion it is necessary to reduce the pH below that provided by the typical chlorine-water solution produced by an ejector. At 1:1 feed ratios by weight, only 60 percent of the chlorite typically reacts (21).

To provide greater yields, several options exist. First, it is possible to produce chlorine-water solutions in excess of 3,500 mg/l using pressurized injection of gas. In this case, however, there will be an excess of unreacted chlorine in the product solution, and the resultant disinfectant will consist of a mixture of chlorine and chlorine dioxide. The second option

(160), consists of acid addition to the chlorine and chlorite solutions; a 0.1 M HCl/M chloride addition enabled the production of a disinfectant solution of 95 percent purity in terms of chlorine dioxide, and achieved a 90 percent conversion of chlorite to chlorine dioxide. A third process, developed by CIFEC (Paris, France), involves recirculation of the chlorinator ejector discharge water back to the ejector inlet to produce a strong chlorine solution (5-6 kg/m^3) (0.04-0.05 lb/gal), typically at pH below 3.0, and in this manner to increase the efficiency of chlorite conversion (21,24). It has been reported that this last option is capable of producing 95 to 99 percent pure solutions of chlorine dioxide (24).

In any of these generation processes, the additional equipment needed (over and above a chlorinator-ejector system) includes the reactor vessel, metering pumps for the chlorite (and, if used, acid) solutions, and ancillary piping.

In the acid-chlorite method of generation, sodium chlorite and hydrochloric acid are used as reagents. There appears to be substantially less operating experience with this process in water or wastewater treatment (21,24); however, it appears to be conceptually simpler than the chlorine-chlorite process. In one plant in Europe (21), a batch reactor is used to mix the reagents in amounts necessary to produce a 20 percent final chlorine dioxide concentration; 15 minute reaction time has been found to be sufficient. There are also reported to be continuous generation processes available for use with the acid-chlorite method (24).

Due to the instability of strong chlorine dioxide solutions, it is not practical to store disinfectant solution for any significant period of time. Hence, the rate at which the chlorine dioxide is generated must be capable of being coupled in some manner directly to process sensors (e. g., flow meters or residual analyzers). This is generally done by varying reagent flow rates.

In addition, with both of the generation processes, a chlorine dioxide solution is produced directly, rather than a chlorine dioxide gas stream. Since chlorine dioxide gas at high concentrations may produce an explosive mixture, its handling as a solution is preferable. This chlorine dioxide solution may be pumped directly to a diffuser or may be diluted in a solution ejector (in which the strong solution is sent to a vacuum ejector similar to that used for gaseous chlorine application for the purpose of dilution) and then piped to a diffuser.

5.7.2 Mixing Systems

The necessity for rapid mixing of the disinfectant solution exiting the diffuser ports with the bulk wastewater flow was graphically shown by Heukelekian and Day (126). In their studies, the vertical gradient in distribution of chlorine residual in a contactor downstream of a simple dropped hose diffuser was measured and was found to be substantial unless adequate turbulence was present at the point of application of the chlorine solution. Unless mixing is present, there will exist zones of high and low disinfectant concentration, and the surviving microorganisms present in the zones of low concentration will act to diminish the process efficiency.

Longley determined that the degree of mixing at the point of disinfectant application has a pronounced effect upon the initial rate of inactivation of a variety of microorganisms over the range of G (root mean square velocity gradient) values of $100\text{-}10^5\ s^{-1}$ (161). White recommends rapid mixing with a G value of $500\text{-}1{,}000\ s^{-1}$ and a residence time of 5-15 seconds (24).

5.7.2.1 In-Line Diffuser

Four major options for mixing exist. In the first case, the diffuser may be placed in the center of a pipe running full at a turbulent Reynolds number. For this system, the mixing intensity may be computed from hydraulic relationships as follows:

Computation of G value in pipes:

The metric form of Manning's equation may be written as:

$$S = (nv/R^{2/3})^2 \qquad (5\text{-}67)$$

where S is the hydraulic gradient (dimensionless), n is Manning's coefficient, v is the flow velocity in m/s, and R is the inside pipe diameter assuming a full-flowing pipe (or hydraulic radius) in m.

Camp and Stein's relationship for the G value is given as:

$$G = (P/uV)^{1/2} \qquad (5\text{-}68)$$

where P is the power dissipation (Watts), V is the volume over which such dissipation occurs and u is the viscosity.

For a closed conduit in which power dissipation occurs by internal fluid friction, the power dissipation is given in terms of the headloss (h_1) by:

$$P = S_w g h_1 Q \qquad (5\text{-}69)$$

where S_w is the density of water, g is the gravitational acceleration, and Q is the volumetric flow rate. By multiplying S obtained from Equation 5-67 by the length of pipe (L), the headloss (h_e) may be obtained. The power dissipation then obtained from Equation 5-69 is substituted into Equation 5-68, and the volume is obtained via the geometric relationship for a volume of a cylinder. The resulting relationship for the G value in a pipe flowing full is given by:

$$G = (S_w g S v/u)^{1/2} \qquad (5\text{-}70)$$

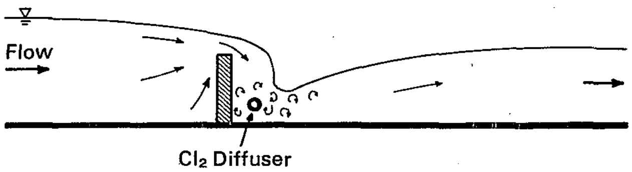
Finally, to ensure turbulent flow, it is necessary to check the Reynolds number from:

$$Re = D v S_w/u \qquad (5\text{-}71)$$

If Re is in excess of 2,000, turbulent flow may be assumed.

Using these relationships, given the design flow, the velocity may be computed for various pipe diameters. If Re (given by Equation 5-71) indicates turbulent flow, and if the G value, given by Equations 5-67 and 5-68 is in the acceptable range, then the design is satisfactory. The headloss may be computed directly from Equation 5-67.

Figure 5-17. **Details of a submerged weir mixing structure.**



5.7.2.2 Submerged Hydraulic Structures

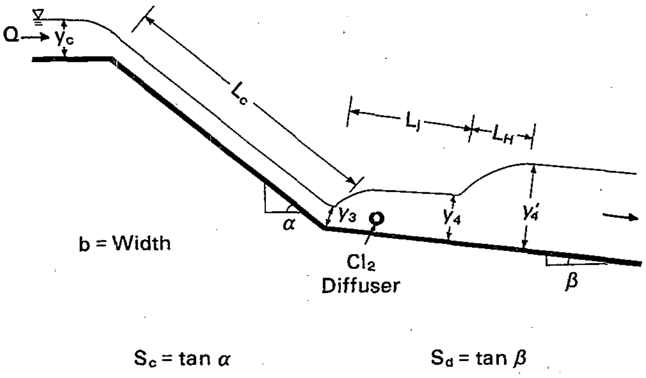
The second mixing option involves the use of a hydraulic structure at which turbulence is induced, as the point of application of chlorine solution. Two such possible structures are the submerged weir and the hydraulic jump (24).

A submerged weir consists of an open channel in which a rectangular projection arises from the channel base in such a manner that both the free liquid surfaces above and downstream from the weir are at a greater elevation than the weir itself. The diffuser inlet should be located at a point of maximum turbulence downstream from this structure. Figure 5-17 details the design of a submerged weir mixing structure.

The major variables necessary to design the submerged weir mixing structure are the distance between the weir and the point of maximum turbulence, at which the diffuser should be located, and the anticipated headloss across the weir.

The hydraulic jump following a chute and a sharp change in channel slope may also be used as a mixing device. Figure 5-18 details the general characteristics of a hydraulic jump used as a mixing structure. To design such a structure it is necessary to ensure that shooting flow (i.e. flow at a sub-critical depth) occurs at the bottom end of the chute, and that a hydraulic jump is located downstream of the diffuser but within the downstream channel.

Figure 5-18. **Details of a hydraulic jump mixing structure.**



The various computations for designing a hydraulic jump mixing device are outlined below. For more detailed discussions, standard hydraulic texts should be consulted (162).

Step 1—Decide upon the entering jump Froude number.

The Froude Number (F) is defined by Equation 5-72, where g is the gravitational acceleration (9.8 m/s^2), Q is the volumetric flow rate, y_4 is the liquid depth and b is the channel width.

$$F = (Q/by_4)/(gy_4)^{1/2} \qquad (5\text{-}72)$$

F must be greater than 1 for a jump to occur (supercritical flow), and Chow indicates best performance in terms of stability and insensitivity to downstream conditions for F between 4.5 and 9.0 (162).

To illustrate these calculations, a hydraulic jump will be designed to serve as a mixing device for a wastewater of average flow 26.5 MGD (1.16 m^3/s). For a Froude number of 6.0, and assuming a channel width (w) of 1 m, the value of y_4 from Equation 5-72 is 0.156 m. From the depth of the water immediately upstream of the jump, the final depth immediately downstream (y_4') may be determined by:

$$(y_4'/y_4) = 0.5\,(-1 + (1 + 8F^2)^{1/2}) \qquad (5\text{-}73)$$

Step 2—Determine chute slope (S_c) and y_3.

For the hydraulic jump to form downstream of the termination of the chute, it is necessary that the depth at the chute outlet, y_3, be some value less than the depth immediately prior to the hydraulic jump (y_4).

One can assume a chute slope and calculate y from Manning's equation (Equation 5-67) or vice-versa. Both the hydraulic radius and the flow velocity are functions of y_3.

Assuming a value of y_3 of 0.12 m (0.4 ft), the hydraulic radius equals 0.0968 m ($wy_3/(w+2y_3)$), and the flow velocity equals 9.67 m/s (Q/wy_3). From Equation 5-67, assuming a Manning's coefficient of 0.015 (typical of finished concrete), the frictional headloss (or required channel slope for uniform flow at the indicated depth) is 0.473 (i.e., the channel must have a rise of 0.473 m/m horizontal). Hence, S_c must be 0.473, and thus the angle of inclination is 28°.

Step 3—Determine length until jump (L_j).

The water depth increases from the base of the chute (y_3) until the onset of the jump (y_4) due to frictional losses that reduce the energy (and since the flow at these points is subcritical, increase the depth). To determine the length of travel between the chute exit and the jump onset, numerical integration of the energy conservation equation in conjunction with Manning's equation must be performed.

This process will be illustrated in tabular form. The basic relationships are Manning's equation (Equation 5-67) and the following rearrangement of Bernoulli's equation accounting for changes in channel elevation and frictional losses:

$$L_{12} = \frac{(d_2 - d_1) + (v_2^2 - V_1^2)}{2g(h_{e12} - S_{12})} \qquad (5\text{-}74)$$

In Equation 5-74, L_{12} is the channel length between two points (1 and 2) having indicated water depths (d) and velocities (v), and between which there is a channel floor slope of h_{e12} (dimensionless rise/run) and a frictional headloss of S_{12} (dimensionless).

The computation of L_j is executed in Table 5-13. First, the range of depths between y_3 and y_4 is subdivided into a number of intervals arbitrarily (however, the number of intervals will influence the precision of the calculation) as indicated in column (1). In columns (2) and (3), the hydraulic radius and the velocity, respectively, associated with each depth (at the known flow of 1.16 m^3/s and width of 1.0 m) are computed. Column (4) represents the frictional slope (S) associated with each depth, and is obtained by the use of Equation 5-67 at each point. Column (5) represents the arithmetic average of the frictional slope in the range defined by an upstream and a downstream depth. Finally, column (6) tabulates the calculated length increment from Equation 5-73 associated with the change in depth—a channel bottom slope of 0.01 has been assumed; however, this has only a very minor effect upon the result. By addition of each of the length increments, a total length from the chute bottom to the initiation of the jump (L_j) of 6.12 m is obtained.

Table 5-13. Computation of Length to Jump (L_j). (Assumed $H_E = 0.01$)

(1) d (m)	(2) R_H (m)	(3) v (m/s)	(4) S	(5) S_{avg}	(6) L_{12} (m)
0.12	0.0968	9.67	0.473	0.421	1.71
0.13	0.1032	8.92	0.369	0.333	1.68
0.14	0.1094	8.29	0.296	0.267	1.74
0.15	0.2154	7.73	0.239	0.226	0.99
			Total L_j		6.12

Therefore, in the example case, the headloss dissipated in the jump is 1.68 m. By application of Equation 5-69, the power dissipation equals 19.1 kW. It is not readily feasible to estimate the G value, since the volume of the liquid contained within the jump is not easy to obtain.

Step 4—Determine length of jump (L_H).

The ratio of jump length to height after the jump (L_H/y) has been observed to be a function of the Froude number alone. For Froude numbers between 4.5 and 9, this ratio is relatively constant at 6.15 (162). In the example problem, since y has been determined to be 1.25 m, the value for L_H is 7.69 m.

Step 5—Determine minimum chute length L_c.

A sufficient length of chute is needed so that the water velocity attained at the base of the chute is its normal value as predicted by Manning's equation. At the onset of the chute, the depth of flow may be assumed to be equal to the critical depth (y_c), predicted from:

$$y_c = (Q^2/gw^2)^{1/2} \qquad (5\text{-}75)$$

Thus, for the example under discussion, the initial depth at the top of the chute, or y_c, equals 0.52 m.

It is necessary that the length of this chute be sufficient so that the depth changes from 0.52 m to the design value of 0.12 m at the chute base. The slope of the chute bottom (h_{e12}) is 0.473. Equations 5-68 and 5-73 may be used in the manner in which they were applied in Table 5-13 to estimate the minimum length of this transition. By this procedure, it can be calculated that the length of the required chute is 6.99 m.

Step 6—Compute energy dissipation in jump.

The amount of energy transferred to the fluid in the hydraulic jump, and thus available for mixing, may be given by the difference in the energy of the fluid immediately prior to and following the jump. Chow gives the headloss in the jump as (162):

$$h_j = (y_4' - y_4)^3/4y_4'y_4 \qquad (5\text{-}76)$$

Step 7—Repeat computations at low and high flows.

Before the above design can be accepted, it is necessary to determine that an acceptable hydraulic jump exists in the given physical system under anticipated low and high flow conditions, and furthermore that the location of the diffuser is upstream of the jump in all likely flow situations. By using the design specifications (chute slope and length, channel width, downstream slope and length) with the expected low and high flows, the values for position of the jump under all circumstances can be calculated by application of the foregoing equations.

Once these computations are completed, the hydraulic profile of the design may be sketched, and the static headloss computed. A scale diagram of the jump illustrating diffuser location is given in Figure 5-19.

5.7.2.3 Mechanical Mixer

The third mixing option involves the use of a mechanical mixer, such as a propeller or turbine mixer, in conjunction with a small residence time mixing chamber containing the disinfectant diffuser. If a G value is specified (for example, 500-1,000/sec as recommended by White (24)), then Equation 5-68 can be used to determine the mixing power to be imparted to the fluid per unit volume.

For example, for the 26.5 mgd (1.16 m^3/s) wastewater flow described above, if a G value of 750 s^{-1} is specified, and the viscosity is assumed to be 0.001 kg/m-s, then from Equation 5-68, a power per unit volume of 866 W/m^3 is required. If the rapid mixing chamber has a residence time of 10 seconds, and hence a volume of 11.6 m, then the required mixing power to be imparted to the fluid equals 1,385 W. To determine the actual electric energy consumption for the drive motor for the necessary mixing device, it is necessary to consult manufacturer's specifications as to the efficiency of conversion of electrical to mechanical mixing energy. Similarly, differing geometries for mixer design exist, and the necessary motor speed will be a function of the specific propeller or turbine utilized.

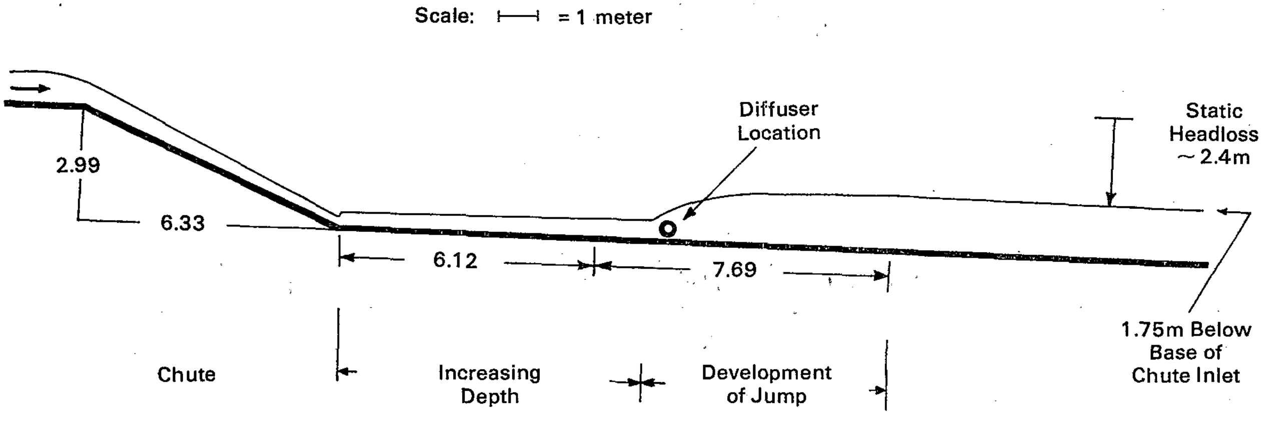
5.7.2.4 Jet Mixer

One additional mixing configuration exists in which the chlorine solution (or in some cases, the gaseous chlorine) coming from the ejector is introduced into a large jet, of similar overall design to a chlorine ejector. This jet mixer carries a substantial volumetric fraction of the influent wastewater through a high velocity nozzle and the disinfectant solution (or the disinfectant gas stream) is injected into the vacuum produced.

This jet mixing system is of proprietary design; however, it has been claimed that a G value in a jet tube of 10 s^{-1} or more, and a dimensionless tube Gt product of 1.5-15 produces satisfactory mixing (163). The jet mixer may be mounted directly in the entrance region of the contact chamber, and a booster pump used to provide the motive power. In one such field test, it was shown that the required dosage of chlorine was less with jet mixing than with poor mixing (164); however more recent tests suggest that such reductions in chlorine dose, and in any necessary dose of sulfur dioxide required for dechlorination may be insufficient to compensate for the additional energy costs of jet mixing (165).

Of the mixing systems, the most common is the use of an in-line diffuser, followed by a rapid mixer, then hydraulic jumps and finally jet mixers. In-line diffusers are particularly applicable to small plants. However at flows greater than 10 MGD it becomes difficult to design a fully flowing turbulent section of pipe. Both in-line diffusers and hydraulic jumps share the advantage of requiring no direct external power input. However, hydraulic jumps are inapplicable unless upstream equalization of flows is practiced, since the horizontal location of the jump is extremely sensitive to flow variations.

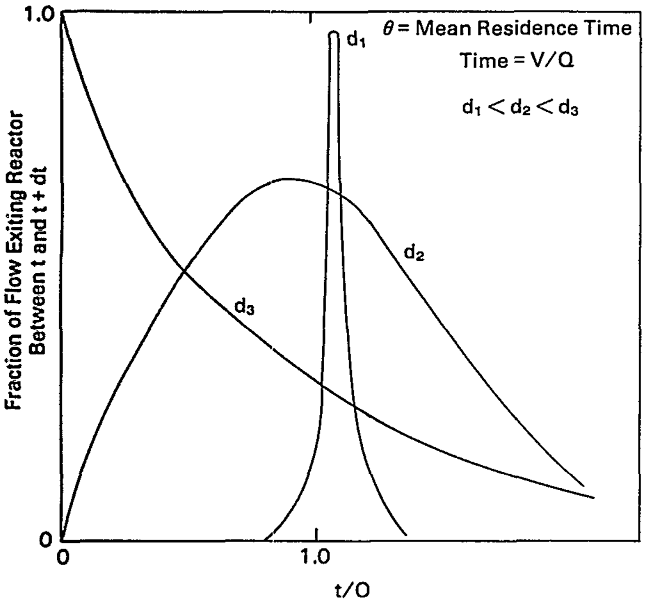
Figure 5-19. Scale diagram of jump as designed.



5.7.3 Contacting Systems

The role of the chlorine contact system is to permit sufficient time to elapse for adequate disinfection to occur. A characteristic measurement of time is the mean hydraulic residence time, θ. However, it is also necessary to ensure that the bulk of the wastewater has had the opportunity to remain in the contact basin prior to release. Figure 5-20 depicts the frequency distribution of fluid residence times in three basins of identical θ. The spread of residence times can be characterized by the dimensionless dispersion index, d, which decreases as the range of individual residence times decreases. See Section 4.2 for a theoretical development of residence time distribution functions.

Figure 5-20. **Residence time distribution functions for contact basins.**



It has been shown that the performance of a chlorine contact chamber declines substantially (i.e., the effluent microbial concentrations increase) when the dispersion in the contactor increases (166), and this finding has been verified experimentally (167-169). There appears to be a point of diminishing return, and Trussell and Chao (166) propose that d = 0.01 is a practical lower limit to be attained to optimize the hydraulic performance of chlorine contact chambers in wastewater disinfection.

Once the average residence time for the contactor has been determined, from the kinetics of microbial inactivation discussed above, several options exist for the configuration of the contactor. When disinfection is the final treatment process (in other words, when no dechlorination is to be practiced), and a sufficiently long distance is available, the outfall pipe itself may be used to provide contact. In other cases, it is necessary to design a separate tank for contacting. In either case, the volume to be contained in the contactor may be determined from the equation $V = Q\theta$, where Q is the design flow, θ is the design contact time, and V is the volume of liquid to be contained in the contactor.

For pipe contactors, diameter influences the headloss in the pipe, which may be determined via Manning's equation, the flow velocity in the pipe, and the dispersion. The velocity in the pipe must be kept above the scour velocity for particles to avoid sedimentation. Generally a 1 ft/s (0.3 m/s) flow velocity will suffice, or the designer may use the scour relationship of Camp to estimate the minimum velocity. The dispersion in a pipe may be estimated by the use of the following equation (166):

$$d = 89{,}500\, f^{3.6}\, (D/L)^{0.859} \qquad (5\text{-}77)$$

In the above equation, f equals the Darcy Weisbach friction factor for pipe flow, which is obtained as a function of the pipe Reynolds number and the relative roughness from a Moody diagram (170), D is the pipe diameter, and L is the length of the pipe. This equation is valid for the case of a straight pipe alone, without bends or other flow disturbances.

If it is necessary to build a separate contact basin, it is beneficial to construct a basin, either rectangular or annular, in such a manner that baffling is present to provide the longest possible pathway for flow in order to minimize the dispersion. Figure 5-21 illustrates various possible configurations for baffling in contact chambers.

In a baffled contact chamber, the flow velocity (defined by the length of the flow path divided by the mean residence time) should be greater than the scour velocity to minimize particle deposition. The dispersion in rectangular baffled contact chambers (either end-around or over and under baffles) may be given by either Equation 5-78 (166) or Equation 5-79 (157):

$$d = 0.14\, K/(L/W) \qquad (5\text{-}78)$$

$$d = 1.15\, (L/W)^{-1.13} \qquad (5\text{-}79)$$

In Equation 5-78, K is a coefficient of nonideality, found to vary between 2.3 and 15.8 for real contactors. In both cases, L is the length of the flow path and W is the width between parallel baffles.

The headloss in baffled contactors is due to frictional losses with the bordering surfaces and dissipative losses associated with changes in flow direction. Frictional losses may be evaluated as in the case of

Figure 5-21. **Types of baffled contact chambers (174).**

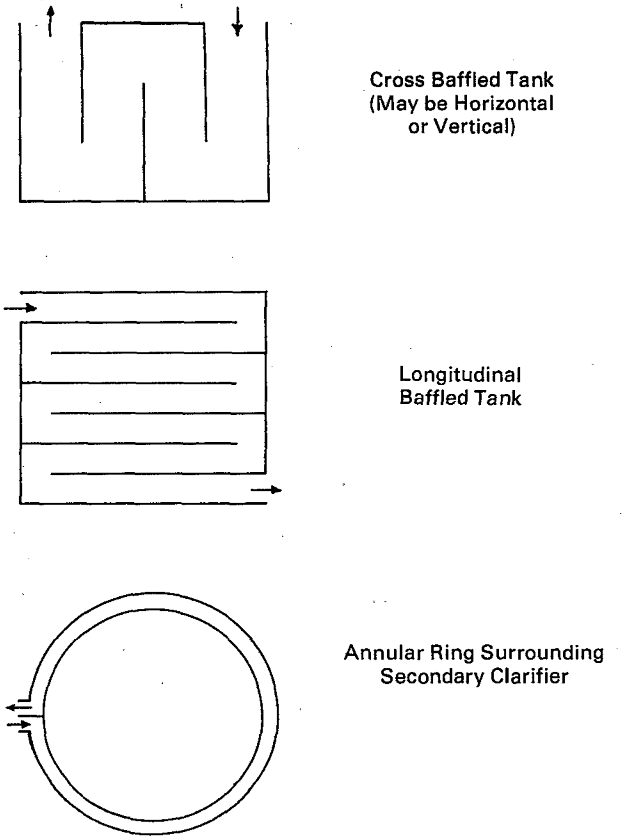
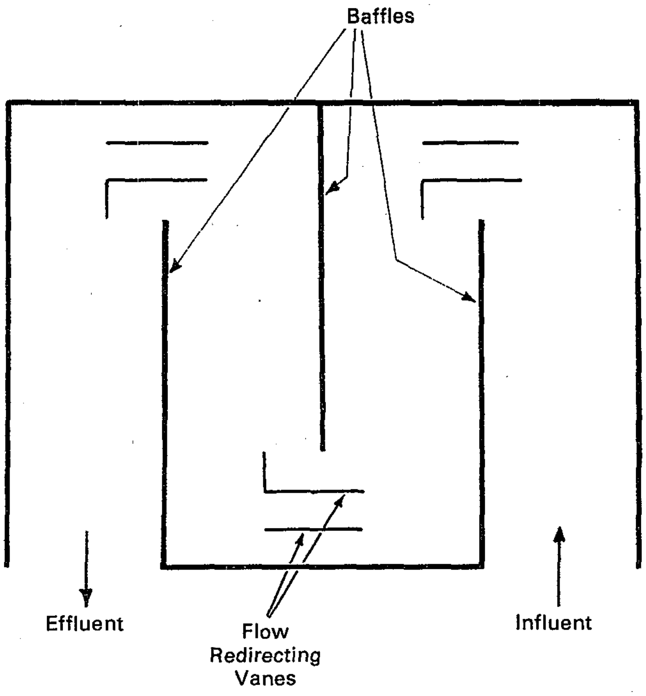


Figure 5-22. **Vaned serpentine contactor design of Louie and Fohrman (172).**



any open channel flow system. Losses due to velocity changes may be estimated using the following (171):

$$h_1 = ((n + 1) v_1^2 + n v_2^2)/2g \qquad (5\text{-}80)$$

In this equation, v_1 and v_2 are, respectively, the flow velocities between parallel baffles and through the baffle slots, and n is the number of baffles. It is possible to reduce the headloss associated with these velocity changes by adding flow redirectors. Louie and Fohrman evaluated a number of such configurations, and concluded that the vaned serpentine design in Figure 5-22 provided minimum headloss and satisfactory reduction in dispersion (172).

It has also been suggested that the performance of baffled rectangular contact chambers may be improved by the use of air scour. This has been demonstrated in scale models (173); however, little full-scale quantitative data exist to evaluate this effect. In addition, such scouring would increase dispersion, possibly to a deleterious degree.

The effluent weir from a contact chamber should extend along the full length of the exit channel from the flow channel. Marske and Boyle (174) found that basins with full sharpcrested weirs performed better than partial length Cipolleti (i.e. trapezoidal) weirs.

Although there has been little direct work in this area, to minimize the potential for short-circuiting associated with entrance effects, the inlet to contact chambers should be designed to introduce flow over as large a fraction of the width as possible, and with minimum velocity. Alternatively, a single pipe inlet impinging upon a momentum-absorbing baffle may be used.

Contact chambers, like many wastewater treatment unit processes, will occasionally foul with a microbial slime. Therefore, at least two parallel contact chambers should be used so that one can be drained and cleaned. The contact chamber should have provision for draining, preferably by gravity, and be in proximity to a high pressure water hose for cleaning.

The required size of a contact chamber is a function of the chlorine dose to be used, the nature of the wastewater to be disinfected, and the required amount of disinfection to be achieved. For a first approximation, manufacturer's data may be consulted for the doses of chlorine to be used in typical situations. As a more exact estimate of the effect of contact time on required chlorine dose, information presented earlier on the kinetics of microbial inactivation and the decay of chlorine residual may be used. This will be illustrated in the design example to be presented below.

5.7.4 Process Control

The function of the process control element is to ensure that there is consistent disinfection performance despite fluctuations in the quality and quantity of wastewater influent to the disinfection process. Due to the scarcity of experience with chlorine dioxide and bromine chloride as wastewater disinfectants, only process control of chlorination will be discussed.

The simplest control strategy involves either the use of flow equalization basins prior to chlorination or extremely conservative design factors. If a sufficiently large equalization basin exists, e.g., a lagoon with multiple day residence time, upstream of disinfection, the wastewater quality and quantity entering the chlorination process will be relatively uniform. Thus, the performance of the chlorination process will be reasonably constant. However, very few treatment plants are able to provide this degree of equalization. Alternatively, the dose of chlorine or hypochlorite applied to the wastewater may be made sufficiently high to ensure that the high flows with the greatest chlorine demands are adequately disinfected. This will ensure that low flows, or flows of lesser chlorine demand, will be at least adequately treated.

For high wastewater flows, it will be impractical to provide equalization or uneconomical to overdose with chlorine. In this case, the amount of chlorine must be varied in some manner so that the dose applied is just sufficient to provide the required degree of disinfection. This may be done manually, by the operator, one or more times per day. The feed rate (kg chlorine per hr) from the gas chlorinator (or the metering pump rate from the hypochlorite storage tank) is adjusted in accordance with the flow and the chlorine demand, usually to attain a target chlorine residual. Typically, this adjustment may be made once per shift.

For larger treatment plants, the savings on chlorine utilization will justify some degree of automatic control. In addition, for treatment plants using dechlorination, automatic control may further be justified by savings on dechlorination chemicals.

The simplest form of automatic control that can be used is flow proportional control. In this case, the instantaneous chlorine application rate is maintained directly proportional to the instantaneous flow rate coming into the disinfection process, resulting in a constant chlorine dosage. If the disinfection influent flow is controlled by a pumping station upstream, the chlorine application rate may be set by a signal from the wastewater pumps to the chlorinator. This modification is referred to as additive rate control. Generally the signal from the pumps is a standard 4-20 ma DC signal (the typical electrical control signal) and may be used to vary the orifice opening on a gas chlorinator vacuum line (with the flow being linear in orifice area) or the metering pump speed on a hypochlorinator to provide a relatively constant chlorine dose. One drawback of this strategy is the slight nonlinear behavior of flow rate versus pump horsepower due to intervening friction losses in the line between the pumps and the chlorine contact chamber. In addition, there will be a lag time in process response due to the hydraulic residence time in the system between the pump station and the point of chlorine application. During this lag, there will be under or over application of chlorine as flow decreases or increases, respectively.

A second form of flow proportional control uses a flow meter (e.g., magnetic in-line, or venturi) to sense the flow entering the chlorine contact chamber. Due to the problem of lags, the plant raw wastewater flowmeter should not be used to supply this control signal. The output of the flowmeter may be a 4-20 ma DC signal, or a vacuum or pressure pneumatic signal. This signal is then applied to either the orifice on the gas chlorinator or the metering pump on a hypochlorinator to provide a constant dose.

While flow proportional control of either type accounts for a major source of variation in chlorine dose requirement, it fails to provide adjustment in chlorine dose with variation in wastewater chlorine demand. For example, if the chlorine demand varies by a factor of 2 over the course of a day it will be necessary to overdose during times of low chlorine demand, even if flow proportional control is used.

To circumvent this difficulty, it is, in principle, possible to measure the chlorine residual after the initial high chlorine demand has been exerted (2 to 5 minutes following rapid mix) and use an automatic measurement of this residual to control the chlorine application rate. This strategy, feedback residual control, is generally not practiced alone, except when flow variations are negligible, due to problems of hydraulic lags in the system.

Instead the flow is most commonly measured using an automatic flow transducer, and the residual after 2 to 5 minutes is measured using an automatic chlorine residual analyzer. These signals are then combined to provide a constant residual. In this manner, the variation in flow is used as the major signal for dose adjustment, and the variation in residual is used to fine tune the dose. This option is designated as compound loop control.

Two major types of compound loop control exist. One type, typified by many gas chlorinators such as those manufactured by Wallace and Tiernan Division of PennWalt Corporation, uses the flow signal to vary the size of the gas orifice, and the residual signal (modified by a square root module) to vary the pressure differential across the orifice. The square root module takes the incoming signal and converts it to its square root—this is done because the flow rate is linear in the square root of pressure drop, and not in pressure drop itself. The second type of compound loop control, implemented in many chlorinators manufactured by Fischer and Porter and Capital Controls, electronically combines the two signals from the flow meter and the analyzer into a single signal to a chlorine valve, which may regulate either the orifice area or the pressure differential across a fixed orifice.

In either case of compound loop control, either or both signals from the flow and residual sensors may be electric (4-20 ma DC), pressure pneumatic (3-15 psig is a common industry standard) or vacuum (5-55 inches of water is common). If pneumatic or vacuum control signals are used, a separate system is needed to produce the vacuum or pressure that drives the control signal lines. In addition, direct digital computer control, using either dedicated distributed or centralized computers, is possible.

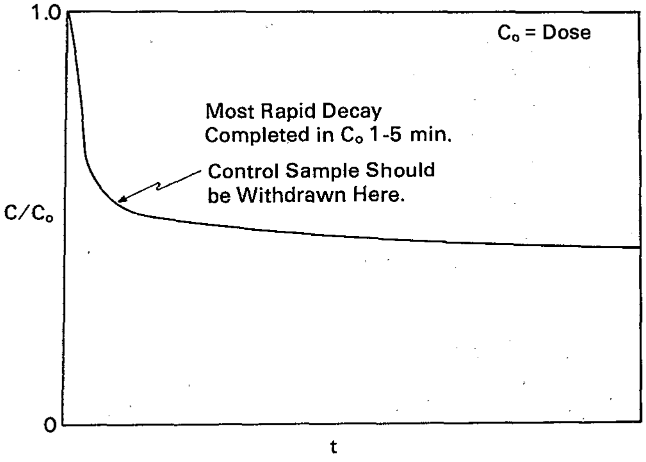
Critical elements in the design of compound loop control are the time lags inherent in the residual measurement system, and delay times between application of the chlorine residual signal to the control device and the attainment of the control action. These time lags include those due to the sample transit time in the sampling lines, the lag time between the ejector and the point of application, and the residence time in the contactor prior to sampling.

The calculation of the lag time in the line between the ejector and the point of application may be calculated as described above once the ejector is sized. The lag time in analysis is of the order of 1-2 minutes. The lag time in the sampling lines is set by the velocity and the length of the line. This velocity should be maintained above 3 m/s (10 ft/sec) to minimize fouling (157). By maintaining the analyzer as close as possible to the chlorine contact basin, this lag may be minimized. The lag time between the point of application and the sampling line should be set so that the rapid initial decay has been completed. This usually occurs 3 to 5 minutes following chlorine application (Figure 5-23).

Due to the existence of lags in the chlorine residual sampling lines, it may be necessary to incorporate a delay timer on the residual control loop. This insures that no additional adjustment in chlorine dose occurs following a preceding adjustment until sufficient time has elapsed for the wastewater receiving the new chlorine dose to be analyzed for its response. This timer should be adjustable and is ordinarily "tuned" during start up of a treatment plant to minimize control instability.

Roop presented a comprehensive review of wastewater chlorine control using the compound loop system (175). For situations in which the rapid phase of chlorine demand is prolonged (5 to 15 minutes), a second chlorine analyzer is recommended at the far point to trim the signal from the first sampling location.

Figure 5-23. **Dissipation of chlorine residual and point of sampling for control.**



The analyzers used for residual may use either the amperometric, or the colorimetric or titrimetric DPD methods, as described above. In either case, manufacturers' designs of continuous analyzers incorporate in-line filters to minimize solids deposition. For continuous analyzers, special reagent solutions are used that differ somewhat from manual analytic techniques in that proprietary additions are incorporated that are said to reduce fouling inherent in continuous systems. The maintenance requirements for continuous analyzers are higher than manual analyzers, and at least daily standardization is required to prevent drifting in the automated analyzer.

5.7.5 Dechlorination

From Equations 5-48 and 5-49, one mole of sulfur dioxide can dechlorinate one mole of either free chlorine or combined chlorine. On a mass basis, therefore, 64/71 grams of SO_2 are required to dechlorinate one gram of chlorine, either in the free or the combined form. On a practical basis, about 1 gram of SO_2 is required per gram of chlorine.

Sulfur dioxide used for dechlorination may be supplied in either 68 kg (150 lb) cylinders or 1,365 kg (3,000 lb) containers. Railroad tank cars for bulk sulfur dioxide shipment are also available. The discharge from sulfur dioxide containers may be taken from the liquid or gaseous phases (106).

For gaseous withdrawal, the maximum safe rate of withdrawal on a continuous basis at 21°C (70°F) is 0.9 kg (2 lb)/hr for a 68-kg (150-lb) container, or 11.4

kg (25 lb)/hr for a "ton" container (106). To attain higher rates of gaseous withdrawal, it is acceptable to immerse the containers or cylinders in a liquid bath, or to surround the containers with strip heaters; in both cases, only the lower 50% of the container should be subject to heating, and heating should be limited to 325°K (106). As in the case of chlorine, the pressure at which sulfur dioxide is withdrawn should be sufficiently below the saturation vapor pressure to provide superheating and prevent reliquifaction (157).

For liquid withdrawal, the maximum rate of removal is limited by hydraulic considerations to 135 kg (300 lb)/hr) from either type of container (106), although it is possible to exceed this by pressurization, or "padding" with dry air or nitrogen up to 515 kPa (60 psig) (157). With liquid withdrawal systems, external evaporators, of similar overall design to those used for chlorine service, are required.

It may be necessary to manifold multiple cylinders of sulfur dioxide for supply to larger dechlorination systems. Considerations for manifold design are similar to those used for gaseous or liquid chlorine systems.

Piping and materials used for chlorine service are generally satisfactory for sulfur dioxide service (157), although the physical systems used for chlorine should not be used for sulfur dioxide, or vice versa, prior to thorough cleaning, to prevent potentially explosive reactions from occurring. In liquid piping systems, expansion chambers are required, as for liquid chlorine service, to prevent hydrostatic expansion rupture from occurring (106).

The gaseous sulfur dioxide exiting from the cylinder or the evaporator may be dispersed into water using vacuum ejector systems of similar design to chlorine ejectors. Design curves for chlorine ejectors must be corrected for different flow rates (on a mass basis) for sulfur dioxide versus chlorine. This is done by multiplying the specified flow rate for a chlorinator by 0.95 (which represents the ratio of the density of sulfur dioxide to that of chlorine). For example, if a gas chlorinator is designed to deliver 45 kg (1,000 lb)/d chlorine at maximum span, it can deliver 43 kg (950 lb)/d of sulfur dioxide.

The mixing at the point of application of sulfur dioxide to the full wastewater flow should be conducted in an intensive manner to allow complete dechlorination of all fluid elements. Injection into the center of a turbulent fully-flowing pipe has proven satisfactory, with complete mixing being attained 10 pipe diameters downstream of the injection point (176).

The kinetics of the reaction between sulfur dioxide and chlorine are sufficiently rapid to preclude the necessity for a separate contact chamber prior to discharge of the effluents (109). If a contact chamber is used, the absence of a chlorine residual may allow for the growth of a microbial slime layer containing coliforms, and possibly other organisms. While prevention of this slime growth does not appear to be feasible, some regulatory authorities (e.g., the State of California) permit the coliform standard to be attained at the chlorine contactor effluent, despite an increase in bacterial numbers during dechlorination (112).

Control of sulfur dioxide dechlorination is generally more difficult than chlorination. According to Chen and Gan: "simple feedforward sulfur dioxide feed control system(s) [i.e., systems which rely solely on a measurement of influent volumetric flow rate] (are) inadequate for most dechlorination installations" (112). This is due to the variation in the chlorine residual at the outlet from the chlorine contact chamber.

Most commonly, control systems for dechlorination consist of either straight flow proportional or feedforward control based on the product of influent flow and chlorine residual. In dechlorination installations, it is necessary to maintain two automatic chlorine residual analyzers—one 3 to 5 minutes from the point of application of chlorine used for chlorine dose control, and the second located at the contact chamber outlet, used for control of the sulfonator (112). Feedback control using measured chlorine residual following sulfur dioxide addition is not currently feasible due to the instability of currently available analyzers measuring low (below 0.1 mg/l) or zero chlorine residuals.

Two types of feedback control systems may be used to circumvent the problems associated with continuous measurement of low chlorine residuals. In the first method, which may be described as split dechlorination, a flow proportional feedback chlorine residual system (i.e. compound loop control) is used to produce an effluent that is dechlorinated to a low residual, but well within the range of commercially available chlorine analyzers—this is usually 90 percent dechlorination of the incoming residual. A subsequent dechlorination step, using simple feedforward control, may then be used to produce a chlorine-free effluent.

The second process uses a chlorine analyzer that is biased with the addition of a constant flow of a side stream containing a constant chlorine concentration (produced by a chlorinator or hypochlorite feed system separate from that used for chlorination) to the sample being analyzed. The setpoint on this analyzer is controlled to maintain the chlorine residual subsequent to dechlorination at the concentration imposed by the biasing stream. In this case, classical compound loop systems may be used (112).

One modification of the biased feedback control strategy has recently been devised (177). In this approach, the chlorine residual immediately prior to dechlorination is measured (C_{pre}). Equal flows of the influent to and effluent from the dechlorination process are blended and the residual measured using a second analyzer (C_{mixed}). The chlorine residual leaving the dechlorination basin may then be calculated as $2C_{mixed}-C_{pre}$, and this value used as a control signal.

This system has been in use in Metro Seattle and is reportedly successful at maintaining a residual below 0.067 mg/l. The principal disadvantage of this approach is the relatively higher complexity of the control system, and the potential for biological slime accumulation in the dechlorinated effluent sample lines; however, this system eliminates the need for a separate chlorinator used to bias the sulfonator analyzer.

In the absence of such control systems, it will be necessary to overdose the wastewater with sulfur dioxide, which may result in deoxygenation of the effluent. In this case, reaeration of the dechlorinated wastewater may be necessary to produce compliance with an effluent DO requirement. In the presence of one of the suggested dechlorination control systems, such deoxygenation and reaeration requirements have been found not to occur (112).

5.7.6 Design Example

The principles of design of chlorination and dechlorination systems will be illustrated by reference to the common design example. A chlorination system is to be designed to treat a non-nitrified activated sludge effluent. The design flows are 7.5 mgd (mean) and 15 mgd (peak); initial flows are 3.5 mgd (mean) and 7.5 mgd (peak). The influent to the disinfection system is expected to contain a geometric mean of 500,000 fecal coliforms/100 ml, with a peak of 2,000,000/100 ml. The effluent coliform standards are a 30-day geometric mean of 200 fecal coliforms/100 ml, and a 7-day geometric mean of 400/100 ml. Other effluent parameters are a 30-day mean for both BOD_5 and suspended solids of 15 mg/l and a maximum daily value of 30 mg/l for each of these parameters. It is also desired to dechlorinate such that the final effluent chlorine residual is less than 0.05 mg/l.

Step 1—Determine the average desired fractional survival.

From the conditions of this problem, the desired fractional survival under average conditions is 200/500,000, or 4×10^{-4}. This will be used as an initial basis for design.

Step 2—Characterize the decay of chlorine residual. This must be done experimentally, or by resort to information from the literature.

For the conversion of free chlorine to combined chlorine in wastewaters where the Cl:N ratio is far below breakpoint, a first order decay relationship may be assumed:

$$C_f = C_o \exp(-k_1 t) \qquad (5\text{-}81)$$

In Equation 5-81, C_f is the free chlorine concentration, C_o is the chlorine dose, and is the apparent first order decay constant, which may be given by (81):

$$k_1 (s^{-1}) = \frac{9.7 \times 10^{10}\, C_N\, K_{DN}\, [H^+]\, \exp(-1510/T)}{([H^+] + K_{KA})\, ([H^+] + K_{DN})} \qquad (5\text{-}82)$$

In Equation 5-82, C_N is the concentration (in M/l) of ammonia nitrogen, K_A is the dissociation constant of HOCl (given by Equation 5-6), and K_{DN} is the dissociation constant of the ammonium cation. T is the Kelvin temperature. With an influent ammonia concentration of 10 mg/l as N (=0.7 mM), a mean temperature of 15°C (288°K), and an influent pH of 7.5, k_1 equals 193 s^{-1}.

The resultant combined chlorine, once formed, decays. It has been found that the kinetics of this decay may be described by two parallel first order reactions (76):

$$C_c = C_o [x \exp(-k_2 t) + (1 - x) \exp(-k_2' t)] \qquad (5\text{-}83)$$

In Equation 5-83, x, k_2, and k_2' are constants characteristic of the wastewater. For a variety of wastewaters, it has been found that values of 0.3, $1.67 \times 10^{-2} s^{-1}$ and $5 \times 10^{-5} s^{-1}$, respectively are typical (76). Since the rates of these two decay reactions are disparate, Equation 5-83 may be simplified to:

$$C_c = C_o (1 - x) + C_o\, x \exp(-k_2 t) \text{ for } t < 1/k_2 \qquad (5\text{-}84)$$

$$C_c = C_o (1 - x) \exp(-k_2' t) \text{ for } t > 1/k_2 \qquad (5\text{-}85)$$

With the particular constants assumed, Equations 5-81, 5-84, and 5-85 may be written as the following, with t in minutes:

$$C_f = C_o \exp(-11{,}580t) \qquad (5\text{-}86)$$

$$C_c = .7C_o + .3C_o \exp(-t) \text{ for } t < 1 \text{ min} \qquad (5\text{-}87)$$

$$C_c = .7C_o \exp(-.003\, t) \text{ for } t > 1 \text{ min} \qquad (5\text{-}88)$$

Step 3—The chlorine decay relationships may now be substituted into the kinetic relationships describing microbial inactivation.

Method 1: Using Equation 5-65, with parameters from Aieta and Roberts (178), the following may be written:

$$N/N_o = (0.25Ct)^{-2.82} \qquad (5\text{-}89)$$

Using either Equation 5-87 (for times below 1 minute) or Equation 5-88 (for times greater than 1 minute) to substitute C as a function of time into Equation 5-89, an equation relating contact time, chlorine dose, and inactivation may be obtained. In other words, it may be shown that:

$$N/N_o = (.175\ C_o t + .075\ C_o t \exp(-t))^{-2.82} \quad t < 1 \text{ min} \tag{5-90}$$

$$N/N_o = (.175\ C_o t \exp(-.003))^{-2.82} \quad t > 1 \text{ min} \tag{5-91}$$

Therefore, since N/N_o has been specified, if a value of t is assumed, C_o may be calculated. Hence, for times of 10, 15, 30 and 60 minutes, C_o (the chlorine dose) equals 9.45, 6.39, 3.34 and 1.83 mg/l, respectively; the effluent residual chlorine concentrations may be determined from Equation 5-88 to be 6.42, 4.28, 2.14 and 1.07 mg/l, respectively.

Method 2: From the Chick-Watson equation, assuming inactivation by free and combined chlorine is additive, the following may be written:

$$\frac{dN}{dt} = -N(k_f c_f n_f + k_c C_c n_c) \tag{5-92}$$

The chlorine demand relationships may be substituted into this equation, and it can be integrated numerically.

For $t > k_2$, the result may be obtained as an analytical integral in the following form:

$$\begin{aligned} -\ln(N/N_o) = {} & (k_f C_o n_f / k_1 n_f)(1 - e^{-k_1 n_f t}) \\ & + (k_c C_o n_c (1 - x)\, n_c / k_2' n_c)^{*} \\ & (e^{-k_2' n_c / k_1} - e^{-k_2' n_c t}) \end{aligned} \tag{5-93}$$

For the inactivation data for *E. coli* at room temperature and relatively neutral pH, values for k_f, k_c, n_f and n_c of 30, .085, 1.46, and 1.25, respectively, may be assumed (Table 5-11). Using these, and the prior assumed constants, the following equation is obtained:

$$\begin{aligned} -\ln(N/N_o) = {} & 2.86 \times 10^{-3} C_o^{1.46} \\ & + 14.4 C_o^{1.25} - 14.5 C_o^{1.25} e^{-0.00375t} \end{aligned} \tag{5-94}$$

Equation 5-94 relates N/N_o, C_o, and t, and since the first parameter has been specified, given a value for t, the requisite value of C_o may be calculated. For t equals 10, 15, 30 and 60 minutes, C_o equals 10.03, 6.92, 3.85, and 2.26 mg/l, respectively. Using these doses and contact times, from Equation 5-88 the effluent chlorine residuals may be calculated to be 6.81, 4.63, 2.46, and 1.32 mg/l, respectively.

Step 4—By two methods, suitable combinations of residence time and chlorine dose have been determined which will satisfy the 30 day coliform standard under average conditions. Based on this information, a contact chamber will be sized. It is now necessary to determine, for a specific volume contact chamber, the performance under various extreme conditions to ascertain the maximum chlorine dose likely to be necessary. While this sensitivity analysis can be performed using either Method 1 or Method 2 (i.e., Equation 5-91 or 5-94), the approach will be illustrated using Method 2 (Equation 5-94), which, based on the computations in Step 3, for this situation, appears to be more conservative.

Based on step 3, it is anticipated that a 60-minute contact tank will be most suitable (obviously, the following can be repeated with different initial assumed tank sizes to generate economically optimized designs). For the design mean flow, this would require a volume of 1,200 m^3 (312,500 gal). Rounding up, the design volume of 320,000 gallons is selected. At the design mean flow and influent coliform concentration, this would require a chlorine dose of 2.26 mg/l. However, at higher flow rates and coliforms, the chlorine dose must be increased if the effluent is to be in compliance with the coliform standard. Based on the assumed tank volume, at each combination of chlorine dose, flow rate, and influent coliform concentration, the anticipated effluent coliform and total residual chlorine concentration may be computed. For example, at a dose of 3.7 mg/l, for the given tank, the following are to be expected:

Assumed Flow	Influent Coliforms	Effluent Coliforms (#/100 ml)	Effluent Total Chlorine Residual (mg/l)
avg design	avg	0.1	2.15
max design	avg	197.4	2.36
avg design	max	0.6	2.15
max design	max	789.7	2.36

This is in compliance with the standards under all conditions except for the combination of maximum design flow and maximum influent coliform concentrations. For design purposes, it may either be assumed that these two factors are unlikely to occur coincidentally more than once in 7 days (and hence the 7-day standard will be complied with), or the dose necessary to produce compliance at this extreme may be computed.

At this point, the chlorinator(s) that is(are) to be specified must be capable of a maximum output of 3.7 mg/l at the maximum design flow of 15 mgd, or 462 lb/d. This is close to 500 lb/d, which is a commonly available chlorinator size; to provide system redundancy, two such units would be specified. Alternatively, the overall plant design might use two parallel chlorination systems, each sized to half the design capacity; in this situation, two 250 lb/d units, plus one spare 250 lb/d unit would be specified.

Step 5—Based upon a 320,000 gal volume, the specific contact basin may now be designed. The two particular options considered will be the use of a single baffled tank, and the use of a long pipe.

Option 1—Long pipe contact:

If a 2-m (78-in) diameter pipe running full is used, the required pipe length may be given by application of the formula of a volume of a cylinder. Thus, a run of 389 meters is necessary to provide the required volume. The velocity in the pipe at mean flow is 389 m/3600 s = 0.11 m/s (21 ft/min). This should be checked against the specific scour velocity of the particles expected to be present to ensure minimum sludge deposition.

The dispersion in this pipe may be estimated using Equation 5-77. The pipe Reynolds number is given by Re = VDR/u (where R is the density of water and u is the viscosity). For turbulent flows, the friction factor 'f' is given by the following:

$$1/f^{0.5} = 2 \log_{10}(D/e) + 1.14 \qquad (5\text{-}95)$$

In Equation 5-95, e is the absolute pipe roughness (0.002 inches is typical for concrete). Thus f = 0.0094. Substituting in Equation 5-87 along with the pipe diameter and the length, the dispersion number is determined as 4.9×10^{-5}, sufficiently low that the plug flow assumption is justifiable.

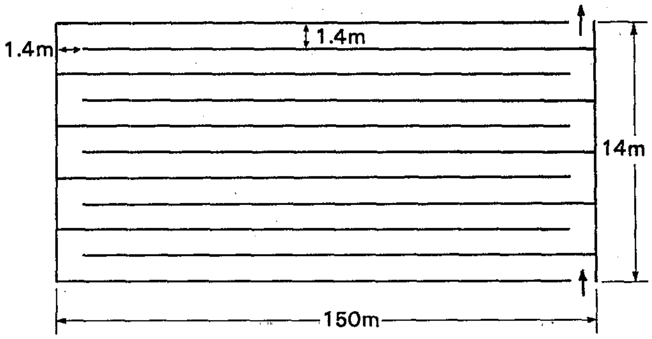
The headloss may be estimated using Manning's equation. With an 'n' value of 0.015, the hydraulic grade is calculated as 9.767×10^{-7} at mean design flow, and 2.127×10^{-7} at peak design flow. Therefore, the estimated headloss is 0.038 cm and 0.151 cm, respectively, under these two conditions.

Option 2—Serpentine contact:

A rectangular baffled contactor with a depth of 2 m is used. The sketch in Figure 5-24 illustrates one configuration providing the required residence time.

The flow path of the system is 40 x 7 = 280 m. The L/W ratio is 280/2.16 = 129. The velocity (at average design flow) between parallel baffles is 0.076 m/s, and through the baffle slots is 0.082 m/s. This should be checked against the specific scour velocity as in the previous option. From Equation 5-78, the estimated 'd' value is 0.0047 (for K = 1), again indicating reasonable approach to plug flow. Finally, the headloss is computed from Equation 5-80 as 0.41 cm at average design flow and 1.65 cm at peak design flow.

Figure 5-24. Definition sketch for rectangular contactor.



Step 6—Based on conditions at maximum design flow, assuming peak chlorine dose of 3.7 mg/l, the chlorine residual leaving the contact basin is estimated as 2.36 mg/l (*see table under step 4*). Thus, at peak design flow, sulfonation capacity (assuming 1 mg SO_2 per mg chlorine residual) of 15 mgd x 2.36 mg/l x 8.34 = 295 lb/d is needed. From this information, the sulfonator(s) can be specified, keeping in mind the desired redundancy.

Step 7—Annual chemical requirements can be estimated using average design conditions. As noted, at mean design flow and coliform concentrations, a chlorine dose of 2.18 mg/l would be satisfactory, and would impart a residual exiting the contact chamber of 1.27 mg/l (using Equation 5-88). Thus, the average annual chlorine requirement under design conditions would be 365 x 7 mgd x 2.18 mg/l x 8.34 = 46,300 lb/yr (say 24 tons/year). The average annual requirement for sulfur dioxide under design conditions (assuming 1 mg SO_2/mg chlorine residual) would be 365 x 7 mgd x 1.27 mg/l x 8.34 = 27,000 lb/yr (say 14 tons/yr). This information could then be used to design the chemical handling facilities.

If chlorine and sulfur dioxide are obtained in ton containers, and if gaseous withdrawal is utilized, the average chemical consumption is 2.4 kg (5.3 lb)/hr and 1.4 kg (3.1 lb)/hr, respectively, for chlorine and sulfur dioxide. Both of these are substantially less than the maximum safe rate of supply from ton containers. Hence, there is no need to manifold more than two cylinders together (one in service and one awaiting) for either chlorine or sulfur dioxide. It would be reasonable for the treatment plant to anticipate ordering 2-ton containers of chlorine and 1-ton container of sulfur dioxide every four weeks. A safe inventory would be 3 containers of chlorine and 2 containers of sulfur dioxide (this would provide for a 28-day reserve for transportation, plus excess for emergencies), and provision should be made for storage of 2 empty chlorine containers and 1 empty sulfur dioxide container.

Step 8—A detailed cost estimate would be made at this point, based on actual catalog costs, or prior bid data. For the purpose of estimating chemical consumption and annual O&M costs during the period prior to attainment of actual design flow, the chlorine dose to achieve inactivation at various interim flows can be computed (using Equation 5-94), and integrated over an estimated flow-duration curve. Additionally, other alternative design contactor volumes may be assumed, and the design computations begun again commencing with step 4.

5.7.7 Economics

The estimation of chlorination, chlorine dioxide or dechlorination costs is highly site specific. Particular concerns include the necessity for separate chlorine contact basins (versus the use of an effluent channel as a contactor), the site-specific chemical costs, and the necessary chlorine dosages. However, preliminary rough estimates of process alternatives may be developed from available literature data based on field experience.

Geisser et al. developed cost equations based on their studies on the disinfection of combined sewer overflows using either chlorine or chlorine dioxide (179). Other useful sources of chlorination and dechlorination cost data are Chen and Gan (112), who have developed estimates for sulfur dioxide dechlorination systems, and Gumerman et al. (180), who provide cost estimates for drinking water chlorination systems.

The design engineer may also use standard estimation manuals for the costing of contact basins and physical structures, plus manufacturers' quotations on equipment (chlorinators, mixers, etc.).

5.8 Safety and Occupational Health Considerations

Halogen disinfectants have certain properties that must be considered at the design stage for the protection of operating personnel from risks that may arise during the plant life. Much detail is available in regard to necessary safety features to be incorporated in wastewater disinfection facilities using gaseous chlorine or hypochlorite as disinfecting agents. Little information is available in regard to adequate precaution using other halogens.

5.8.1 Physical Site Layout

For plants using gaseous chlorine, general space requirements may be estimated as follows (157):

- single chlorinator facilities of less than 200 lb/d (90 kg/d) capacity require at least 64 ft^2 (6 m^2) of space for the chlorinator and ancillary equipment;
- for plants with two chlorinators and feed rates up to 180 kg (400 lb)/d, 15 m^2 (160 sq ft) of area are needed for the chlorinators; and
- for each extra chlorinator above two, an additional 15 m^2 (160 sq ft) area should be provided.

In addition to the above areas needed for the chlorinator and evaporator modules, space is also required for the chlorine containers or cylinders being used to feed the system (or, in the case of sodium hypochlorite plants, the hypochlorite storage tanks) and space for inventory and empty cylinders or containers. The size of gaseous chlorine containers is given in Table 5-14, and, in conjunction with estimates of required inventory, working supply, and storage of empties, can be used to develop site area estimates for chlorine storage.

Table 5-14. Physical Dimensions of Chlorine Gas Containers

Capacity	Diameter (inches)[a]	Length (inches)
150 lbs (70 kg)	10.25 to 10.75	53 to 56
2000 lbs (910 kg)	30	79.75 to 82.5

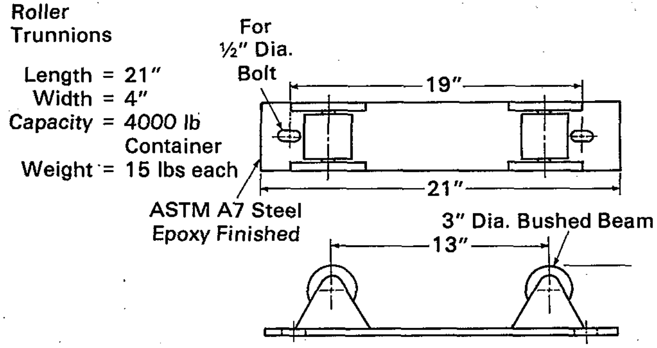
[a] 150 lb cylinders should be used and stored upright, thus requiring approximately 0.6 ft^2 (0.06 m^2) (plus separation area) per cylinder, while ton containers are used and stored in the horizontal position, requiring approximately 17 ft^2 (1.6 m^2) (plus separation area) per container.

The chlorine storage room (which may or may not be separate from the room containing the chlorinators) should be isolated from any other process by use of a separate building, or by use of an isolated room bounded by fire resistant walls (20). If the chlorine cylinders are physically separate from the chlorinators, and if the gaseous chlorine withdrawal method is used, it is necessary that the temperature of the room containing the chlorine gas cylinders be kept below the temperature of the room containing the chlorinator in order to prevent reliquifaction in the lines between the chlorine supply and the chlorinators. For chlorine installations practicing liquid withdrawal from either ton or bulk containers, outdoor storage is acceptable. Climatic considerations will dictate necessary site specific details, such as provision for prevention of icing of valves and shielding containers from direct sunlight.

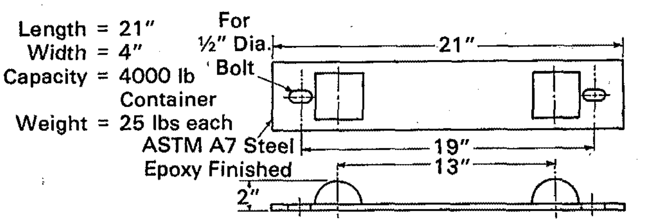
The chlorine supply room must have at least 2 means of egress, with doors opening outwards from the room. The structural designer must consider the dead load of full chlorine containers in his calculations. If ton containers are to be used, floor mounted trunions or scales (Figure 5-25) are needed. Furthermore, for ton containers, the use of overhead crane of at least 2-ton capacity in conjunction with a special lifting bar is needed to facilitate container movement, and therefore the physical site layout must carefully consider ceiling heights. It is highly desirable that the

Figure 5-25. **Ton container mounting trunions. (Courtesy Force Flow Equipment.)**

Storage Roller Trunnions



Storage Cradles



chlorine container(s) in use be placed on a scale or load cell to have a positive record of remaining disinfection.

Chlorine cylinders of 68-kg (150-lb) capacity are emptied in a vertical position, secured to a wall or a sturdy upright by means of a band or chain clamp to the upper portion of each cylinder.

In the chlorine storage room, ventilation must be such to assure a complete air change in 1-4 minutes. Generally, this is achieved by use of an exhaust fan near the floor of an outside wall (20); such fan must have a switch outside the chlorine room itself as well as inside. No ductwork, shafts, or other potential sources of gas travel should exist between the chlorine supply room and any other portion of the plant. The exhaust fan should vent at a level above neighboring buildings, trees, etc., to afford high dilution of any contained contaminant gases with the atmosphere.

No turpentine, ether, ammonia (except for small amounts associated with leak detection equipment), finely divided metals, or other flammable materials are permissible within the chlorine storage room (20).

5.8.2 Leak Detection

In gaseous chlorine installations, the major safety and health concerns involve the possibility for chlorine leakage to occur from a cylinder, a valve, or piping. The threshold limit value (TLV) for worker exposure to chlorine in air is 1 ppm by volume as an 8-hr time weighted average (181). Other atmospheric chlorine concentrations of interest are presented in Table 5-15 (20). It is particularly noteworthy that the minimum chlorine concentration detectable by odor is greater than the above specified TLV. Therefore, in order to provide for the continuous sensing of chlorine leakage, it is necessary to rely upon some chemical or electronic device. These may be of several types.

Table 5-15. **Gas Phase Chlorine Concentrations Evoking Specific Effects (20)**

Response	Concentration (ppm v/v)
Minimum odor threshold	3.5
Minimum 1-hour no serious effect level	4.0
Throat irritation	15.1
Coughing	30.2
30-minute, 1-hour danger level	40 to 60

The older type of continuous chlorine sensors rely upon a version of the iodometric chlorine detection procedure, using starch/iodide or other chemically impregnated paper and measuring the change in color that results when vapor containing chlorine passes over this material. A more modern version of this principle is used in a sensor that measures the current required to electrochemically reduce the chlorine present in a gas (Mine Safety Appliances, Pittsburgh, PA). A third principle used in continuous chlorine gas analyzers uses the change in electrical conductivity of a gas that occurs as chlorine concentrations increase (International Sensor Technology, Irvine, CA).

Once the vapor phase analyzer has signaled the onset of a chlorine leak, it is necessary to determine the cause of such event and proceed towards its repair. Personnel entering the room in which the ambient chlorine levels have exceeded the alarm limit should be clothed in protective equipment (gloves, breathing device, suit). Using portable continuous chlorine sensors, the location of highest chlorine concentration may be found, and thus the site of the leak inferred. Alternatively, it is possible to locate the site of a gaseous chlorine leak by spraying a solution or the vapors of a solution of strong ammonia (commercial, not household strength). A white cloud of NH_4Cl precipitate will form in the region where the chlorine leak exists (181).

5.8.3 Emergency Responses

All breathing devices used for personal protection in chlorine installations must be U.S. Bureau of Mines approved, and should be sterilized after use if it is anticipated that they may be used by another individual. The "buddy" system should be employed in situations where it is necessary to go into a room containing high vapor concentrations of chlorine.

Approved breathing devices may be categorized as follows (20):

Industrial Canister Mask ("gas masks"). This is only suitable for atmospheric chlorine concentrations below 1 percent and oxygen concentrations above 16 percent, and only for short duration use. It should not be relied upon for use during leaks, and rigid adherence to canister replacement time is required.

Self-Contained Breathing Apparatus. These may use a portable cylinder of oxygen or air, or a chemical generation system for continuous production of oxygen. For the latter type of system, it is necessary to wait until oxygen production has commenced prior to entry into a contaminated area. No oxygen system should be used for entry into closely confined areas (storage tanks) or where the danger of sparks or fire exists.

Positive Pressure Blower Mask. A hose is used to supply breathable air from either a remote air tank or a remote compressor. Use is permissible only if immediate safe escape from the contaminated area is possible in the event of a failure in the air supply system. If a compressor is used, its intake should be at least 6 ft (1.8 m) above grade to prevent inadvertent contamination with chlorine emanating from the area under investigation.

Personnel should be trained in the use of such breathing devices, and, in particular, in the location and management of chlorine leaks while using such protective equipment.

When a chlorine leak is located, if possible, the chlorine cylinder or container should be turned so that the defect is shifted at the upper portion of the container. This will assure that gaseous (rather than liquid) chlorine leakage occurs, and, eventually, the leak may be self-limited by evaporative cooling described earlier. All facilities handling gaseous chlorine should have available the appropriate Chlorine Institute emergency kit, and be familiar with its use. Devices in this kit may be used to patch or repair the leaking container, valve, or line.

If the gaseous or liquid chlorine that leaks can be collected, it is possible to neutralize it using alkaline solutions. Table 5-16 describes the required chemicals and water necessary to neutralize the contents of 68-kg (150-lb) and 910-kg (2,000-lb) chlorine containers. In the event of fire in association with a chlorine leak, no water should be used, since this will increase the corrosivity of the mixture. Non-water based extinguishers should be used until the chlorine leak is stopped.

Table 5-16. Neutralization Requirements for Chlorine Containers

Container Capacity	100 Percent Caustic	Soda Ash	Hydrated Lime
150 lb	188 lb in 60 gallon	450 lb in 150 gallon	188 lb in 188 gallon
2000 lb	2500 lb in 800 gallon	6000 lb in 2000 gallon	2500 lb in 2500 gallon

Should any persons come in contact, either by inhalation, or eye or skin contact, with chlorine resulting from a leak, the following first aid measures may be taken, prior to consultation with a physician (156):

General. Remove the person to an uncontaminated area, and remove contaminated clothing, washing any parts of the body exposed to chlorine with water.

Inhalation. If breathing has ceased, commence artificial respiration. When breathing recommences, or if breathing has not stopped, administer oxygen. Keep the person warm and at rest.

Eye Contact. The eye should be flushed with water for 15 minutes, holding the eyelids apart to get complete irrigation. In the design stage, eyewash basins should be provided.

Skin Contact. Wash the exposed parts with soap and water. It would be desirable to provide in the facility design provision for an emergency shower. Installations using sulfur dioxide should adhere to safety precautions regarding handling of a nature similar to those used in chlorine handling. At the design stage, it is particularly useful to specify fittings, valves, etc., for SO_2 service that are compatible with Chlorine Institute emergency kits.

5.9 Operation and Maintenance (O&M) Requirements

In the start-up of chlorination systems initially, or after a shut-down, a careful procedure must be followed to minimize the likelihood of chlorine leaks. Sepp and White outline the following procedure when gaseous chlorine is used (157):

1. Check all joints for proper gasketing and check that all supply valves are closed;

2. Check ejector for proper vacuum;
3. Place automatic chlorinators in manual mode, and set feed rate for about 25 percent of maximum;
4. Open one chlorine cylinder slightly, and check all joints for leaks. If no leaks are found the remaining number of required cylinders can be opened;
5. If a leak occurs, immediately close the cylinder valve, open all other valves fully and increase the chlorine feed setting on the chlorinator to its maximum value. When all chlorine has been ejected, repair the leak, and return to step (1);
6. Check performance at maximum rated capacity. The following are likely causes of capacity loss:
    a. insufficient ejector vacuum,
    b. vacuum line leaks,
    c. insufficient feed water pressure,
    d. solution line friction losses too high due to too small pipe diameter,
    e. air or gas binding in solution line.

The following modifications to the above must be made when the liquid chlorine withdrawal method is used (157):

1. Dry the entire system by heating water in the evaporator bath and passing dry air (-40°C dew point) through the evaporator and all supply lines from the cylinders to the chlorinators. Several hours may be required for this step.
2. Bring the evaporator water to the design temperature. Start the ejector water flow. Start up system with gas phase withdrawal to check leaks as above. If no leaks are found, then liquid withdrawal may be commenced.

The maintenance requirements for the chlorine supply system are as follows (157):

1. Flexible copper tubing reliability must be periodically checked. This can be done by bending the lines slightly. Any "screeching" is indicative of corrosion, and the tubing must be replaced.
2. Minute leakage from lines and fittings, undetectable by odor or standard leak detection procedures, may be ascertained by inspecting for signs of moisture accumulation or metal discoloration, both of which are signs of incipient leak development.
3. In some cases, a one inch flat file may be used to reface chlorine cylinder valves to ensure more precise seating.
4. Evaporator vessels should be inspected for sludge accumulation every year or each 200 tons of chlorine. If superheating capability declines, this may be indicative of either sludge accumulation or heater failure. Piping and connections to the evaporator should be inspected every six months. Sacrificial anodes used for corrosion control in the evaporator water bath should be inspected, and, if necessary, replaced, every six months.
5. The sludge accumulation in the evaporator may be cleaned by flushing with cold water until the effluent water runs clear. At this time, the vessel can also be inspected for pitting, and, if this is severe, replaced. Before placing the evaporator back into service, it must be dried by holding the external water temperature at 82°C (180°F) and maintaining a vacuum of 25 inches of mercury (85 kPa) for 24 hours.
6. The chlorine gas filter should be inspected every six months, at which time the filter element should be replaced. The sediment trap should be washed and dried at this time, and the lead gaskets disposed and replaced.
7. The chlorine pressure reducing valves may be cleaned of any deposits with a soft cloth, or, in more severe cases, with isopropyl alcohol or trichloroethylene. The valve spring should be replaced every two to five years.

The chlorinator system itself is subject to the following maintenance requirements (157):

1. The ejectors should be disassembled and cleaned every six months, and iron and manganese deposits removed with muriatic acid.
2. Booster pumps are subject to similar maintenance requirements as any other pump.
3. The chlorinator rotameters and floats should be removed and cleaned every six months, and the metering orifice inspected. All valve stems and seats should be cleaned and inspected once a year, and all valve springs should be replaced every two years.

The chlorine analyzer is probably the single system component requiring most careful attention. The following operation and maintenance requirements are essential (157):

1. Sample lines must be inspected daily for solids accumulation and filters and screens cleaned daily. This is particularly important in lines from dechlorination systems that lack residual, which could reduce line fouling.

2. In amperometric systems, electrodes should be cleaned weekly or biweekly, or more frequently if erratic or drifting calibrations occur.

3. The effluent line pH should be checked daily with pH paper. If it is not within the range of 4.5 to 5.0, a stronger buffer solution may be required.

4. Periodically, preferably at least daily, the analyzer reading should be compared with the results from a manual analysis. A control chart graphing the difference in results as a function of day should be maintained. If there is a consistent deviation, the span control on the analyzer may be adjusted on a weekly basis. A trend to increasing deviation may be indicative of instrument malfunction or sample line or filter fouling.

5. The adequacy of KI addition may be checked by adding slightly more KI than recommended to the feed buffer solution. If there is a change in cell response, this is indicative of insufficient KI. Continue increasing the KI concentration until no further change in the analyzer output occurs.

Culp and Heim present a useful guide to troubleshooting of chlorination systems (182). This is reproduced in Table 5-17.

5.10 Case Studies

In this section, several treatment plants that use chlorination and dechlorination will be described, indicating their design details, operating experience, and features of their facility that influence their ability to achieve satisfactory performance.

5.10.1 Stony Brook Regional Sewerage Authority Treatment Plant #1, Princeton, NJ

This treatment plant is located in Princeton, New Jersey and services a metropolitan area consisting of the Borough and Township of Princeton, West Windsor Township, Hopewell Borough and Township, Pennington Borough, and South Brunswick Township, NJ. Wastewater is collected in separate sewers, and the service area is essentially 100 percent domestic. Within the service area is the campus of Princeton University.

The current treatment plant was completed in 1977. The average daily design flow is 11 mgd (482 l/s), and the current average daily flow is 5 mgd (219 l/s). The discharge is to the Millstone River, a tributary to the Raritan River.

Permit requirements include a BOD of 8 mg/l, a 2 mg/l ammonia nitrogen requirement, a suspended solids requirement of 10 mg/l, a maximum chlorine residual of 0.05 mg/l, and a bacteriological standard of 20 fecal coliform/100 ml (enforced year round). The bacteriological standard is enforced at the chlorine contact tank effluent. In addition, a minimum dissolved oxygen requirement of 6.0 mg/l is enforced at the final effluent.

The treatment system consists of primary and preliminary treatment, activated sludge, separate stage biological nitrification, multi-media filtration, chlorination, dechlorination with sulfur dioxide, and reaeration.

The filtered, nitrified effluent routinely contains less than 1 mg/l suspended solids, and 1-3 mg/l total Kjeldahl nitrogen (occasional spikes of 10-20 mg/l occur). Nitrites have not been found in this effluent.

The disinfectant used is gaseous chlorine, supplied in ton cylinders. Chlorine usage is 6.5 kg/h (350 lb/d), on average (average dose of 8.4 mg/l) and two cylinders may be manifolded for withdrawal. The chlorine supply system was originally designed for liquid phase withdrawal, and the plant has evaporators; however, due to the low hydraulic loading, gas withdrawal has been found to be satisfactory, and no problems with reliquifaction have been noted. Approximately 1 cylinder in 30 is defective, and an inventory of about 6 cylinders is maintained, with new orders received when only two full cylinders remain (this is about a 10 day supply).

The chlorine solution is dispensed through solution feed chlorinators. The plant was originally designed to operate on a residual control mode; however, the operators noted that the analyzers were difficult to maintain, and that the control system was unstable (possibly due to a 200-500 ft length of sampling line, or to the location of the point of sampling prior to the end of the rapid decay phase of chlorine demand). The chlorination process is currently operated in manual mode, with residual monitored 3 times/shift at the end of the chlorine contact chamber. A residual of 1.5-2 mg/l is generally used as a control point (some of this residual is free).

The chlorine solution is applied at a separate stilling well through submerged nozzles. Two parallel, baffled (3 baffles of the end-around type) contact chambers are used. The contact time at design average flow is 18 minutes, making the current contact time at average flow about 40 minutes. Slime accumulation has occurred, and the contact tanks must be cleaned by high pressure water hoses every 3-4 months. Some difficulty in cleaning these tanks due to lack of provision for isolated clean outs for wasting of the loosened debris has been noted.

The dechlorination system uses gaseous sulfur dioxide, also supplied as "ton" containers, with gas withdrawal. The SO_2 utilization rate is 89 kg (197

Table 5-17. Trouble Shooting Guide Adapted (181)

Indicators/Observations and Action

1. Low chlorine gas pressure at chlorinator.
    A. Reduce feed rate and note if pressure rises appreciably after short period of time.
        (i) so, it is likely that more cylinders must be manifolded together to avoid exceeding the maximum safe withdrawal rate.
        (ii) if not, but if icing or cooling effect on lines continues, it is likely that there is a stoppage or flow restriction between the cylinders and chlorinators. Disassemble the header system, locate the blockage, and clean with solvent.

2. No chlorine gas pressure at chlorinator.
    A. Visually inspect to verify that chlorine cylinders are connected, and that they are not empty.
    B. Inspect the pressure reducing valve for plugging or damage, and repair, if necessary, following emptying chlorine gas from system be certain that a sediment trap is upstream of the valve.

3. Chlorinator will not feed any chlorine.
    A. Visually inspect the chlorinator valve stem and seat for dirt, and clean if necessary; precede the valve with a sediment trap.
    B. Measure the temperature in the chlorine cylinder storage area. If this area is warmer than the chlorinator room, reduce the temperature (shading, ventilation).

4. Chlorine gas escaping from chlorine pressure reducing valve (CPRV).
    A. Place ammonia bottle near termination of CPRV vent line to confirm leak.
        (i) disassemble valve and diaphragm, repairing if necessary. Inspect chlorine supply system for moisture intrusion.

5. Inability to maintain chlorine feed rate without icing of chlorine system.
    A. Reduce feed rate to about 75 percent of evaporator capacity. If this eliminates the problem, then there is an insufficiency of evaporatory capacity.
    B. Inspect CPRV cartridge and flush and clean if necessary.

6. Chlorination system unable to maintain water-bath temperature sufficient to keep external CPRV open.
    A. Remove and replace water bath heating element.

7. Inability to obtain maximum feed rate from chlorinator.
    A. Check chlorine gas pressure; if inadequate, remove and replace empty cylinders.
    B. Check water pump on injector for clogging, and clean, if necessary, with acid.
    C. Check for leaks in the vacuum relief valve; disassemble and replace all springs.
    D. Check for leaks in the vacuum lines, joints, gaskets associated with the chlorinator system by using an ammonia solution, or moistened starch-iodide indicator paper. Repair all leaks, and replace all leaking gaskets, tubing, etc.

8. Inability to maintain adequate chlorine feed rate.
    A. Inspect water feed pump and overhaul, if indicated. If a turbine pump is used, try closing the needle valve to maintain proper discharge pressure.

9. Wide variation in chlorine residual produced in effluent.
    A. Check chlorine meter capacity against plant flow meter.
        (i) replace with higher chlorination capacity meter.
    B. Check automatic controls, and request manufacturer's service, if indicated.
    C. Check for solids accumulation in the contact chamber, and clean, if necessary.
    D. Check zero and span of flow control device on chlorinator.
        (i) re-zero and span the device in accordance with manufacturer's instructions.

10. Chlorine residual analyzer recorder controller does not control chlorine residual properly.
    A. Inspect electrodes for fouling, and clean if necessary.
    B. Check loop-time, and reduce, if necessary, by one of the following actions:
        (i) move injector closer to point of application.
        (ii) increase velocity in sample line to analyzer cell.
        (iii) move cell closer to sample point.
        (iv) move sample point closer to point of application.
    C. Check that sufficient KI is being added for the anticipated amount of chlorine residual, and increase KI if indicated.
    D. Check pH of analyzer cell, and replace buffer solution, or increase buffer strength if the pH is out of range.
    E. Disconnect analyzer cell and apply a simulated signal to recorder mechanism. If recorder works, contact authorized service personnel for repair of analyzer cell.
    F. Analyze a series of samples taken after the point of mixing manually for chlorine residual under conditions of constant chlorine feed rate. If there is a wide variation in residual, inadequate turbulence at the point of mixing is indicated—enhance mixing efficiency.
    G. Check that the rotameter tube on the chlorinator is in the proper range, and replace, if necessary.

lb)/d. Based on a 2 mg/l chlorine residual, this is 2.4 mg SO_2/mg Cl_2. The sulfonator was designed for manual control and is controlled by monitoring the influent chlorine residual 3 times/shift. No particular problems with the sulfonator system were noted.

There is no data available on the dissolved oxygen at the effluent from the sulfonator mixing point; however, the plant was designed with reaeration due to the effluent DO permit condition. The operating personnel only use the aerators sporadically, and believe that the necessity for reaeration stems from the oxygen demand from nitrification rather than from any demand from sulfur dioxide.

Both the chlorine and sulfur dioxide containers are stored in a single room isolated from the other rooms in the plant. Temperature control and ventilation is provided. Necessary safety equipment, including self-contained breathing systems, and leak detectors are on-hand. The two cylinders each of sulfur dioxide and chlorine which are in use are mounted on hydraulic load scales. The operating personnel feel that the necessary handling and safety features appropriate to chlorine are also appropriate to sulfur dioxide. No corrosion or leak problems, other than occasional defective cylinders, have been noted.

Due to the substantial overcapacity, no significant down times have been recorded, except for an instance of failure of the sulfonator due to water intrusion. Spare vacuum regulator parts and valve springs are maintained. However, there is no periodic maintenance program.

The operating personnel estimate that 3 hr/d of operator time are required for the manual sampling and dose adjustments for the chlorinators and sulfonators, and about 0.5 hr/wk for cylinder switchovers. Some complaints as to the lack of manufacturer training for maintenance and troubleshooting of the chlorinators and sulfonators were voiced.

The 1983 annual chemical costs for the chlorination-dechlorination system were $14,000 for the chlorine ($0.24/kg) and $6,500 for sulfur dioxide ($0.20/kg).

5.10.2 Southeast Water Pollution Control Plant, City and County of San Francisco, CA

This treatment plant is located in San Francisco, CA and services a portion of the combined sewer system of the City and County of San Francisco. The flow is predominantly domestic.

The current treatment plant was completed in 1981. Design average flow for secondary treatment is 85 mgd (3,700 l/s) with a peak flow of 210 mgd (9,200 l/s). Current flows are 20-200 mgd (875-8,750 l/s). Discharge is to San Francisco Bay. Permit requirements are monthly average BOD/SS of 30/30 and a 5-day median total coliform requirement of 240/100 ml (with a geometric mean coliform requirement over 30 days of 200 fecal coliform/100 ml). The bacteriological requirements are met at the chlorine contact chamber exit, although problems with high counts after dechlorination have been noted. The instantaneous maximum chlorine residual is 0 mg/l.

Treatment consists of preliminary and primary treatment, closed tank oxygen activated sludge (non-nitrifying), chlorination and dechlorination. Chlorination is achieved using 14 to 15 percent sodium hypochlorite stored in fixed storage tanks. An approximate six day storage capacity exists. Chlorine dose is generally 12 mg/l.

While compound loop control hypochlorinators are installed, they have not been used, since it is believed (by the operators) that before the secondary treatment process went on line, the use of primary effluent fouled up the analyzers. Additionally, the automatic control system was originally designed for gaseous chlorine service, and only mid-way through construction were modifications made to switch to hypochlorite service. Due to more pressing operational difficulties, the shakedown of the automatic disinfection control system has not yet been accomplished. Flow-paced manual chlorine control is practiced in which the residual at the end of the contact chamber is measured every hour, and feed rate adjusted to maintain a 6-8 mg/l residual in dry weather and 3-5 mg/l in wet weather.

The hypochlorite is supplied to diffusers, and mixing at the point of application is supplied by two 30-hp (22-kW) turbine mixers. The chlorine contact system consists of an underground channel (2 identical parallel channels, each 680 m (2,200 ft) long, 3 m (10 ft) high and 2.8 m (9 ft) wide) providing 50 minutes contact at 85 mgd (3,700 l/s). No solids accumulation or sliming problems have been noted.

The dechlorination system uses sodium bisulfite supplied as a 22 percent (as SO_2) solution. Tankage for 1 week's supply exists. The dose of dechlorinating chemical is also controlled in a flow-paced manual mode after hourly analyses just after the point of sulfite addition to provide about a 1-2 mg/l excess dose. Mixing at the point of application is provided using two 22-kW (30-hp) mixers.

The plant appears to have substantial problems in achieving automatic chlorine control. However, these are traceable to the initial use of the chlorination equipment for treatment of primary effluent, and the need to modify the disinfection/dechlorination system from gaseous chlorine and sulfur dioxide to sodium hypochlorite and sodium bisulfite late in the design and construction process.

5.10.3 Sacramento Regional Wastewater Treatment Plant, Sacramento, CA

This treatment plant is located in Sacramento County, CA and services a metropolitan area including the City and County of Sacramento and the City of Folsom. Except for a small portion of the downtown Sacramento area, the facility services a separate sewer collection system. About 10 percent of the flow is due to industrial sources, primarily canneries, which contribute a high BOD, low nitrogen wastewater.

The current treatment plant was completed in 1982. The design flow is 136 mgd (6,000 l/s) (daily average) with a design hourly peak flow of 240 mgd (10,500 l/s). The current daily average flow is 130 mgd (5,700 l/s), with a peak hourly flow of 240 mgd (10,500 l/s). The discharge is to the Sacramento River.

Permit requirements include a BOD monthly average of 30 mg/l, 30 mg/l monthly average suspended solids, monthly average total coliform MPN of 23/100 ml, with a daily maximum of 500/100 ml, a maximum chlorine residual of 0.1 mg/l. The bacteriological standard applies at any location at the choice of the facility, and is currently achieved at the chlorine contact tank effluent.

The treatment system consists of primary and preliminary treatment, closed tank oxygen activated sludge (non-nitrifying, although an average of 0.4 mg/l nitrite is present in the secondary effluent), chlorination and dechlorination.

The secondary effluent contains an average of 12 mg/l suspended solids, 9 mg/l ammonia, and 22 mg/l total Kjeldahl nitrogen.

Gaseous chlorine is supplied in 90 ton tank cars delivered by rail. The chlorine dose is 13 mg/l for an average daily use of 260 kg/hr (14,000 lb)/d (giving a 12-day life for a tank car). Two tank cars are located on site, with one in service and one full tank car awaiting use, or on order. Standby equipment to use ton containers also exists. The tank car loading facilities are located away from the chlorinator building, and liquid chlorine is withdrawn into an intermediate storage tank with a level sensor. The drop in the liquid level sensor in the storage tank is used as a signal for switching to another tank car. No load scales exist on the tank cars and this is recognized as a design deficiency—the plant computer control system maintains a cumulative log of chlorine withdrawals from the tank car as a means of determining when the tank car is nearly exhausted. Withdrawal of chlorine is accomplished with a compressed air padding system.

From immediate storage, chlorine is fed to evaporators, and the gaseous chlorine is then fed to chlorinators. Chlorinators are controlled using a compound loop with residual trim algorithm. The residual at the end of the contact system is maintained at 9 mg/l.

The chlorine solution is applied through diffusers at a point where two two-speed 36/72 kW (35/70 hp) turbine mixers are located. The initial residual analyzer sampling point is located a short distance (probably less than 1 minute) from this point of mixing. The contact system consists of a 3,000 m (10,000 ft) pipe 2.6 m (8.5 ft) in diameter, that has never been inspected (in approximately 1 year of service) for slime or solids accumulation; however it has not appeared to be a problem. The second analyzer, providing the residual trim signal, is located at the end of this pipe.

The dechlorination system uses gaseous sulfur dioxide. The SO_2 is supplied either by truck or by rail and off loaded into a fixed 150 ton storage tank. The fixed storage tank was modified such that its fittings are compatible with the Chlorine Institute emergency kit specifications. There is no load scale on the storage tank, and inventory is monitored using the control computer system. Withdrawal of liquid SO_2 is accomplished under compressed air padding, and the compressor system is separate from that used for the chlorine withdrawal system. Sulfur dioxide dose is typically 1 mg/mg chlorine residual, plus an excess of 2 mg/l for a daily use of 5,450 kg (12,000 lb).

The sulfur dioxide is fed to evaporators that supply sulfonators. Gaseous sulfur dioxide under vacuum is then piped 3,000 m (10,000 ft) to the end of the chlorine outfall, where it is mixed in ejectors and then, using four circumferentially-mounted pipe diffusers, injected to the wastewater flow. The control system is compound loop using a biased chlorine residual signal using a hypochlorite biasing system. The set point is at a 2 mg/l sulfur dioxide excess, partially due to the requirement for maintenance of an undetectable chlorine residual, and partially due to the inability to achieve precise sulfonation control in the long vacuum line (a lag of 5 minutes in the SO_2 vacuum supply line was cited). No problems with dissolved oxygen depression or pH were cited, although continuous pH sensors are used.

No problems with slime growth or recontamination subsequent to dechlorination were expressed, although the total coliforms measured after dechlorination are somewhat erratic.

The plant has been fully operational for only about one year, so evaluations of corrosion and machine reliability are difficult. Some problems noted to date are the corrosion of the trim chlorine analyzers from the chlorine vapors emanating at the end of the contact pipe, the breakdown of PVC solution in sunlight (painting of PVC is suggested), and the lag time in the sulfonation control system noted above.

The dechlorination analysis system is cleaned and calibrated twice daily, and it is estimated that total operator time required for operation of the chlorination/dechlorination system is 4 hr/d. Table 5-18 summarizes the overall O&M schedule for the major components of the chlorination-dechlorination system at Sacramento.

5.11 References

When an NTIS number is cited in a reference, that reference is available from:

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Table 5-18. O&M Schedule, Sacramento Regional Wastewater Treatment Plant

Component	Action
Cl_2 leak indicator	Adjust sample flow rates (M)
Cl_2 pressure reducing shutoff valves	Remove and clean valve seats (SA)
SO_2 ejector	Check and clean (if necessary) (M).
NaOCl biasing pump	Check and adjust V-belts (Q). Overhaul (A).
Chlorine Unloading platform	Exercise emergency baths and all valves (M). Check for leaks (M). Check: alarm switch lights, emergency repair kit maintenance tools and masks, catwalks and rails, safety chains, area lighting, tank chocks, safety signs, and padding air quick discharge (M).
Air padding system	Check oil level, air intake, drain condensate trap, separator (W). Tighten bolts (M). Check pressure reducing valves (M). Change oil frame oil (Q). Clean cooling coil (Q). Inspect intake air cleaner and valves (Q). Check-motor lubrication (A).
Chlorine room exhaust fan	Lubricate fan bearings (M). Check V-belt tension (M). Check fan wheel (Q). Check fan motor and bearings, V-belt alignment (SA).
Air padding dryer	Check purge rate and temperature, pressure and flow rate, filter pressure drop, solenoid valves, cycle timer (M). Check outlet dew point and blow down relief valves (Q). Change air filters (SA). Inspect exhaust mufflers (SA). Inspect dessicant, check valve seats, and solenoid valves (A).
Chlorinators and sulfonators	Remove organic residues with wood alcohol and inorganic residues with hydrochloric acid (A).
Cl_2 and SO_2 emergency expansion tanks	Visual inspection (A).
Flash mixers	Check motors (M).
Sulfonator ejector	Check motor condition and lubrication, pumps packings, and pump lubrication and condition (M).
Hypochlorite mix tanks	Clean (M).
Chlorine analyzers	Check and clean sample intake lines, electrodes, reagent reservoir and sample pH (D). Check calibration, sample flow, and liquid level (W). Clean constant head tank (W).
Chlorine evaporators	Clean and inspect vaporization chamber (A).

(D) = daily; (W) = weekly; (M) = monthly; (Q) = quarterly; (SA) = semiannually; (A) = annually.

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Chapter 6

Ozone Disinfection

6.1 Introduction

6.1.1 General

The planning, design, construction and operation of a wastewater treatment process involves at least five separate parties: owner, design engineer, contractor, equipment manufacturer, and regulatory agency. Because the use of ozone for wastewater disinfection is relatively new, the responsibility for development of process and equipment design has been largely the equipment manufacturer's. However, many design features are not related to a given manufacturer's piece of equipment, like process flexibility considerations, contact basin size and configuration, and influent water quality characteristics. The information presented in this manual is intended to provide necessary design capability and understanding to all parties so that each party can be properly involved in the development of an ozone disinfection system.

During the development of this design manual, site visits to seven operating ozone installations were conducted in order to obtain information on operating experiences at existing facilities. The information obtained from these site visits is incorporated in the manual.

6.1.2 History of Ozone

A history of ozone has been documented by others (1,2). This work was based on surveys of operating ozone installations in Europe and the United States and is briefly described.

Experiments conducted in 1886 showed that ozonized air "will effect the sterilization of polluted water." The first drinking water plant to use ozone was built in 1893 at Oudshoorn, Holland. The French studied the Oudshoorn plant, and after pilot testing constructed an ozone water plant at Nice, France in 1906. Because ozone has been used at Nice since that time, Nice is often referred to as "the birthplace of ozonation for drinking water treatment."

In Europe there is a strong commitment to attain a water of the highest chemical quality. Currently, there are over 1,000 European drinking water plants that use ozone at one or more points in the treatment process.

In contrast to the widespread use of ozone for water treatment in Europe, very few European wastewater ozone disinfection systems exist. Currently, there are more ozone disinfection systems in use at U.S. wastewater plants than at U.S. water plants or European wastewater plants.

The first U.S. wastewater plant to use ozone for disinfection was Indiantown, Florida, which began operation in 1975 (3). By 1980 about 10 wastewater treatment plants using ozone for disinfection had been constructed. Unfortunately, some of these earlier ozone disinfection facilities have chosen to abandon ozone disinfection for one or more of the following reasons: excessive high cost of operation, equipment problems, excessive maintenance cost, and inability to attain performance objectives without major modifications (4). Despite these early setbacks, many facilities have proposed to utilize ozone for wastewater. A list of wastewater plants that have used or are using ozone disinfection is presented in Table 6-1 (5).

6.2 Ozone Properties, Chemistry and Terminology

Ozone is a molecule that can co-exist with air or high purity oxygen, or can dissolve in water. It is a very strong oxidizing agent and a very effective disinfectant.

6.2.1 Ozone Properties

Ozone (O_3) is an unstable gas that is produced when oxygen molecules are dissociated into atomic oxygen and subsequently collide with another oxygen molecule (6). The energy source for dissociating the oxygen molecule can be produced commercially or can occur naturally. Some natural sources for ozone production are ultraviolet light from the sun and lightning during a thunderstorm.

Ozone may be produced by electrolysis, photochemical reaction, radiochemical reaction, or by "electric discharge" in a gas that contains oxygen (7). The electric discharge principle has been used in most commercial applications and in all known water and wastewater treatment applications. The electric discharge method is presented in this manual.

Table 6-1. U.S. Municipal Wastewater Treatment Plants Using Ozone (5)

Location	Feed Gas	Primary Purpose of Ozone	Startup Date	Average Flow Rate mgd	Average Flow Rate m^3/d
Indiantown, FL	air	disinfection	1975	0.5	1,890
Woodlands, TX	air	disinfection	1976	1.5	5,680
Upper Thompson Sanitation District, CO	air	disinfection	1977	1.5	5,680
Hunter, NY	air	disinfection	1977	0.1	368
Harriman, NY	air	disinfection	1977[a]	0.36	1,360
Chino Basin, CA	air	suspended solids	1978	5	18,930
Palo Alto, CA	air	organics	1978	4	15,100
Collegeville, MN	air	disinfection	1978	0.22	830
Mahoning County, OH*	O_2	disinfection	1978	4	15,100
Hunter Highlands, NY	air	disinfection	1978	<1	<3,780
Cotter Gasville, AR	air	disinfection	1978[a]	1	3,780
Springfield, MO*	O_2	disinfection	1978[a]	30	113,600
Bull Shoals, AR	air	disinfection	—	0.19	720
Sebring, FL	air	disinfection	—	0.52	1,960
Maryland City, MD	air	disinfection	—	0.58	2,200
Oak Ridge, NY	air	disinfection	1980[a]	0.12	450
Norton AFB, CA	air	phenols	1980[a]	0.25	950
Carmel, NY	air	disinfection	1980[a]	1	3,780
Potomac Heights, MD	air	disinfection	1980[a]	0.2	760
Murphreesboro, TN*	O_2	disinfection	1980[a]	8	30,280
Pensacola, FL*	O_2	disinfection	1980	20	75,700
Hercules, CA	air	disinfection	1980[a]	0.4	1,510
Marion, NY	air	disinfection & BOD flotation	1980[a]	0.125	470
Brookings, SD	air	disinfection	1980[a]	6	22,700
Concord, NC*	O_2	disinfection	1980[a]	25	94,600
Delaware County, OH	air	disinfection	1980[a]	1.5	5,680
Frankfort, KY	air	disinfection	1980[a]	7	26,500
Ocean City, MD*	O_2	disinfection	1980[a]	12	45,400
Madisonville, KY*	O_2	disinfection	1980[a]	5	18,900
Little Valley, NY	air	disinfection	1981	0.28	1,060
Yaphank, NY	air	disinfection	1981	0.12	450
Granby, CO	air	disinfection	1981	<1	<3,680
Rocky Mount, NC*	O_2	disinfection	1982[a]	40	151,400
West Knoxville, TN	air	disinfection	1982	2	7,570
Vail, CO	air	disinfection	1982[a]	2.7	10,200
Hagerstown, MD*	O_2	disinfection	1983	8	30,300
Olympia, WA*	O_2	disinfection	1983	14	53,000
Indianapolis, IN					
Belmont Plant*	O_2	disinfection	1983	120	454,200
Southport Plant*	O_2	disinfection	1983	125	473,200
Twining, NM	air	disinfection	1983	0.095	360
Somero, NY	air	disinfection	1983	0.1	368
Alburndale, FL	air	disinfection	1984	2	7,570

[a] Plants operating per 1982 EPA survey

*Plants using oxygen activated sludge process

Ozone is used in relatively low concentrations in water and wastewater treatment applications. The properties of pure ozone are presented for general background information in Table 6-2 (7).

6.2.1.1 Ozone Color and Odor

At ordinary temperatures ozone is a blue gas, but at typical concentrations its color is not noticeable unless it is viewed through considerable depth (7).

Ozone has a very distinct odor and it owes its name to its odor. The word ozone is derived from the Greek word 'ozein', which means to smell. Ozone can be detected at concentrations of only 2x10^{-5} to 1x10^{-4} g/m^3 (i.e., 0.01 to 0.05 ppm by volume) (1). The ability to smell ozone at small concentrations is considered a safety feature because an ozone odor can be detected before ozone related health considerations develop. The present allowable 8-hour exposure concentra-

Table 6-2. Properties of Pure Ozone (O_3) (7)

melting point, °C	−192.5 ± 0.4
boiling point, °C	−111.9 ± 0.3
critical temperature, °C	−12.1
critical pressure, atm	54.6
critical volume, cm³/mole	111

density and vapor pressure of liquid

Temperature, °C	Density, g/cm³	Vapor Pressure, Torr
−183	1.574	0.11
−180	1.566	0.21
−170	1.535	1.41
−160	1.504	6.73
−150	1.473	24.8
−140	1.442	74.2
−130	1.410	190
−120	1.378	427
−110	1.347	865
−100	1.316	1605

density of solid ozone, g/cm³, at 77.4°K	1.728
viscosity of liquid, cP³, at 77.6°K	4.17
at 90.2°K	1.56
surface tension, dyn-cm, at 77.2°K	43.8
at 90.2°K	38.4
parachlor* at 90.2°K	75.7
dielectric constant, liquid, at 90.2°K	4.70
dipole moment, debye	0.55
magnetic susceptibility, cgs units, gas	0.002×10^{-6}
liquid	0.150
heat capacity of liquid from 90 to 150°K	$C_P = 0.425 + 0.0014\ (T-90)$
heat of vaporization, kcal/mole, at −111.9°C	3410
at −183°C	3650

heat and free energy of formation

	ΔH_f, kcal/mole	ΔG_f, kcal/mole
gas at 298.15°K	34.15	38.89
liquid at 90.15°K	30.0	
hypothetical gas at 0°K	34.74	

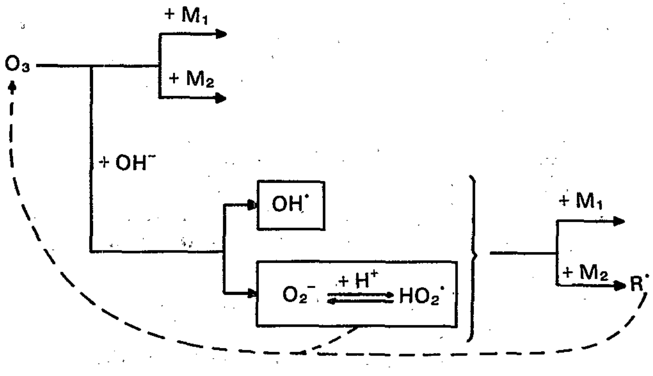
*$M_{\gamma}^{1/4}$ (D − d) where M = molecular weight; Γ = surface tension; D = liquid density; d = vapor density

tion is $2x10^{-5}$ g/m³ (i.e., 0.1 ppm by volume), which is 2 to 10 times higher than the concentration at which ozone can be smelled. Additional safety considerations for ozone are discussed in Section 6.6

6.2.1.2 Ozone Stability

The stability of ozone is greater in air than in water but is not excessively long in either case. The half-life of residual ozone in water is reported to range from 8 minutes to 14 hours depending on the phosphate and carbonate concentration of the water (8). With no phosphates or carbonates and the water adjusted to pH 7.0 with sodium hydroxide, the half-life was 8 minutes. Hoigne and Bader (9) found that ozone will react directly with solutes in the water, and that hydroxide ions and hydroxyl radicals will provide a catalyst for the decomposition of ozone into intermediate compounds that are also reactive, such as peroxide ions and hydroxide ions. Their findings are summarized in Figure 6-1.

Figure 6-1. **The direct reaction of ozone with solutes (M_1, M_2) and a hydroxide ion (or radical) catalyzed decomposition reaction, leading to reactive intermediates, compete for ozone (9).**



Hoigne and Bader's results suggest that ozone disinfection is influenced by raw water chemistry characteristics, in addition to the more well known influences of wastewater pollutants. The water chemistry influences are generally not utilized in ozone system design, except as developed in pilot plant studies of the specific wastewater to be disinfected. However, the water chemistry influences are important to keep in mind when a comparison is made of ozone disinfection performance at different facilities.

The residual ozone concentration in water is decreased rapidly by aeration or agitation of the liquid (7). A wastewater sample with residual ozone must not be agitated or collected after a period of agitation has occurred, or the measurement of residual ozone will be inaccurate. Also, because ozone may be released from the water, a plant effluent that contains a high residual ozone concentration that is used as a source of non-potable water, may release excessive amounts of ozone and contaminate the ambient environment.

The stability of ozone in air or oxygen is significantly affected by the temperature of the gas. In a clean vessel at room temperature the half-life of ozone may range from 20 to 100 hours (7). At 120°C (248°F) the half-life is only 11 to 112 minutes and at 250°C (482°F) only 0.04 to 0.4 seconds. This characteristic of ozone is important for design because cooling of the ozone generators is necessary. Also, good room ventilation is necessary in case an ozone leak occurs, and ozone contained in the off-gas must be destroyed.

6.2.1.3 Ozone Physical Characteristics

Gaseous ozone is explosive at an ozone concentration of 240 g/m^3 (20 percent wt in air) (7). Fortunately, the maximum gaseous ozone concentration typically found in water or wastewater ozone disinfection systems does not exceed 50 g/m^3 (4.1 percent wt in air). If, however, a medium that can adsorb and concentrate ozone is inappropriately located in the system, then explosive ozone concentrations could develop.

6.2.1.4 Ozone Solubility

Ozone solubility in water is important because ozone disinfection is dependent upon the amount of ozone transferred to the wastewater. Henry's law relative to ozone systems states that the mass of ozone that will dissolve in a given volume of water, at constant temperature, is directly proportional to the partial pressure of the ozone gas above the water (10). Using Equation 6-1 the maximum solubility of ozone in water at various temperatures and feed-gas ozone concentrations has been developed, as shown in Table 6-3.

$$H = Y/X \qquad (6\text{-}1)$$

where:

Y = partial pressure of the gas above the liquid atmospheres
X = molar fraction of the gas in the liquid at equilibrium with the gas above the liquid
H = Henry's law constant (varies with temperature), atm/mole fraction

Table 6-3. Solubility of Ozone in Water

Water Temperature (°C)	Henry's Constant Atm/Mole	Ozone Concentration		Ozone Solubility mg/l
		mg/l	ppm-vol	
0	1,940	12.07	6,044	8.31
5	2,180	12.07	6,044	7.39
10	2,480	12.07	6,044	6.50
15	2,880	12.07	6,044	5.60
20	3,760	12.07	6,044	4.29
25	4,570	12.07	6,044	3.53
30	5,980	12.07	6,044	2.70
0	1,940	18.11	9,069	12.47
5	2,180	18.11	9,069	11.09
10	2,480	18.11	9,069	9.75
15	2,880	18.11	9,069	8.40
20	3,760	18.11	9,069	6.43
25	4,570	18.11	9,069	5.29
30	5,980	18.11	9,069	4.04
0	1,940	24.14	12,088	16.62
5	2,180	24.14	12,088	14.79
10	2,480	24.14	12,088	13.00
15	2,880	24.14	12,088	11.19
20	3,760	24.14	12,088	8.57
25	4,570	24.14	12,088	7.05
30	5,980	24.14	12,088	5.39
0	1,940	36.21	18,132	24.92
5	2,180	36.21	18,132	22.18
10	2,480	36.21	18,132	19.50
15	2,880	36.21	18,132	16.79
20	3,760	36.21	18,132	12.86
25	4,570	36.21	18,132	10.58
30	5,980	36.21	18,132	8.09

Note: The concentration of the ozone gas is determined at a standard temperature of 68°F (20°C) and a standard pressure of 1 atmosphere (101 kPa).

6.2.2 Ozone Chemical Reactions

Ozone is a very strong oxidizing agent, having an oxidation potential of 2.07 volts (1). Ozone will react with many organic and inorganic compounds in the wastewater. These reactions are typically called "ozone demand" reactions. They are important in ozone disinfection system design because the reacted ozone is no longer available for disinfection. Wastewaters that have high concentrations of organics or inorganics may require high ozone dosages to achieve disinfection. It is very important to conduct pilot plant studies on these wastewaters during ozone disinfection system design in order to determine the ozone reaction kinetics for the level of treatment prior to ozone disinfection.

In most instances, the oxidation reactions produce an end product that is less toxic than the original compound (1,2). Numerous studies have been completed describing the reactions with ozone and various inorganic and organic compounds (11,12). A brief summary of these reactions is presented.

6.2.2.1 Reactions with Inorganic Compounds

The inorganic compounds that most commonly react with ozone in a wastewater treatment plant are sulfide, nitrite, ferrous, manganous, and ammonium ions. Other reactions may also occur if the wastewater characteristics are affected by an industrial contribution or by in-plant recycle loads. Ozone reactions with various inorganic compounds have been analyzed by several researchers (2). These reactions are summarized below:

Sulfide. The degree of oxidation of sulfide depends upon the amount of ozone used and the contact time. Organic sulfides will oxidize to sulfones, sulfoxides and sulfonic acids at slower rates than the sulfide ion itself. The sulfide ion will oxidize to sulfur, to sulfite and to sulfate.

Nitrogen Compounds. Organic nitriles, nitroso compounds, and hydroxylamines will be oxidized depending on the amount of ozone used and the contact conditions. The oxidation reaction of ammonia is first-order with respect to the concentration of ammonia and is catalyzed by OH^- over the pH range 7-9 (13). At an initial ammonium concentration of 28 mg/l as N, a pH of 7.0, and a contact time of 30 minutes, an 8 percent reduction of ammonium was reported. At a pH of 7.6, a 26 percent, 8.4 a 42 percent, and 9.0 a 70 percent reduction was observed. Narkis reports that total oxidation of organic nitrogen and ammonia was never achieved, even at a pH of 12, and at a pH of 6 nitrates were not produced (14).

Nitrite ion is oxidized very rapidly to nitrate ion. This reaction can have a significant effect on ozone disinfection capability when incomplete nitrification occurs. Venosa reported that as much as 2 mg/l of ozone was required to oxidize 1 mg/l of nitrite-nitrogen (15).

Iron and Manganese. The reaction with ozone and the ferrous and manganous ions will form an insoluble precipitate. The ferrous ion will be oxidized to ferric, which will react with OH^- to form an insoluble precipitate. Similarly, manganeous ions will form manganic ions which will react with OH^- to form a precipitate.

Cyanide. Toxic cyanide ions are readily oxidized by ozone to the much less toxic cyanate ion. At low pH, cyanate ion hydrolyzes to produce carbon dioxide and nitrogen.

6.2.2.2 Reactions with Organic Compounds

An in-depth analysis of the reactions with ozone and various organic compounds were developed by several investigators. These reactions were described by Miller et al. (1), and are summarized below.

Aromatic Compounds. Phenol reacts readily with ozone in aqueous solution. Oxalic and acetic acids are relatively stable to ozonation in the absence of a catalyst such as ultraviolet light or hydrogen peroxide. Cresols and xylenols undergo oxidation with ozone at faster rates than does phenol. Pyrene, phenanthrene, and naphthalene oxidize by ring rupture. Chlorobenzene reacts with ozone slower than does phenol.

Aliphatic Compounds. There is no evidence that ozone reacts with saturated aliphatic hydrocarbons under water or wastewater treatment conditions. There is no evidence that ozone oxidizes trihalomethanes. Ozone combined with ultraviolet radiation does oxidize chloroform to produce chloride ion, but no identified organic oxidation product. Unsaturated aliphatic or alicyclic compounds react with ozone.

Pesticides. Ozonation of parathion and malathion produces paraoxon and malaoxon, respectively, as intermediates, which are more toxic than are the starting materials. Continued ozonation degrades the oxons, but requires more ozone than the initial reaction. Ozonation of heptachlor produces a stable product not yet identified. Aldrin and 2,4,5,-T are readily oxidized by ozone, but dieldrin, chlordane, lindane, DDT, and endosulfan are only slightly affected by ozone.

Humic Acids. Humic materials are resistant to ozonation, requiring lengthy times of ozonation to produce small amounts of acetic, oxalic, formic and terephthalic acids, carbon dioxide, and phenolic compounds. Ozonation of humic materials followed by immediate chlorination (within eight minutes) has been shown to reduce trihalomethane formation in some cases. Ozonized organic materials generally are more biodegradable than the starting, unoxidized compounds.

6.2.3 Ozone Disinfection Reactions

Transfer of ozone into the wastewater is the first step in meeting the disinfection objective, since ozone must be transferred and residual oxidants produced before effective disinfection will occur (16). Once transferred, the residual oxidants, such as ozone, hydroxide, or peroxide, must make contact with the organisms in order for the disinfection action to proceed. Therefore, similar requirements and kinetic relationships used for chlorine disinfectants can also be used for ozone disinfection.

Contact time has been the subject of much controversy for ozone disinfection. Studies have shown that effective disinfection can occur at contact times as short as one minute (17), but most existing ozone disinfection systems have contact times that are 10 to 15 minutes. Effective ozone disinfection is due to the combined results of high transfer efficiency, good mixing, adequate contact time, and minimal short-circuiting in the contactor. These factors are all interrelated, thus isolating contact time as a particular variable during kinetic analysis has not been practical. As a result, kinetic relationships in terms of transferred ozone or residual concentrations have been utilized.

The transferred ozone dosage versus disinfection performance kinetic relationship is further discussed in Section 6.5.1.2. The relationship between disinfection performance and residual oxidants is shown in Figure 6-2 (17). These types of relationships can be used graphically or empirical equations can be developed and used to predict disinfection performance at various dosage rates. All these relationships assume that effective gas/liquid contacting occurs.

Studies have shown that ozone disinfection can also be related to contactor off-gas ozone concentration (15,18,19). Additional work has shown that it was possible to empirically relate effluent fecal coliform concentration to the product of contactor off-gas concentration and liquid contact time (20). The data shown in Figures 6-3 and 6-4 indicate that a better correlation occurred for the product of off-gas ozone concentration times time than occurred for off-gas ozone concentration alone. Additional information regarding the use of off-gas ozone concentration as a process control parameter is presented in Section 6.3.3.

Figure 6-2. **Effluent total coliform concentration versus total residual oxidants and residual ozone (17).**

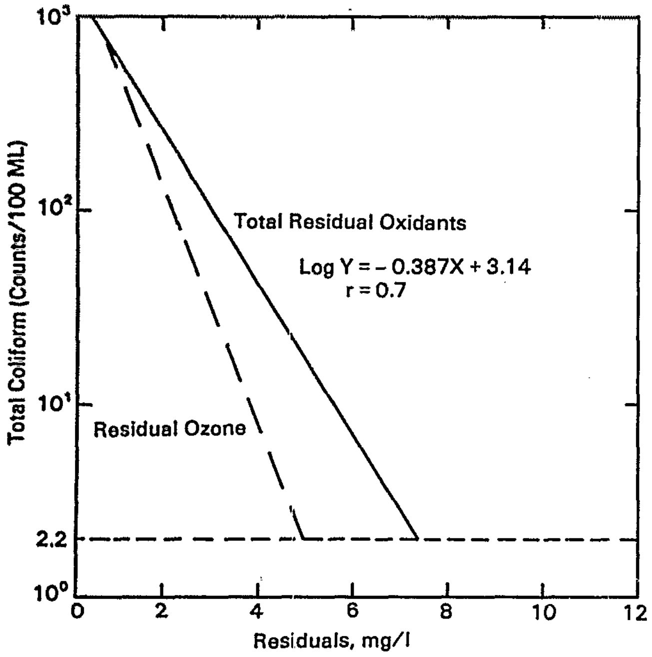


Figure 6-3. **Effluent fecal coliform concentration versus off-gas ozone concentration (20).**

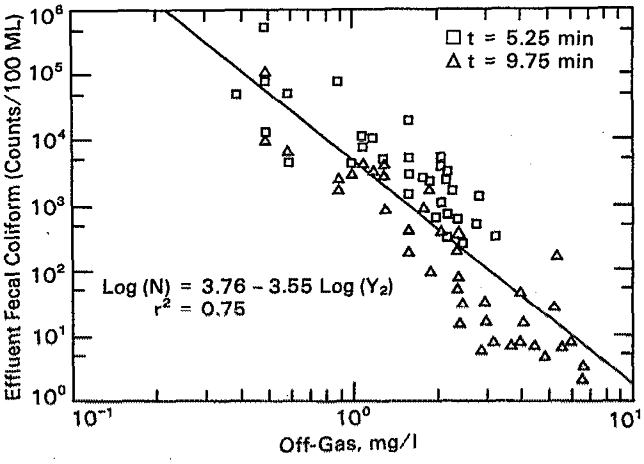
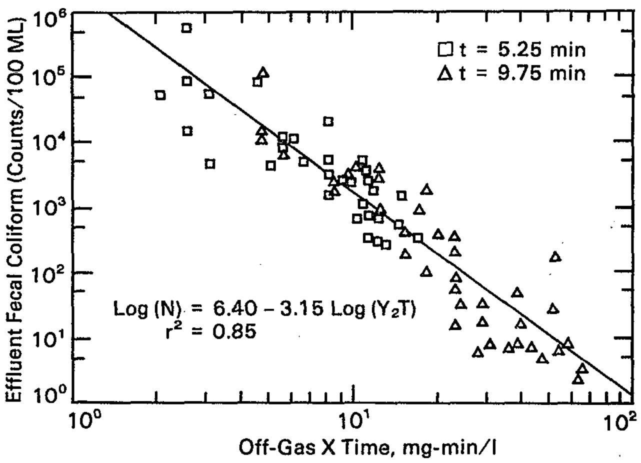


Figure 6-4. **Effluent fecal coliform concentration versus product of off-gas ozone concentration times time (20).**



6.2.4 Ozone Terminology

Throughout this manual the most common units of expression are used. The standard temperature and pressure are 20°C (68°F) and 101 kPa (1 atmosphere), respectively. All gas volume values are corrected to standard temperature and pressure, unless otherwise noted.

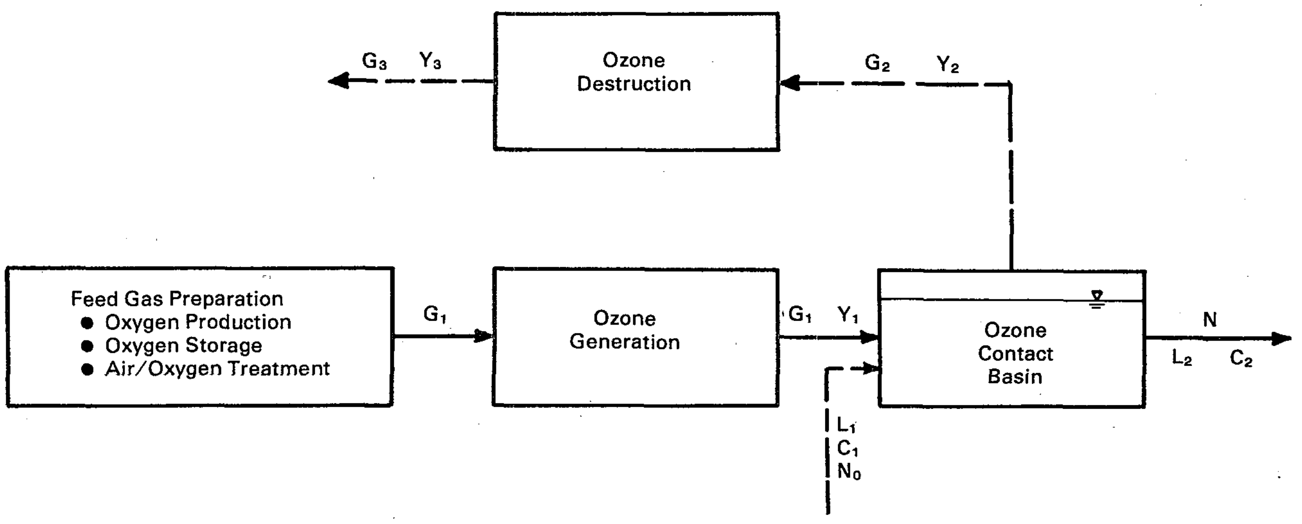
6.2.4.1 Measured Ozone Parameters

A simplified line diagram of a wastewater ozone disinfection system is shown in Figure 6-5. The system consists of the four components of feed-gas preparation, ozone generation, ozone contacting and ozone destruction. The feed-gas source may be either air or high purity oxygen. In unique applications recycled oxygen has been utilized. The feed-gas flow rate (G_1) is usually measured before the ozone generator so more precise instrumentation may be used without the added expense of providing ozone resistant materials.

The concentration of ozone in the feed-gas (Y_1) is measured before the ozone contact basin. In the contact basin most of the ozone is transferred to the wastewater and reacts with the ozone demanding constituents in the wastewater or decomposes to hydroxyl radicals (9). Some of the transferred ozone may not react or decompose and will exist in the contactor as residual ozone (C_2). Ozone not transferred exits the contact basin in the off-gas flow (G_2).

The off-gas flow rate (G_2) can be quite different from the feed-gas flow rate (G_1). When oxygen is used as the feed-gas, from 5 to 10 percent of the oxygen may dissolve in the wastewater and cause the off-gas flow rate to be lower than the feed-gas flow rate. If an exhaust blower is used to pull the off-gas out of the contact basin, then the off-gas flow rate could be higher than the feed-gas flow rate due to air getting into the tanks through cracks, etc. The off-gas flow rate may be measured before or after the ozone destruct unit since the two flows are identical. The ozone concentration in the off-gas (Y_2) is measured prior to the ozone destruct unit and the ozone concentration in the exhaust-gas (Y_3) is measured after the ozone destruct unit. The wastewater flow into (L_1) and out of (L_2) the contact basin is identical, and may be measured at either point. The influent coliform concentration (No) is sampled prior to the

Figure 6-5. Simplified ozone process schematic diagram.



Legend:

G_1 = Feed-Gas Flow Rate (m^3/min)
Y_1 = Feed-Gas Ozone Concentration (g/m^3)
G_2 = G_3 = Off-Gas and Exhaust-Gas Flow Rate (m^3/min)
Y_2 = Off-Gas Ozone Concentration (g/m^3)
Y_3 = Exhaust-Gas Ozone Concentration (g/m^3)
L_1 = L_2 = Wastewater Flow Rate
C_1 = Residual Ozone Concentration in Wastewater Influent (mg/L)
C_2 = Residual Ozone Concentration in Wastewater Effluent (mg/L)
N_0 = Influent Coliform Concentration (#/100 ml)
N = Effluent Coliform Concentration (#/100 ml)

ozone contact basin and the effluent coliform concentration (N) is sampled after the contact basin.

The metric units of expression and English unit conversion factors for typical ozone related measurements are presented in Table 6-4. The English unit conversions for ozone concentration in the gas are dependent upon the standard temperature and pressure conditions selected. The conversion factors shown are for a standard temperature of 20°C (68°F) and a standard pressure of 101 kPa (1 atmosphere). If other standards are used, then the conversion factors may be changed using the formulas shown in Table 6-4. The English unit equivalents for typical ozone concentrations encountered in ozone disinfection facilities are presented in Table 6-5.

6.2.4.2 Calculated Ozone Parameters

Ozone measurements described in the previous section can be used to calculate ozone production (P), applied ozone dosage (D), ozone transfer efficiency (TE), and transferred ozone dosage (T). The terminology, symbols, units of expression and formulas for the calculated ozone parameters are shown in Table 6-6.

The formulas shown assume that metric units of expression are used for the individual parameters. The most common English units of expression are percent wt for Y_1 and Y_2, scfm for G_1 and G_2, mgd for L_1 and L_2, and lb/d for P. To calculate P, the gas flow rate is typically converted to lb/d and is multiplied by the ozone concentration in percent wt.

P in lb ozone/day = G_1 in lb/d * Y_1 in % wt/100

The applied ozone dosage, in English units, is equal to the ozone production in lb/d divided by liquid flow in mgd divided by a conversion factor of 8.34 lb/gal.

D in mg/l = P in lb ozone/d/L_1 in mgd/8.34

Table 6-4. Terminology for Measured Ozone Parameters

Parameter	Symbol	Metric Units	English Conversions
Flow Measurements			
Feed-gas	G_1	m^3/min	cfm × 0.02832
Off-gas	G_2	m^3/min	cfm × 0.02832
Exhaust-gas	G_3	m^3/min	cfm × 0.02832
Wastewater	L_1 & L_2	m^3/min	mgd × 2.629 gpm × 0.003785
Ozone Concentration Measurements			
			Air Feed-gas
Feed-gas	Y_1	g/m^3	%Vol × 19.96
Off-gas	Y_2	g/m^3	%Wt × 12.10
Exhaust-gas	Y_3	g/m^3	ppm Vol × 0.001996
			ppm Wt. × 0.001210
			Oxygen Feed-gas
			%Vol × 19.96
			%Wt × 13.35
			ppm Vol × 0.001996
			ppm Wt × 0.001335

Note: The ozone concentration conversion factors are based upon a standard temperature of 20°C and standard pressure of 1 atmosphere. The conversion factors may be changed and the ozone concentration may be calculated using the formulas shown below:

Ozone concentration in g/m^3 = (%Vol)(W)(10)
= (M)(W)(1000)/((M-48) + (4800/% Wt))
= (ppm Vol)(W)(0.001)
= (ppm Wt)(W)(M/48)(0.001)

where:

%Wt and % Vol are expressed as percent and not as a decimal (i.e., 1% and not .01)
M = Gram molecular weight of the feed-gas (Assumed)
  Oxygen = 32 g/mole
  Air = 29 g/mole
W = Maximum weight of ozone per unit volume of gas at standard temperature and pressure (see example calculation below)
  Example calculation at 1 atmosphere standard pressure:
    Ozone gram molecular weight = 48 g/mole
    Molar volume = 0.08205 L/mole/°K
  Therefore:
    Molar volume at temperature of 20°C = (273.15 + 20°K)* (0.08205 L/mole°K)
    = 24.503 L/mole
    W = (48 g/mole)/(24.053 L/mole) = 1.996 g/L

Table 6-5. English Unit Equivalents for Ozone Concentration

Standard Pressure	14.696 psi (101 kPa)
Standard Temperature	68°F (20°C)
Gram Molecular wt. of air	29 g/mole
Gram Molecular wt. of oxygen	32 g/mole
Gram Molecular wt. of ozone	48 g/mole
Molar Volume	0.08205 L/mole/°K

Metric	English					
Wt Ozone to Vol Gas (g/m³)	Percent Weight		Percent Volume (%)	PPM Weight		PPM
	Air (%)	Oxygen (%)		Air (ppm)	Oxygen (ppm)	Volume (ppm)
2	0.17	0.15	0.10	1,658	1,503	1,002
4	0.33	0.30	0.20	3,313	3,004	2,004
6	0.50	0.45	0.30	4,967	4,503	3,007
8	0.66	0.60	0.40	6,618	6,001	4,009
10	0.83	0.75	0.50	8,267	7,498	5,011
12	0.99	0.90	0.60	9,914	8,993	6,013
14	1.16	1.05	0.70	11,559	10,486	7,015
16	1.32	1.20	0.80	13,201	11,978	8,018
18	1.48	1.35	0.90	14,842	13,469	9,020
20	1.65	1.50	1.00	16,480	14,958	10,022
22	1.81	1.64	1.10	18,116	16,446	11,024
24	1.98	1.79	1.20	19,750	17,932	12,026
26	2.14	1.94	1.30	21,382	19,417	13,029
28	2.30	2.09	1.40	23,012	20,900	14,031
30	2.46	2.24	1.50	24,640	22,381	15,033
32	2.63	2.39	1.60	26,265	23,862	16,035
34	2.79	2.53	1.70	27,889	25,340	17,038
36	2.95	2.68	1.80	29,510	26,818	18,040
38	3.11	2.83	1.90	31,129	28,294	19,042
40	3.27	2.98	2.00	32,746	29,768	20,044
42	3.44	3.12	2.10	34,362	31,241	21,046
44	3.60	3.27	2.20	35,974	32,712	22,049
46	3.76	3.42	2.31	37,585	34,182	23,051
48	3.92	3.57	2.41	39,194	35,651	24,053
50	4.08	3.71	2.51	40,801	37,118	25,055

Table 6-6. Terminology for Calculated Ozone Parameters

Parameter	*Symbol*	*Units*	*Calculation Formula*
Ozone Production	P	g/min	$G_1 \times Y_1$
Applied Ozone Dose	D	mg/l	$G_1 \times Y_1/L_1$
Transfer Efficiency			
Precise	TE	%	$100 * [(G_1 * Y_1) - (G_2 * Y_2)]/(G_1 * Y_1)$
Approximate	TE	%	$100 * (Y_1 - Y_2)/Y_1$
Transferred Ozone Dose	T	mg/l	$Y_1 \times TE * G_1/L_1$

To accurately calculate ozone TE several parameters must be measured (See Table 6-6). However, a simplified and acceptable estimation can be obtained if the off-gas flow rate is assumed to approximate the feed-gas flow rate, as shown in Table 6-6.

The error that occurs using the approximate TE calculation is less than two percent for most design situations (i.e., actual TE is usually greater than 85 percent and off-gas flow rate is usually within 10 percent of the feed-gas flow rate). In most applications the approximate calculation is adequate and eliminates the need for installing a separate flow meter for off-gas flow measurement. However, when precise data are required, for example for verification of design specifications, the precise method of calculating TE should be used.

6.2.4.3 Auxiliary Measurements and Calculations

Other important parameters in ozone terminology are temperature, pressure, energy, power, moisture content, and dew point. Moisture content and dew point of the feed-gas are most important when air or recycled oxygen is used because the feed-gas moisture content must be reduced to very low levels in order to prevent damage to the internal components of the ozone generator. Moisture removal is typically not required when high purity oxygen is the feed-gas, because its moisture content is usually very low.

The relative moisture content of the feed-gas is called dew point temperature, or "dew point." Dew point is the temperature at which a gas (at a specific pressure condition) is saturated with water. If the temperature of the gas decreases or pressure increases from those conditions, the water will condense. Calculating dew point is an important element in ozone system design because the size of some of the air treatment equipment is a function of the feed-gas dew point temperature.

The relationship between moisture content in air and dew point temperature at standard pressure is shown in Table 6-7 (21). The data is for the range of dew point temperatures typically encountered in ozone process design. The moisture content increases dramatically as the dew point temperature increases. For example, the initial moisture content of an air feed-gas with a standard dew point of 20°C (68°F) is 17.7 g/m^3 (1.11 lb/1,000 ft^3), whereas the desired moisture content of the treated feed-gas is only 0.008 g/m^3 (0.000507 lb/1,000 ft^3) (dew point temperature of -76°F (-60°C)). The dew point temperatures shown are for a standard pressure of 101 kPa (14.7 psi), and must be adjusted for operating pressure conditions. The precise calculation for dew point and moisture content is by means of psychrometric formulas or by use of psychrometric charts (21). However, the calculation may be approximated by an inverse proportional relationship of absolute pressures (gauge plus atmospheric pressure). A summary of atmospheric pressures at different elevations is shown in Table 6-8 to assist in the determination of absolute pressure.

In ozone process design the moisture loading to the desiccant dryer should be calculated. The desiccant dryer is used to reach the required dryness of the feed-gas, and is a very important piece of equipment in the air treatment process. An example calculation of moisture loading to the desiccant dryer is shown in Example 6-1.

Example 6-1. Calculation of Moisture Loading to Desiccant Dryer

Assume that a plant is at an elevation of 915 m (3,500 ft) above sea level (atmospheric pressure is 89.1 kPa (12.93 psi)); the feed-gas flow rate is 600 scfm (17 m^3/min); the gauge pressure is 93 kPa (13.5 psig); the inlet air temperature is 30°C (86°F); the relative humidity is 60 percent; and the compressed air temperature is cooled with an after-cooler to 20°C (68°F). Determine the ambient air moisture content;

Table 6-7. Moisture Content of Air for Air Temperature from −80 to 40°C (21)

Air Temperature (°C)	Air Temperature (°F)	Moisture Weight lb H_2O/lb air	Weight[a] Volume lb/ft^3	Moisture[b] Content lb H_2O per 1,000 ft^3
−80	−112	0.0000003168	0.07526	0.00002384
−75	−103	0.0000007713	0.07526	0.00005805
−70	−94	0.000001640	0.07526	0.0001234
−65	−85	0.000003342	0.07526	0.0002515
−60	−76	0.000006743	0.07526	0.0005074
−55	−67	0.00001311	0.07526	0.0009866
−50	−58	0.00002464	0.07526	0.001854
−45	−49	0.00004455	0.07526	0.003352
−40	−40	0.00007925	0.07526	0.005964
−35	−31	0.0001381	0.07526	0.010393
−30	−22	0.0002344	0.07526	0.017640
−25	−13	0.0003903	0.07526	0.029373
−20	−4	0.0006731	0.07526	0.050665
−15	5	0.001020	0.07526	0.07676
−10	14	0.001606	0.07526	0.1209
−5	23	0.002485	0.07526	0.1870
0	32	0.003788	0.07526	0.2851
5	41	0.005421	0.07526	0.4080
10	50	0.007658	0.07526	0.5763
15	59	0.01069	0.07526	0.8045
20	68	0.01475	0.07526	1.1100
25	77	0.02016	0.07526	1.5171
30	86	0.02731	0.07526	2.0552
35	95	0.03673	0.07526	2.7642
40	104	0.04911	0.07526	3.6959

Air weight/volume is corrected to 68°F and 1 atmosphere pressure

[a] Air weight/volume = 1,205 g/m^3
[b] g/m^3 = lb/1,000 ft^3 * 16.012

Table 6-8. Atmospheric Pressure at Different Altitudes.

Altitude		Atmospheric Pressure		
Feet	Meters	psi	kPa	Atm
0	0	14.70	101.325	1.000
500	152	14.43	99.49	0.982
1,000	305	14.16	97.63	0.964
1,500	457	13.91	95.91	0.947
2,000	610	13.66	94.18	0.930
2,500	762	13.41	92.46	0.912
3,000	914	13.17	90.80	0.896
3,500	1,067	12.93	89.15	0.880
4,000	1,219	12.69	87.49	0.864
4,500	1,372	12.46	85.91	0.848
5,000	1,524	12.23	84.32	0.832
6,000	1,829	11.78	81.22	0.802
7,000	2,134	11.34	78.19	0.772
8,000	2,438	10.91	75.22	0.742
9,000	2,743	10.50	72.39	0.714
10,000	3,048	10.10	69.64	0.687

the compressed, cooled feed-gas moisture content; and the moisture loading of the feed-gas to the desiccant dryer.

Ambient air moisture content. From Table 6-7 the moisture content of the air at standard pressure and a temperature of 30°C (86°F) is 32.1 g/m^3 (2.055 lb/1,000 ft^3). This value must be adjusted for ambient pressure and relative humidity conditions, and may be approximated as follows:

2.055 lb/1,000 ft^3 * 14.70 psia/12.93 psia * 0.60 = 1.401 lb/1,000 ft^3

or

24.29 g/m^3 * 1 atm/.88 atm * .60 = 16.56 g/m^3

Compressed, cooled feed-gas moisture content. From Table 6-7 the moisture content of the feed-gas at a temperature of 20°C (68°F) is 17.8 g/m^3 (1.110 lb/1,000 ft^3). The absolute pressure is 182 kPa (26.43 psia) (i.e., 13.5 psi + 12.93 psi). The approximate moisture content is:

1.110 lb/1,000 ft^3 * 14.70 psia/26.43 psia = 0.617 lb/1,000 ft^3

The moisture content of 9.88 g/m^3 (0.617 lb/1,000 ft^3) is the maximum amount of water the feed-gas can "hold" at 20°C (68°F) and 182 kPa (26.43 psia). This moisture content is 41 percent of the moisture content of the ambient air. The remaining 59 percent of the moisture will condense out of the air stream and will be removed through a "moisture trap." Although not part of the question, the standard dew point of the compressed air can be approximated using Table 6-7 as about 10°C (50°F).

Moisture loading to the desiccant dryer is approximated as follows:

0.617 lb/1,000 ft^3 * 600 scfm * 60 min/hr/1,000 = 22.21 lb/hr

or

9.88 g/m^3 * 17 m^3/min = 167.9 g/min

6.3 Process Flow Schematics

The ozone disinfection process consists of a liquid and a gas flow scheme, as shown in Figure 6-6. The feed-gas is typically air or high purity oxygen. A small amount (from 1 percent wt in air to 4 percent wt in oxygen) of the feed-gas is converted to ozone within the ozone generator. The ozone containing feed-gas is directed to a contact basin where it combines with the wastewater. Typically, a high percentage of the ozone, but only a small percentage of the total feed-gas flow, is transferred to the liquid in the contact basin. The disinfection action proceeds in the contact basin. The gas and liquid flows separate after the contact basin. The off-gas is treated to remove the un-reacted ozone and is reused, recycled, or discharged to the atmosphere. The disinfected wastewater is discharged to the receiving water. Any residual ozone in the wastewater is reacted or is reverted back to oxygen.

6.3.1 Feed-Gas Flow Schematic

Ozone disinfection processes are typically distinguished by the type of feed-gas used. The most common types are air-fed and oxygen-fed. Oxygen-recycle systems have been used on some occasions. A diagram showing the components of these processes is presented in Figure 6-7. The air-fed system is most common in plants where oxygen is not available. Oxygen-fed systems are typically used in conjunction with an oxygen activated sludge treatment system, where the unused oxygen from the ozone disinfection system is used in the biological treatment process.

The quality of the feed-gas is critically important for the electric discharge ozone generators. The feed-gas must be oil-free, particle-free, and dry. To achieve these characteristics the air-fed (and oxygen-recycle) systems must pre-treat the gas to remove moisture and particulates, and oil, if present. The once-through, oxygen-fed system typically pretreats for particulates only, because the oil and moisture content of the high purity oxygen from the oxygen production facilities is negligible.

The ozone generation equipment and power requirements are about 50 percent lower for an oxygen-fed system than for an air-fed system because approximately twice as much ozone is produced with high purity oxygen at a given input of power to the generator. However, this lower cost of producing ozone is often offset by a higher cost of obtaining high purity oxygen to feed the generator. On the other hand, when the oxygen can also be used in the biological system the economics are more favorable using high purity oxygen.

The ozone disinfection contacting units are similar for all feed-gas systems shown in Figure 6-7. In the contact basin some disinfection action occurs through direct contact of the microorganism with the ozone in the gaseous phase; however, "effective" disinfection appears to exist only when residual oxidants or an ozone residual is present (16,17). In order to obtain residual oxidants or residual ozone, a sufficient amount of ozone must be transferred to the wastewater.

Several investigators have documented that an excellent correlation exists between the log-transferred ozone dosage (log(T)) and log-coliform survival ratio (log(N/N_o)) (17,18,22-25). Stover (17) reported

Figure 6-6. Ozone disinfection process gas and liquid flow diagram.

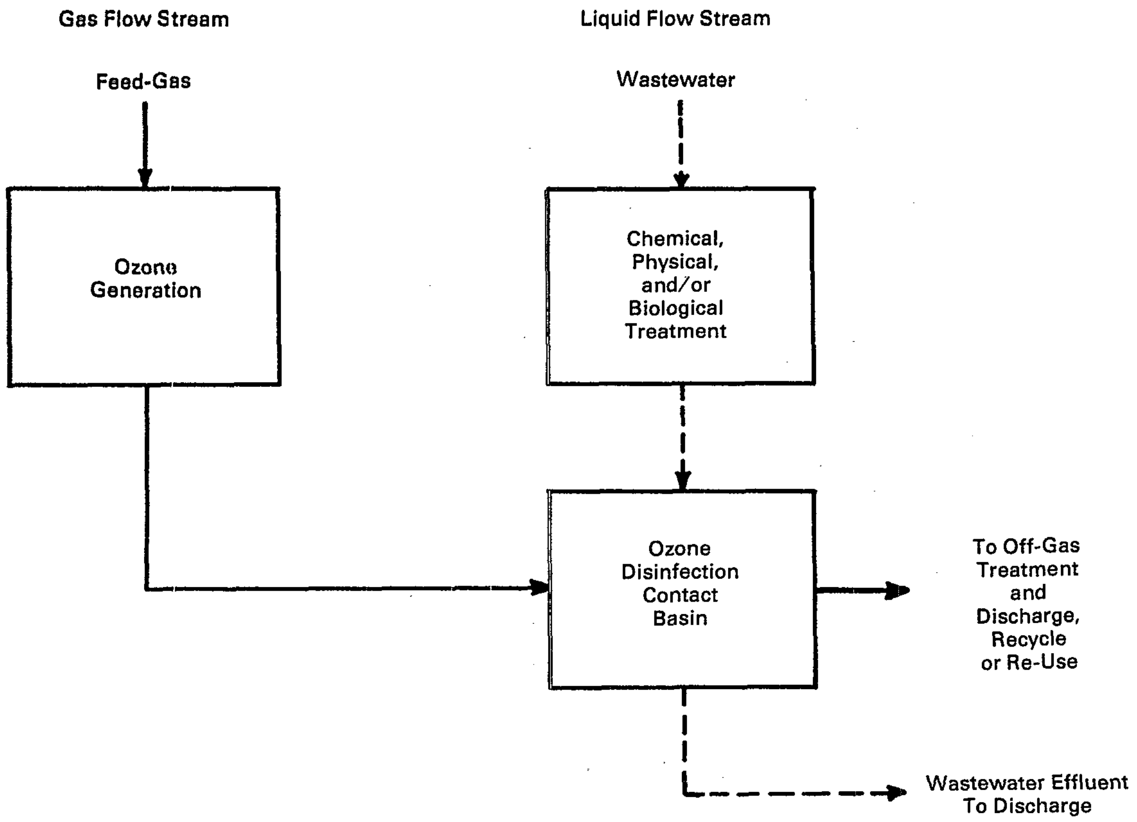
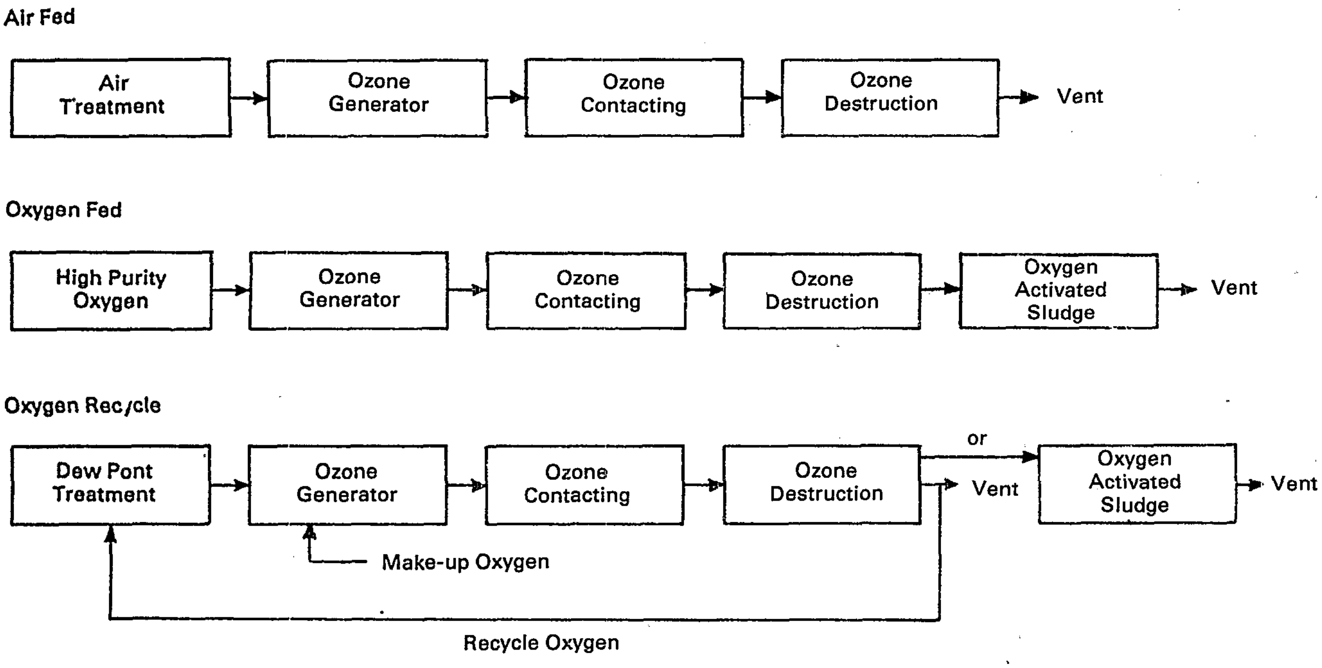


Figure 6-7. Diagrams showing feed-gas flow of typical ozone disinfection processes.



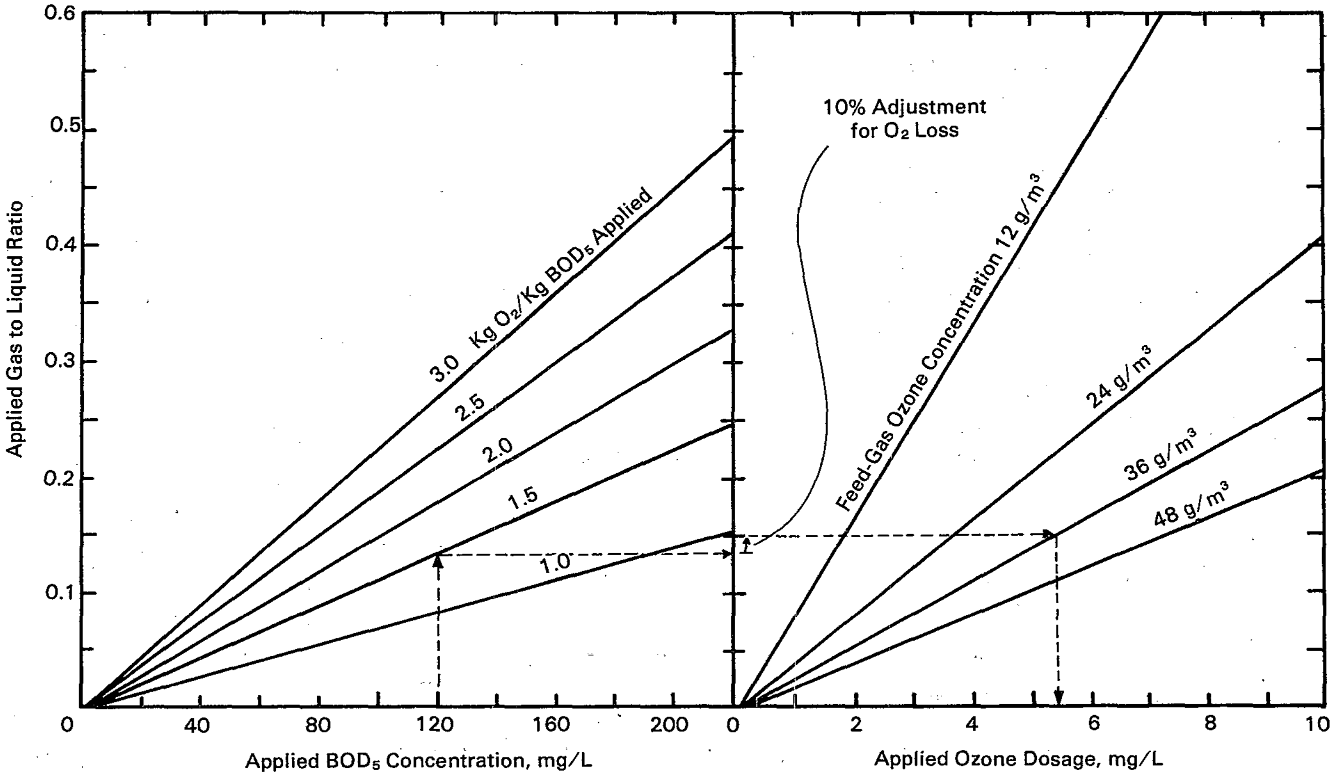
that this relationship exists irrespective of the type of feed-gas used; thus, the decision to use air or oxygen as the feed-gas is typically based on economic considerations of providing the oil-free, particle-free dry gas to the generator and not on the disinfection performance capability associated with the feed-gas.

A common application for oxygen-fed ozone disinfection is in conjunction with an oxygen activated sludge process where the oxygen is reused. Only about 10 percent of the oxygen in the feed-gas is "used-up" in the ozone process; thus, 90 percent or more is available for the biological treatment process. The major design consideration is to balance the oxygen requirements of both processes. This can be addressed by balancing the oxygen gas (G_1) to wastewater liquid (L_1) flow rates to both processes.

An example approach to balancing the gas to liquid ratios to both processes is shown in Figure 6-8. The first step is to estimate the applied BOD concentration to the oxygen activated sludge process. The activated sludge oxygen gas to wastewater liquid ratio can then be approximated by estimating the applied oxygen requirement. The example in Figure 6-8 shows that the activated sludge G_1/L_1 ratio is 0.135 for an applied BOD concentration of 120 mg/l and an oxygen requirement of 1.5 kg O_2/kg BOD. The example also shows that the activated sludge gas to liquid ratio balances with the ozone process G_1/L_1 ratio at a feed-gas ozone concentration of 40 g/m^3 (3 percent wt) and an applied ozone dosage of 6 mg/l, assuming 10 percent loss in the ozone contactor.

The G_1/L_1 ratio required for the ozone system is shown as 10 percent greater than required for the activated sludge process, because about 10 percent of the oxygen is used up in the ozone process through transfer of ozone and oxygen to the wastewater. For the example in Figure 6-8, the adjusted G_1/L_1 ratio for the ozone system is 0.15. Knowing the G_1/L_1 ratio for ozone, either the applied ozone dosage can be determined for a given feed-gas ozone concentration, or the required feed-gas ozone concentration can be estimated for a given applied ozone dosage. If the oxygen balancing analysis indicated that significantly more oxygen was required for ozone disinfection because the ozone concentration was too high, then an oxygen recycle process could be provided. If the analysis indicated that more oxygen was required for the activated sludge process, then an oxygen bypass pipe around the ozone disinfection system could be provided.

Figure 6-8. Oxygen requirement for ozone disinfection compared to oxygen requirement for activated sludge.



When oxygen is not used in the biological portion of the wastewater treatment process, then the cost of an air-fed ozone process may be compared with the cost of a total oxygen-recycle system to determine the most economical alternative. The main disadvantages of the oxygen-recycle system were the cost of obtaining oxygen-rich feed-gas and the additional attention required to handling an oxygen-enriched gas. The main advantages were the reduced cost of the ozone generation equipment and the reduced energy consumption for ozone generation. Most oxygen activated sludge plants in the United States that use ozone use the once-through, oxygen-fed system. As indicated in Table 6-1, all wastewater treatment plants that use ozone disinfection, but do not use oxygen activated sludge, use the air-fed system.

The ozone destruct units shown in Figure 6-7 are fundamentally the same for all feed-gas systems. The size of the ozone destruct unit is based on the volume of gas directed to the unit.

6.3.2 Wastewater Flow Schematic

Ozone will disinfect to a very high degree if a sufficient amount of ozone is applied to the wastewater and a properly designed disinfection contact basin is provided (17). However, there appears to be economic considerations for the ozone disinfection process associated with the type of wastewater treatment processes selected prior to the ozone disinfection system. The main effect of the wastewater treatment processes used is the impact on the amount of ozone required to obtain the desired disinfection level.

Wastewater treatment schemes prior to ozone disinfection that have been evaluated by various researchers are shown in Figure 6-9 (17,24,25). These schemes range from simple "fine screening" to biological systems followed by chemical treatment. Given and Smith et al. (24) evaluated ozone disinfection using the effluent from four types of wastewater treatment process schemes, including fine screening effluent, rotating biological contactor effluent, anaerobic lagoon effluent, and a "strong" waste flow stream. The coliform survival results obtained over a range of transferred ozone dosages is shown by the dose/response curves in Figure 6-10.

The major factor influencing the amount of ozone required to achieve a desired reduction in fecal

Figure 6-9. Example treatment schemes using ozone disinfection (17,24,25).

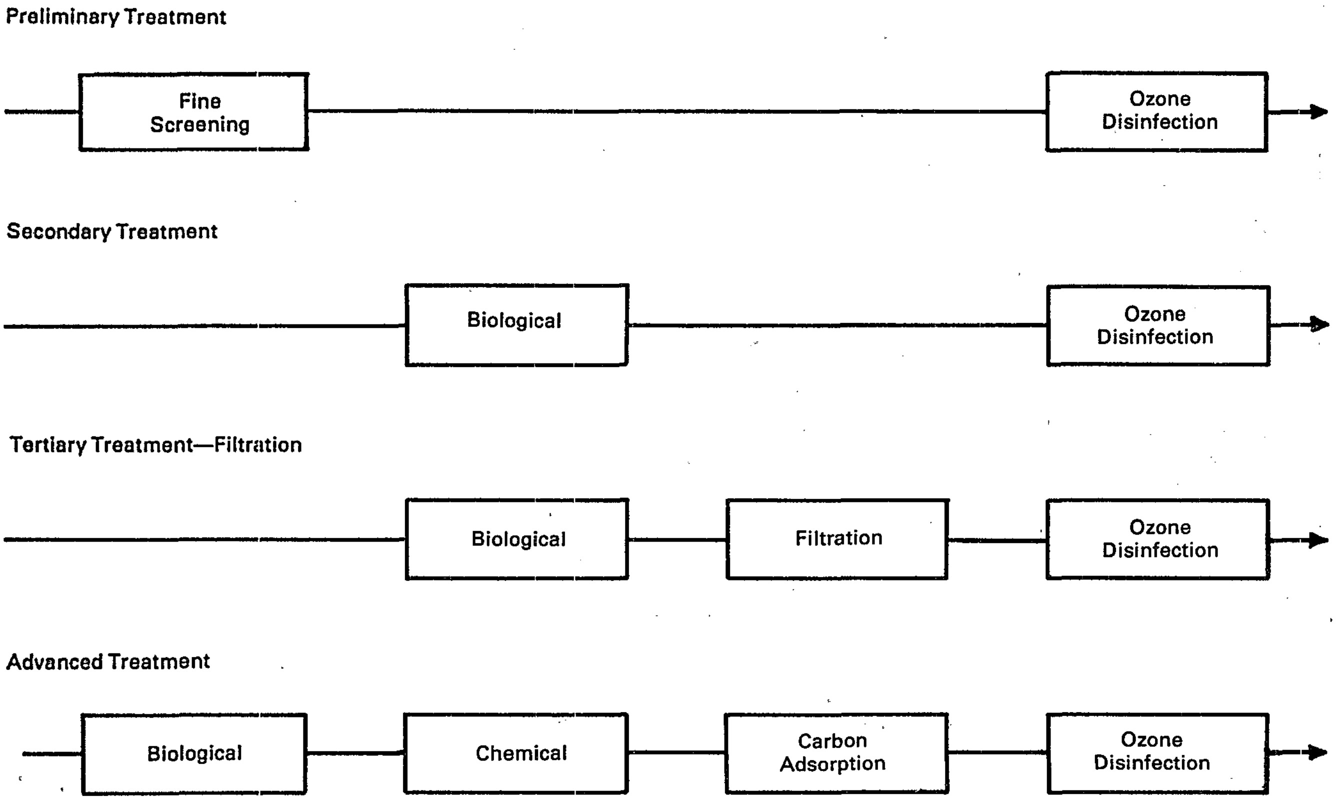
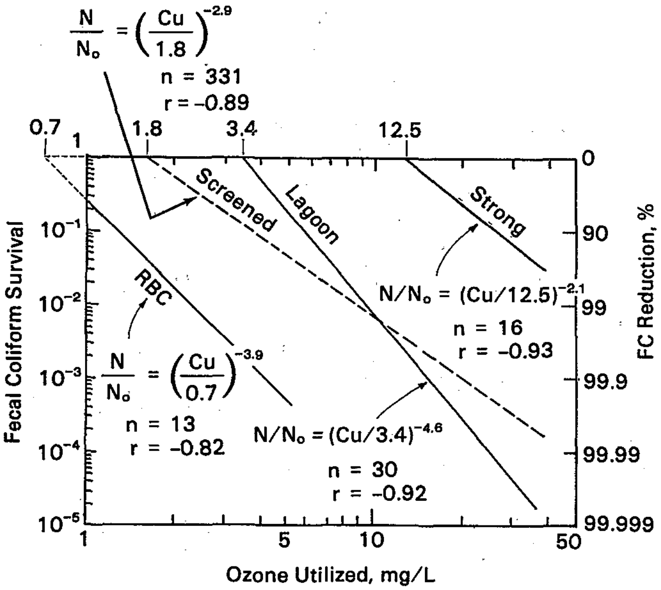


Figure 6-10. **Fecal coliform survival for rotating biological contactor effluent, screened effluent, anaerobic lagoon effluent, and strong wastewater (24).**



Initial Conditions

	Type of Waste			
Parameter	RBC	Screened	Lagoon	Strong
FC	6.5×10^4	1.3×10^6	3.8×10^5	2.8×10^6
TEMP	8.9	8.6	6.9	5.4
TURB	5	42	67	151
pH	7.9	7.8	7.6	—
BOD	13	95	92	—
SS	13	102	121	1010
VSS	8	57	47	620

coliform concentration was the "initial ozone demand" of the wastewater, which was 0.7, 1.8 , 3.4 , and 12.5 mg/l for the RBC, screened, anaerobic lagoon and "strong" effluents, respectively (24). The slopes of the dose/response curves varied from -2.9:1 to -4.6:1. Some difference in the slope may be due to the type of wastewater treated, but the data set for all wastes except the screened waste was too limited (13 to 16 data points) to confirm a cause for the variation in slope.

Gan et al. (25) evaluated ozone disinfection performance of an activated sludge effluent and three tertiary treatment schemes following activated sludge treatment including: coagulation, sedimentation, and filtration (Scheme A); coagulation and filtration (Scheme B); and carbon adsorption (Scheme C). The carbon adsorption data set was further divided into Scheme C1 and Scheme C2. Scheme C1 had a nitrite-nitrogen concentration greater than 1.0 mg/l and Scheme C2 a nitrite-nitrogen concentration less than 1.0 mg/l. The major factors influencing the amount of ozone required for disinfection were the influent dissolved chemical oxygen demand (DCOD), the influent nitrite-nitrogen concentration, and the target or desired effluent coliform concentration. The effect of the target effluent coliform concentration was evaluated by comparing the ozone dosage required to meet the former EPA standard of 200 fecal coliforms per 100 ml and the more stringent California standard of 2.2 total coliforms per 100 ml. The results are shown in the bar graph in Figure 6-11.

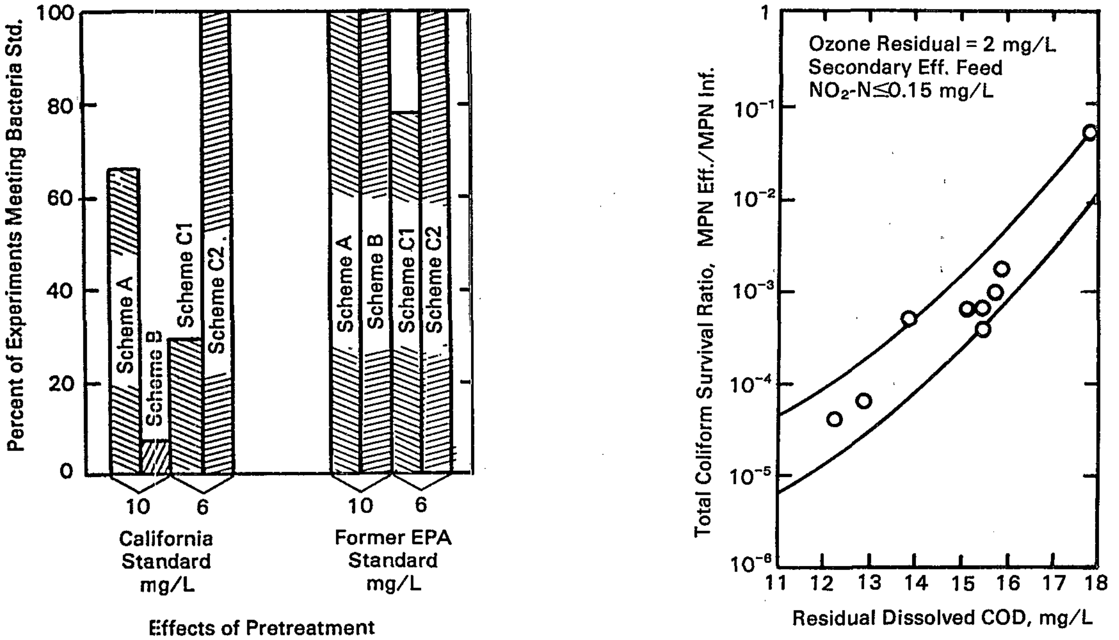
The carbon adsorption effluent met the more stringent California standard 100 percent of the time at an ozone dosage of 6 mg/l. The other pre-treatment effluents did not meet this high coliform standard even at ozone dosages of 10 mg/l. The major effect of carbon adsorption pre-treatment was to reduce the DCOD concentration of the wastewater and thus reduce its ozone demand.

The less stringent disinfection standard of 200 fecal coliform/100 mL was achieved 100 percent of the time for all wastewater treatment schemes evaluated, except when the nitrite-nitrogen concentration was greater than 1.0 mg/l (Scheme C1). The high nitrite-nitrogen concentration of the wastewater influent to the contact basin was caused by an upset in the activated sludge plant. The ozone reacted with the nitrite causing an increase in ozone demand and a significant reduction in disinfection performance (25).

The specific relationships between disinfection performance and the DCOD and nitrite-nitrogen concentrations are shown in Figure 6-11. At a given ozone residual in the contact basin effluent of 2.0 mg/l, the best disinfection performance was achieved with an influent nitrite-nitrogen concentration less than 0.15 mg/l and a DCOD concentration less than 12 mg/l. Based on these results, to keep the ozone dosage less than 10 mg/l, tertiary treatment of the wastewater may not be necessary to meet the former EPA standard (25). However, tertiary treatment to reduce the ozone demand due to DCOD and nitrite-nitrogen appears necessary to meet the stringent California standard of 2.2 total coliforms per 100 ml.

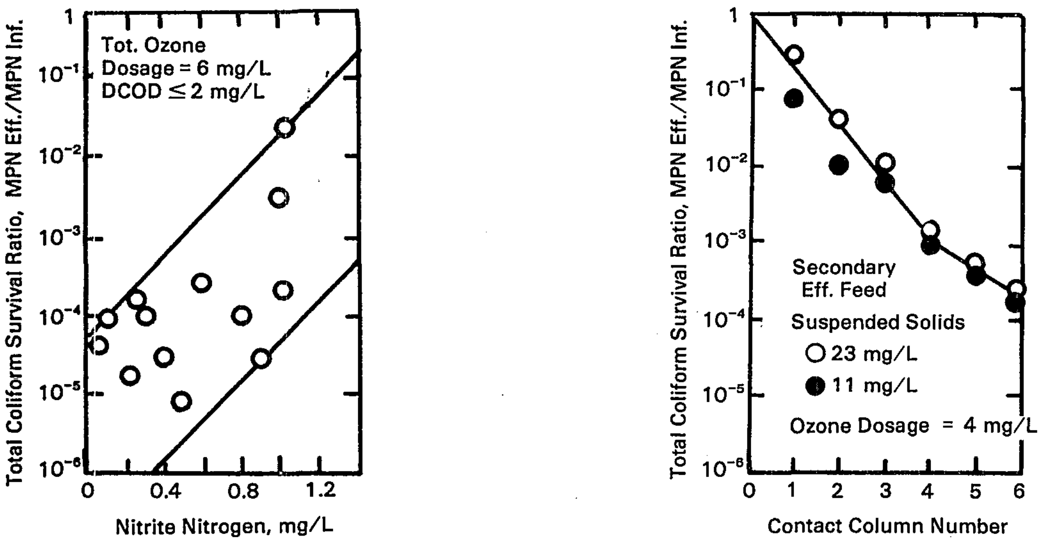
Stover et al. (17) also evaluated the ozone dosage requirements to meet two different disinfection standards of 2.2 and 70 total coliforms per 100 ml. The evaluation was completed for three different wastewater treatment schemes consisting of a filtered activated sludge secondary effluent, a nitrified effluent, and a filtered nitrified effluent. The most significant factors affecting the dosage requirement were the organic quality of the wastewater as

Figure 6-11. Effect of water quality and performance criteria on ozone dosage requirement (25).



Effects of Pretreatment on Meeting Coliform Bacteria Standards

Effect Residual DCOD After Ozonation on Total Coliform Survival Ratio



Effect of Nitrite-Nitrogen on Disinfection with Ozone

Comparison of Disinfection Performance Between the High and Low Suspended Solids Effluent

measured by COD concentration, the target effluent coliform concentration and the influent coliform concentration. The magnitude of the effect of these factors is illustrated in Figure 6-12, where the effluent total coliform concentration is shown to decrease as the transferred ozone dosage increases.

Stover et al. (17) evaluated the effect of the organic concentration of the wastewater influent to the contact basin by comparing the transferred ozone dosage requirement for secondary and nitrified effluents to obtain similar effluent total coliform concentrations. For both disinfection standards evaluated, significantly more transferred ozone dosage was required for the filtered secondary effluent (COD of 40 mg/l) than was required for the nitrified effluents (COD of 20 mg/l).

Stover also showed that a more stringent effluent total coliform concentration standard caused a significant increase in the transferred ozone dosage required. For example, Figure 6-12 shows that for the filtered nitrified wastewater treatment scheme five times more transferred dosage was required to meet the 2.2 per 100 ml standard (15 mg/l) than was required to meet the 70 per 100 ml standard (3 mg/l).

Stover analyzed the effect of the influent total coliform concentration by comparing the disinfection results for the nitrified and filtered nitrified effluents. The data in Figure 6-12 show that less ozone was required to reach a given effluent coliform concentration for the filtered nitrified effluent than was required for the nitrified effluent. However, these data do not consider the fact that the influent total coliform concentration of the filtered, nitrified wastewater was lower. The influent total coliform concentration was incorporated in the analysis in Figure 6-13, where the log reduction of total coliform (Log N_o/N) is plotted against the log transferred ozone dosage. The results indicate no difference in performance. Stover's conclusion was that the benefit of filtration in removing total coliform may be a more significant factor in reducing the ozone dosage requirement than is the benefit of removal of suspended solids, at least within the small range of suspended solids tested (average 2.5 mg/l for the filtered nitrified and 5.7 mg/l for the nitrified effluents).

Gan et al. (25) also addressed the subject of filtration for suspended solids removal prior to an ozone disinfection process. The results of Gan's comparison between a high (23 mg/l) and a low (11 mg/l) suspended solids influent to the ozone disinfection contact basin are shown in Figure 6-11, where total coliform survival is shown as a function of contact column or stage in the disinfection contact basin. As shown, the total coliform survival ratio was lower (i.e., better kill of coliforms) when the suspended solids concentration was lower. However, the effect of suspended solids diminished as disinfection progressed through the six-stage contact basin. The effect of suspended solids was greatest in the first two stages, and minimal effect was evidenced by the sixth stage. *Gan concluded that the removal of* suspended solids (via flotation) that occurred in the first two stages "indicates that removal of suspended solids prior to ozonation may not be essential to achieve disinfection of wastewater."

Figure 6-12. Total coliform concentration versus transferred ozone dosage for various effluents (17).

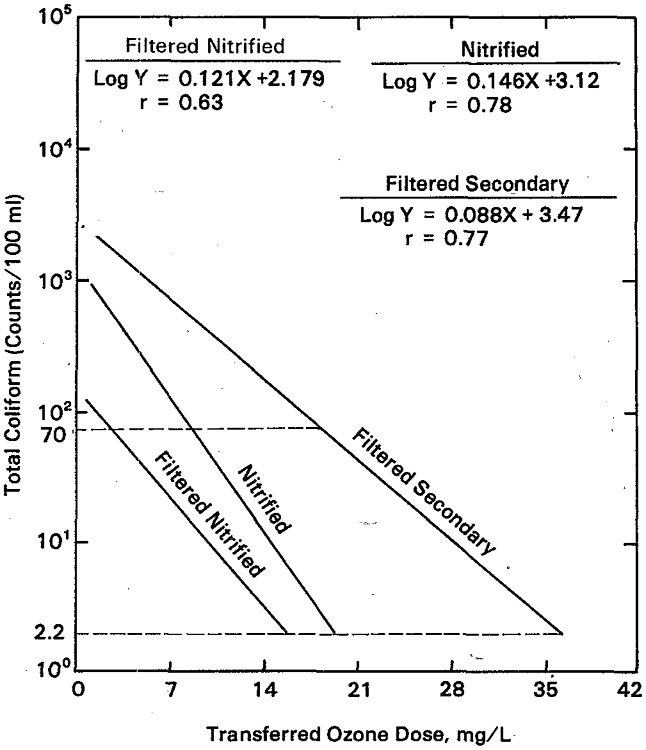
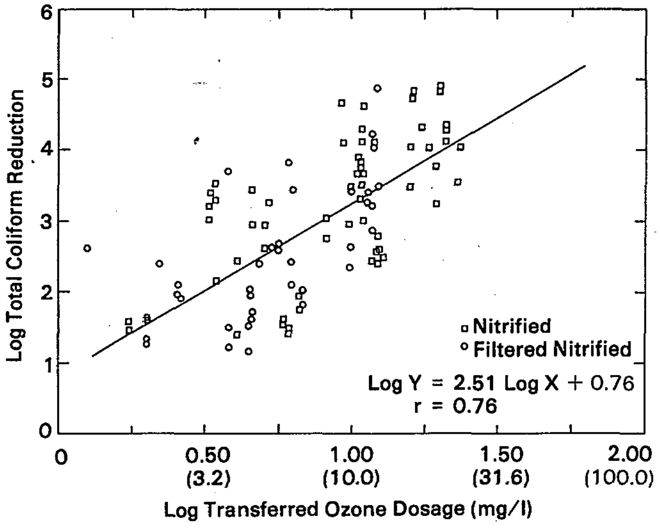


Figure 6-13. Total coliform reduction versus log transferred ozone dosage for nitrified effluents (17).



Venosa et al. (26) also studied the effect on disinfection of filtered and non-filtered secondary effluent. The data indicated that improved disinfection efficiency occurred as a result of filtration, but was more related to the reduction in total chemical oxygen demand (TCOD) than to removal of TSS. The conclusion was that filtration to remove TSS may not be necessary when the TCOD concentration is low.

The EPA Water Engineering Research Laboratory has also evaluated several secondary treatment plant effluents to determine the relationship between ozone dosage and total coliform reduction (27). The most significant factor influencing the ozone dosage requirement to achieve a desired effluent total coliform concentration was the TCOD concentration of the effluent. At five plants where the TCOD of the secondary effluent was less than 40 mg/l, a total coliform concentration of 1,000 per 100 ml could be achieved with ozone dosages between 4 and 7 mg/l. However, when Meckes et al. (27) evaluated a plant which treated a significant amount of industrial waste (TCOD = 74 mg/l), a dosage greater than 12 mg/l was projected in order for the process to meet the 1,000 total coliforms per 100 ml limit.

Based on the data presented above, it appears that there is no technical basis for excluding the use of ozone following any treatment scheme. However, depending upon the type of wastewater treated and/or the effluent disinfection requirement, the wastewater treatment scheme may be an important economical consideration. A summary of the issues to consider when selecting the liquid flow schematic prior to ozone disinfection is presented below:

1. Required Effluent Target
   - To meet the former EPA standard of 200 fecal coliforms per 100 ml, tertiary treatment may not be necessary.
   - To meet more stringent standards, such as 14 fecal coliforms per 100 ml, tertiary treatment should be considered.
   - To meet a standard of 2.2 total coliforms per 100 ml, advanced treatment unit processes prior to the ozone disinfection process may be required.
2. Influent Coliform Concentration
   - Coliform removal is a function of transferred ozone dosage; thus, the influent coliform concentration will affect the amount of ozone dosage required to meet specific effluent criteria.
   - Treatment processes that reduce the influent coliform concentration (such as filtration) will decrease the ozone dosage required to acheive a specific effluent standard.
3. Wastewater Quality Characteristics
   - The ozone demand of the wastewater significantly increases the ozone dosage requirements. A plant with a large industrial contribution may have a large ozone dosage requirement. Pilot testing to establish ozone dosage requirements in these plants is highly recommended.
   - Incomplete nitrification and a high concentration of nitrite-nitrogen will significantly increase the ozone demand and thus the ozone dosage requirement. The nitrite-nitrogen concentration preferably should be less than 0.15 mg/l to optimize disinfection performance.

6.4 Ozone Equipment Design Considerations

Ozone generation is an established process, but its use in wastewater disinfection is relatively new. In this section the current state-of-the-art equipment design considerations to develop an ozone disinfection system that consistently achieves desired levels of performance is discussed.

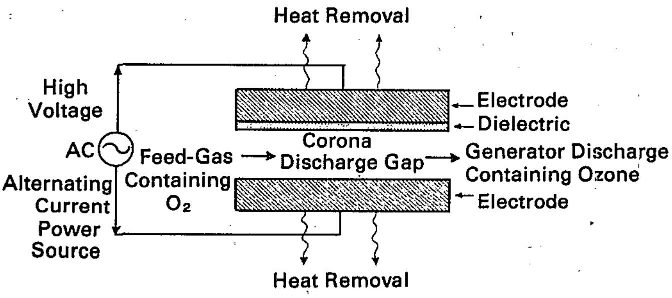
6.4.1 Ozone Generation Equipment

The basic components that comprise an ozone generator are depicted in Figure 6-14. The components include an electrical source that supplies high voltage, alternating current across a discharge gap where the oxygen containing feed-gas passes; a dielectric material that prevents electrical short-circuiting; and a heat removal mechanism that prevents rapid decay of the ozone molecule back to an oxygen molecule. Heat removal is required because 85 to 95 percent of the electrical energy supplied to the ozone generator produces heat (1,7).

6.4.1.1 Ozone Generation Theory

Electrical power used to generate ozone is received from a voltage regulator, and in some generators a frequency regulator is also used. The altered current

Figure 6-14. **Cross-section view of principal elements of a Corona discharge ozone generator.**



is supplied to a number of ozone generation cells connected in parallel, as shown in Figure 6-15. Each cell acts as a capacitor, as illustrated in Figure 6-15b. The capacitance of the cell is a function of the width of the gas space and the electrical conductivity of the dielectric material. Because the dielectric material (glass or ceramic) is the major component of the cell, the ozone generating cell is often called a "dielectric."

Ozone is produced when the ionization potential of the dielectric is reached. The system's electrical characteristics before and after the ionization potential is reached are quite different. Prior to the dielectric reaching its ionization potential the voltage is insufficient to allow an electric discharge. When the ionization potential is reached a flow of electrons (corona) will occur across the discharge gap and the electrical circuit will be completed. The minimum voltage required to meet the ionization potential of a dielectric is about 10,000 volts (1,7,28).

When the voltage is greater than the ionization potential of the dielectric the electrons travel from one electrode to the other electrode within the discharge gap and collide with the oxygen molecules in their path. Upon collision the reactions noted in Figure 6-16 occur. The number of ozone molecules formed will vary from none at all to a maximum of two for every free electron discharged. The number formed is highly dependent upon the temperature of the ozonized gas. At higher temperatures the ozone rapidly decomposes back to oxygen. A major consideration in the design of an ozone generation system is cooling of the ozone generator.

An alternating electrical current must be used in ozone generation; thus, the voltage will cycle above and below the ionization potential of the cell (1). Ozone production occurs when the voltage is greater than the ionization potential. Ozone production is terminated in the portion of the cycle when the voltage is below the ionization potential, as shown in Figure 6-17. The amount of time ozone is formed is dependent upon the frequency of the power supply.

The number of free electrons discharged is a function of the applied, peak voltage and its electrical frequency. If the peak voltage is significantly higher than the ionization potential of the dielectric, a proportionately greater number of electrons will be released. The result is an increase in the ozone production rate if the temperature of the ozonized gas is acceptable. Similarly, an increase in the frequency of the power supply increases the amount of time that the ionization voltage will be reached, which also increases the ozone production rate if the temperature of the ozonized gas is acceptable.

Figure 6-15a. Schematic diagram of a typical power supply to an ozone generator (1).

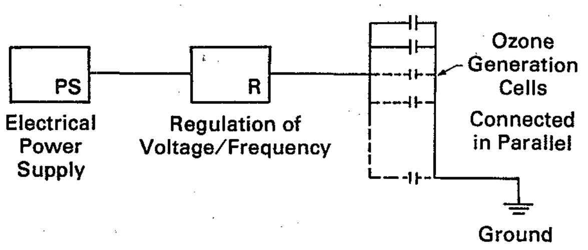


Figure 6-15b. Schematic diagram of an ozone producing cell, a "Dielectric" (1).

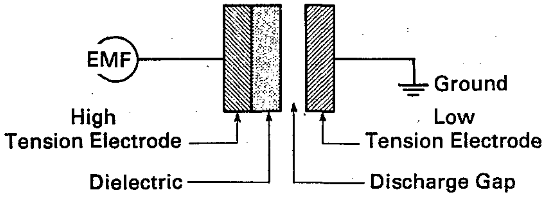
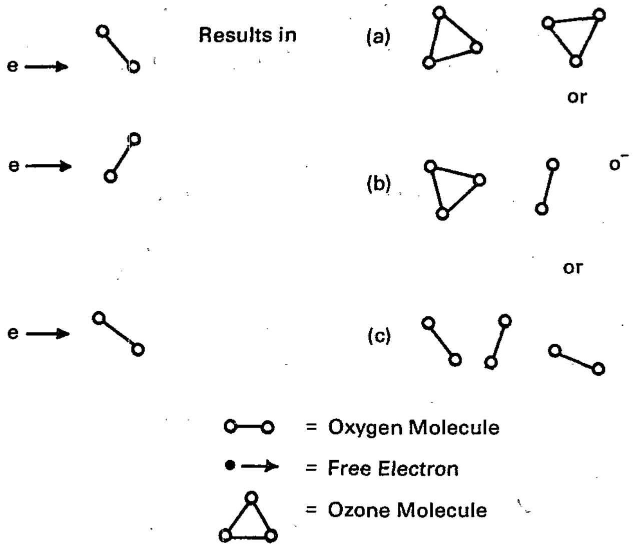


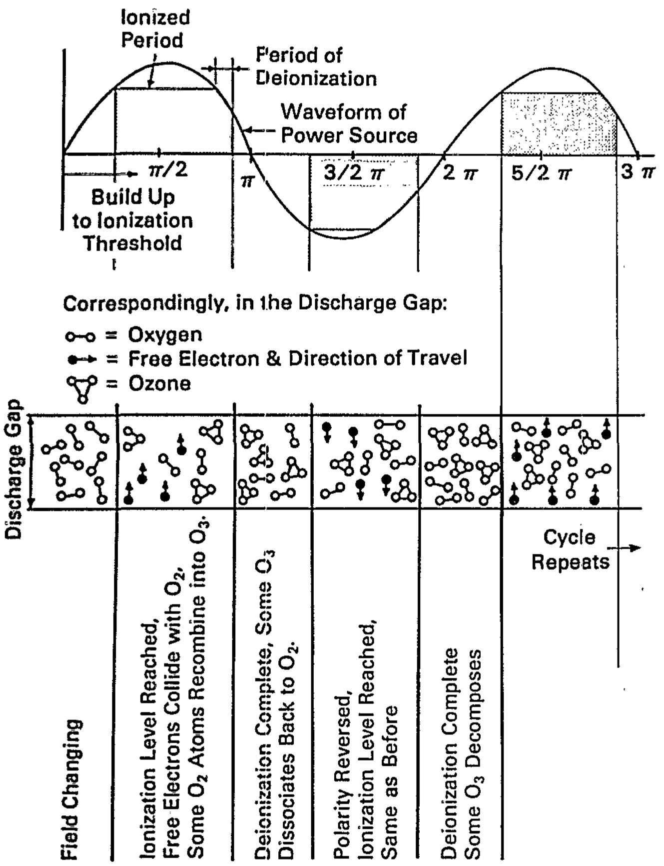
Figure 6-16. A free flow of electrons in the discharge gap causes various reactions with the oxygen molecule (1).



6.4.1.2 Design Considerations for Ozone Generation Equipment

The amount of ozone produced by an ozone generator is affected by the physical characteristics of the equipment, the power supply to the generator; the

Figure 6-17. **Ozone formation occurs when the voltage level is sufficient to create a free flow of electrons within the discharge gap (1).**



moisture content and dust content of the feed-gas; the temperature of the ozonized gas, and the feed-gas oxygen content. Each factor is further discussed.

Physical Characteristics of Ozone Generator. The relationship of the factors affecting the ozone production rate of the generator dielectric is shown in Equation 6-2 (7).

$$P \sim\sim 4k\ f\ Va\ [Cd(Vo - Va) - VaCd] \qquad (6\text{-}2)$$

where:

P = Ozone production
Va = Voltage applied to the discharge gap
Vo = Extreme or peak value of the applied voltage
Cd = Capacitance of dielectric
Ca = Capacitance of the discharge gap
f = Frequency of current supply
k = Constant

The most important point in the equation is that the ozone production rate increases with frequency and the square of the voltage. Frequency or voltage changes are used to adjust the ozone production rate for a given ozone generator. The production rate also increases with a greater number of dielectrics in service. Because of these relationships, almost any desired production rate can be supplied by the ozone equipment manufacturers.

The dielectric constant and the size of the discharge gap (typically 2 to 3 mm) between the high and low tension electrodes also affect the ozone production rate (7). These factors are typically defined by the individual ozone generator manufacturer and are proprietary items. The dielectric constant and width of the discharge gap may vary slightly from generator to generator, even with generators of the same manufacturer. A very slight change in either parameter has a large influence on the ozone production rate. As such, the ozone production rate may vary slightly for similar sized ozone generators, and from generator to generator.

Ozone Generator Power Supply. Three primary ways in which the power is supplied to the ozone generator are shown in Figure 6-18 (1). Method A is a low frequency (typically 60 hertz), variable voltage system. This system is most common because high voltage transformer technology preceded the technology in high frequency transformers. Method B is a medium frequency (up to 600 hertz), variable voltage power supply system. This process has been used to increase the production rate of installed ozone equipment (1). Method C is a variable high frequency, constant voltage unit. Method C is used by various ozone generator manufacturers in lieu of the variable voltage process. No one method has a clear advantage over the other.

The most common ways in which voltage and frequency to the ozone generator are controlled are shown in Figure 6-19 (1). The voltage controlled system uses single phase power, which will cause an imbalance in the amperage of a three-phase power supply system unless three or multiples of three ozone generators are operating simultaneously. However, this capability is seldom available; thus, the impact of an unbalanced electrical load on the overall plant three-phase power supply should be evaluated. To balance the electrical load a Scott transformer is typically used (29).

The power supply to the ozone system is typically at a frequency of 60 cycles and potential of 480 volts. For a variable voltage system the voltage may be increased by applying the line current to one of a series of tappings on the primary side of the potential transformer, thereby changing the transformer ratio and the resultant secondary voltage (1). Another method is to use a variable autotransformer to feed

Figure 6-18. **Schematic diagram of three power supply systems typically used for ozone generation (1).**

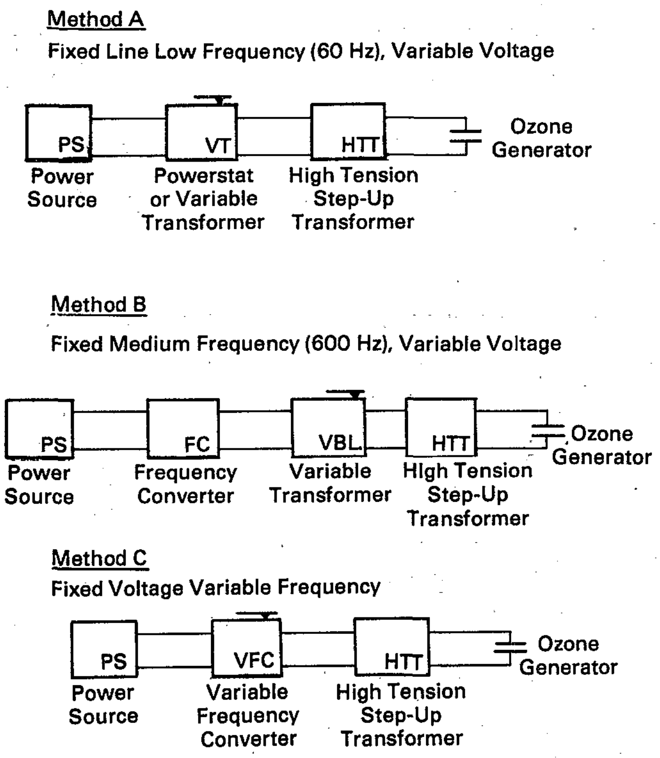
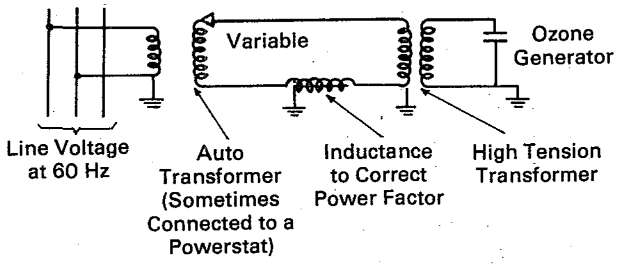
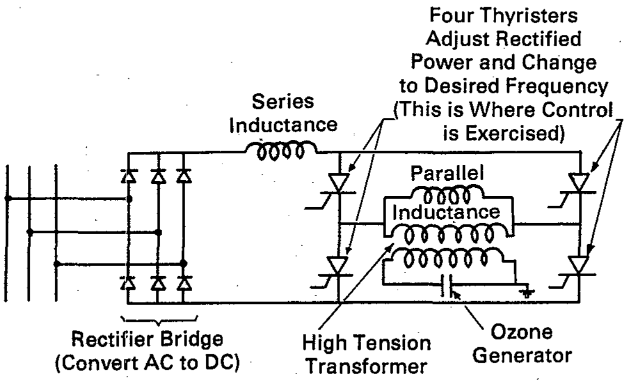


Figure 6-19. **Typical ways for varying voltage and frequency to an ozone generator (1).**



Variable Voltage



Variable Frequency

the primary side of the main transformer (1). When a variable frequency system is used both the frequency and voltage must be increased. The voltage must be increased to a level above the ionization potential of the dielectric (about 10,000 volts). The frequency is increased up to a maximum of 2,000 hertz (1).

Because the ozone generator uses high voltage and in some cases also high frequency electrical current, special electrical design considerations must be implemented. For example, special insulation must be provided for the electrical wire; a cool environment for the high voltage transformers should be provided; and the electrical transformers should be protected from ozone contamination due to minute ozone leaks that could occur on a periodic basis.

The electrical considerations for an ozone system should receive special attention. For example, a number of problems have been reported with dry-type potential transformers (1). Oil-cooled transformers apparently have performed more reliably. In view of the dependence of ozone generation on high frequency or high voltage electrical energy, the ozone generator supplier should be responsible for designing and supplying the electrical subsystems. However, the specifications should require that the frequency and voltage transformers be high quality units designed for ozone service. The ozone generator supplier should be requested to provide a record of successful electrical equipment performance.

Another item to consider in the design of the ozone generation system is power factor. An operating ozone generator can decrease the power factor to 0.3 to 0.5, depending on the generator setting (29). Corrections will normally be cost effective, since utilities that supply electrical power typically impose penalties for a low power factor.

Power factor is the ratio between the apparent power (kW) measured by a watt-hour meter and actual power (kVA) measured in terms of voltage and amperage. This relationship is shown in Equation 6-3.

$$\text{Power Factor} = \text{Apparent power}/\text{Actual power}$$

$$pf = kW/kVA \tag{6-3}$$

The power factor is UNITY when the voltage and current of an alternating current power supply are "in-phase" with each other, for example in a purely resistive circuit like a heating element. In a purely capacitive circuit, like an ozone generator that has not reached the ionization potential of the dielectric, the voltage and current are 90 degrees out of phase. In this case the power factor is ZERO. For a generator producing ozone the voltage and amperage will be somewhere between 0 and 90 degrees out of phase; thus, the power factor will be less than 1.0. The actual power factor will vary depending on the power supply to the ozone generator and the amount of electrical resistance developed within the electrical circuit.

The power factor may be corrected by installing inductors in the electrical circuit or by using the inductance created by the operation of motors in other areas of the treatment plant. However, caution must be exercised when using other plant equipment for power factor correction, because of the variable operating conditions of the equipment from hour to hour and from day to day. The inductance of other equipment should only be used to control the low power factor of the ozone generator when consistent equipment operation can be assured.

Moisture and Dust Content of the Feed-Gas. The moisture content of the feed-gas to the ozone generator has a two-fold influence on ozone production. A high moisture level not only decreases the ozone production rate, it also increases the rate of contamination of the ozone generator dielectrics. Contamination occurs whether the feed-gas is high purity oxygen or air. If oxygen is the feed-gas, hydrogen peroxide is formed in the presence of water vapor and forms deposits on the dielectrics; these deposits can be removed by scrubbing with soapy water (7).

If air is the feed-gas, about one mole of nitrogen pentoxide (N_2O_5) will develop for every one hundred moles of ozone formed (30). Nitrogen pentoxide can decompose to nitrogen dioxide (NO_2), which interferes directly with ozone output (30). When nitrogen pentoxide is in the presence of water vapor, nitric acid will form (7). The nitric acid will be deposited on solid surfaces inside the ozone generator and piping, and will enhance corrosion of metal surfaces (7). In addition, the nitric acid can create a heat sink on the glass or ceramic dielectrics and will increase the potential for dielectric breakage.

In normal practice (i.e., air dew point less than -50°C (-58°F)), about 3g to 5g of nitric acid are formed for every 1,000 g of ozone formed (30). Nitric acid can be easily removed by scrubbing with soapy water. This task should be considered as routine maintenance on a 12-month interval. Significantly greater quantities of nitric acid will form when the dew point temperature of the air is higher than -50°C (-58°F). If a malfunction occurs and excessive moisture is discharged to an operating ozone generator, the dielectrics must be cleaned before the generator is placed back into service.

A high moisture content of the feed-gas not only causes damage to ozone generator dielectrics and increased generator maintenance, but also will decrease the ozone production rate at a given power setting of the generator. Ozone production begins to decrease when the dew point temperature exceeds -51°C (-60°F)(31,32). To prevent the negative effects of moisture content on ozone production and generator maintenance, several authors suggest a minimum dew point temperature of the feed-gas.

Manley and Niegowski (7) suggest that the feed-gas moisture content should be less than 0.02 to 0.03 grams of water per cubic meter of air (i.e., dew point temperature less than -53°C (-63°F)). Varas et al. (33) recommend that the dew point temperature be less than -60°C (-76°F). Robson (34) suggests that less maintenance and prolonged dielectric life occurs when the dew point temperature of the feed-gas is -60°C (-76°F) or lower. Gerval (35) states that the "highest value to be guaranteed for dew point should be -50°C (-58°F)." Chapsal (36) recommends that the dew point temperature range between -50 and -80°C (-58 and -112°F). Damaz (37) recommends a dew point of -60°C (-76°F) to protect the generator equipment.

Based on these sources it appears that the dew point temperature should not be higher than -50°C, in order to achieve maximum output from the ozone generator. However, in order to reduce maintenance requirements and prolong dielectric life, a dew point temperature equal to or less than -60°C appears warranted. To achieve either dew point condition, extensive feed-gas treatment is required for air-fed and oxygen recycle ozone generation systems.

Damaz (38) presents other design considerations to prevent moisture contamination of the ozone generator. Moist air back-flow from the contact basin should be avoided by installing a leak-proof check valve in the piping from the ozone generator to the ozone contact basin. Also, the ozone generator must be purged with dry feed-gas prior to start-up of the ozone generator. Most operators purge with dry feed-gas for a minimum of eight hours prior to start-up of the ozone generator. It should be noted that this time period does not include the time required to obtain the dry feed-gas from the dessicant dryers.

Dust and organics in the feed-gas can also create operating problems with the ozone generator (35). The dust can collect on the dielectrics, decrease generator efficiency, increase dielectric stress and cause unnecessary dielectric breakage. A filter should be installed prior to the ozone generator to capture the dust and organics. According to Gerval (35), the filter(s) should be able to remove 99 percent of the particles greater than 1 micron in diameter and 98 percent of the particles greater than 0.4 micron in diameter. Rakness et al. (18) also recommend two filters in series, the first a 1 micron filter and the second a 0.3 micron filter. It is emphasized that filters should be installed for oxygen as well as air feed-gas systems.

Ozone Generator Cooling. Ozone production increases when the temperature of the ozonized gas is minimized. In addition, reduced heat build-up will increase life expectancy of the dielectrics. The feed-gas flow rate and the heat removal capability of the ozone generator are factors influencing the temperature of the ozonized gas. The feed-gas flow rate is typically established by the design engineer through the selection of the maximum ozone concentration. The heat removal capability of the ozone generator is governed by the ozone generator equipment manufacturer and by the temperature of the cooling medium.

Because cooling is a major aspect of the energy efficiency of an ozone generator, the cooling methods utilized by the ozone generator manufacturers are highly competitive. The discussion of generator cooling presented in this manual addresses general design considerations. The various ozone generator manufacturers must be contacted to determine the specific cooling requirements for their individual ozone generators.

Depending on the type of ozone generator, cooling is accomplished with either water, oil or freon plus water, or air. In order to optimize electrical energy efficiently, the cooling water temperature should be 20°C (68°F) or less. Approximately 3 to 4 liters (0.75 to 1.0 gal) of 20°C (68°F) cooling water is required for each gram of ozone produced. Different ozone generators will vary on the amount of cooling water required. A common design parameter is that the temperature rise of the cooling water should not exceed 5°C (41°F). It is noted that some ozone generators have operated successfully with cooling water temperatures as high as 40°C (104°F). However, the design engineer must be aware that ozone production capacity is lowered, electrical energy efficiency is reduced, and dielectric glass is stressed during operation with warm water temperatures. Equipment capacity must be de-rated for operation with warm temperatures.

The cooling water source can be potable water or good quality non-potable water. The cooling system may be closed-loop or once through and discharge. Typically, potable water in a closed-loop cooling system is used with the potable water re-cooled with water from the plant non-potable water supply. The ozone generator cooling water is often treated to obtain "boiler" quality water, in order to prevent scaling or corrosion in the cooling loop. For example, water with a high chloride concentration has been reported to "attack" stainless steel. A cost-effective analysis can be completed to determine if it is economically feasible to recycle the cooling water, or to simply purchase, use and discharge the ozone generator cooling water. In instances where the source of cooling water is too hot, a refrigeration cooling system has been used (1).

Air cooled ozone generators require that the generator room ventilation system be an integral part of the ozone system design. Experience from operating plants indicates that the cooling air must be dust-free and oil-free in order to prevent electrical short-circuiting on the high voltage electrical supply system. The heated air may be used for plant heating, but care must be taken to avoid ozone contamination of the plant buildings in case of ozone leakage.

Feed-Gas Oxygen Content and Flow Rate. The feed-gas flow rate and oxygen content can be used to change the ozone concentration from the ozone generator, and thus change the rate of ozone production. When the oxygen concentration of the feed-gas is greater, a higher concentration of ozone can be obtained for a given generator power setting. The ozone concentration increases because the potential increases for collisions between the electrons released across the discharge gap and the oxygen molecules within the gap. In some cases oxygen enriched air has been used to upgrade the capacity of an existing ozone generator (39). However, the most common use of an oxygen-fed ozone disinfection system is direct use of high purity oxygen and re-use of the oxygen in the activated sludge process.

A lower feed-gas flow rate will increase the ozone concentration. However, the specific energy consumption will be greater because of less cooling capability by the air flow. This relationship is shown in Figure 6-20 for a typical air-fed ozone generator. As shown, the specific energy consumption gradually increases from an average 12.1 Wh/g (5.5 kWh/lb) at an ozone concentration of 4.8 g/m^3 (0.4 percent wt) to an average 23.2 Wh/g (10.5 kWh/lb) at an ozone concentration of 30.2 g/m^3 (2.5 percent wt). It should be noted that operation at a lower ozone concentration tends to reduce ozone generation specific energy, but will require a higher feed-gas flow rate and higher air treatment cost. A balance between these two opposing factors may be evaluated. The maximum ozone concentration is typically 1.5 percent wt, where the average specific energy is 17.6 Wh/g (8 kWh/lb).

In Figure 6-20 a range of specific energy consumptions is shown to account for different cooling capabilities of various ozone generators and different feed-gas temperatures that may be encountered. It should be noted that the data shown is for a high quality feed-gas through clean ozone generators; thus, Figure 6-20 can be used to estimate energy consumption for typical ozone generation systems. The manufacturer should be contacted for more precise information on performance characteristics of individual ozone generators.

The information from Figure 6-20 may be rearranged to show an ozone "generator mapping" curve by selecting different gas flow rates. The mapping curve may be used to evaluate specific energy requirement relative to ozone production and assist in the selection of feed-gas flow rate. An example generator mapping curve based on actual test results is shown in Figure 6-21 (17). The approach to developing the generator mapping curve is as follows: (1) Select a desired gas flow rate, (2) Calculate the ozone production at a selected ozone concentration, (3) Plot the data point (i.e., calculated ozone production at specific energy requirement for corresponding ozone concentration), (4) Repeat steps 1-3 for various ozone concentrations, and (5) Select another gas flow rate and repeat steps 1-4.

The design ozone concentration for an air-fed ozone generator typically ranges from 12 to 24 g/m^3 (1 to 2 percent wt). A concentration of 18 g/m^3 (1.5 percent wt) is most common. While individual ozone generators may achieve concentrations greater than design, it is recommended that the discharge ozone concentration from all ozone generators be maintained at or below design conditions in order to maximize generator cooling effectiveness. Also, it is recommended that individual ozone generators not be operated at power settings greater than 75 percent of maximum, unless necessary. For example, rather than operate one ozone generator at 100 percent of capacity, 2 ozone generators should be operated at 50 percent capacity. These operating practices will cause less stress on generator dielectrics, will decrease generator maintenance problems, and will usually minimize electrical consumption.

The typical range of ozone concentration and specific energy consumption for an oxygen-fed ozone generator is shown in Figure 6-22. The typical (40 g/m^3) design ozone concentration for an oxygen-fed ozone generator at 3 percent wt, is twice the concentration for an air-feed system. Also, the specific energy is about half of an air-feed system.

It should be noted that the energy consumption information shown in Figure 6-20 and Figure 6-22 is for the ozone generator only. These data must be coupled with the energy consumption of the auxiliary equipment in order to determine the overall energy consumption of the ozone disinfection system. More information on energy consumption on auxiliary equipment is discussed in Section 6.5.2.

Figure 6-20. **Specific energy consumption versus ozone concentration for an air-fed ozone generator.**

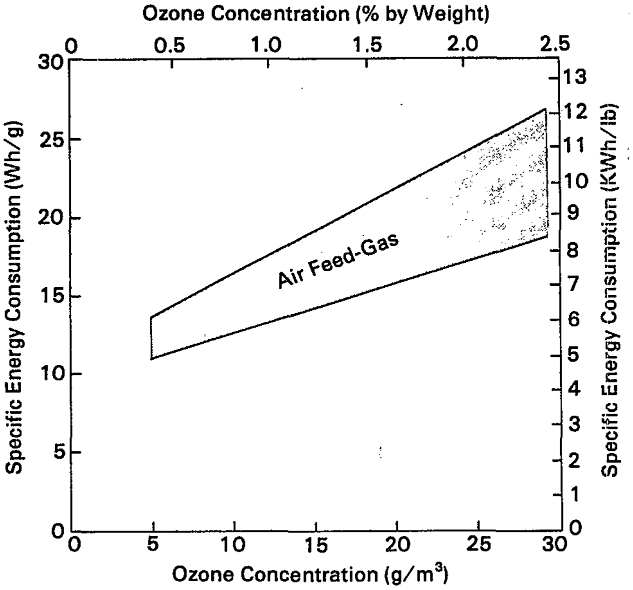
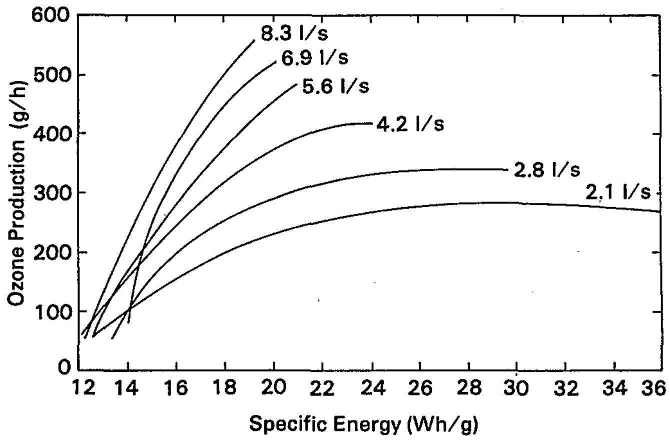


Figure 6-21. **Example ozone generator mapping curve using air feed-gas (17).**

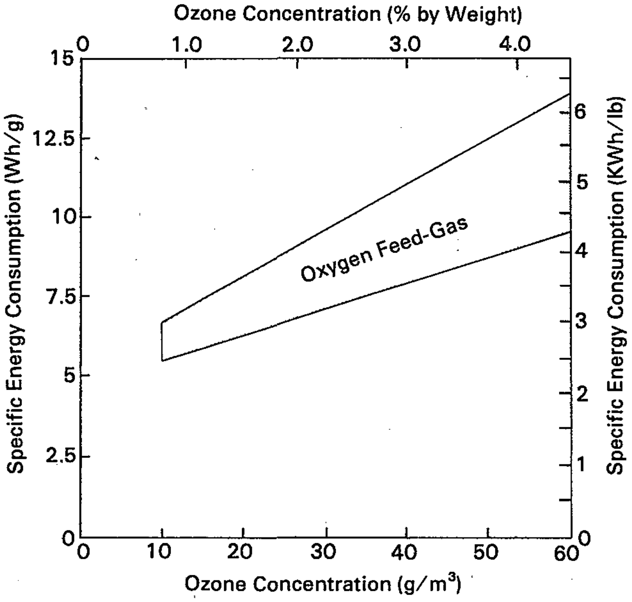


Ozone Generator Monitoring and Alarms. The following are required to adequately monitor ozone generation equipment:

a. Inlet feed-gas flow rate—Monitor system loading.

b. Inlet feed-gas temperature—Monitor system operation. Also provide alarm and shut-down device.

Figure 6-22. **Specific energy consumption versus ozone concentration for an oxygen fed ozone generator.**



c. Inlet pressure—Monitor system operation.

d. Discharge ozonized gas temperature—Monitor system operation. Also provide alarm and shut-down.

e. Discharge ozone concentration with in-line ozone meter—Monitor system performance. Also provide a recorder. Monitor may be considered optional for smaller installations.

f. Discharge ozone concentration using wet-chemistry procedures (40)—Monitor system concentration.

g. Inlet voltage—Monitor system loading.

h. Inlet amperage—Monitor system loading.

i. Inlet frequency—Monitor system loading.

j. System watt-hour meter—Monitor system energy. May be considered optional for smaller installations.

The operator should know the various aspects of system loading, performance and operation to be able to determine production levels and operating characteristics. The alarm and shut-down on the inlet and discharge gas temperature is used to protect the generator from damage. The recorder on the discharge ozone concentration meter is used to monitor generator performance. This meter and recorder may be considered optional equipment on very small systems, but the operators should at least be provided with the equipment and procedures for determining ozone concentration using wet-chemistry methods.

k. The cooling water system should be designed to achieve adequate cooling for the worst case conditions.

l. Cooling water system monitoring equipment should provide detailed information regarding system operation:

- Water temperature—Monitor system operation. Also provide alarm and shut-down.
- Inlet water flow rate—Monitor system operation. Also provide alarm and shut-down.
- Outlet water temperature—Monitor system operation.
- Heat exchanger cooling water inlet and outlet temperature and inlet water flow rate—Monitor system operation.

The cooling water system is vital to the operation of the water-cooled ozone generators. The operators should be provided with the instrumentation to be able to make control adjustments, as necessary, to maximize generator cooling. The alarm and shut-down on the water flow rate and temperature are provided to protect the generator dielectrics from breakage due to heat stress.

m. Cooling air system monitoring equipment should provide detailed information regarding system operation.

- Ambient air temperature—Monitor system operation.

The air cooling system is primarily dependent upon the room temperature. The "cooling" fans typically operate at a constant speed. The cleanliness of the cooling air is important to proper cooling, although no direct measuring is typically done to monitor this parameter. The air should be as cool as possible to maximize generator cooling, and as clean as possible to prevent dust accumulation on the cooling surface. Also, no oil should be used, because an oil film will develop on the cooling surface.

n. Inlet feed-gas temperature with alarm and shut-down—Monitor system operation and protect dielectrics.

o. Inlet feed-gas dew point with alarm and shutdown—Monitor system operation and protect dielectrics.

p. Safety inter-lock to prevent power energizing during generator cleaning or repair.

6.4.1.3 Types of Ozone Generators

Ozone generators are typically classified by their control mechanism, cooling mechanism and physical arrangement of the dielectrics. Another method of describing an ozone generator is by name of the inventor. The control mechanism may be a voltage or frequency unit. The cooling medium may be water, water plus oil or freon, and air. The physical arrangement of the dielectrics is typically either vertical or horizontal.

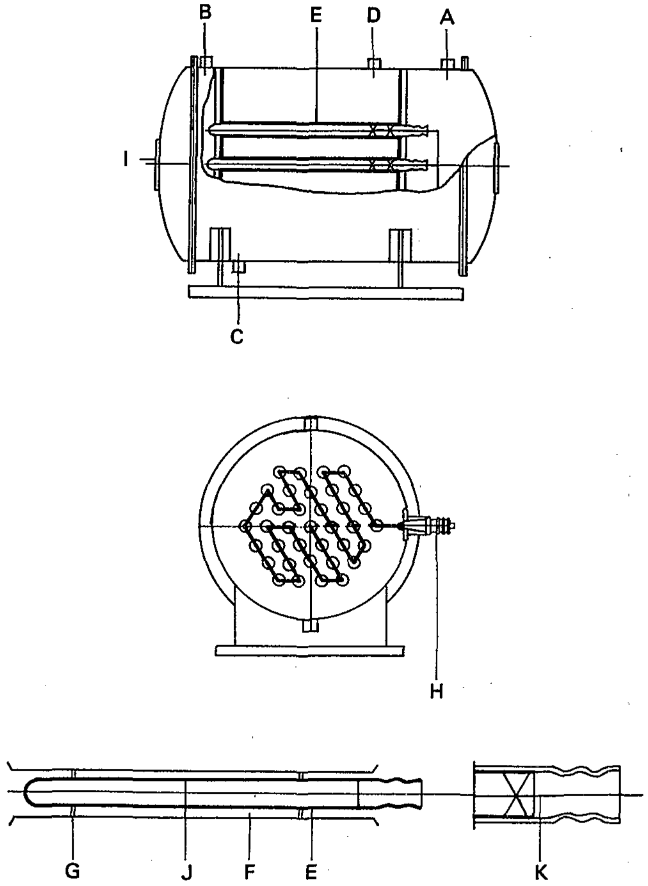
Horizontal Tube, Voltage Controlled, Water Cooled. The horizontal tube, voltage-controlled, water-cooled ozone generator is the most commonly used (1). Figure 6-23 shows details of the ozone generation equipment. The feed-gas enters one end of the ozone generator and the ozonized gas exits the opposite end. The stainless steel jacket acts as the low tension electrode, and contains multiple, cylindrical tubes where glass dielectrics are inserted. The internal side of the glass dielectric is coated with a metallic coating which acts as the high tension electrode. The feed-gas passes between the external side of the glass dielectric and the stainless steel jacket.

The equipment is normally designed to operate at pressures up to 103 kPa (15 psig). The majority of these generators operate at an electrical frequency of 60 hertz, although operation at frequencies from 600 to 800 hertz has been practiced (1). The horizontal tube, voltage-controlled, water-cooled ozone generators are installed at a number of wastewater treatment plants including: Frankfort, Kentucky; Vail, Colorado; Pensacola, Florida; Murfreesboro, Tennessee; Brookings, South Dakota; and Olympia, Washington.

Vertical Tube, Voltage Controlled, Water-Cooled. The vertical tube, voltage controlled, water-cooled ozone generator utilizes the cooling water both as the grounding electrode and the coolant. Details of the generation system are shown in Figure 6-24 (1). Input gas is "pulled" through the dielectrics by means of a vacuum created in the ozone contacting system.

The ozone generator consists of three compartments. The feed-gas is drawn into the upper compartment where it enters hollow metal tubes which are the high tension electrodes. The gas is drawn downward within the tubular metal electrode to emerge into the closed end of a glass dielectric tube. The feed-gas then passes upward through the corona discharge, which is created between the tubular stainless steel, high tension electrode and the glass dielectric, low tension electrode. The ozonized gas is discharged into the middle compartment, from where it is drawn to the ozone contactor.

Figure 6-23. **Details of a horizontal tube, voltage controlled, water cooled ozone generator (1).**



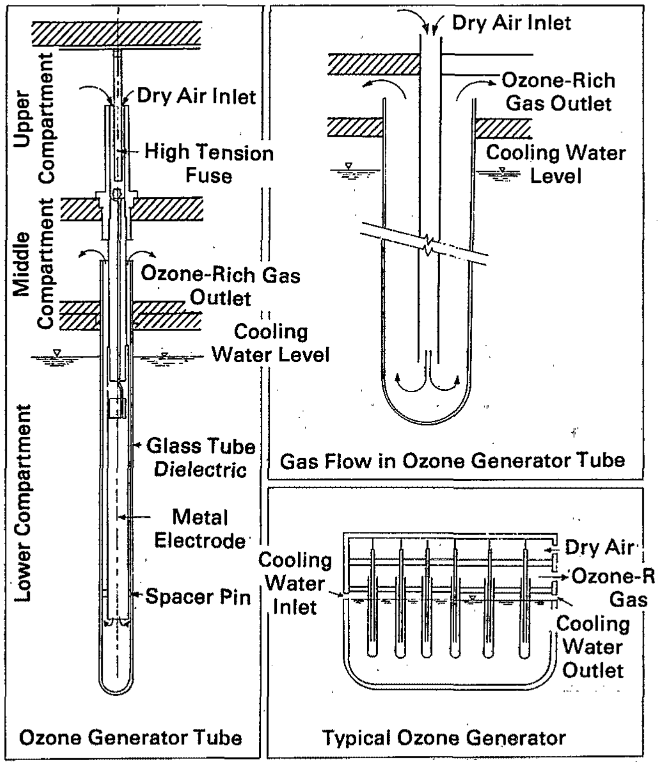
Legend

A–Air Inlet
B–Ozonized Air Outlet
C–Coolant Inlet
D–Coolant Outlet
E–Dielectric Tube
F–Discharge Zone

G–Tube Support
H–H.V. Terminal
I–Port
J–Metallic Coating
K–Contact

The vertical tube, voltage-controlled, water-cooled ozone generator is coupled with a proprietary, aspirating turbine mixer contacting system. The process was installed at the first wastewater treatment plant to use ozone disinfection in the United States (Indiantown, Florida), and was evaluated in an EPA research study (17).

Figure 6-24. **Details of vertical tube, voltage controlled, water cooled ozone generator (1).**

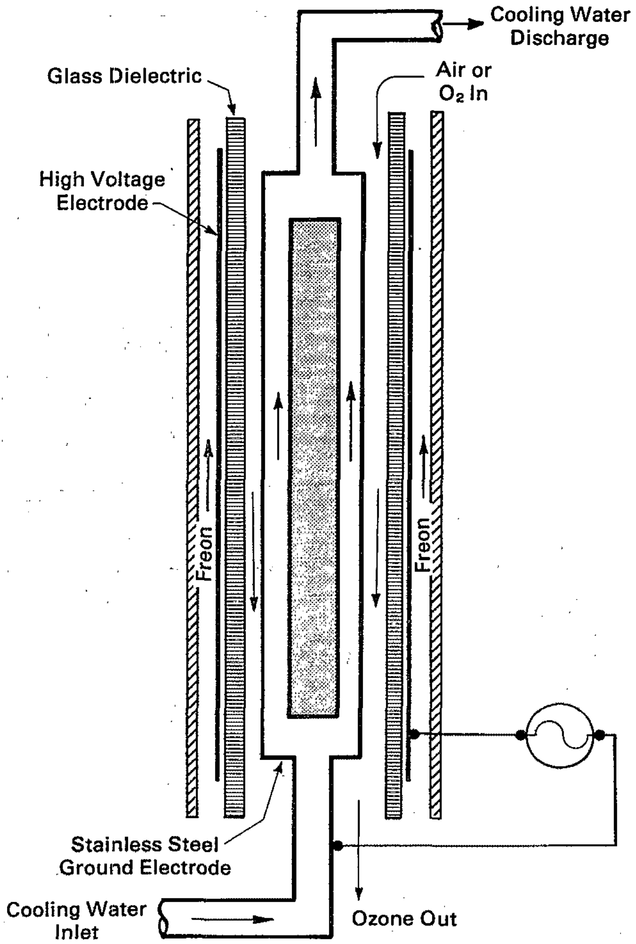


Vertical Tube, Frequency-Controlled, Double-Cooled. The vertical tube, frequency-controlled, double-cooled ozone generator was invented by Bollyky (U.S. Patents 3,766,051; 3,833,492). The generator utilizes both water and a non-conducting liquid (oil or freon) for cooling. The unit incorporates a complex dielectric design to achieve additional cooling, as illustrated in Figure 6-25 (1). The high tension electrode is cooled by the non-conducting liquid that is contained in a closed-loop cooling system. The low tension electrode is cooled by water.

The dual-cooled generator is typically operated at 10,000 volts, a frequency of 2,000 hertz, and a pressure up to 137 kPa (20 psig). A disadvantage of this unit is the higher capital cost associated with more involved cooling equipment. However, the generator energy efficiency is reported to be better due to the additional cooling provided and is reported to require fewer electrodes.

Wastewater treatment plant installations for the vertical tube, frequency-controlled, double-cooled ozone generator include Indianapolis, Indiana; Granby, Colorado; Alburndale, Florida; West Knoxville, Florida; Summers, New York; and Hunter, New York.

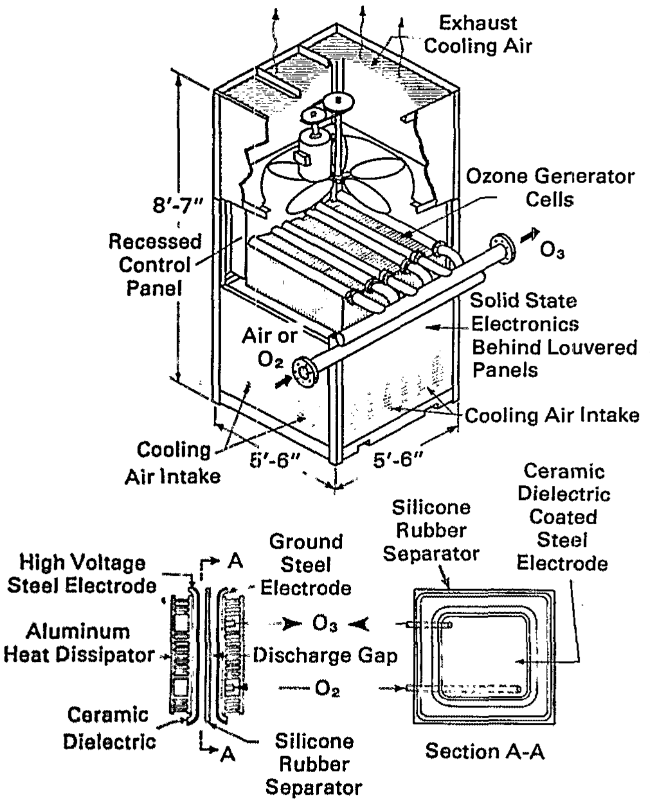
Figure 6-25. **Details of a vertical-tube, frequency controlled, double-cooled, ozone generator (1).**



Lowther Plate, Frequency-Controlled, Air-Cooled. The basic elements of the Lowther plate, frequency-controlled, air-cooled ozone generator are shown in Figure 6-26 (1). The dielectric consists of an aluminum heat dissipator, a steel electrode coated with a ceramic material, a silicone rubber spacer to establish the discharge gap, a second ceramic coated steel electrode with gas inlet and ozonized gas outlet, and a second aluminum heat dissipator. Thirty to forty dielectrics are manifolded to form a module. A typical Lowther plate module is also shown in Figure 6-26. The unit uses ambient air for cooling and operates at an upper frequency range of 2,000 hertz, at a potential of 9,000 volts. The maximum operating pressure is 103 kPa (15 psig).

A Lowther plate ozone generator was installed at Springfield, Missouri; Holland, Michigan; Moorhead, Minnesota; and Concord, North Carolina. Another type of air-cooled generator was installed at the Mission Viejo plant near Denver, Colorado (41).

Figure 6-26. Details of an air-cooled, Lowther plate type ozone generator (1).



6.4.1.4 Specific Manufacturers' Ozone Generators

Ozone generators manufactured by different companies have unique characteristics, but also have some common requirements. For example, all ozone generators should have a similar quality feed-gas, and most ozone generators can operate independent of the type of contact basin used. A listing of ozone generator manufacturers can be obtained from the International Ozone Association (IOA).

6.4.2 Feed-Gas Supply and Treatment Equipment

Feed-gas to an ozone generator may be air, once-through oxygen, or recycled oxygen, as discussed in Section 6.3.1. Typically, either air or once-through oxygen is used. Once-through oxygen systems typically do not require further moisture removal. However, filters to remove particulates should be provided. Air and oxygen recycle feed-gas ozone generation systems must provide treatment to remove both excess moisture and particulates.

Inadequate air or oxygen recycle feed-gas treatment, especially inadequate moisture removal, has caused several ozone system problems and failures of ozone treatment processes (4,42). As such, air treatment is a critical aspect in ozone system design. The major equipment design considerations are the quality of the treated air and the reliability of the treatment equipment.

6.4.2.1 Quality of Feed-Gas

The quality of the feed-gas for ozone generation is described by its moisture content and particulate content. Moisture content is mass of water per unit volume of gas (lb/1,000 ft^3), but the design specification for moisture content is typically expressed as dew point temperature. The relationship between moisture content and dew point temperature is discussed in Section 6.2.3.3.

Excessive moisture content in the feed-gas to the ozone generator can decrease ozone production, can increase generator maintenance, and can cause damage to internal components of the generator. To maximize ozone production and minimize maintenance problems, it is recommended that the dew point temperature of the feed-gas to all corona discharge ozone generators not exceed -60°C (-76°F), unless absolutely necessary. Operation at a dew point temperature greater than -50°C (-58°F) should be avoided altogether.

The primary function of the air treatment system is to remove moisture from the air. Moisture can be removed by increasing pressure, by cooling, or by adsorption techniques (21). Equipment used to remove moisture is further described. The procedure for calculating moisture content at various temperature and pressure conditions is discussed in Section 6.2.3.3.

The particulate content of the feed-gas should also be minimized in order to optimize generator production efficiency and reduce maintenance. Prior to the ozone generators, filters which remove particles greater than 1 micron in diameter followed by filters which remove particles greater than 0.3 to 0.4 micron in diameter are recommended (18,35).

6.4.2.2 Air Feed-Gas Treatment Systems

Air feed-gas systems are typically classified by their operating pressure. The most common is a low pressure system, which operates at a pressure ranging from 69 to 103 kPa (10 to 15 psig); although pressures up to 275 kPa (40 psig) have been reported when the pressure is reduced prior to the ozone generator. High pressure systems operate at a pressure ranging from 480 to 690 kPa (70 to 100 psig), reduce the pressure prior to the ozone generator, and are typically used in "small to medium" sized applications. Either system may be used in conjunction with most of the ozone generation equipment discussed in Section 6.4.1.3, and all of the ozone contacting systems discussed in Section 6.4.3. It should be noted that the decision to use a high or low pressure system is often based on a qualitative evaluation of potential maintenance requirements, in addition to the quantitative capital cost evaluation. Some of the issues to consider are listed below:

a. The high pressure air pre-treatment equipment generally has a higher maintenance requirement for the air compressors.

b. The high pressure air pre-treatment equipment generally has a lower maintenance requirement for the desiccant dryers.

c. The high pressure air pre-treatment system generally has a lower capital cost. At small to medium sized installations this lower capital cost may offset the additional maintenance required for the air compressors and associated equipment, such as filters for the oil type compressors. The engineer should investigate the potential maintenance associated with the high and low pressure systems rather than evaluating the design on capital cost alone.

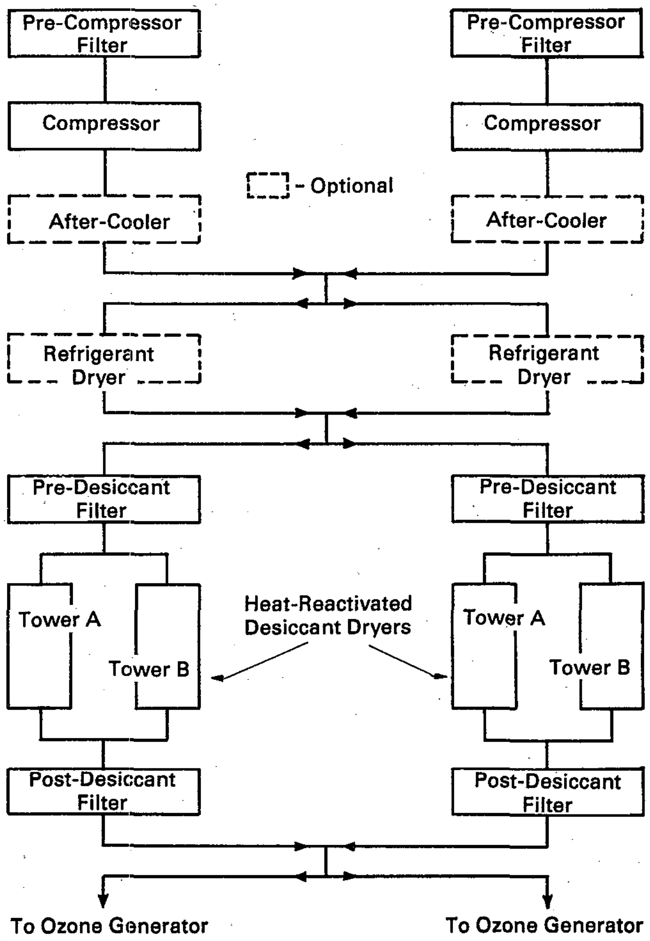
Another type of air feed-gas treatment system is the "nominal pressure" system, which typically operates at a negative or in some cases a slightly positive pressure. The nominal pressure process is a proprietary process used in conjunction with the Kerag ozone generator and aspirating turbine mixer ozone contactor.

Pressure System Flow Schematic. A schematic diagram of a low pressure air treatment system is shown in Figure 6-27. The diagram illustrates a dual component process, and shows desired flexibility for the equipment provided. More information on process flexibility design considerations is discussed in Section 6.5.1.3.

The pre-compressor filters are provided to protect the air compressors from damage due to large particles. The air compressors are typically positive displacement, oil-less units. Positive displacement compressors are used in order to obtain constant air flow at variable operating pressures. Variable pressures are often encountered due to variable pressure losses in downstream equipment and processes such as filters and ozone contact basins. Oil-less compressors are used to eliminate oil contamination of the downstream desiccant dryer medium and ozone generator dielectrics. Liquid-seal and rotary lobe compressors have been used most frequently.

The compressors may be followed by an after-cooler or a refrigerant dryer. These components are depicted by dotted lines in Figure 6-27, which indicate that they are optional. Typically, either one or the other option is provided. These cooling mechanisms are used to remove moisture in the air at minimal operating expense.

Figure 6-27. Example low pressure air feed-gas treatment schematic.



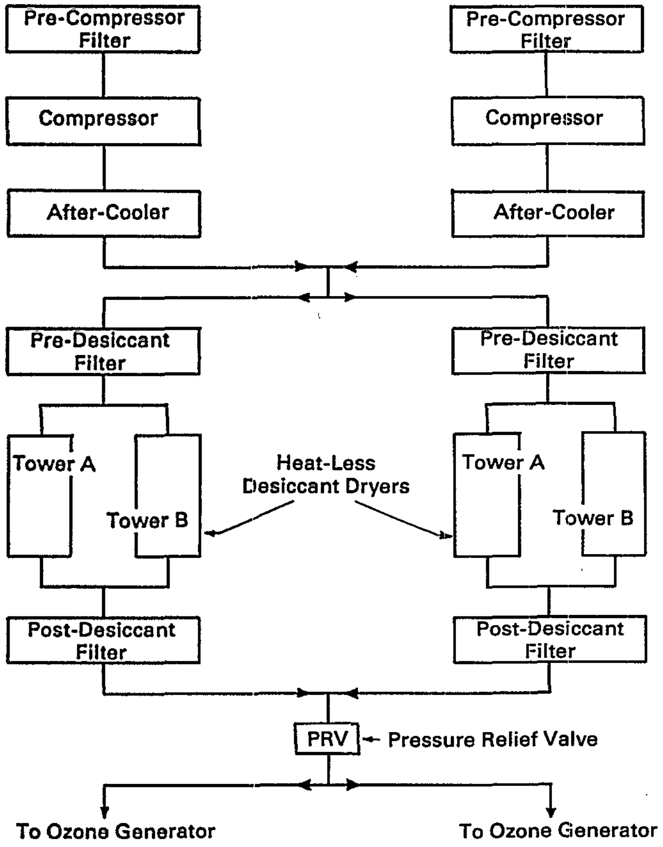
The compressed, cooled air is directed to a pre-desiccant filter, which is used to remove dust and dirt particles greater than 3 to 5 micron in diameter. Particulate removal prior to the desiccant dryers reduces plugging in the desiccant medium.

Probably the most important component of the air treatment process is the desiccant dryer. The desiccant dryer consists of two towers containing moisture adsorbing media. One tower operates while the other tower is regenerating. The low pressure system desiccant dryer uses heat for reactivation of the desiccant.

The post-desiccant filters are installed to remove particulates less than 0.3 to 0.4 micron in diameter. Two-stage filtration is preferred. The first stage filter removes particulates greater than 1 micron and the second stage removes particulates less than 0.3 to 0.4 micron in diameter (18,33,35).

High Pressure System Flow Schematic. A schematic of a high pressure air treatment system is shown in Figure 6-28. The pre-compressor filters are used to remove larger particulates and protect the air compressors. The compressors are typically oil-less units; however, oil-seal compressors followed by extensive oil removal equipment have been utilized.

Figure 6-28. **Example high pressure air feed-gas treatment schematic.**

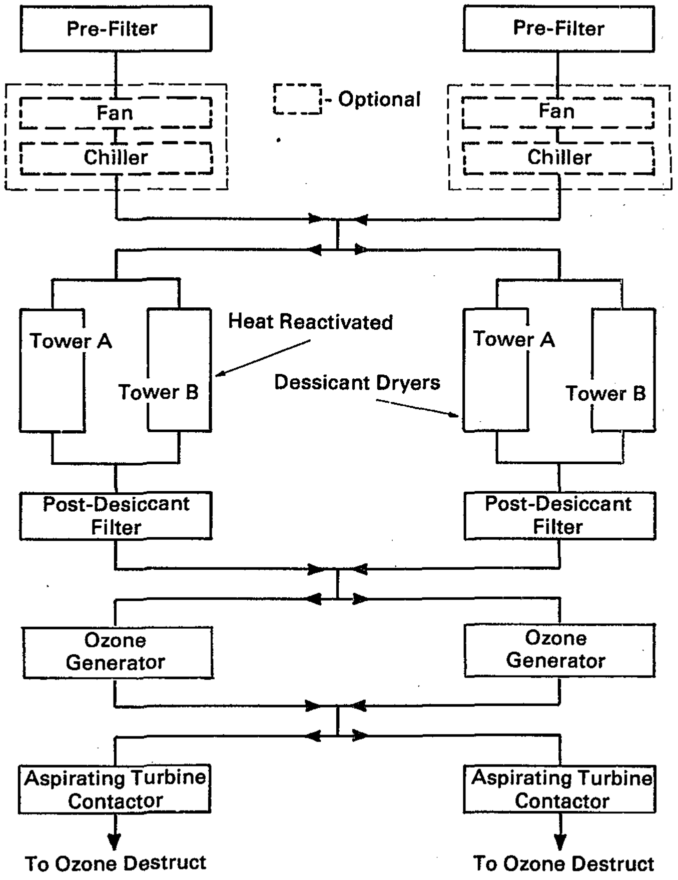


The after-coolers following the high pressure compressors are essential, as they are used to remove the heat of compression. The filter(s) before the desiccant dryer are used to remove particulates less than 3 to 5 micron in diameter when oil-less compressors are used. When oil-seal compressors are used, filtration to remove oil droplets less than 0.03 micron is provided (33).

The high pressure system desiccant dryer consists of two towers with moisture adsorbing media. One tower operates while the other tower is regenerating. Regeneration is accomplished without additional heat; thus, the high pressure desiccant dryers are called heat-less units. The post-desiccant filters remove particulate matter less than 0.3 to 0.4 micron in diameter. The high pressure system also has a pressure reducing valve to regulate operating pressures in the ozone generator.

Nominal Pressure System Flow Schematic. A schematic diagram of the nominal pressure air treatment system is shown in Figure 6-29. The nominal pressure system typically operates at a pressure slightly below ambient pressure conditions. In cases where a chiller is used to cool the air, a blower may be used to overcome the pressure drop within the chiller and the system may then operate under a slightly positive pressure. The nominal pressure system is a proprietary process used in conjunction with an aspirating turbine mixer contacting unit.

Figure 6-29. **Example nominal pressure air feed-gas treatment schematic.**



The primary difference between the nominal pressure system and the other air treatment process schematics is the method of moving the air. With the nominal pressure system the air is drawn through the air treatment process and through the ozone generators with the vacuum created by the operation of the turbine mixer. The pre-desiccant filter is used to remove particulates greater than 5 microns in diameter, and the post-filter removes particles greater than 1 micron in diameter (33). The nominal pressure system uses a heat-reactivated desiccant dryer.

The energy input and air supply to the turbine mixing contacting unit can be controlled by a variable speed drive on the mixer to adjust its pumping action, and by adjusting the orifice size of the water inlet to the mixer. The specific details of achieving this flexibility can be arranged with the manufacturer.

6.4.2.3 Air Compressor Design Considerations

Air compressors are used for the low and high pressure air treatment processes. Typically, oil-less compressors are used. However, for high pressure applications oil-seal compressors have been used when they are followed with extensive oil removal equipment.

Low Pressure Air Compressors. Low pressure air compressors operate at a pressure ranging from 69 to 103 kPa (10 to 15 psig), although operation up to 276 kPa (40 psig) has been reported when the pressure to the ozone generator is reduced. The following design considerations are applicable for low pressure air compressors:

a. Positive displacement compressors have been used most frequently because good control over air flow rate is achieved at variable operating pressures encountered.

b. The air compressors should be located in a remote area to avoid noise disruption, or should be located in a sound-proofed housing.

c. Liquid-seal compressors often have been used for ozone, despite a higher capital cost and in some cases a slightly higher energy requirement. These disadvantages were acceptable considering an overall design objective of reliable, continuous performance. It should be noted that when liquid-seal compressors are used the size of the downstream moisture removal equipment does not change because that equipment is designed to treat saturated air no matter what type of compressor is used.

d. Air compressors should be followed by an after-cooler or by refrigerant dryers. Water-to-air after-coolers are generally used, rather than air-to-air after-coolers.

e. Instrumentation that should be provided to monitor and control air compressor equipment operation is:
   - Discharge pressure—Monitor system operation.
   - Discharge temperature—Monitor system performance. Also include system alarm and shut-down.
   - Discharge flow—Monitor system performance.
   - Seal water temperature, when liquid-seal compressors are used—Monitor system operation.
   - Seal water flow—Monitor water availability. Also include system alarm and shut-down.

f. *Instrumentation that should be provided to* monitor and control after-cooler operation is:
   - Feed-gas temperature—Monitor system loading.
   - Discharge temperature—Monitor system performance.
   - Cooling water inlet and outlet temperature —Monitor system operation.
   - Cooling water flow rate—Monitor system operation.

High Pressure Compressors. The following design *considerations are applicable for high pressure air* compressors:

a. Non oil-lubricated, high pressure compressors (piston type compressors that use a Teflon seal) should be designed with a duty cycle less than 60 percent and should be coupled with preplanned methods of replacing the Teflon seal on a periodic interval. A replacement such as once per year may occur.

b. High pressure compressors may be oil-lubricated if followed by a sophisticated oil removal process. Successful performance has occurred with a cyclonic moisture/oil separation device, followed by an impingement filter, followed by a coalescing filter. The final filter should remove oil particles greater than 0.03 micron in diameter. At the same time, it should be anticipated that the medium in the desiccant dryers may need to be replaced quite frequently (e.g., every 12 months).

c. Instrumentation should be provided for process monitoring and control, as discussed previously in the Low Pressure Air Compressor section.

6.4.2.4 Refrigerant Dryer Design Considerations

The value of refrigerant dryers has been questioned at some operating installations. Some problems that have occurred with their use include sensitivity to icing because operating temperatures are quite close to the freezing point of water and non-availability of *maintenance expertise required to repair the refrig-*

eration equipment (42). If these problems can be overcome, refrigerant dryers can be very effective because they can remove excess moisture in the air for a small amount of energy consumption. The decision to use a refrigerant dryer is typically based on cost savings that can result in reducing the size of the desiccant dryer because of the moisture removed by the refrigerant dryer (21).

The following design considerations are applicable for refrigerant dryers:

a. If the desiccant dryers are reduced in size such that operation of the refrigerant dryer is essential to the overall treatment process, back-up equipment and spare parts must be provided to insure continued operation of this important, sensitive-to-operate piece of equipment.

b. Plant maintenance personnel must be trained on refrigeration equipment, or refrigeration equipment service and repair must be available on a responsive contract basis.

c. Instrumentation and control equipment that must be provided to prevent overload of the downstream desiccant dryers are:

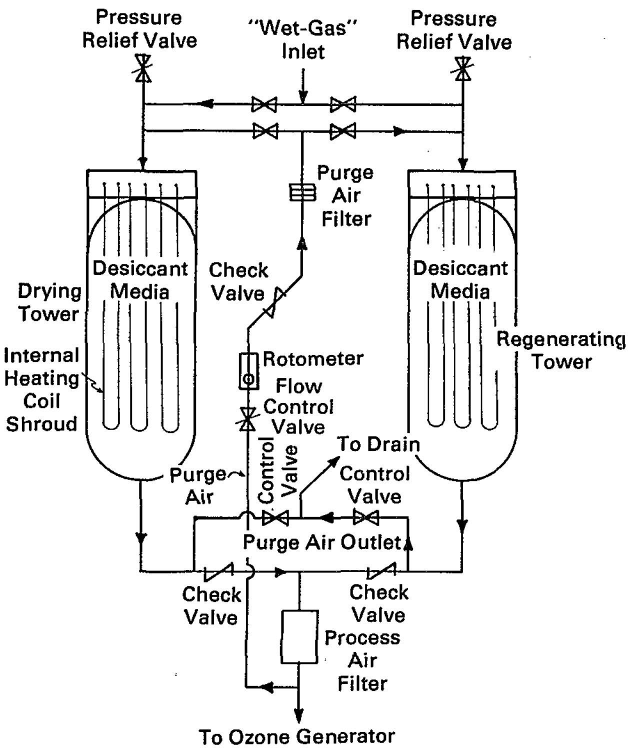
   - Inlet temperature—Monitor unit loading.
   - Outlet temperature—Monitor unit performance. Also provide an alarm and shut-down for high temperature.
   - Feed-gas flow rate—Monitor system loading.

6.4.2.5 Desiccant Dryer Design Considerations

The desiccant dryer is the most important unit in the air treatment system. Poor performance of the desiccant dryer will reduce ozone production and damage internal components of the generator. Desiccant dryers used for ozone systems are special application units, since "off-the-shelf" dryers typically are capable of achieving dew points in the range of only -40°C (-40°F) or higher. Dew points of -60°C (-76°F) and lower are required for ozone generation. It should be noted that dew point temperature is a function of pressure, and recorded dew point temperatures always should be referenced to pressure. In this manual all dew point temperatures are based on standard pressure conditions, unless otherwise noted. Refer to Section 6.2.4.3 for the procedure to adjust dew point temperature to standard conditions.

Heat-Reactivated Desiccant Dryers. Heat-reactivated desiccant dryers are used in nominal pressure and low pressure air treatment systems. A schematic of a heat-reactivated desiccant dryer is shown in Figure 6-30. Wet-air is directed to the operating tower where moisture in the air is adsorbed onto the desiccant. The desiccant is typically activated alumina and molecular sieves (21). In some instances silica gel is used. After several hours of operation the desiccant becomes saturated with moisture and is unable to maintain the desired dew point. At that time the desiccant must be regenerated. The cycle valve(s) switch the operating tower so the "used" medium can be regenerated. It should be noted that the switch valve is an important component in the dessicant dryer. Stainless steel valves have been reported to work best. Also, routine maintenance is necessary to make sure the valves switch every time.

Figure 6-30. **Diagram of a heat-reactivated desiccant dryer with internal heating coils.**



Regeneration involves heating the desiccant to a temperature between 90 and 260°C (200 and 500°F) and purging with dry air (21). The typical design is for heating to a temperature between 120 and 170°C (250 and 350°F) for 1 to 2 hours, and cooling for a minimum of 6 hours. Heating enhances moisture evaporation. The evaporated moisture is removed by purge air. After heating is completed the purge air will continue to flow through the tower and cool the desiccant before the regenerated tower is placed back into service. It is extremely important that the desiccant be cooled before being placed back into service so that a "temperature spike" does not occur. An aftercooler may be considered after the dessicant dryer and before the dew point analyzer, with alarm, in order to insure that a "temperature spike" does not occur. The dew point analyzer is used to prevent possible moisture contamination due to a leak in the aftercooler or excessive moisture from the dryer.

From 5 to 20 percent of the dry air is recycled (called purge air) to the regenerating tower. If the purge air flow is inadequate, hot air and excess moisture will remain in the regenerated tower. This moisture laden air will be discharged to the ozone generator when the regenerated tower begins its operating cycle, and will cause damage to generator components. To prevent unexpected or unnoticed loss of purge air flow, it is essential that the purge air flow be monitored and controlled. It is important to note that this is typically an option that must be specified in detail.

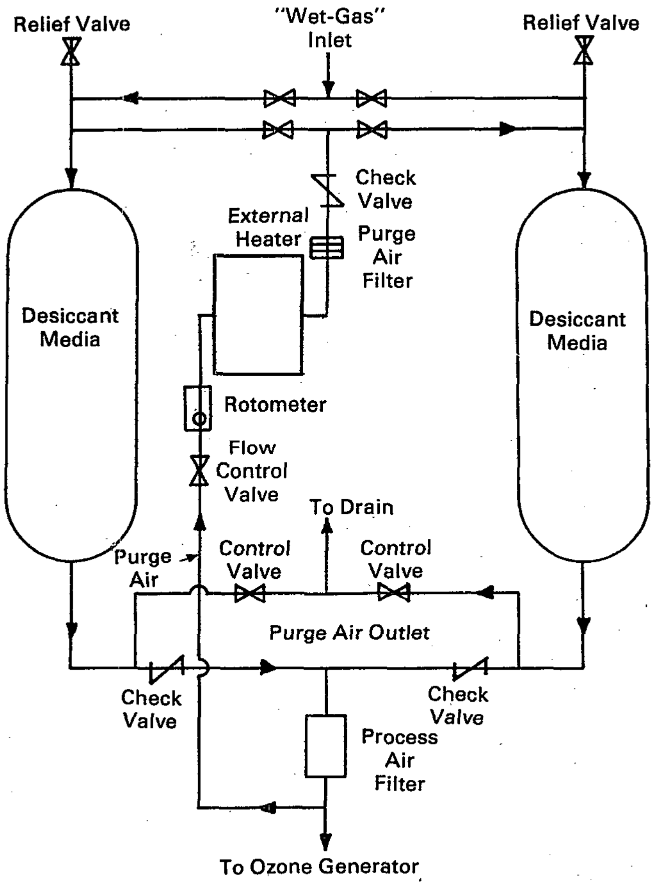
The purge air is filtered to protect the equipment, particularly the air supply control or to weir switching valves. The schematic diagram in Figure 6-30 shows that the source of the purge air is subsequent to the process air filters, which remove particles greater than 0.3 to 0.4 micron in diameter. This piping scheme provides initial filtration of the purge air and decreases the frequency of plugging of the smaller filters on the purge air line. Eliminating plugging is a very important operational consideration, given the importance of constant availability of purge air flow.

After months of operation the desiccant will lose its moisture adsorbing capability. A life expectancy of 36 months is considered normal but can range from 12 to 60 months. The factors affecting the desiccant life are frequency of heating and cooling, plugging, initial capacity (i.e., pounds of desiccant per lb/d of moisture loading), and method of heating.

An internal heating mechanism is shown in Figure 6-30. Internal heating is typically used in smaller installations because of capital cost considerations. However, internal heating coils often do not result in even heating of the desiccant. The desiccant closest to the heating elements will be over-heated and will deteriorate faster. In the desiccant dryer system shown in Figure 6-31, external heating capability is illustrated. The purge air flow is heated and directed to the regenerating tower. External heating is preferred because better distribution of heat throughout the desiccant is provided (37), but this system is more complicated because of the additional equipment requirements.

The amount of desiccant used and the regeneration cycle time are the most important design considerations. The cycle time must be long enough to obtain sufficient cooling of the regenerating tower, and short enough to maintain the desired dew point by the operating tower. The recommended minimum cycle time, because of cooling considerations, is eight hours (i.e., time from beginning of regeneration mode to end of operating mode). However, the design cycle time should be longer (e.g., preferred 16-hr and minimum 12-hr) because the desiccant loses its moisture adsorbing capability in a few months of operation. At initial start-up the cycle time should be longer in order to give the operator flexibility to reduce the cycle time as the desiccant loses its moisture adsorbing capacity. When the minimum, 8-hr cycle time is reached, the operator must replace the desiccant and can then re-set the cycle time to the original design settings.

Figure 6-31. Schematic of a heat-reactivated desiccant dryer with external heating equipment.



The regeneration cycle may be controlled on a timed or on a demand basis. The timed basis uses a pre-established time to initiate the regeneration cycle, whether or not the operating (drying) tower has exhausted its capability to remove moisture and achieve the desired dew point. The timed cycle control is adjusted by changing the timer settings.

A regeneration cycle based on demand is more expensive to install because of the controls involved. However, the system has potential O&M cost savings.

Using the demand regeneration system the cycle is initiated when a pre-set maximum value for the dew point of the discharge air from the dryer is reached. Dew point temperature is monitored on a continuous basis. Energy savings occur because the cycle time, and corresponding heating time would be activated only when the tower has exhausted its drying capability. Further, the life expectancy of the desiccant is prolonged because it would not be subjected to the stress of reactivation (i.e., heating and cooling) as often.

Demand regeneration control has some definite advantages, but has important operational considerations that must be addressed, namely the reliability of the dew point monitor and sensitivity of the control logic. Ozone equipment suppliers, desiccant dryer manufacturers, and existing ozone plant operating personnel may be contacted to determine which meters have been successfully used and how sensitive the control logic must be. Successful performance has been obtained from various systems, but problems have been encountered with others. All in-line dew point monitors should be routinely checked by performing a "dew point cup" test (32,42).

A word of caution is noted in operation of all dew point probes. When the measuring probe gets wet due to a moisture spike, it takes several hours of operation under dry conditions before the probe will accurately reflect the correct dew point temperature reading.

The amount of desiccant and the operating pressure (i.e., absolute pressure) controls the moisture removal capability of the process. Once the type of air treatment process is selected (i.e., low or high pressure), operating pressure is not considered further in design. This leaves amount of desiccant as a most important factor. Inadequate desiccant can cause operating problems in a short period of time. At one plant with insufficient desiccant, dew point breakthrough occurred within 5.5 hours of an 8-hr drying time and the desiccant was only 3 months old. The performance objective was a dew point temperature of -50°C (-58°F). At this plant the desiccant moisture loading was 13 lb desiccant/lb of moisture loading in an 8-hr period.

Apparently, 13 lb desiccant/lb of moisture loading is insufficient. Varas suggests that at least 18 lb desiccant be provided per lb of moisture received during one cycle of operation (33). Damez suggests that at least 5.5 lb of desiccant be provided per lb/d of moisture received, which corresponds to 16.5 lb desiccant/lb of moisture received during an 8-hr regenerating cycle (37). These suggested desiccant amounts may be used as guidelines for the design of desiccant dryers, but should not be considered the final design criteria because the type and quality of desiccant are also very important.

A recommended design approach is to specify that a minimum dew point of -60°C (-76°F) (corrected to standard pressure conditions) will be achieved during a minimum cycle time of 16 hr while being operated at the maximum expected moisture loading conditions (i.e., maximum temperature, maximum flow and minimum pressure). In addition, a minimum amount of desiccant for each operating tower should be specified.

A summary of the design considerations for heat-reactivated desiccant dryers is listed below:

a. The tower switching mechanisms should be designed for long life and must be maintained per the manufacturer's recommendations to insure continued, reliable performance.

b. The desiccant dryers should be designed with sufficient room around the units to allow for inspection or repair of the heating elements, for maintenance and repair of the tower switching mechanism, and for ease of replacement of the desiccant.

c. The heating coils used in internal heating, heat-reactivated desiccant dryers should not be contacted directly with the desiccant. A protective shroud should be provided.

d. The decision to use a timed or demand reactivation cycle control system should be based upon economic considerations. Demand reactivation cycle systems are generally desireable, especially for larger systems. For either control system the operators must routinely check and calibrate the in-line dew point monitor to assure reliable performance from this critically important instrument.

e. The minimum operating cycle time of a heat-reactivated desiccant dryer should be 8 hr, in order for the regenerated bed to cool down before being placed back into operation. The design cycle time should be 16 hr (minimum 12 hr) to allow for decreasing this time as the desiccant gradually deteriorates.

f. The process air compressors must be sized to also handle the desiccant dryer purge air flow rate, which may be as much as 20 percent of the ozone generation process air flow rate. Reliable process monitoring and control instrumentation that should be provided are:

- Feed-gas flow rate—Monitor system loading.
- Inlet temperature—Monitor system loading.
- Outlet temperature—Monitor system per-

formance. Also include alarm and system shut-down capability.

- Inlet and outlet pressure—Monitor system operation.
- Purge air flow rate—Monitor system operation.
- Discharge feed-gas dew point, in-line meter—Monitor system performance. Also include recorder and system alarm and shut-down.
- Discharge feed-gas dew point, dew point cup measurement—Measure dew point by manual methods to check on the reliability of the in-line meter.

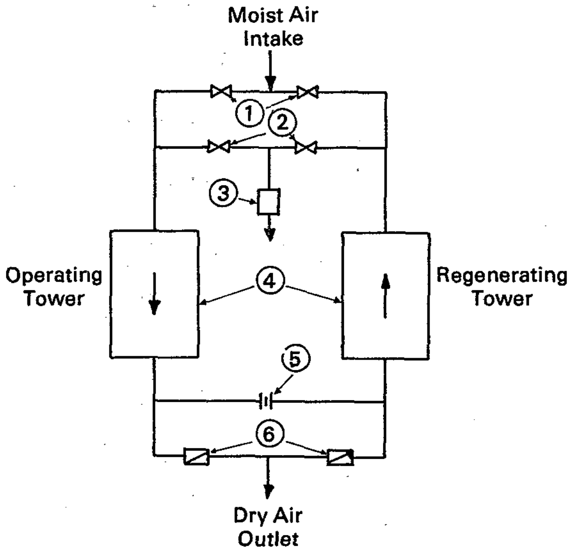
Heat-Less Desiccant Dryers. The heat-less desiccant dryers typically operate at a pressure ranging from 480 to 690 kPa (70 to 100 psig) (21). Another name for a heat-less desiccant dryer is a pressure swing desiccant dryer because the unit operates with varying pressures ranging from high pressure to low pressure. Heat-less desiccant dryers are considered a viable alternative for moisture removal for small to medium sized ozone disinfection applications.

A schematic diagram of the heat-less desiccant dryer is shown in Figure 6-32. The desiccant used is typically activated alumina and molecular sieves. The principle of operation of the heat-less desiccant dryer is adsorption of the moisture onto the desiccant under a high pressure. After a period of time ranging from 1 to 5 minutes the drying towers are switched. The tower to be regenerated is reduced to atmospheric pressure conditions and purged with "dry" air from the operating tower. The moisture that has adsorbed onto the desiccant is evaporated into and carried away by the "dry" purge air because of the lower pressure air's capacity to hold much more moisture (21).

The purge air flow rate for a heat-less desiccant dryer is normally 15 to 25 percent of the process air flow rate. It is important to note that the compressors must be sized to accommodate this additional flow requirement. The amount of moisture that is removed and the corresponding dew point is lowered when the amount of desiccant is increased and/or the cycle time is decreased.

The heat-less desiccant dryers have a history of reliable performance, if properly sized and maintained. The tower switching mechanism must be maintained per the manufacturer's instructions, and the amount of desiccant must be adequate. Over a period of time (1 to 5 years) the desiccant will lose its moisture adsorbing capability and will have to be replaced. This process will occur gradually and must be monitored closely. The design considerations for heat-less desiccant dryers are similar to those for heat-reactivated desiccant dryers discussed in the previous section.

Figure 6-32. Pressure swing (heat-less) high pressure desiccant dryer in purging mode.



① Wet Air Inlet
② Regenerated Air Outlet Valve
③ Silencer
④ Adsorption Desiccant Media
⑤ Orifice Plate
⑥ Check Valve

6.4.3 Ozone Contacting

The capability of the ozone contacting unit is critical to the successful performance of the ozone disinfection system. It is also important to the economical operation of the process. Two important ozone contact basin design considerations are the capability to achieve good disinfection and capability to achieve good ozone transfer. In a well designed ozone contact basin, good ozone transfer will exist if good disinfection occurs. However, the opposite may also occur if the contactor is poorly designed; good ozone transfer can exist when disinfection is less than satisfactory.

At some of the plants visited during development of this design manual, poor disinfection was noted despite good ozone transfer. Excessive short-circuiting was considered the primary cause of the poor disinfection performance. Design considerations for both ozone transfer and ozone disinfection are discussed in this section of the manual.

6.4.3.1 Ozone Transfer Design Considerations

The bubble diffuser contactor is the most common type of contacting system used. Other ozone contactors, such as the aspirating turbine mixer contactor, are usually coupled with a proprietary ozone generator (Kerag), and ozone transfer efficiency (TE) information can be obtained from the manufacturer. The characteristics of the bubble diffuser ozone contactor are presented herein.

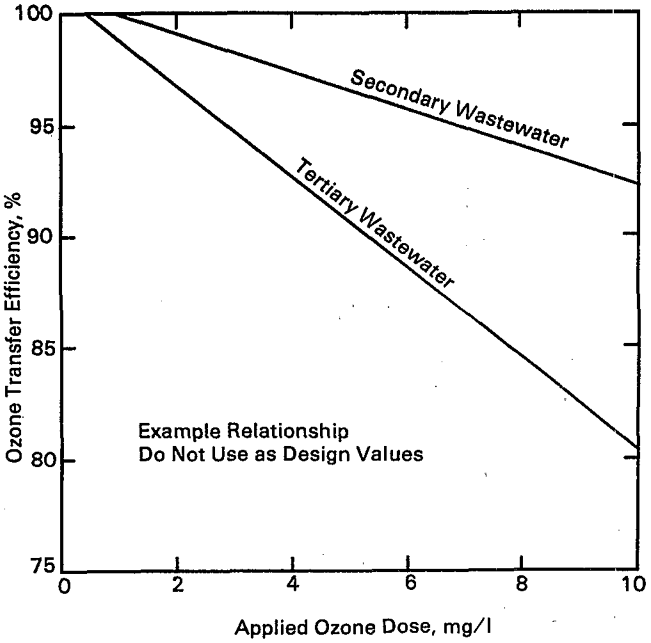
The transfer of ozone into wastewater has been evaluated by several researchers (43-45). All conclude that ozone transfer into wastewater can be described by the two-film theory. In this theory the mass transfer of ozone per unit time is a function of the two-film exchange area, the exchange potential, and a transfer coefficient. The exchange area for the bubble diffuser contactor is the surface area of the bubbles, which is discussed later. The exchange potential is called the "driving force" and is dependent upon the difference between saturation ozone concentration minus residual ozone concentration.

Ozone transfer can be described by the two-film theory, but in practice contactor basins have not been designed utilizing this theory. The theoretical basis has been avoided because the design coefficients have not been well documented, not because the theory is unsound. When the design coefficients are well documented, contactor design indeed may be established using the two-film transfer model. In this section of the manual the current state-of-the-art of ozone contactor design is used. The most important factors affecting design are discussed, and the design rationale is presented.

Ozone TE is primarily influenced by the physical characteristics of the contactor and the quality of the wastewater. At a given applied ozone dosage, a wastewater of poor quality will have a high ozone demand and the contactor will exhibit a high TE. The high TE is due to the disappearance of ozone in oxidation reactions (i.e., ozone demand reactions). An example of the effect of water quality on transfer efficiency is illustrated in Figure 6-33 (17). The TE of the same contactor was higher when treating secondary quality wastewater than when treating tertiary quality effluent. The differences in TE were more pronounced as applied ozone dosage increased.

The chemical quality of the wastewater also affects ozone TE, especially pH and alkalinity. A high pH and/or a low alkalinity will cause a lower ozone residual (i.e., other factors being constant) because the hydroxyl radicals will be maximized (refer to Section 6.2.1.4). The lower residual will increase the exchange potential, or driving force, and will increase TE.

Figure 6-33. Ozone transfer efficiency decreases as applied ozone dosage increases and as ozone demand of the wastewater decreases (17).



Wastewater quality will affect ozone TE, and is important to keep in mind when evaluating the performance of existing contactors, pilot-scale contactors, and newly installed contactors. Wastewater quality is typically not used as a basis to modify the physical characteristics of the contactor. A summary of the important water quality considerations on ozone TE design is listed below:

a. Ozone TE will decrease as applied ozone dosage increases. A specified minimum design TE should be coupled with a specified applied ozone dosage.

b. Ozone TE will increase as wastewater quality deteriorates (i.e., ozone demand increases). A specified minimum design TE should be coupled with a specified description of the wastewater quality.

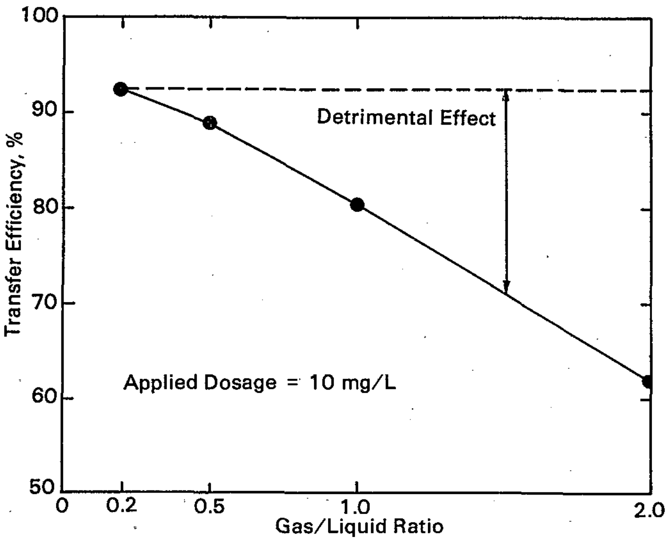
c. Ozone TE will increase as wastewater chemical quality favors the presence of hydroxyl radicals such as a high pH or low alkalinity. A comparison of TE of existing full-scale and pilot-scale results should consider differences in wastewater chemical quality.

The physical characteristics of the ozone contactor are the most important considerations for the design engineer because the engineer controls this element of the process. The most important physical characteristics for optimizing ozone TE are depth of the contactor and type and location of diffusers. The contactor physical characteristics for optimum disinfection performance are discussed in Section 6.4.3.2.

Ozone transfer theory states that the mass transfer of ozone is dependent upon the exchange area of the gas/liquid film. The exchange area for the bubble diffuser contactor is primarily a function of the diameter of the bubble. The bubble diameter can be controlled through proper selection of diffuser material. Best TE results are obtained when the bubble size is between 2 and 5 mm in diameter, and preferrably between 2 and 3 mm in diameter (30). The primary factors affecting bubble size include diffuser permeability and air flow rate. A lower permeability and lower air flow rate will decrease the size of the bubble.

Another consideration to optimize ozone TE is to optimize the ozone gas to wastewater liquid ratio (G_1/L_1) and ozone gas to contactor volume ratio (G/V). Stover et al. (17) observed a decrease in TE as the G_1/L_1 ratio increased, as depicted in Figure 6-34. A minor decrease in TE occurred as the G_1/L_1 ratio increased from 0.2 to 0.5. The decrease in TE was more pronounced as the G_1/L_1 ratio approached 1.0. Grasso (46) and Given and Smith (24) reported a decrease in TE as the G/V ratio increased from 0.005 to 0.05. They reported that for best results the G/V ratio should not exceed 0.03.

Figure 6-34. An increase in the gas to liquid ratio causes a decrease in ozone transfer efficiency.



The depth of the diffusers in the contactor is probably the most important design issue because it represents a cost consideration as well as a TE consideration. Deep diffusers will increase TE because the saturation ozone concentration and associated driving force will increase. However, a deep contactor will also increase capital cost. A shallow contactor will reduce capital cost, but will decrease TE. A high elevation of the contactor will also decrease TE. Hegg et al. (42) reported that a 3.6-m (12-ft) deep contactor at an elevation of 2,290 m (7,500 ft) above sea level achieved a TE of 50 percent while disinfecting a tertiary effluent.

Most of the bubble diffuser ozone contactors at plants visited during development of this manual had a diffuser depth between 4.9 and 6.1 m (16 and 20 ft) deep. These contactors achieved a TE between 80 and 95 percent when treating a high quality secondary effluent at an applied ozone dosage equal to or less than 6 mg/l.

A summary of important design considerations for the physical characteristics of the ozone contactor are listed below:

a. The diffusers in the bubble diffuser ozone contactor should be at least 4.9 m (16 ft) deep for a plant located near sea level treating a high quality secondary effluent at an applied ozone dosage less than 6 mg/l. The contactor should be deeper if the wastewater is of higher quality, if the applied ozone dosage is higher, or if the plant is located at a higher elevation.

b. Both rod- and disc-shaped porous stone diffusers have been used in bubble diffuser ozone contactors.

c. The air flow rate to the diffuser must be within the range recommended by the manufacturer.

   - For the 6.3-cm (2.5-in) diameter by 61-cm (24-in) long rod-shaped diffuser the air flow rate per diffuser should not exceed 1.9 l/s (4 cfm).

   - For the approximately 260-cm^2 (40-in^2) disc diffusers the air flow rate per diffuser should not exceed 0.6 l/s (1.25 cfm).

   - For the approximately 390-cm^2 (60 in^2) disc diffusers the air flow rate per diffuser should not exceed 0.8 l/s (1.8 cfm).

d. Each porous stone diffuser should be secured with stainless steel holders and sealed with ozone-resistant gaskets such as Hypalon, Viton, Teflon or Silicon.

e. The diffusers should be able to achieve a estimated bubble size of 2 to 3 mm in diameter.

f. Porous stone diffusers that have the following criteria have typically been used in bubble diffuser ozone contactors.

   - Permeability of the porous media typically has ranged from 12 to 20 cfm/ft^2/in at 2 inches water column. Note: Permeability is defined by a test of a porous stone plate that is 12 in by 12 in by 1 in thick, tested under 2 in water column pressure.
   - Porosity of the porous stone diffuser has been between 35 and 45 percent.

g. Diffusers have typically been located from 15 to 30 cm (6 to 12 in) from the bottom of the contactor.

h. All diffusers should be installed at the same elevation in order to evenly distribute the gas flow to each diffuser.

i. Diffusers have typically been installed in the liquid's downflow stages of the contactor in order to maximize ozone TE. The first stage of the contactor typically has more diffusers in order to satisfy the initial ozone demand of the wastewater. For example, a three-stage ozone contactor has contained 50 percent of the diffusers in the first stage.

6.4.3.2 Disinfection Performance

The primary factors affecting disinfection efficiency assuming ozone demand is not excessive (see Section 6.5.1.2 for discussion of zone demand), are: transfer efficiency, short-circuiting, mixing, and contact time. Transfer efficiency should be optimized as described in the previous section, in order to allow effective use of the ozone produced as a disinfectant and to minimize costs.

Short-circuiting is also a most important consideration. The effect of short-circuiting on disinfection performance is illustrated in Table 6-9. A small amount of short-circuiting significantly increases the effluent coliform concentration because the short-circuited wastewater contains a very high concentration of organisms relative to the concentration desired in the wastewater effluent.

Table 6-9. Effect of Short-Circuiting on Disinfection Performance

Influent Coliform #/100 ml	Effluent Coliform Target #/100 ml	Percent Short-circuiting (%)	Resulting Effluent Coliform #/100 ml
100,000	200	0.00	200
100,000	200	0.10	300
100,000	200	1.00	1,200
100,000	200	2.00	2,200

Mixing is necessary to bring the residual oxidants into contact with the microorganisms, but back-mixing can increase the potential for short-circuiting. For example, in a bubble diffuser ozone contact basin back-mixing in each stage of the contactor is sufficient to change the liquid flow characteristics from a plug flow to a near complete mix pattern in that stage. Therefore, to minimize the effect of short-circuiting, multiple stages that are positively isolated from each other should be provided. The optimum number of stages has not been documented.

Wastewater detention time in the contact basin is also a factor affecting disinfection performance with a wide range of contact times reported to achieve acceptable disinfection. Gan reported a 6-log reduction in coliform organisms for a high quality activated carbon effluent with 2 minutes liquid contact time and an applied ozone dosage of 12 mg/l. However, only a 1-log reduction was achieved with the same conditions when secondary effluent was treated. Gan concluded that to achieve better disinfection of poorer quality wastewater additional contact time would be necessary. He further recommended that multiple stages be used to provide the additional contact time (25).

Farooq et al. (16) reported a 3-log reduction in *Escherichia coli* after a detention time of only 6 seconds when a residual ozone concentration was present. This suggests that the disinfection action of ozone is very rapid. However, Farooq further suggested that the contact time be on the order of several minutes rather than several seconds since the microorganism must come into contact with the ozone.

Bollyky and Siegel (47) reported that disinfection action continued when the ozonized wastewater was held for a period up to 10 minutes, with minimal improvement after another 10 minutes of holding time. Perrich et al. (48) reported that the disinfection action was a function of the product of residual ozone concentration times liquid contact time. Venosa, et al. (20) also evaluated the effect of contact time on disinfection performance, and reported that disinfection efficiency correlated with the product of off-gas ozone concentration times liquid contact time. In all of these studies contact time was reported to be an important factor in achieving desired disinfection performance.

Stover et al. (17) evaluated the application of ozone to secondary and tertiary wastewater to achieve high levels of disinfection. His conclusion was that high levels of disinfection could be achieved at contact basin liquid detention times ranging from 1 to 10 minutes, but extremely high applied ozone dosages and a high level of residual oxidants (including high residual ozone concentration) were required. Stover was not able to evaluate the effect of a longer contact time on achieving equal disinfection at lower applied ozone dosages. However, he suggested that a substantial improvement in disinfection may be realized after initial ozone contacting is complete when the residual oxidants concentration is relatively high.

Legeron (49) and Miller et al. (1) discuss contact time relative to chemical reaction and mass transfer limited ozone processes. Both propose that the required contact basin liquid detention time for chemical reaction rate limited processes is dependent upon the rate of the chemical reaction. For example, the contact time can be very short (0.5 to 1 minute) for removal of free iron. However, for a mass transfer limited process, such as disinfection, they suggest a longer contact time. Legeron recommends a minimum contact time of 6 minutes. Miller reported that the contact time used for disinfection in European water treatment plants ranged from 6 to 10 minutes.

For purposes of this manual it is concluded that data are not available to define an optimum liquid contact time in the ozone contact basin, or to define an optimum configuration to prevent short-circuiting. However, general guidelines are considered applicable to address disinfection efficiency from ozone contact basins.

a. Multiple staged ozone contactors should be provided to minimize the effect of short-circuiting. A minimum of 3, and preferably more stages should be provided. Each stage should be positively isolated from the other to simulate plug flow characteristics and minimize the potential for short-circuiting.

b. The liquid contact time should be adjusted based on the desired disinfection target.

   - To achieve the former EPA standard of 200 fecal coliforms per 100 ml, the minimum contact time should be 6 minutes and preferably at least 10 minutes at design flow rates (See Section 6.5.1 for discussion of design flow).
   - To achieve more stringent standards the contact time should be lengthened to obtain the benefit of the disinfection potential of the residual oxidants produced. Pilot testing should be completed to determine the optimum contact time.

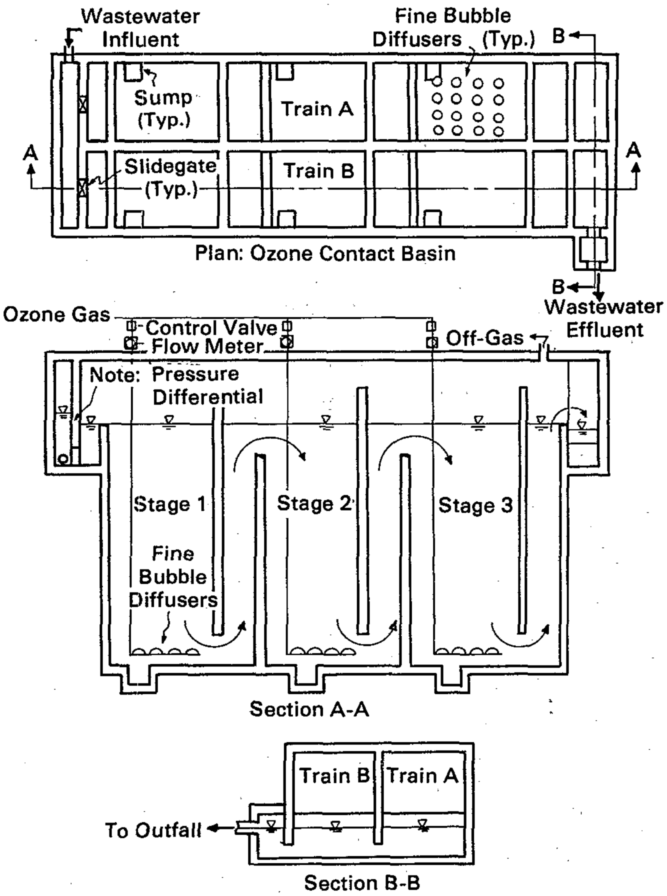
6.4.3.3 Types of Ozone Contactors

Various types of ozone contact basins have been proposed or used for wastewater disinfection including: positive pressure injectors, packed columns, spray towers, turbine mixers and bubble diffusers (22,28,50). The spray tower contactor is generally not used because ozone transfer efficiency is quite poor (56).

The packed column reactor, positive pressure injector and bubble diffuser contactor were evaluated by Venosa et al. (22). Results indicated that all contactors achieved equal disinfection performance at equivalent levels of transferred ozone dosage; however, their ozone transfer efficiency was significantly different. The bubble diffuser achieved better ozone transfer, especially as the ozone dosage increased. Nebel reported similar results (50).

The bubble diffuser ozone contactor is the most common reactor for ozone disinfection. Many different types of bubble diffuser configurations are available. A schematic of one type is shown in Figure 6-35. Several important design considerations that maximize ozone transfer and disinfection performance are illustrated in this schematic. These considerations are discussed below.

Figure 6-35. **Schematic of a 3-stage, bubble diffuser ozone contact basin.**



a. The contact basin should be as deep as practical, preferably greater than 5 m (16 ft) at sea level and deeper at a higher elevation, such as 6 m (20 ft) at 2,440 m (8,000 ft). The maximum depth may be limited by the maximum pressure in the ozone generator, which is usually 103 kPa (15 psig).

b. The bubbles formed by the porous stone diffusers should range between 2 and 3 mm in diameter. (See text for detail on permeability and porosity.)

c. The contactor should have at least two independent trains with isolated off-gas compartments to allow for continuous operation during inspection and cleaning.

d. The contactor should have features that simulate plug flow and reduce short-circuiting.

   - A minimum of 3, and preferably more, separate stages should be provided.
   - Each stage should be positively separated from the other stages. No chance for short-circuiting should exist: for example, through drain holes at the bottom of the walls separating the stages.
   - Each stage should be provided with a separate drain pit to aid in cleaning on a routine basis (e.g., once or twice per year).

e. The contactor should have from 1.2 to 1.8 m (4 to 6 ft) of "head" space to allow for foaming.

f. Each set of diffusers should have a flow control valve on the ozonized air piping and separate flow measurement. More diffusers should be located in the first stage to meet the higher demand for ozone in that stage, and thus provide capability to maintain a uniform residual oxidants concentration throughout all stages of the contact basin. The rest of the diffusers can be equally spaced in the remaining stages.

g. The wastewater flow should be counter-current to the ozonized air flow to maximize ozone transfer efficiency.

h. The contact basins should be made of typical construction grade concrete, with ozone resistant (e.g., Hypalon) water stops.

i. The contact basins should be covered and sealed as much as possible. Sealing with Sika-flex 1-A compound covered with coal tar epoxy or teflon sheeting has been used in some cases (18,51). However, basin sealing is difficult to maintain, and periodic leaks through the ceiling of the contact basin may occur (51). It is suggested that the ozone contact basin be placed in a location where the entire roof of the basin is exposed to the open atmosphere. Also, the basin should have the capability to operate under negative pressure.

j. Stainless steel piping for ozonized gas flow must be provided for positive pressure ozone systems.

   - Tungsten Inert Gas (TIG) welding is recommended.
   - Schedule 10 or better and type 304L or 316L stainless steel is recommended.
   - Flange-to-flange fittings, rather than threaded fittings, should be used in applications where welded connections are not made.

k. Ozonized feed-gas and contact basin off-gas sample lines should be stainless steel tubing. Teflon tubing may be considered for short runs.

The turbine reactor contacting system was evaluated by Stover et al. (17) and Venosa et al. (22). In both studies the ozone transfer efficiency and disinfection performance were shown to be comparable to that of the bubble diffuser contactor. A schematic diagram of the aspirating turbine mixer ozone contactor is shown in Figure 6-36. The turbine draws ozonized gas into the unit, where it is mixed with the wastewater and pumped outward through the impeller tips. The speed of the impeller and the size of an orifice controls the amount of water "pumped." The pumping rate controls the amount of mixing within the contactor, the amount of ozone dissolution or transfer, and the amount of ozonized-gas that is received from the ozone generator. At the same time these controls affect the amount of energy consumed by the process.

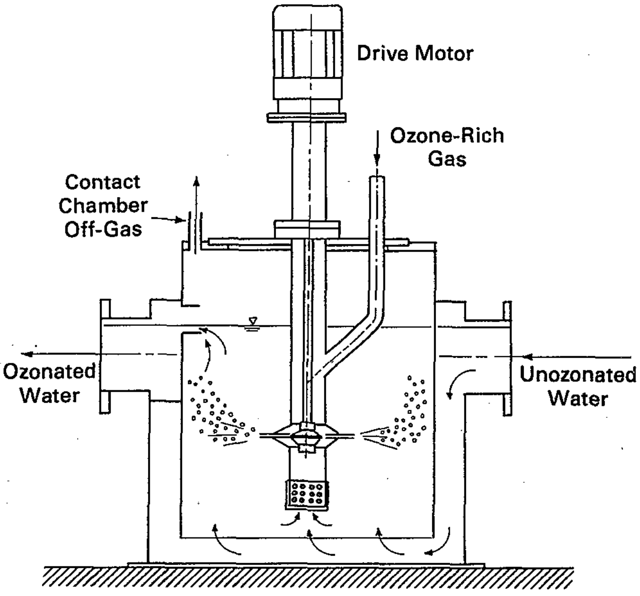
The turbine mixer contactor is typically used in conjunction with the nominal pressure ozone generation system and is capable of operating at shorter wastewater detention times because of the intensive mixing provided. The detailed design of size of the turbine mixer, amount of water pumped and feed gas flow rate should be determined in conjunction with the manufacturer of the nominal pressure system. In addition, the flexibility for controlling the energy consumption to match varying operating conditions should be incorporated into the design of the process. For example, flexibility should be provided to adjust the energy consumed by the turbine mixer at different wastewater flow rates and different applied ozone dosages.

6.4.4 Ozone Destruction Equipment and Unit Sizing

Ozone destruction is used to remove excess ozone in the contact basin off-gas prior to venting, or prior to recycle or reuse of the off-gas. Safety is the major consideration. The maximum allowable ambient ozone concentration for an 8-hr working day is 0.0002 g/m^3 (0.1 ppm by volume). (See Section 6.1). This concentration is significantly less than the ozone concentration in the off-gas, which is normally greater than 1.0 g/m^3 (500 ppm by volume).

The primary methods for treating excess ozone in the off-gas are: thermal destruction, thermal/catalyst destruction, and catalyst destruction (52-54). Activated carbon destruction has also been used, but the reaction with activated carbon causes the formation of powdery activated carbon which may be explosive (53). The use of activated carbon for ozone destruction is not recommended.

Figure 6-36. ***Schematic of a turbine mixer ozone contactor.***



6.4.4.1 Thermal Destruction

Thermal destruction is typically not used with oxygen feed-gas systems because of the high oxygen concentration and potential for uncontrollable fires. The thermal destruct method of reducing ozone in the off-gas involves heating the off-gas to a high temperature and maintaining this temperature for a period of time. From 50 to 100 percent ozone destruction has been reported at operating temperatures between 250 and 350°C (480 and 660°F) that are maintained from 1 to 3 seconds (52-56). A temperature between 300 and 350°C (570 and 660°F) for 3 seconds is required to achieve greater than 99 percent ozone destruction.

Because of the high temperatures involved, heat recovery units are typically provided on thermal ozone destruct systems. The outlet gas temperatures for a heat recovery thermal destruct unit range from 70 to 110°C (160 to 230°F), and typically between 90 and 100°C (195 and 210°F) (52,54-56).

A schematic of a thermal destructor with heat recovery is shown in Figure 6-37. The ozone contact basin off-gas passes through a pressure/vacuum relief valve and demister prior to entering the heat exchanger. The pressure/vacuum relief valve is provided to protect the contact basin from structural damage due to excessive pressure or vacuum build-up within the basin. The demister (i.e., stainless steel wire mesh) is provided to reduce foam accumulation within the heat exchanger and heating elements. Different types of heat exchangers can be used, including cross-flow, shell-and-tube, or plate-type (52). A fan is shown as an optional piece of equipment, depending on whether or not the contact basin is to be operated under a pressurized or vacuum condition. The option to operate under a vacuum condition is highly desirable.

The energy requirement of the thermal destruct system and the size of the thermal destruct heating element can be estimated by knowing the off-gas flow rate, the off-gas temperature rise and the heat loss of the equipment (i.e., energy efficiency of the unit). Assuming no heat loss, the amount of energy required to raise 1.0 m^3 of gas 1.0°C is 0.37 Wh (i.e., assume a specific heat of 0.2454 kcal/kg/°C, a gas density of 1.293 kg/m^3, and a conversion factor of 861.29 kcal/kWh (52,53). The heat loss of a thermal destructor is reported to be about 30 percent (53,55). Combining these parameters, the specific energy requirement (Wh/m^3) for a thermal destruct unit was determined for various operating conditions of temperature rise, as shown in Figure 6-38. This specific energy value can be coupled with the off-gas flow rate to determine the energy consumption and power requirement of the thermal destruct unit.

The following design considerations are summarized for a thermal ozone destruction process:

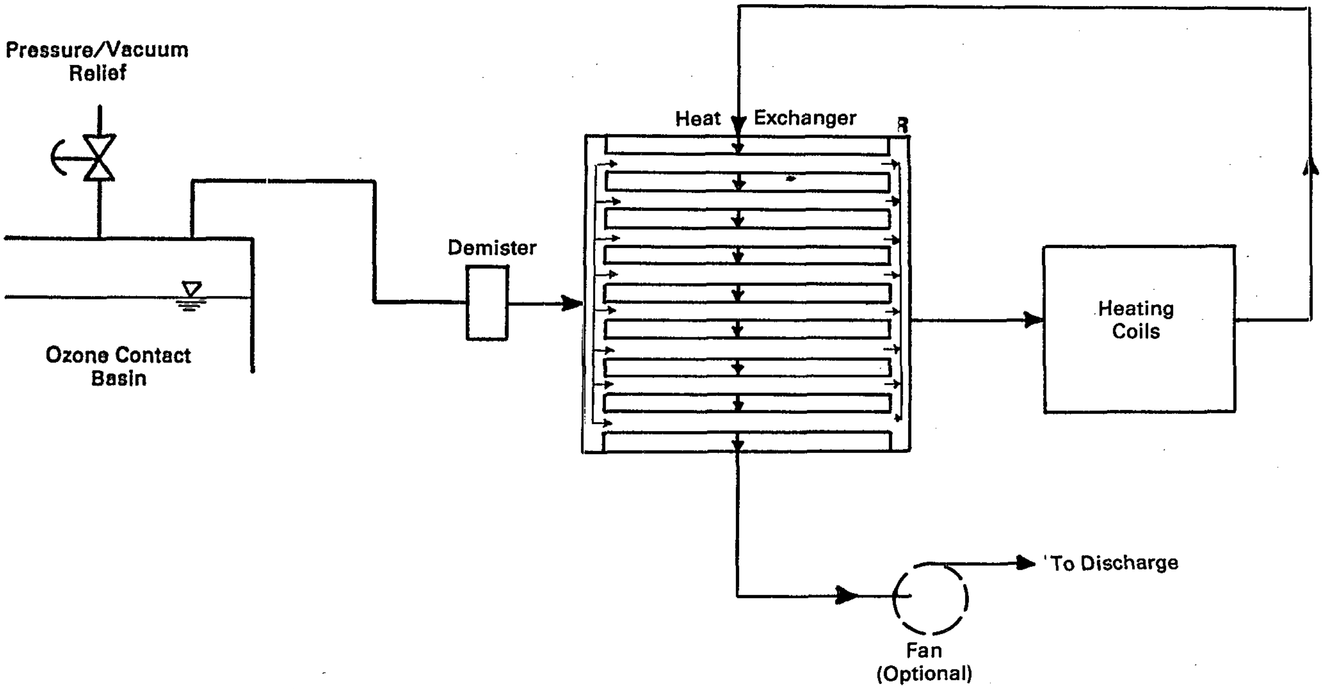
a. The thermal destruct equipment and piping should be well insulated to minimize heat loss.

b. A stand-by unit should be provided to allow for continuous operation during repair and maintenance.

c. A pressure/vacuum relief valve should be located on the contact basin to protect the contact basin from structural damage due to excessive pressure or vacuum.

d. A demister (i.e., stainless steel wire mesh) should be located prior to the ozone destruct unit to reduce foam accumulation on the heating elements.

Figure 6-37. Example diagram of a thermal destruct unit with a heat-exchanger.



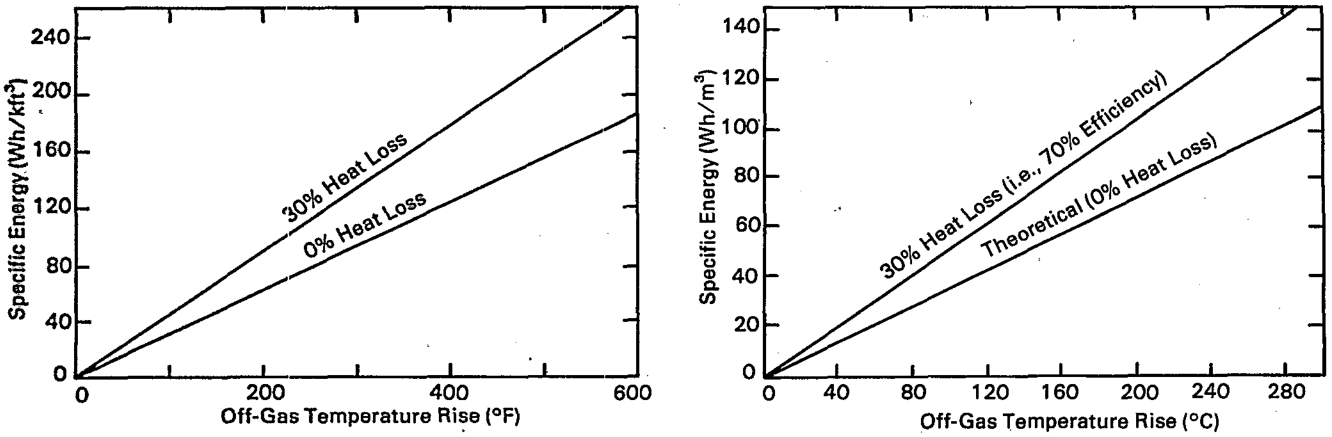
e. The operating temperature should be 300 to 350°C (570 to 660°F), and the contact time at least 3 seconds.

f. A heat recovery system should be considered to reduce operating cost.

g. Instrumentation for process monitoring and control that must be provided are:

- Inlet and outlet gas temperature—Monitor system performance.
- Inlet gas flow rate—Monitor system loading.
- Inlet ozone concentration—Monitor system loading.
- Outlet ozone concentration using meter—Monitor system performance.

Figure 6-38. Specific energy consumption versus off-gas temperature rise through the thermal destruct unit.



6.4.4.2 Thermal/Catalyst and Catalyst Destruction

The use of catalysts for ozone destruction is fairly recent (52). Specific information about the type and quantity of material in the catalyst is not available since it is proprietary information. Several general classifications of catalysts are known to destroy ozone including: metal catalysts, metal oxides, hydroxides and peroxides (53).

A schematic of a metal catalyst ozone destruct unit is shown in Figure 6-39. The contact basin off-gas passes through a demister prior to entering the unit. It is very important that contaminants such as foam be kept away from the catalyst. The size of the metal catalyst should be obtained from the manufacturer, due to the proprietary nature of the catalyst. A flow rate per volume of catalyst of 20 scfm/ft^3 has been successfully used (18). Flow rates as high as 50 scfm/ft^3 have been reported (52). A fan is shown as an optional piece of equipment, depending on whether or not the contact basin is to be operated under a pressure or vacuum condition.

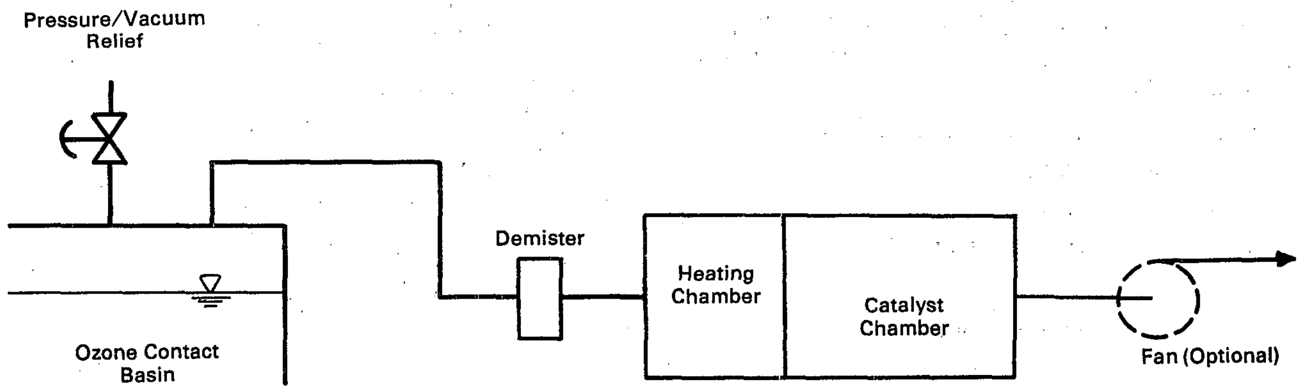
Metal catalysts are now being used most often, because they are more active than metal oxide catalysts (54). Metal catalysts, such as finely divided platinum or palladium, can operate at temperatures as low as 29°C (85°F) (54). The use of metal catalysts is advantageous, primarily because of lower operating cost. The required temperature rise is much lower than for thermal destruction and the catalyst life expectancy is about 5 years (53). In some instances the catalyst is operated without any temperature increase. However, moisture condensation on the catalyst can blind the catalyst and render it ineffective; therefore, it is generally advisable to increase the gas temperature to prevent moisture condensation on the catalyst.

A metal oxide catalyst, such as aluminum oxide that contains palladium, operates at a temperature ranging from 50 to 70°C (120 to 160°F) (53). A disadvantage of the metal oxide catalyst is that this is more sensitive to chemical reactions with nitrogen oxides, chlorine and its derivatives, and sulfides, which destroy the catalyst (54). The hydroxide and peroxide catalysts have not been used on field-scale plant equipment.

The following design considerations for the catalyst ozone destruction process should be considered in addition to the design considerations for the thermal destruct unit discussed in the previous section.

a. The catalyst and heating elements should be easily reached for maintenance.
b. The gas pressure differential across the catalyst should be monitored.
c. The size of the metal catalyst must be obtained from the manufacturer, given the proprietary nature of the material. An accurate projection of inlet and desired outlet ozone concentration must be presented to the manufacturer.

Figure 6-39. **Example diagram of a thermal/catalyst ozone destruct unit.**



6.5 Ozone Disinfection Process Design Considerations

Ozone disinfection process design involves sizing of ozone generation equipment and basins to meet disinfection objectives economically over the anticipated range of operating conditions. Aspects of the ozone system that affect disinfection performance and cost-effectiveness include: process energy efficiency, process flexibility, effluent disinfection criteria, wastewater quality, wastewater flow variations, plus others. Guidelines for ozone disinfection process design are presented in this section of the manual.

6.5.1 Ozone Production Requirements

Proper sizing of the ozone generation equipment is important for meeting desired effluent criteria without excessive capacity that results in high capital costs. To properly establish the ozone production capacity the wastewater flow rate and applied ozone dosage must be properly selected.

6.5.1.1 Wastewater Flow Design Considerations
The design wastewater flow rate for the disinfection process may be dictated by State or other regulatory design criteria, such as peak hourly flow rate (10-State Standards, 1973 ed.), or may be determined based on an analysis of local conditions. An analysis of local conditions may involve an evaluation of the frequency of the flow exceeding a specified value that may be determined by developing a probability curve using existing flow data. The design flow rate can be based on a selected probability of occurrence. Flow equalization should be evaluated to optimize the size of ozone generation equipment needed to achieve disinfection.

At plants surveyed during development of the design manual, the ozone disinfection system design flow rate was typically two to three times the average daily flow rate. Ozone generation capacity was provided to achieve a specified applied ozone dosage at the peak design flow rate. Stand-by ozone generation capacity was not provided at peak conditions. It was anticipated that ozone equipment maintenance that would require removal of equipment from service could be completed during expected low flow conditions. Where peak flows are frequent and unpredictable, back-up or stand-by equipment during peak flow conditions should be provided. As such, stand-by equipment must be evaluated on a case-by-case basis. In high growth areas or for a long-term plant design life (i.e., 20 years), the most economical alternative may be to provide the space for additional ozone generation equipment in the original design and purchase the equipment when needed.

6.5.1.2 Ozone Dosage Design Considerations
Both applied (D) and transferred (T) ozone dosage are important in ozone process design. Transferred ozone dosage is typically used for establishing the relationship between ozone dosage and disinfection performance. Once T and TE are defined, D can be established. A determination of applied ozone dosage is required to determine ozone production capacity. In this section of the manual an approach for determination of T and D is presented. Refer to Section 6.4.3.1. for a discussion of ozone transfer efficiency.

Determination of Transferred Ozone Dosage. The transferred ozone dosage (T) required to achieve disinfection is dependent upon the quality of the wastewater (i.e., potential for chemical reaction with ozone), the plant discharge criteria, and the disinfection performance capability of the ozone contact basin. Because of the variables involved, selection of transferred ozone dosage is probably the most difficult process design consideration. The preferred approach to establishing a design-transferred ozone dosage is to conduct a pilot plant evaluation on the treated wastewater to be disinfected. The type of pilot-scale ozone generator used is not critical to overall results; however, the type of pilot-scale ozone contact basin must duplicate the proposed full-scale basin for the results to be applicable to full-scale design.

In practice, pilot testing has not been routinely accomplished. Dosage requirements have often been based on published pilot plant or existing full-scale plant operating data. However, these data are site specific and may not be directly applicable to other installations. In this section both reported data from existing plants and a rational approach for determination of transferred ozone dosage are discussed.

The transferred ozone dosage (T) requirement to achieve various levels of disinfection performance were evaluated by several investigators and were discussed in detail in Section 6.3.2. Typically, transferred ozone dosages between 4 and 10 mg/l met the former EPA fecal coliform standard of 200 per 100 ml (27) when the total COD concentration of the treated wastewater was less than 40 mg/l. Transferred ozone dosages greater than 10 mg/l were projected when the wastewater had a large industrial contribution and a COD concentration greater than 70 mg/l. To meet a stringent standard of 2.2 total coliforms per 100 ml, a transferred ozone dosage between 36 and 42 mg/l was required when secondary treatment plant effluent was disinfected (17). A transferred ozone dosage between 15 and 20 mg/l was required when nitrified wastewater was disinfected.

The design transferred ozone dosages were not available for existing plants using ozone disinfection, but the design applied ozone dosage for seven plants ranged from 3 to 14 mg/l, as shown in Table 6-10 (4). Most of these plants were required to achieve a concentration of 200 fecal coliforms per 100 ml. Operating data were not reported at these plants; however, applied ozone dosage operating data obtained from site visits conducted during the development of this design manual and are shown in Table 6-11. The reported data for three plants appear to be in line with the design criteria for the other plants. However, at two plants visited the disinfection standard of 200 fecal coliforms per 100 ml could not be met even though the applied ozone dosage was significantly higher (i.e., greater than 10 mg/l) than at the other plants (i.e., 3 to 6 mg/l) and the treated wastewater was of similar quality. The poor performance at these plants was believed to be due to poor contact basin disinfection capability, especially excessive short-circuiting.

Table 6-10. **Reported Design Applied Ozone Dosages for Various Wastewater Treatment Plants (4)**

Name of Plant	Location	Design Applied Dosage mg/l	Permit Limitations BOD/TSS/NH$_3$ mg/l	Fecal Coliform #100/ml
Rocky River Regional	Concord, NC	14	20/30/13	N/R
Upper Thompson Sanitation District	Estes Park, CO	6	30/30/20	200
Frankfort	Frankfort, KY	3	10/10/1	200
Southwest	Springfield, MO	4	N/R	N/R
Brookings	Brookings, SD	3	20/30/2	1,500
Murphreesboro	Murphreesboro, TN	6	N/R	N/R
Madisonville	Madisonville, KY	6	10/30/1	200

N/R = Not Reported

Table 6-11. **Reported Operating Applied Ozone Dosages for Various Wastewater Treatment Plants**

Name of Plant	Location	Operating Applied Dosage mg/l	Effluent Quality BOD/TSS/NH$_3$ mg/l	Fecal Coliform #/100 ml
Southport	Indianapolis, IN	5.0	5/5/NR	17
Southwest	Springfield, MO	5.4	5/5/NR	10
Vail	Vail, CO	2.5	5/5/NR	1,000

Note: At two other plants visited an applied ozone dosage greater than 10 mg/l reportedly does not allow consistent achievement of the 200 fecal coliform per 100 ml limit, despite the fact that the wastewater quality was similar.

Data obtained during site visits to these plants.
N/R = Not Reported.

The transferred ozone dosage that is required to achieve a desired concentration of coliform organisms in the effluent is dependent upon the disinfection performance capability of the ozone contact basin, the demand for ozone in reactions not associated with disinfection, the influent coliform concentration, and the discharge coliform requirement. A change in any of these parameters can cause a significant change in the discharge coliform concentration. The approach to design presented in the remainder of this chapter of the manual allows for an independent evaluation of the effect of each parameter on transferred ozone dosage requirement. Data to support recommended design criteria are presented. These criteria should be modified as data become available to justify an adjustment.

The rational approach to design uses the relationship between coliform removal and transferred ozone dosage reported by several investigators. An example relationship was presented in Section 6.3.2, Figure 6-13 (17). In Figure 6-13 total coliform removal (log (N_o/N)) increased as the transferred ozone dosage (log T) increased. A linear-log relationship was indicated for the approximately 100 data points over a range of transferred ozone dosage from 1.5 mg/l to 22 mg/l. The slope of the line was 2.51, intercept, 0.76; and correlation coefficient, 0.76. Other investigators have reported a similar relationship between coliform reduction and transferred ozone dosage, although the slope and intercept of the individual lines are quite variable (18,22,24,27). The regression line of best fit of Figure 6-13 was rearranged, as shown in Figure 6-40, in order to depict the dose/response curve in a form that can be more readily used for design. The equation of the line in this form becomes:

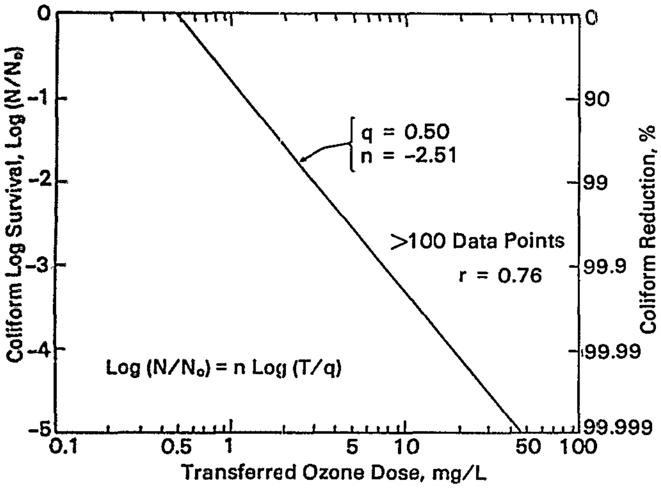
$$\text{Log}\,(N/N_o) = n * \text{Log}(T/q) \qquad (6\text{-}4)$$

where:

T = transferred ozone dosage (mg/l)
N = effluent coliform concentration (#/100 ml)
N_o = influent coliform concentration (#/100 ml)
n = slope of dose/response curve
q = X-axis intercept of dose/response curve, which is the amount of ozone transferred before measurable kill is observed.

By mathmatical rearrangement the slope (n) of the dose/response curve for Stover's results (see Figure 6-13) was calculated as -2.51, and the X-axis intercept (q) was 0.50 mg/l. These results are presented in Table 6-12, along with results obtained by other investigators. A wide range of slope and intercept data are indicated, but individual results

Figure 6-40. Dose response curve for nitrified effluent at Marlborough (17).



appear to accurately describe each operating condition as indicated by the good correlation coefficient obtained in each study. Therefore, specific differences are assumed to contribute to the variation in results for the X-axis intercept and slope. Possible differences are further discussed.

Table 6-12. Summary of Dose/Response Curve Slopes and Intercepts for Various Ozone Disinfection Research Studies

Name of Plant	Slope	Intercept	Data Information Corr	No. of Points	Effluent COD mg/l	Reference
Pilot Scale Test Results						
Muddy Creek (fecal)	−3.96	0.95	−0.80	25	29	27
Cincinnati (fecal)	−5.50	1.00	−0.99	>100	30	22
Indian Creek (fecal)	−4.14	1.04	−0.92	25	26	27
Loveland (fecal)	−4.62	1.05	−0.83	25	39	27
Fairfield (fecal)	−4.97	1.68	−0.78	25	39	27
Sycamore (fecal)	−6.65	1.93	−0.95	25	38	27
Mill Creek (fecal)	−3.08	4.95	−0.81	25	74	27
RBC Effluent (fecal)	−3.90	0.70	−0.82	13	13*	24
Lagoon Effluent (fecal)	−4.60	3.40	−0.92	30	95*	24
Screened Effluent (fecal)	−2.90	1.80	−0.89	331	92*	24
Screened Effluent (total)	−2.60	2.10	−0.85	329	95*	24
"Strong" Effluent (fecal)	−3.10	12.50	−0.93	16	—	24
Field-Scale Test Results						
Marlborough (total)	−2.51	0.50	−0.76	>100	21	17
Marlborough (total)	−3.14	1.05	−0.72	>140	40	17
Vail (fecal)	−3.15	0.79	−0.89	12	25	18

*Indicates BOD concentration

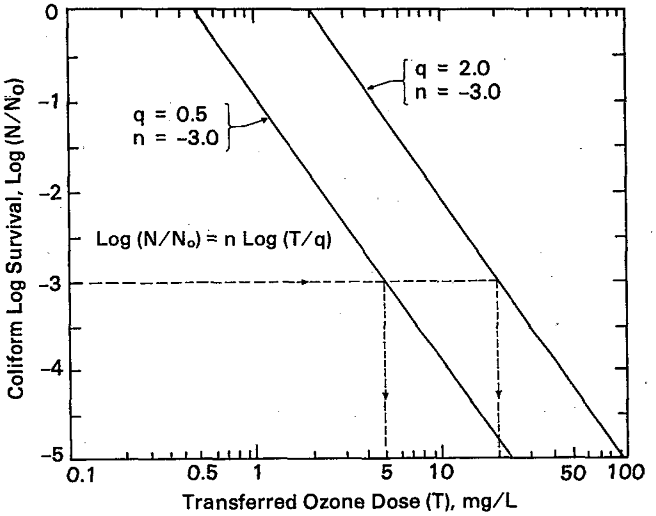
It should be noted that the X-axis intercept (i.e., transferred ozone dosage at 100 percent coliform survival) of the dose/response curve is calculated from the data; it is not a measured value. It is improbable that a straight-line relationship occurs near the X-axis intercept because some degree of coliform reduction would be expected to occur immediately as ozone is transferred to the wastewater. However, from a practical standpoint this reduction is insignificant and the X-axis intercept is defined as the transferred ozone dosage where effective disinfection begins to occur.

Note that the term, log-coliform survival, can also be described as percent reduction of coliform organisms. In Figure 6-40 both percent reduction and log-coliform survival are shown.

In the presentation of this design approach the X-axis intercept is called the initial ozone demand of the wastewater. Generally, initial ozone demand will increase as the quality of the wastewater deteriorates. Factors affecting initial ozone demand are organic and inorganic materials in the wastewater that are readily oxidized by ozone, such as iron, nitrite nitrogen, and manganese; materials that affect the COD concentration; and other materials. A combination of these materials typically affects the initial ozone demand. Limited data are available to be able to quantify the ozone demand of a particular wastewater; however, from Table 6-12 some general trends are indicated based on wastewater COD concentration. The X-axis intercept for the studies of the wastewaters with a low COD (20 to 30 mg/l) were between 0.5 and 1.0 mg/l; with a moderate COD (30 to 40 mg/l) between 1.0 and 2.0 mg/l; and with a high COD (74 mg/l) about 5 mg/l. These data may be used to estimate an X-axis intercept for ozone process design, but conservative estimates may be appropriate considering the limited data base that is available. It is recommended that pilot or bench-scale testing be completed to better define the X-axis intercept (i.e., initial ozone demand).

The selection of the X-axis intercept will affect the transferred ozone dosage requirement, as shown in Figure 6-41. For a high quality wastewater with an initial ozone demand of 0.5 mg/l a projected transferred ozone dosage of 5 mg/l would be required to achieve a 3-log reduction in coliform organisms when the slope of the dose/response curve is -3.0. For a wastewater with an initial ozone demand four times greater (2.0 mg/l), the projected transferred ozone dosage is four times greater (20 mg/l) to meet the same level of disinfection. Ozone disinfection effectiveness is highly dependent upon the initial demand for ozone. Wastewaters with a potential high initial ozone demand may not be good candidates for ozone disinfection systems.

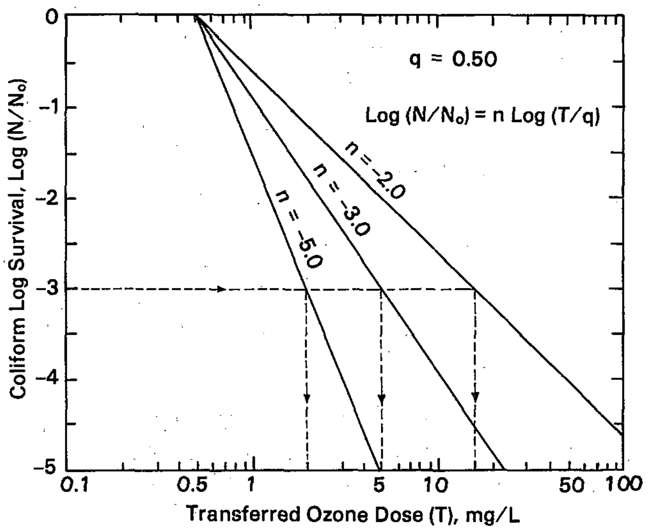
Figure 6-41. **Example curve showing the effect of different X-axis intercepts on transferred ozone dosage requirement.**



The slope of the dose/response curve represents the change in coliform survival per mg/l transferred ozone dosage. The effect of slope on transferred ozone dosage required to achieve a 3-log reduction in coliform organisms is shown in Figure 6-42, assuming the X-axis intercept is 0.5 mg/l. At a relatively steep slope of -5.0, only 2 mg/l transferred ozone dosage is required. At a flatter slope of -3.0, a projected transferred dosage of 5 mg/l is required and at a slope of -2.0 a dosage of 16 mg/l would be necessary.

The slope of the dose/response curve will become flatter when the disinfection performance capability of the ozone contact basin is poorer, or when ongoing chemical reactions with ozone reduce the effectiveness of the disinfectant. For wastewaters that have similar water quality characteristics and a similar initial ozone demand, it is anticipated that the long-term ozone reactions would be similar. Therefore, for these conditions and in the rational design approach, the different slopes are considered to be primarily a function of the disinfection performance capability of the contactor.

Figure 6-42. **Example curve showing the effect of different slopes on transferred ozone dosage requirement.**



The slope of the dose/response curves reported in Table 6-12 vary considerably, ranging from -2.51 to -6.65. A wide variation also exists when the wastewater quality is similar (i.e., initial ozone demand less than 1.1 mg/l and COD concentration less than 40 mg/l), ranging from -2.51 to -5.5. However, in all cases the pilot-scale contact basin disinfection performance capability was much better than the capability of the field-scale units, as evidenced by the consistently steeper slopes of the pilot-scale units. Better plug flow capability in the small scale reactors and corresponding reduced potential for short-circuiting are believed to account for the improved disinfection performance of the pilot-scale systems. The potential superior performance of pilot-scale contact basins should be considered when scaling-up pilot plant results to full-scale design.

An overall review of the dose/response data presented indicates general design criteria that may be used for a rational approach to the determination of the design transferred ozone dosage. For a good quality secondary treatment plant effluent (COD less than 40 mg/l), an initial ozone demand of 1.0 mg/l appears reasonable. If a poorer quality wastewater is anticipated, a higher initial ozone demand should be selected. Conversely, a lower initial ozone demand can be selected if a high quality wastewater is to be disinfected. Pilot plant results may be used to obtain a reasonably good estimation of the initial ozone demand.

The slope of the dose/response curve is more difficult to establish. Pilot plant results were generally better than full-scale performance capability; however, the pilot plants had three stages, while the full-scale plants had only one and two. If the field-scale ozone contact basins are designed to match the performance capability of the pilot-scale units (i.e., multiple stages), then the steeper slopes, -4.0 to -5.0, may be used in design. Disinfection efficiency design considerations for the ozone contact basin are discussed in Section 6.4.3.2. Otherwise, a flatter slope of -3.0 appears justified.

A summary of applicable guidelines for determining the transferred ozone dosage requirement is presented below:

a. The approach for determination of transferred ozone dosage may be used for ozone process design.

   - The initial ozone demand can be estimated based on the quality of the wastewater treated. For a good quality secondary treatment plant effluent (COD less than 40 mg/l and negligible nitrite nitrogen), an initial ozone demand of 1.0 mg/l appears reasonable.
   - The slope of the dose/response curve can be based on design features that enhance contact basin disinfection capability. For a contact basin with good design features that emulate reported pilot scale performance, a slope of -4.0 to -5.0 may be used. Otherwise, a flatter slope should be used.
   - Influent coliform concentration should be determined based on existing data, if available, or on reported concentrations for similar plants.
   - Effluent coliform concentration should be based on the most stringent design limitations.

b. To properly establish the transferred ozone dosage requirement, pilot testing should be conducted for all wastewaters and especially for unique ozone disinfection applications such as:

   - Disinfection of "strong" or highly industrial wastewaters.
   - Disinfection to achieve permit standards more stringent than the former EPA standard of 200 fecal coliforms per 100 ml.
   - Disinfection using a type of ozone contact basin that does not have a proven record of performance.

c. Literature-reported ozone dosages may be used for conventional applications of ozone disinfection.

   - A transferred ozone dosage between 4 and 10 mg/l appears satisfactory to meet the former EPA standard of 200 fecal coliforms per 100 ml, when disinfecting a good quality secondary or tertiary treatment plant effluent in a properly designed ozone contact basin.
   - A transferred ozone dosage between 15 and 20 mg/l reportedly meets the stringent standard of 2.2 total coliforms per 100 ml, when disinfecting good quality tertiary plant effluent in a properly designed ozone contact basin.
   - A transferred ozone dosage between 36 and 42 mg/l reportedly meets the stringent standard of 2.2 total coliforms per 100 ml, when disinfecting highly polished secondary treatment plant effluent in a properly designed ozone contact basin.

Determination of Applied Ozone Dosage. The applied ozone dosage is the mass of ozone from the generator that is directed to a unit volume of the wastewater to be disinfected. Design considerations for determining transferred ozone dosage (T) were discussed in the previous section, and for determining ozone transfer efficiency (TE) in Section 6.4.3.1. The following equation can be used to determine applied ozone dosage.

$$D = T * 100/TE \qquad (6\text{-}5)$$

By determining the applied ozone dosage and the design wastewater flow rate, the design ozone generation system production capacity can be established. (See Section 6.2.3.2 for the procedure to calculate ozone production).

6.5.2 Feed-Gas Supply and Process Flexibility

The design feed-gas flow rate to the ozone generator is dependent upon the design ozone production rate and design ozone concentration. Determination of ozone production rate was discussed in Section 6.5.1. Determination of ozone concentration was discussed in Section 6.4.1.2. From these data the design feed-gas flow rate can be determined using the equations discussed in Section 6.4.1.2.

The peak design feed-gas flow rate is a fairly straightforward calculation. However, most ozone disinfection systems seldom operate at the peak ozone production capacity. Requirements at start-up are usually less than design conditions, and requirements at average design conditions are typically less than at peak conditions. Process flexibility must be installed to provide economical operation at variable ozone production rates. Flexibility in ozone generation equipment and feed-gas supply equipment are most important. Also, flexibility must be provided to obtain reliable, continuous operation. Design considerations to address process flexibility are further discussed.

Ozone production by most ozone generators can be reduced to at least 20 percent of the peak production rate, and often lower, by reducing the power supply to the generator. Therefore, flexibility is available for the ozone generator to reduce power consumption while operating between 20 and 100 percent of peak capacity.

Typically, at least two ozone generators of similar capacity are provided to allow for continuous operation during routine maintenance, such as cleaning the generator. Under these conditions the minimum ozone production rate is 10 percent of the peak production rate. If the projected minimum ozone production requirement is less than the minimum production rate, additional flexibility in ozone generation equipment must be provided. This flexibility may be obtained with additional, smaller generators or by providing capability to remove some of the dielectrics from service (18).

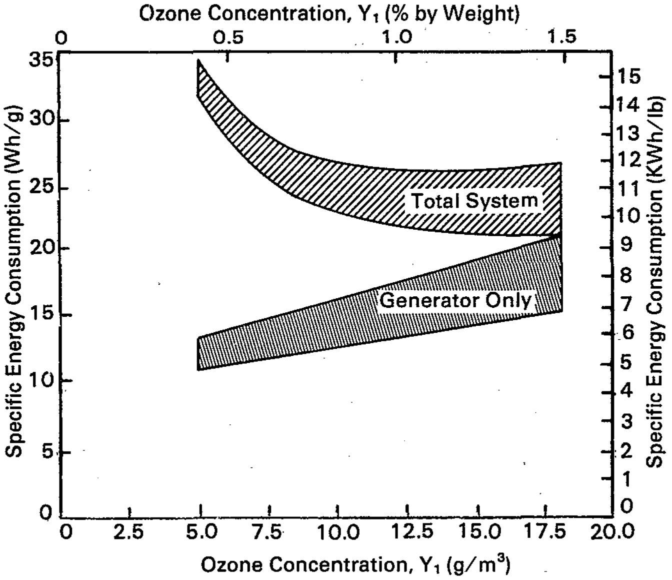
Flexibility in the feed-gas supply and treatment process is necessary for several reasons, but it is especially important for energy conservation and process reliability considerations. However, providing flexibility in this auxiliary equipment is typically more involved and more difficult to obtain than providing flexibility in the ozone generator's air pre-treatment process. Flexibility is more involved because it includes design considerations for piping as well as equipment size and type. It is more difficult to achieve because of equipment capabilities, especially the capability to vary the air flow rate from the air compressors and at the same time reduce power consumption.

When oxygen is the feed-gas, the oxygen requirement of the ozone disinfection system must be coordinated with the oxygen requirement of the biological treatment process. Balancing these oxygen needs was discussed in Section 6.3.2. Generally, fine-tuning the system to achieve a precise oxygen supply balance is difficult in full-scale applications because of the difficulty in rapidly adjusting the oxygen supply from the oxygen production facilities. Before an oxygen supply control system is designed, it is recommended that existing oxygen-fed ozone disinfection plants be visited to obtain information on problems encountered with the oxygen supply control approaches that have been attempted.

When air is the feed-gas, energy efficient operation may be improved by providing flexibility in the air supply equipment. The air compressor(s) are typically the second largest consumer of energy in the ozone disinfection system, and may even exceed the consumption of the ozone generator in some instances. The compressor(s) are usually sized to meet the design ozone concentration at the ozone generator's peak production rate. At these peak design conditions the total or overall system-specific energy consumption is best, as shown in Figure 6-43. The total system-specific energy consumption averaged about 23 Wh/g (10.5 kWh/lb) at the typical design ozone concentration of 18 g/m^3 (1.5 percent wt). However, when the ozone generator power level was at its lowest setting and the feed-gas flow rate was not adjusted (i.e., ozone concentration was 5 g/m^3 (0.4 percent wt), the total system-specific energy consumption was significantly higher at 33 Wh/g (15 kWh/lb).

In Figure 6-43 a range of specific energy consumption data is shown. The variability is due to type of equipment selected and operating conditions, such as dew point, dielectric cleanliness, temperature, etc. The range shown is typical of most air-fed ozone generation systems and may be used to estimate energy consumption for proposed ozone disinfection processes. A summary of applicable design considerations to achieve energy efficient operation and process reliability is listed below.

Figure 6-43. **Specific energy consumption for a typical air-fed ozone generation system.**



a. Specific energy consumption should be lowest at expected operating conditions, rather than only at peak ozone production rates.

b. Multiple air compressors and/or variable speed compressors should be considered to reduce the air flow rate and reduce energy consumption.

c. The ozone concentration should not be greater than the design concentration in order to avoid excess heat build-up in the ozone generator and damage to generator dielectrics.

d. Independent skid-mounted generation systems should be avoided. If skid-mounted units are provided, then the equipment should be interconnected with appropriate piping so that each component can be operated independent of other components on the skid.

6.5.3 Ozone Process Control and Automation

The amount of ozone system control and automation has ranged from almost negligible capability to very extensive provisions. The purpose of addressing control and automation has been to reduce energy consumption rather than to balance disinfection capability with water quality effects, as is the case with chlorine disinfection systems. Overdosing with ozone does not create a water quality problem because the ozone will simply "degrade" back to oxygen. Factors to consider in evaluation of process control and automation requirements are discussed in this section.

The primary consideration in an evaluation of process control and automation needs is the potential for payback of the capital investment. The issues to consider are the complexity of the system for the size of the plant and the potential for savings as affected by anticipated variations in ozone production (i.e., variations in wastewater flow and quality). Generally, operating flexibility to minimize energy consumption should be provided, as discussed in the previous section. Operating flexibility not only provides energy savings potential, but also improves process reliability and enhances system maintenance. Process automation, on the other hand, should be evaluated on a case-by-case basis.

Automation of the auxiliary equipment, especially air supply and treatment equipment, is typically more complex than automation of the ozone generator. Unless the plant is quite large or variation in ozone production extensive, it is generally cost effective to control the auxiliary equipment manually. Manual adjustments might include adding or deleting an air compressor or increasing or decreasing the compressor speed at certain times of the day, week, season, or year, in order to operate in an acceptable range of ozone concentration from the ozone generator. If automated control is provided, the number or speed of the air compressor(s) and number of ozone generators on-line is directly controlled by the ozone concentration.

Typically, the ozone production requirement to achieve disinfection is going to vary on a routine basis. This production requirement can be adjusted by changing the power setting of the ozone generator(s). Three approaches to analyzing and controlling the generator power settings have been used, including applied ozone dosage control, residual ozone concentration control, and off-gas ozone concentration control. The approach, advantages, and disadvantages of each method are further discussed.

6.5.3.1 Applied Ozone Dosage Control

The simplest method of ozone system process control is to make adjustments to the ozone production rate so that the applied ozone dosage is maintained at a constant value. Once the applied ozone dosage has been selected the only other variable involved is the wastewater flow rate. The adjustments to ozone production to maintain a specified applied ozone dosage may be completed by manual means, or can be completed automatically. The primary advantage of this method of process control is the simplicity of the system. The primary disadvantage is that the control method is completely unresponsive to changes in water quality.

Manual control using the applied ozone dosage method requires that the operator know the wastewater flow rate being treated and the ozone production rate of the ozone generator. The operator then adjusts the generator power setting to maintain the desired applied dosage as the wastewater flow rate varies.

The applied ozone dosage method may be fairly simple or can be quite complex. A simple method of automating the applied dosage control technique involves varying the power supply to the ozone generator in proportion to the wastewater flow rate. This method is best used when the ozone production rate is linear to the generator power setting. A more common method is to use a microprocessor control system that receives information for wastewater flow rate, feed-gas flow rate and ozone concentration; calculates the applied ozone dosage; and compares this figure with the set-point value. The output signal increases or decreases the ozone generator power supply so that the calculated applied ozone dosage is equal to set-point. This will cause some changes in the ozone concentration since the feed-gas flow rate remains the same. Large variations in ozone demand will require adjustments to the air supply rate.

The microprocessor method of automated process control is fairly reliable, if reliable instrumentation is provided. The applied ozone dosage may be maintained at the desired set-point over a fairly wide range of wastewater flow rates. The set-point may be changed as disinfection performance indicates, although the operator is usually not aware that a change is needed until the bacteriological test results are available. The method works best when the wastewater quality and disinfection performance do not vary considerably.

6.5.3.2 Wastewater Ozone Residual Control

The ozone residual control method is effective because disinfection performance is closely related to ozone residual. However, the method is generally unreliable because it is difficult to maintain calibration of the residual monitors as liquid characteristics change (57). Also, the probes tend to foul-up in the wastewater environment, although improvements in probes are continually being made. Attempts to use the residual ozone control method must be coupled with a commitment to keep the monitors calibrated. According to Grunwell and Gordon, the Indigo method or the arsenic (III) method of chemical analyses may be used as a laboratory check of in-line ozone residual meters (8,57).

The ozone residual control method is fairly straightforward from an instrumentation standpoint. The signal from the residual monitor is sent to a microprocessor, which compares the signal received to the value set by the operator. The output signal is sent to the ozone generator power supply which changes ozone production until the residual ozone concentration is equal to the set-point value. A disadvantage of this control scheme is the time delay between the residual monitor signal to the microprocessor due to the wastewater contact time. Also, improperly designed systems could produce excessive "hunting" of the equipment.

6.5.3.3 Off-Gas Ozone Concentration Control

The off-gas ozone concentration control method was recently presented as a possible control approach for wastewater disinfection (15,19,58). It was proposed based upon an evaluation of pilot-scale disinfection performance data. A trend of improved disinfection occurred as the off-gas ozone concentration increased. Initially, the method was proposed as applicable when the ozone feed-gas to wastewater liquid flow ratio (G_1/L_1) was constant. However, more recent testing has indicated that strict control over the G_1/L_1 ratio is not required, if the G_1/L_1 ratio is kept between 0.1 and 0.5 (15,18,58).

The off-gas control method has been used at the Vail, Colorado wastewater treatment plant (18). However, the system was not automated. The plant operators simply check the monitor reading on a routine basis, and make adjustments to the ozone generator power setting to maintain a pre-established range of off-gas ozone concentration.

Currently, an automated off-gas control system has not been installed or tested. However, the method shows promise of being an effective process control device. The biggest advantage is that instrumentation seems to be quite reliable, while at the same time responsive to water quality and wastewater flow rate changes.

6.5.4 Ozone Disinfection Design Example

The example presented in this section of the manual is intended to illustrate the design considerations involved with development of an ozone disinfection system. The wastewater treatment design information for the example design problem is shown in Table 6-13. The plant is an air activated sludge plant located at an elevation of 3,500 ft (1,067 m) above sea level. The influent to the ozone disinfection system is the effluent from the secondary clarifiers.

Step 1—Determine the Transferred Ozone Dosage, Applied Ozone Dosage, and Ozone Production Design Values.

Method A. Based on a literature search of existing facilities (See Section 6.5.2.2) the design applied ozone dosage ranged from 4 to 10 mg/l at plants required to meet the 200/400 fecal coliforms per 100 ml standard. The design transfer efficiency ranged from 85 to 90 percent. Compare these data with the rational approach to design presented in Method B.

Method B. Refer to Section 6.5.1.2 and develop the design ozone dosage based on a rational approach to design. In developing the rational design values, two important assumptions are required including the initial ozone demand and the slope of the dose/

Table 6-13. Ozone Disinfection System Criteria for Design Example Problem

Average Daily Wastewater Flow	7.5 mgd
Peak Daily Wastewater Flow	15.0 mgd

NOTE: The daily peak flow rate will not exceed 15.0 mgd because of storm flow equalization facilities.

Start-up Daily Average Wastewater Flow	3.5 mgd
Start-up Peak Daily Wastewater Flow	7.5 mgd
Average Effluent BOD/TSS	15/15 mg/l
Maximum Daily Effluent BOD/TSS	30/30 mg/l
Design Required Effluent Fecal Coliform	
Weekly Maximum Limitation	400 per 100 ml
Geometric Mean Monthly Limitation	200 per 100 ml
Disinfection System Influent Fecal Coliform	
Geometric Mean Concentration	500,000/100 ml
Maximum Concentration	200,000/100 ml
Chlorine Residual	≤0.05 mg/l

response curve. According to the discussion in Section 6.5.1.2, the initial ozone demand at most secondary treatment plants was about 1.0 mg/l. Therefore, in this design example the initial ozone demand is assumed to be 1.0 mg/l.

The discussion in Section 6.5.1.2 indicated that the slope of the dose/response curve was more difficult to establish. Evaluations using pilot-scale contactors reported a slope between -4.0 to -5.0, while reported field-scale data indicated a flatter slope of -3.0. The transferred dosage for a slope of -3.0, initial ozone demand of 1.0 mg/l, and coliform log survival of -4.0 is calculated below using Equation 6-6, which is a rearrangement of equation 6-4.

$$T = q * 10^{[\mathrm{Log}(N/N_o)]/n} \qquad (6\text{-}6)$$

where

q = 1.0 mg/l
n = –3.0
N = 200 per 100 ml
N_o = 2,000,000 per 100 ml

then

T = 21.5 mg/l

The transferred ozone dosage for a slope of -4.0 and -5.0 was also calculated for an initial ozone demand of 1.0 mg/l and log survival of -4.0, and was compared with the dosage determined for the slope of -3.0.

Slope of Dose/ Response Curve	Transferred Dosage
-3.0	21.5 mg/l
-4.0	10.0 mg/l
-5.0	6.3 mg/l

As shown, the slope of the dose/response curve has a significant impact on the resulting transferred ozone dosage. The flatter slope reported for field-scale evaluations may be due to "scale-up" factors, or may be due to the shape of the contactors. The two field-scale contactors had only one and two stages of contacting, while the pilot-scale contactors had three stages of contacting. Contactor staging has a significant impact on disinfection performance, as discussed in Section 6.4.3.2.

It should be noted that two alternative design considerations exist relative to obtaining desired disinfection performance at reasonable ozone dosages. One alternative is to provide a tertiary filter, which reportedly will reduce the influent coliform concentration by a factor of 10 (i.e., 1 log reduction) (17). If filters were installed and the maximum influent coliform concentration were reduced from 2,000,000 to 200,000 per 100 ml, then the log survival is -3.0 instead of -4.0. The resulting transferred ozone dosage for a dose/response curve slope of -3.0 is 10.0 mg/l, which is less than half of the dosage required at a log survival of -4.0. Further, the 10 mg/l dosage is equal to the dosage required when the log survival was -4.0 and the slope was -4.0. Therefore, the effect of reducing the influent coliform concentration by one log is similar to the effect of improving the slope of the dose/response curve from -3.0 to -4.0.

The second alternative to obtaining desired disinfection performance at reasonable ozone dosages is to provide contact basin design features that improve disinfection performance and achieve a steeper dose/response slope of -4.0 to -5.0 instead of -3.0. For this design example it is assumed that a tertiary filter is not an available option. Therefore, a coliform log survival of -4.0 will be used to establish the required transferred ozone dosage.

Currently, data are not available to document the optimum number of stages for an ozone contact basin. Pilot-scale data indicate that three stages consistently achieved a log survival between -4.0 and -5.0. However, the stages must be designed to simulate plug flow conditions (i.e., eliminate short-circuiting), which apparently was the case with the pilot-scale contactors. For this design example it is proposed that a 4-stage contactor will be used and that the slope of the dose/response curve will be -4.0. Also, it is assumed that the contactor diffuser depth will be 5.5 m (18 ft), the ozone transfer efficiency will be a minimum of 85% at the plant elevation of 1,070 m (3,500 ft), and the detention time will be 15 minutes at the average flow rate. It should be noted that the conditions for applied ozone dosage obtained from these criteria range from 3 to 11.6 mg/l, and therefore are consistent with the literature reported dosages of 4 to 10 mg/l presented in Method A above.

Based on the discussion above the following design basis is used for the example design problem.

Initial Ozone Demand	1.0 mg/L
Slope of Dose/Response Curve	-4.0
Ozone Transfer Efficiency	85%
Type of Contactor	Bubble Diffuser
Number of Stages	4
Number of Basins	2
Contact Basin Detention Time at Average Flow	15 min
Diffuser Depth	18 ft (5.5 m)

Step 2—Determine the Design Ozone Production Rate.

Calculate the transferred ozone dosage, applied ozone dosage, and ozone production rate for the various operating conditions that may occur, as shown in Table 6-14.

Table 6-14. Transferred Ozone Dosage Calculations for Design Example

Trial Number	Wastewater Flow (mgd)	Effluent Fecal Coliform (#/100 mL)	Influent Fecal Coliform (#/100 mL)	Log Coliform Survival	Trans Dose (mg/L)	Applied Dose (mg/L)	Applied Mass (lb/d)
1	15	200	10,000	−1.70	2.66	3.13	391
2	15	200	500,000	−3.40	7.07	8.32	1,041
3	15	200	2,000,000	−4.00	10.00	11.76	1,472
4	7.5	200	10,000	−1.70	2.66	3.13	196
5	7.5	200	500,000	−3.40	7.07	8.32	520
6	7.5	200	2,000,000	−4.00	10.00	11.76	736
7	3.5	200	10,000	−1.70	2.66	3.13	91
8	3.5	200	500,000	−3.40	7.07	8.32	243
9	3.5	200	2,000,000	−4.00	10.00	11.76	343

Shown below is an example calculation for Condition 1, and shown in Figure 6-44 is a graph of the production results for all conditions.

$$T = q * 10^{[\mathrm{Log}(N/N_o)]/n}$$

where

q = 1.0 mg/l
n = −4.0
N = 200 per 100 ml
N_o = 10,000 per 100 ml

then

T = 2.66 mg/l

$D = T * 100/TE$

where

TE = 85%

then

D = 3.13 mg/l

$P = D * L * 8.34$

where

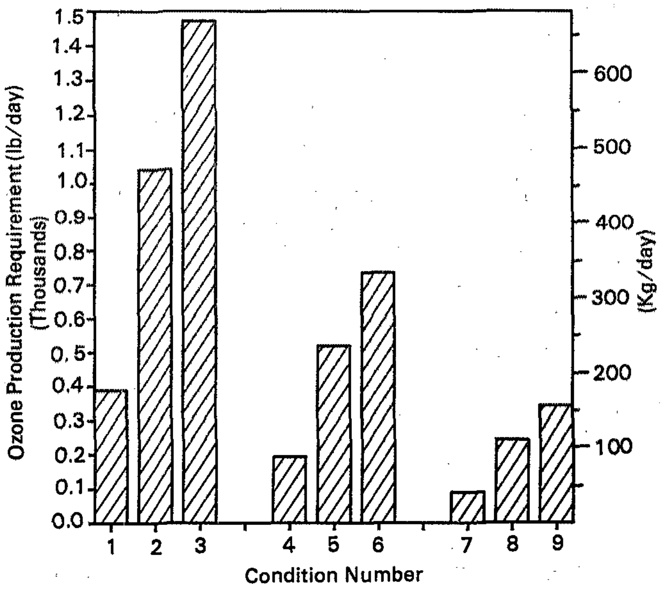
L = 15 mgd

then

P = 391 lb/d (123 g/min)

From Figure 6-44 and Table 6-14, the ozone production rate that would satisfy all assumed design conditions is 463 g/min (1,472 lb/d). Therefore, for this design example the design ozone production rate is assumed to be 473 g/min (1,500 lb/d).

Figure 6-44. Design example projected ozone production rate for various operating conditions.



Step 3—Select the Number of Ozone Generators

The considerations for selection of the number of ozone generators is to address both the maximum and the minimum expected ozone production rates. If three ozone generators each with a capacity of 158 g/min (500 lb/d) were provided, the maximum production rate would be met. However, if the maximum production rate were expected to occur on a frequent basis, then a fourth generator or additional capacity of the three generators should be provided so that the generators are not required to operate at peak production rates for extended periods of time (See Section 6.4.1.2).

The minimum production rate from an ozone generator is 10 to 20 percent of its maximum production capability. From Figure 6-44 and Table 6-14 the minimum expected ozone production rate at start-up conditions is 29 g/min (91 lb/d). If the capacity of each generator were 158 g/min (500 lb/d), then the minimum ozone production rate would be 16 to 32 g/min (50 to 100 lb/d) and would match the projected requirements at start-up conditions.

For this design example three ozone generators each with a capacity of 158 g/min (500 lb/d) are proposed. Also, room for a fourth generator will be provided in case the design peak ozone production rates are required for extensive periods of time. Since air activated sludge is used for secondary treatment (i.e., oxygen activated sludge is not used; thus a readily available source of oxygen is not present), an air-fed ozone system is proposed.

Step 4—Determine the Size and Number of Air Compressors

Proposed Design Basis:

1. The maximum ozone concentration is 18 g/m^3(1.5 percent wt).
2. A low pressure air-fed system is used.
3. The desiccant dryer purge air flow is 20 percent.

Calculate the generator air flow rate and total system air flow rate at both peak design and average start-up conditions.

At Peak Design Conditions

Generator air flow rate
= 1,500 lb/d/(0.015 lb O_3/lb air)/(0.0753 lb air/ft^3)/1,440 min/d
= 922 scfm (21.1 m^3/min)
Total system air flow rate
= 922 scfm/0.8
= 1,153 scfm (32.7 m^3/min)

At Average Start-Up Conditions

Generator air flow rate
= 243 lb/d/(0.015 lb O_3/lb air)/(0.0753 lb air/ft^3)/1,440 min/d
= 149 scfm (4.2 m^3/min)
Total system air flow rate
= 149 scfm/0.8
= 187 scfm (5.3 m^3/min)

Three air compressors each with a capacity of 11.3 m^3/min (400 scfm) would provide the peak design flow rate. However, in order to provide flexibility to minimize power at start-up conditions (See Section 6.5.2), at least four air compressors should be provided (i.e., 2 at 11.3 m^3/min (400 scfm) and 2 at 5.7 m^3/min (200 scfm). In addition, at least one of the larger plus one of the smaller compressors should be provided with the capability to vary output and reduce power using a variable speed drive. Determine the compressor pressure rating using hydraulic calculations for maximum air flow rate and include all appurtenances such as valves, filters, dryers, meters, diffusers, water depth in contactor, etc.

Step 5—Determine the Size and Number of Desiccant Dryers

Proposed Design Basis:

1. The operating pressure is 93 kPa (13.5 psig)
2. The inlet air temperature is 30°C (86°F)
3. With an after cooler the desiccant dryer inlet air temperature is 20°C (68°F)
4. With a refrigerant dryer the desiccant dryer inlet air temperature is 4.4°C (40°F)
5. The required dew point temperature (@ 1 atm pres) is 60°C (76°F)
6. The desiccant design cycle time is 16-hr
7. The maximum desiccant loading is 8 kg (18 lb) of water received during one complete drying cycle per pound desiccant
8. The design gas flow rate is 34 m^3/min (1,200 scfm)

Step 5A—Case without a Refrigerant Dryer and with an Aftercooler

See Section 6.2.3.3, Example 6.1, for the procedure to calculate the moisture loading to the desiccant dryer for the above conditions. From Example 6.1, moisture loading was 10.1 kg/hr (22.2 lb/hr) when the gas flow rate was 17 m^3/min (600 scfm). Therefore, at a gas flow rate of 34 m^3/min (1,200 scfm), the moisture loading is 20.2 kg/hr (44.4 lb/hr). Using this moisture loading, determine the amount of desiccant required for all desiccant towers that are drying.

Desiccant/tower = 18 lb/lb * 44.4 lb/hr * 16 hr
= 12,790 lb/towers that are drying (5,800 kg/towers)

Consider using three desiccant dryers, each with two towers. Determine the desiccant amount per tower. NOTE: The amount for the dryer is twice the amount per tower, since there are two towers per dryer.

Desiccant/tower/dryer = 12,790 lb/tower/3 dryers
= 4,265 lb/tower/dryer (1,935 kg/tower/dryer)

Step 5B—Case with a Refrigerant Dryer

Calculate moisture loading from refrigerant dryer.

Interpolating from Table 6-7 the moisture content at 4.4°C (40°F) is 0.394 lb/1,000 ft^3.

Operating absolute pressure is:

Pressure at altitude of 1,070 m (3,500 ft) is 12.93 psi
Operating gauge pressure is 13.5 psi
Operating absolute pressure is 13.5 + 12.93 = 26.43 psia (89 kPa)

Calculate moisture content at operating pressure

= (0.394 lb/1,000 ft^3) * (14.7 psia/26.43 psia)
= 0.219 lb/1,000 ft^3

Calculate moisture loading

= (0.219 lb/1,000 ft^3) * (1,200 ft^3/min * 60 min/hr)
= 15.8 lb/hr (7.2 kg/hr)

Calculate amount of desiccant for all operating towers

Desiccant/tower = 18 lb/lb * 15.8 lb/hr * 16 hr
= 4,550 lb/towers that are drying (2,065 kg/towers)

Consider using three desiccant dryers, each with two towers. Determine the desiccant amount per tower. NOTE: The amount for the dryer is twice the amount per tower, since there are two towers per dryer.

Desiccant/tower/dryer = 4,550 lb/towers/3 dryers
= 1,517 lb/tower/dryer (690 kg/tower/dryer)

The analysis of the amount of desiccant required for the case with and without a refrigerant dryer indicates that significantly less desiccant is required (i.e., desiccant dryers would be smaller) when a refrigerant dryer is provided. An option exists as to whether or not to provide the refrigerant dryer in conjunction with the desiccant dryer, or to eliminate the refrigerant dryer from the process because of its sensitive operation (See Section 6.4.2.4). In some instances, the desiccant dryer has been sized to operate without the refrigerant dryer, yet a refrigerant dryer was installed to provide flexibility for a case when problems may occur with the desiccant dryer (18). If one desiccant dryer were out of service, and normally three would be required, the refrigerant dryer could be operated to reduce the moisture loading to the other two operating desiccant dryers. In this regard, the refrigerant dryer could be considered a back-up unit for the desiccant dryers. A most important consideration is to absolutely never minimize the importance and impact of the desiccant dryer. The dryer should never be undersized.

The design calculations shown above are a portion of the overall design considerations of an ozone disinfection process. Other considerations include instrumentation, layout, flexibility, access for maintenance, type of materials, alarm systems, type of equipment, control features, automation, etc. The engineer should include all this design related information in the contract documents so that all equipment manufacturers can bid the ozone equipment on an equal basis. The various sections within this manual are intended to assist the engineer in making these design decisions.

6.6 Safety

Ozone is a toxic gas, and like chlorine can cause severe illness and death if inhaled in sufficient quantity. However, ozone systems have safety advantages not available with the chlorine disinfection process. Ozone is generated on-site, thus eliminating transportation hazards. Also, the generation system can be shut down if an ozone leak develops. Another safety advantage is the physical characteristic of ozone that allows it to be detected (smelled) at concentrations much lower than harmful levels.

In addition to safety precautions against exposure to ozone, protection against noise and electrical hazards should be incorporated into the design and operation of an ozone disinfection system.

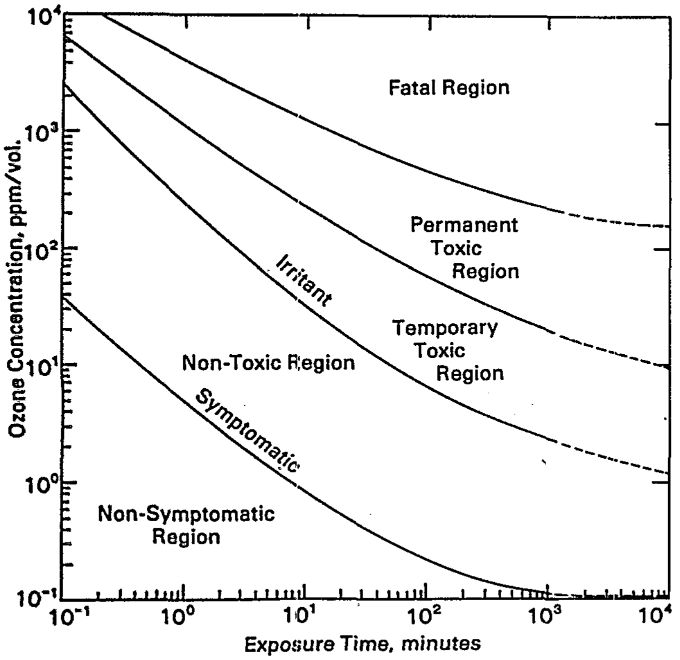
6.6.1 Recommended Exposure Limit to Ozone

A study of the health effects of ozone exposure was conducted by the United States Air Force (59). The results were summarized in the graph shown in Figure 6-45. Another summary of the health effects of ozone was compiled by the American Society for Testing and Materials (ASTM) in support of their recommended standard for limiting human exposure to ozone. The reported biological effects range from dryness of mouth and throat, coughing, headache, and chest restrictions at concentrations near the recommended limit, to more acute problems at higher concentrations.

The recommended ambient ozone exposure levels have been proposed by the Occupational Safety and Health Administration (OSHA), the American National Standards Institute/American Society for Testing and Materials (ANSI/ASTM), the American Conference of Government Industrial (ACGI), and the American Industrial Hygiene Association (AIHA) as follows (1):

> Control occupational exposure such that workers will not be exposed to ozone concentrations in excess of a time weighted average of 0.2 mg/m^3 (0.1 ppm by volume) for eight hours or more per workday, and that no worker be exposed to a ceiling concentration of ozone in excess of 0.6 mg/m^3 (0.3 ppm by volume) for more than 10 minutes.

Figure 6-45. **Human tolerance for ozone (59).**



These recommended limits for ozone concentration are much higher than the concentrations at which ozone can typically be smelled. Generally, an individual can detect ozone at concentrations ranging from 0.02 to 0.1 mg/m^3 (0.01 to 0.05 ppm by volume) (1). The more often a person is exposed to ozone the higher the required concentration for detection.

6.6.2 Ambient Ozone Concentration Monitors

The subject of safety in the design and operation of an ozone system should receive a high priority. All ozone systems should be provided with an ambient ozone monitor or monitors which are set up to measure the ozone concentration at potential ozone-contaminated locations within the plant (e.g., at various places in the room housing the ozone generators). A single monitor may be installed, and the air from different locations pumped to the monitor for detection of ozone concentration. The monitors should be set up to sound an alarm when the ozone concentration reaches 0.2 mg/m^3 (0.1 ppm by volume), and should be set up to shut down the ozone system when the concentration exceeds 0.6 mg/m^3 (0.3 ppm by volume). However, if the ozone equipment is located in an area where "smoggy" days due to ozone levels in the atmosphere are common, higher values may be necessary to be able to detect ozone leakage by the ozone generation system.

A listing of reliable ozone monitors used for ambient monitoring purposes may be obtained from the ozone generation equipment manufacturers and from operating plants. These monitors may be checked and calibrated using extensive calibration procedures (60), but loss of calibration typically has not been a problem. Under normal operating conditions the monitor does not "alarm" and the operators cannot smell ozone. Therefore, on a periodic basis the operators should check the operation of the monitor by directing a small volume of ozone from the generator to the monitor to test the meter's responsiveness. In this manner the operators can be assured that the monitor will respond in case of an ozone leak.

6.6.3 Miscellaneous Safety Considerations

Some ozone generation systems have been labeled as "noisy" installations. However, the source of the noise is usually the feed-gas compressors and not the ozone generators. The ozone generators themselves have a slight "hissing" sound that is typically not objectionable. High frequency generators may give off a high frequency pitch.

Ideally, the feed-gas compressors should be isolated in a room that has some degree of sound proofing. However, in smaller installations this alternative may be economically unattractive. In these situations the operators should wear ear protection equipment when operating or maintaining the ozone equipment.

As with any toxic chemical, the operators should be trained concerning the potential hazards involved and the emergency operating procedures required if a problem occurs. Equipment should be provided to assist the operator. Applicable equipment is listed below:

a. A self-contained breathing apparatus should be provided, and should be located at a place where access is not restricted by ozone in case an ozone leak occurs.

b. An eye-washing sink should be provided to enable the operator to rinse ozone from the eyes, if needed.

c. Safety manuals on performing artificial respiration should be provided.

d. Separate ladders should be provided to enable the operator to enter the ozone contact chamber. Fixed steps in the contact basin should not be relied on.

6.7 References

When an NTIS number is cited in a reference, that reference is available from:

National Technical Information Service
5285 Port Royal Road
Springfield, VA 22161
(703) 487-4650

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Chapter 7

Ultraviolet Radiation

7.1 Introduction

The use of ultraviolet (UV) radiation for the disinfection of wastewaters, relative to the established technologies of chlorination and ozonation, is an emerging process application which has been developed over the past 10 years. As a perspective, three major committee reports were issued in the mid to late seventies (1-3). All effectively described the use of ultraviolet radiation as "potentially" advantageous for the disinfection of relatively high quality treated wastewater. The U.S. Environmental Protection Agency (USEPA) Task Force Report on the Disinfection of Wastewater (1) concluded that: "although ultraviolet light has not been widely used to disinfect wastewater, there is limited information that indicates it may become a potentially desirable alternative. It is the only physical process whereas all the disinfectants are chemical processes. On-going research will provide answers as to its applicability to adequately disinfect wastewater." The Task Force went on to recommend that "the use of alternate disinfectants should be further pursued because of recent findings of the potentially hazardous halogenated organics in drinking water."

This chapter presents current state-of-the-art knowledge on the design of the UV disinfection process, much of which represents information developed over the past decade through research and demonstration of large scale applications of UV to the disinfection of wastewater. As a new application of the technology, the process design procedures are still formative, and in-field experience in the operation and maintenance of UV facilities is limited, but growing as new plants come on line. The underlying conclusion which should be stated at the beginning of this chapter is that the potential which had been foreseen earlier has been confirmed; the recommended investigations into UV and the demonstration of its application on a full scale basis have shown the process to be viable, feasible for application to a wide range of wastewater qualities, effective in the inactivation of pathogens, capable of complying with disinfection goals, and cost-effective. Its advantages lie in its relative simplicity and in the absence of both a residual and any chemical intermediates.

7.1.1 Chapter Description

It is reasonable to state that the UV disinfection process has reached a state of development where the mechanisms are understood and the critical design parameters have been identified and generally demonstrated. Field experience is limited, but gaining. There is not, however, a clear and concise compilation of this information, including the procedures by which a UV system can be designed or evaluated. This manual attempts to provide this, in addition to O&M considerations, which have been identified and demonstrated by direct field experience.

The objective, then, is to bring together the knowledge and experiences with UV as it is applied to the disinfection of municipal wastewaters. It is not the intent of these discussions to give an exhaustive teaching on the technical aspects of the components which make up the UV hardware, e.g., lamps, ballasts, etc.; rather the approach will be to discuss the basic concept and status of these components. References are provided if the reader wishes to pursue these aspects in more detail. The chapter will focus its attention on primary considerations for the design of a system such that it will meet both its performance requirements and will be amenable to efficient operations and maintenance (O&M).

Introductory Section (7.1). This gives an overview of the technology and its current status relative to wastewater disinfection application. Of particular interest are descriptions of UV reactors; these will give the reader a visual perspective of the system, which will be helpful in the subsequent sections on design. The chapter also provides listings of UV plants in Tables 7-2, 7-3, and 7-4, which will be helpful to the designer or operator who wishes to learn about UV disinfection experience at other installations.

Background Discussions (7.2). This section is not critical to the designer. It is useful, however, if one wishes to gain a perspective on the mechanisms of UV inactivation and the evolution of the process as it is applied to wastewater disinfection.

Process Design Considerations (7.3). This is the section most important to the designer. It details the process design elements critical to effective design and offers guidance on defining specific design parameters. These particularly address:

a. Hydraulics (7.3.2); of interest are discussions on dispersion, reactor layout, headloss, and residence time distribution.

b. Intensity (7.3.3); a calculation technique is used, solutions are presented to give the intensity of any practical lamp reactor configuration, as a function of the UV density and the wastewater UV absorbance coefficient.

c. Wastewater characteristics (7.3.4); this gives the designer guidance on the important wastewater parameters. These include the flow, initial density, suspended solids, UV absorbance coefficient, and the inactivation rates. Existing data are compiled for these parameters.

Process Design Example (7.4). A design example is given to demonstrate the design protocol, incorporating the considerations discussed in 7.3. The designer can use this as a stepwise outline for developing the design of a UV reactor.

O&M and Facilities Design Considerations (7.5). This last section should be used by both the designer and operator. It provides guidance related to effective O&M. Of particular interest are the cleaning aspects of the reactors. This is the single most important element for effective reactor performance.

7.1.2 General Description of the UV Process

Disinfection by ultraviolet radiation is a physical process relying on the transferance of electromagnetic energy from a source (lamp) to an organism's cellular material (specifically, the cell's genetic material). The lethal effects of this energy result primarily from the cell's inability to replicate. The effectiveness of the radiation is a direct function of the quantity of energy, or dose, which was absorbed by the organism. This dose is described by the product of the rate at which the energy is delivered, or intensity, and the time to which the organism is exposed to this intensity.

The basic kinetics of disinfection have been discussed as part of Chapter 4. The ideal UV disinfection model follows first order kinetics, whereby:

$$N = N_0 e^{-kIt}$$

where:

N = bacterial density remaining after exposure to UV

N_o = initial bacterial density

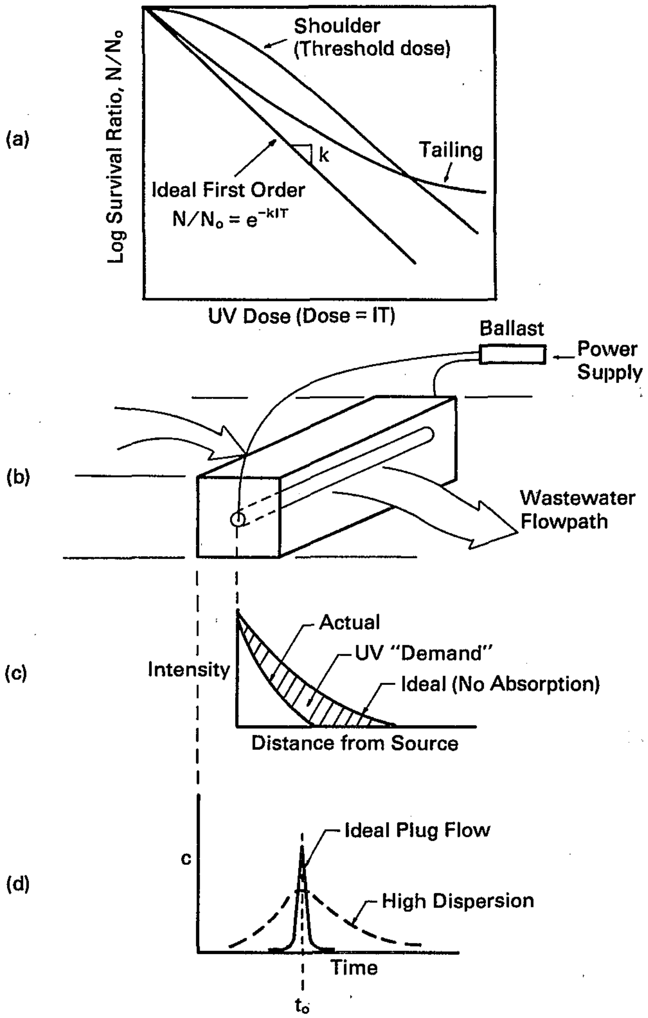
k = rate constant

I = intensity of UV radiation

t = time of exposure

The product It is the UV dose. Thus the response, noted by the log of the survival ratio, N/N_o, can be plotted against dose; the slope is the rate coefficient, k. This is shown on Figure 7-1(a). Deviation from this model is generally manifested by "shoulders," whereby minimal response is noted below a "threshold" dose; and by tailing effects, often attributed to occlusion (shadowing) of bacteria by particulate matter.

Figure 7-1. General description of UV design.



The primary artificial source of UV energy, at present, is the low pressure mercury arc lamp. It is almost universally accepted as the most efficient and effective source for disinfection systems application. The primary reason for its acceptance is that approximately 85 percent of its energy output is nearly monochromatic at the wavelength of 253.7 nanometers (nm), which is within the optimum wavelength range of 250 to 270 nm for germicidal effects. The lamps are long (standard lengths are typically

0.75 and 1.5 m (2.5 and 4.9 ft) arc lengths) thin tubes (typically 1.5 to 2 cm (0.6 to 0.8 in) in diameter). The radiation is generated by striking an electric arc through mercury vapor; discharge of the energy generated by excitation of the mercury results in the emission of the UV light.

These lamps can be suspended outside the liquid to be treated or submerged in the liquid; the intent is to get the energy into the liquid as efficiently as possible. Typically, if the lamp is to be submerged into the liquid, it is inserted into a quartz sleeve to minimize the cooling effects of the water. Figure 7-1(b) is presented to schematically represent the principal concerns when considering UV disinfection. In this example, the lamp is placed in the liquid, with the lamp perpendicular to the direction of flow. Other configurations may have the lamp parallel to flow, or the lamp may be suspended above the flowing liquid. Referring to Figure 7-1(c), as the lamp emits radiation, the intensity will attenuate as the distance from the lamp increases; this is due simply to the dissipation or dilution of the energy as the volume it occupies increases. A second attenuation mechanism involves the actual absorption of the energy by chemical constituents contained in the wastewater. This, analogous to the chlorine demand, is the "UV demand" of the wastewater.

The UV demand of a wastewater is quantified by a spectrophotometric measurement at the key wavelength of 253.7 nm; this expresses the absorption (or transmittance) of energy per unit depth. The output is absorbance units/cm, or a.u./cm. The percent transmittance can be determined from this unit by the expression:

$$\% \text{ Transmittance} = 100 \times 10^{-(\text{a.u./cm})}$$

The term most often used for design purposes is the UV absorbance coefficient, α, expressed in base e:

$$\text{UV absorbance coefficient, } \alpha = 2.3 \text{ (a.u./cm)}$$

The unit for α is cm^{-1}.

Although wastewater characteristics will be different site to site, ranges of the UV demand can be described for different levels of treatment:

	UV Absorbance Coefficient α (cm^{-1})	Percent Transmittance	Absorbance (a.u./cm)
Primary Treatment	0.4 to 0.8	67 to 45	0.174 to 0.35
Secondary Treatment	0.3 to 0.5	74 to 60	0.13 to 0.22
Tertiary Treatment	0.2 to 0.4	82 to 67	0.087 to 0.174

A second major concern is the provision of adequate exposure time to the microorganisms in order to meet the dose requirement at a given intensity. This was also generally discussed in Chapter 4; the key is to have plug flow through the system (see Figure 7-1(d)) such that each flow element resides in the reactor for the same amount of time. Perfect plug flow is not going to be achieved, of course; some dispersion will exist, such that there will be a distribution of exposure times about the ideal, theoretical exposure time. A design objective will be to minimize this distribution.

7.1.3 Current System Designs

In all, the design of a UV system must accommodate a few simple considerations: satisfy the UV demand of the wastewater; maximize the use of the UV energy being delivered by the lamps; and provide the conditions which encourage plug flow. Before proceeding with the detailed discussions of various technical aspects of the UV process, it is appropriate to first gain a perspective of UV system configurations. This is best done by reviewing design configurations which are currently being used at full-scale plants. This is done to enable the reader to better "visualize" the subsequent discussions. The use of these figures does not suggest that the configurations represent optimal designs; in fact, certain design configurations have been demonstrated to be inefficient.

Two basic generic reactors encompass current designs. The first is a contact reactor in which the lamps are submerged at all times in the wastewater; the submerged systems have the lamps encased in quartz sleeves which are only slightly larger in diameter than the lamp itself. The second reactor design does not allow contact of the water with the lamp (i.e., the quartz sleeve), but rather suspends the lamp above the liquid or surrounding conduits carrying the liquid. These conduits are transparent to the UV radiation.

Let us first consider the so-called submerged quartz systems. These can take on any number of configurations, generally described by the arrangement of the lamps relative to the direction of flow and to the hydraulic design of the lamp reactor. A common approach is the encasement of the lamp battery in a sealed reactor shell, as shown by the schematic on Figure 7-2. Flow enters the unit through an inlet pipe, typically perpendicular to the lamps, redirects and flows parallel to the lamps, finally exiting the reactor through the outlet pipe. A modification of this arrangement was provided at the Vinton Water Pollution Control Plant, Vinton, Iowa, as shown on Figure 7-3 (4). A steel plate was installed to split the cylinder in half lengthwise. Flow is directed down the unit on one side, then turns and flows down the second half before discharge. This encourages a plug flow condition by increasing the length of travel

Figure 7-2. Example of closed vessel UV reactor, with flow parallel to lamps (Courtesy of Ultraviolet Purification Systems, Inc., Bedford Hills, NY).

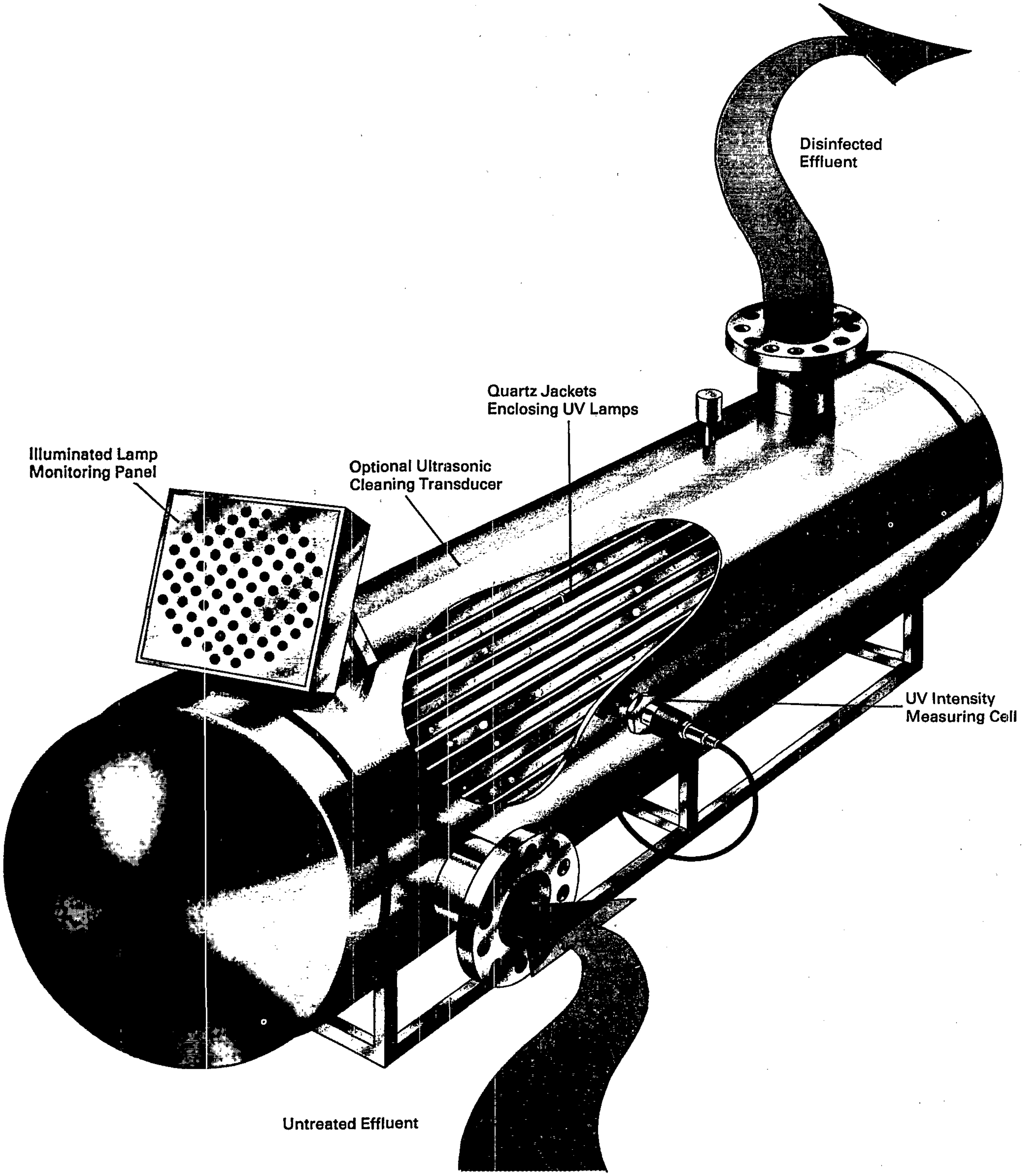
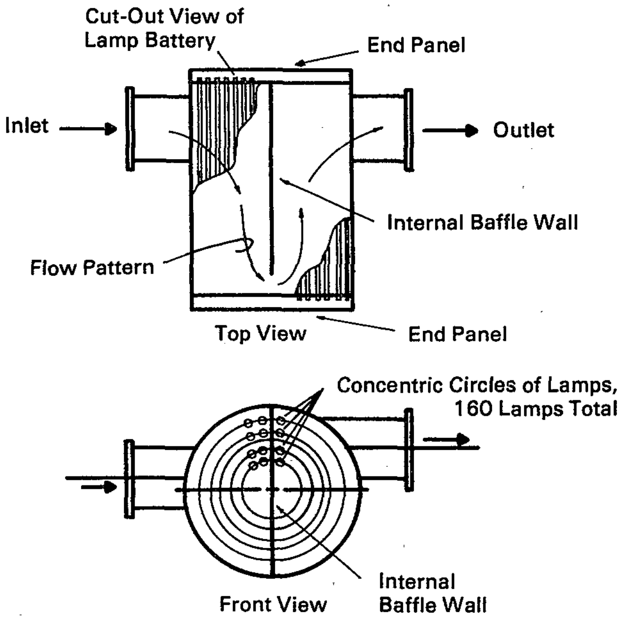


Figure 7-3. **Schematic of quartz UV unit in Vinton, Iowa (4).**

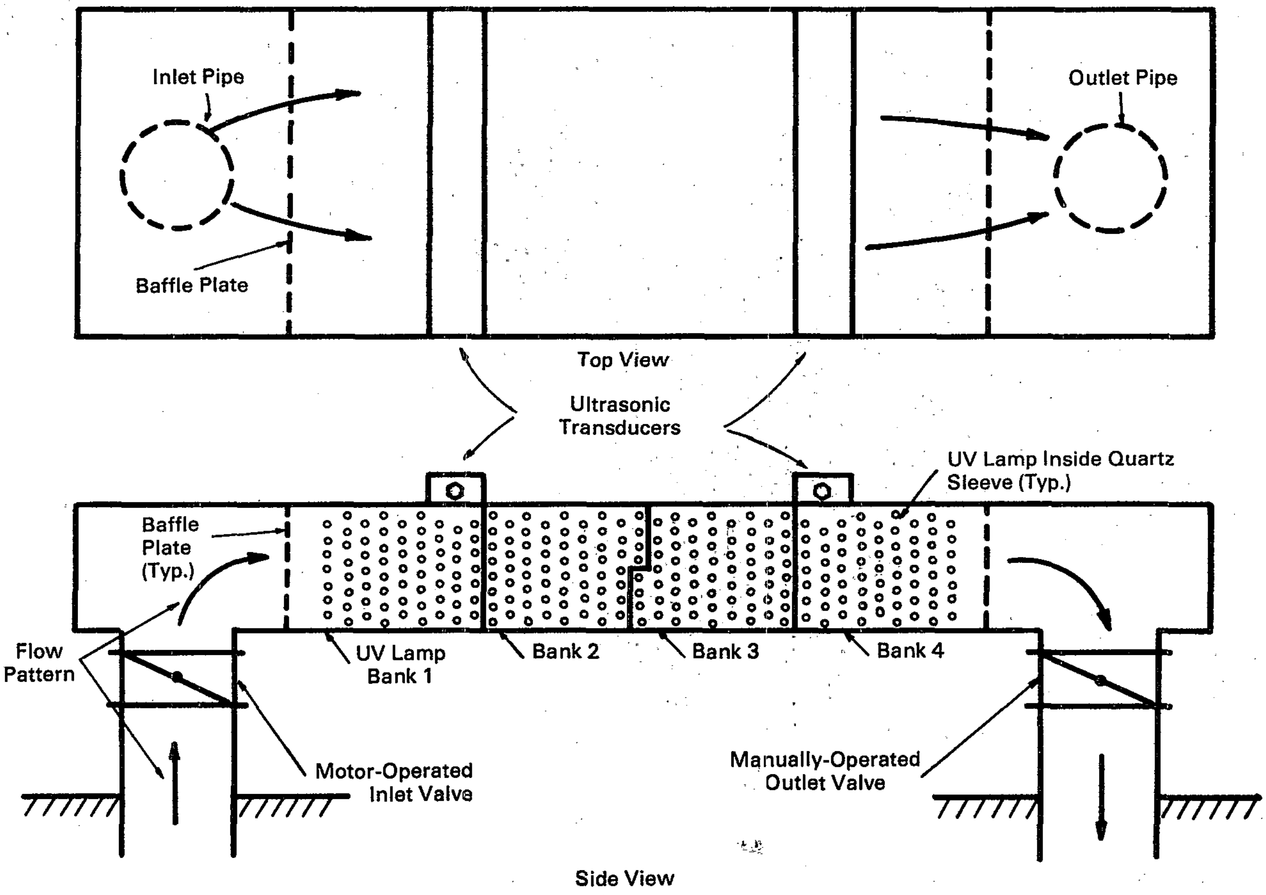


relative to the unit's hydraulic radius. Subsequent testing of this unit indicated that shortcircuiting occurred within the reactor and that its effective volume was significantly reduced. This is discussed further in Section 7.3.2.

The lamps can also be arranged perpendicular to the direction of flow in the same type of cylindrical reactor shell. Baffle plates distribute the wastewater along the length of the lamp battery; the flow is then directed upward through the lamp battery and over an internal overflow weir which runs the length of the reactor.

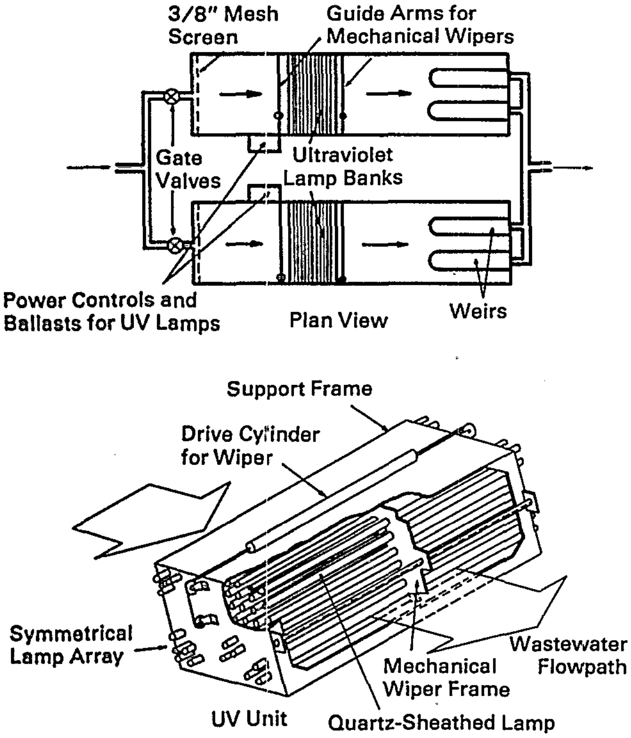
The sealed reactors can also be arranged to simulate channel flow. An example is provided on Figure 7-4, which is a schematic cross section of the UV units installed at the Suffern Water Pollution Control Plant, Suffern, New York. The lamps in this case are arranged in a symmetrical array, perpendicular to the direction of flow. The wastewater is pumped to the inlet chamber; a perforated baffle plate separates the chamber from the lamp battery to distribute the flow across the inlet plane of the lamp battery. A second plate is installed on the outlet side of the lamp section

Figure 7-4. / **Schematic of quartz UV unit in Suffern, NY.**



before the liquid enters the outlet chamber for discharge through the effluent pipe. The lamps are staggered to encourage turbulance and the system is arranged with a long path length to influence a plug flow condition.

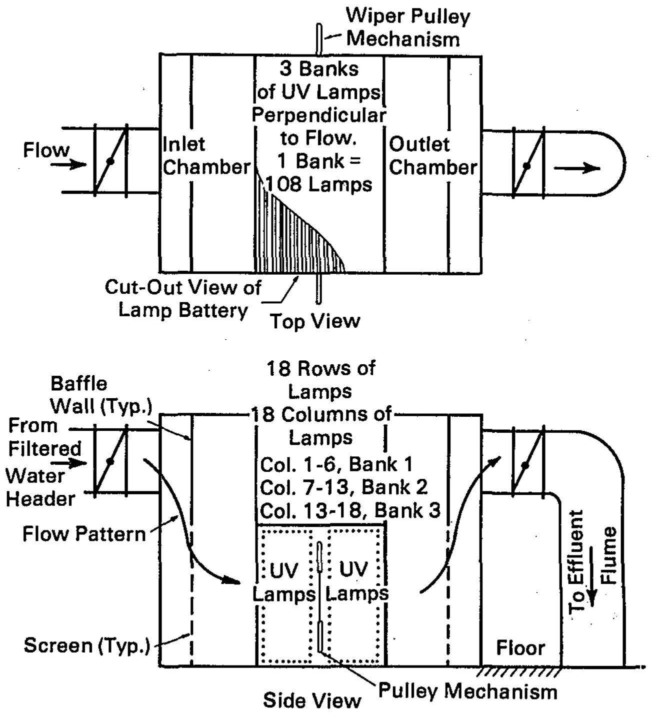
Figure 7-5. Example of open channel unit at Pella, Iowa, with flow directed perpendicular to lamps (5).



The submerged quartz systems are also arranged as open channel systems operating under gravity flow. An example is provided in Figure 7-5, which schematically presents the UV system installed at the Sents Creek Water Pollution Control Plant, Pella, Iowa (5). In this case the lamps are arranged in a symmetrical array, in the fashion of an open rectangular box as shown on the lower panel of Figure 7-5. The lamp battery is inserted into an open channel with the lamps perpendicular to the direction of flow (upper panel). Downstream of the lamp battery, the wastewater is collected in effluent launders for final discharge. This open-channel effect is also simulated by encasing the same type of lamp battery between open influent and effluent tanks. An example of this is the Albert Lea Water Pollution Control Plant, Albert Lea, Minnesota; a schematic of the units installed at this plant is presented on Figure 7-6.

The UV system for the Tillsonburg, Ontario (Canada) Water Pollution Control Plant is also installed as an open channel system. In this case, however, the lamps are arranged parallel to the direction of flow. The lamps were installed in the plant's existing secondary effluent channels. Each vertical module in these units is independent and can be removed by simply lifting them from the channel. A schematic of this modular, open channel design is shown on Figure 7-7.

Figure 7-6. Schematic of quartz UV unit in Albert Lea, Minn.



The non-contact UV systems are represented primarily by the system design in which the liquid is carried by thin-walled Teflon conduit transparent to the UV. The lamps are placed outside and parallel to the conduit. A schematic of this system arrangement is presented on Figure 7-8. The flow enters the influent chamber and is split to all or a portion of the Teflon tubes; the wastewater recollects in the effluent chamber and is discharged. The lamps, as shown on the figure, are typically inserted between the Teflon rows on removable racks (either vertically or horizontally). Pressure systems are also manufactured with the Teflon tubes; in this case the tubes are arranged in a serpentine pattern to reduce the overall size of the reactor.

The systems presented in Figures 7-2 through 7-8 represent, reasonably well, the various configurations of germicidal lamps utilized in current UV disinfection systems. The total UV system must also accomodate the lamp ballasts. The ballast is placed in series with the lamp to provide a starting voltage and to maintain constant current. These are generally

Figure 7-7. Schematic of open-channel, modular UV system (Courtesy of Trojan Technologies, Inc., London, Ontario, Canada).

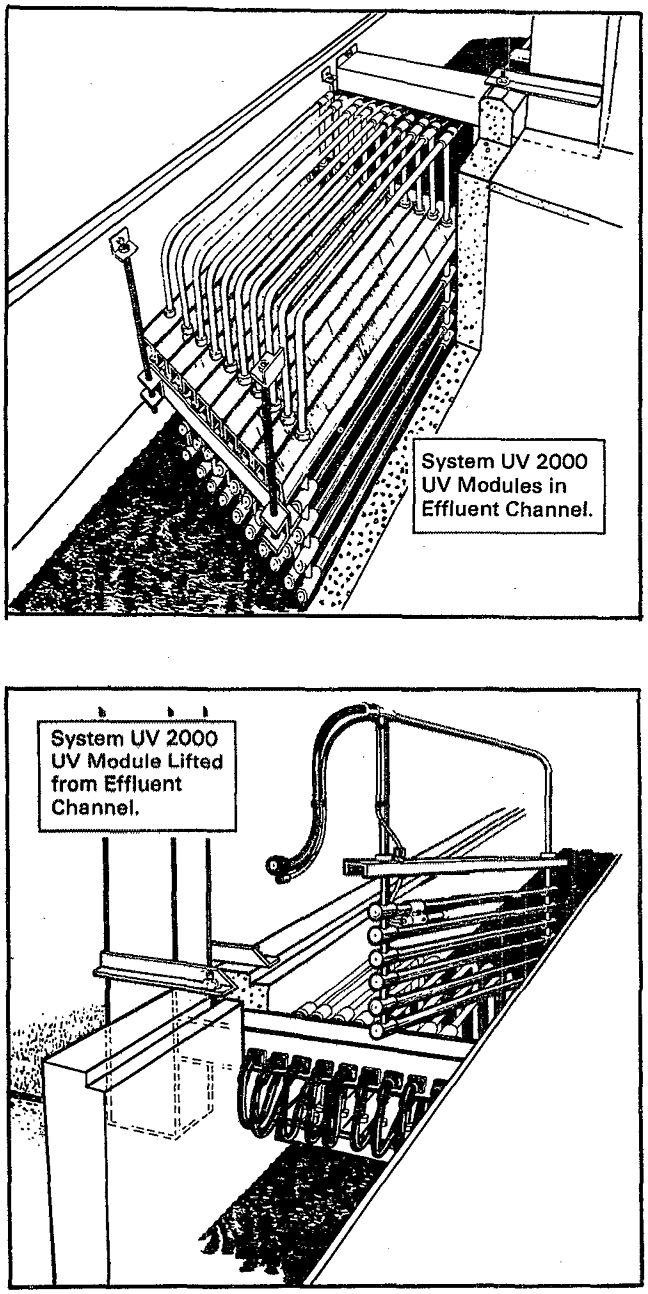
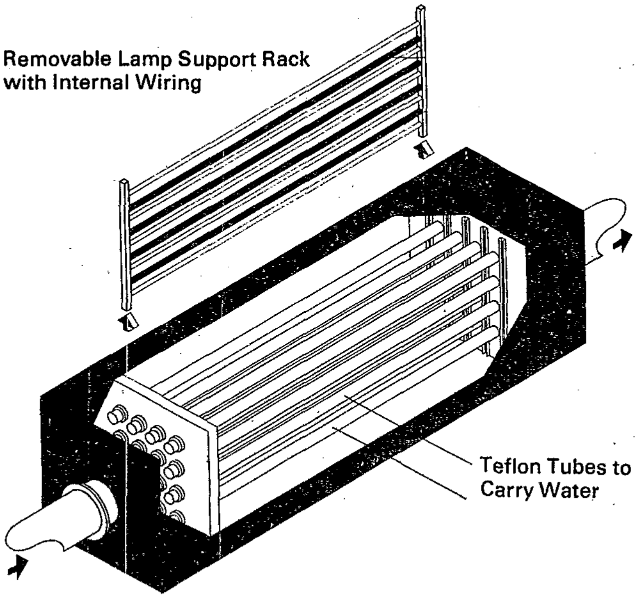


Figure 7-8. Example of UV system utilizing Teflon tubes (Courtesy of Ultraviolet Technology, Inc., Rancho Cordova, California).



held in enclosures above the lamp battery (Figure 7-8), or are remote from the reactors in separate power panels (Figures 7-2 through 7-7). Instrumentation generally entails UV intensity monitors and individual lamp operations circuitry. Control of the system can be on a manual basis, often involving the selective operation of modules, or banks of lamps within a module, as a function of the hydraulic load to the system. Automatic controls generally slave the lamp bank (or in some cases the lamp voltage) operations to the flow rate and/or the water quality.

A major element in the operation of UV systems is the cleaning of the surfaces which must be kept transparent to the UV radiation for efficient performance. These include the quartz sheaths and the Teflon tubes. Most commercial systems include accessory equipment to assist in this cleaning task; these include the use of mechanical wipers, ultrasonic transducers, and the provision to chemically restore the surfaces.

7.1.4 Current Technology Status

As suggested by the foregoing discussions, the UV process is relatively simple. Not unlike chlorination, an agent is added to the wastewater in sufficient quantity to effect the inactivation of bacteria. Time in this case is not provided to allow for a specific reaction to take place, but rather to accomplish the necessary dose. The effect on the microorganism is not in itself lethal to the microorganism; the main effect is to impose sufficient damage such that the organism is unable to replicate. The process, like ozonation, requires on-site generation of the germicidal agent; the generator (the UV lamp), however, is far simpler in concept and operation than that required to produce ozone. Like both chemical processes, UV must also satisfy a "demand" of energy exerted by the wastewater itself.

Other than the simplicity of the process, UV also offers the advantages of system flexibility and a capability of responding quickly to changes in demand. There is relatively little complexity to the hardware, and maintenance generally requires low skill levels. The hazards of the process are low, principally related to the high electrical loads and the personal exposure to the UV radiation; these are conditions which are easily safeguarded. A major advantage of the process is the absence of a residual in the wastewater and any subsequent impact on the receiving water. A corollary to this is the ability to "overdose" with UV and still not affect the receiving water. This allows for a less rigorous control requirement than associated with the use of chlorine. The absence of a residual can also be viewed as a disadvantage when considering the operational control of the process. There is no immediate monitor of performance analogous to the chlorine residual. Since the energy levels are not high enough to affect chemical reactions, there are no significant intermediates formed by the process, even at overdose levels. This is clearly an advantage over the chemical addition processes.

Given its advantages, UV was still not seriously considered as an alternative to chlorination for wastewater disinfection until the mid-seventies. At that point the process was considered more in response to the negative aspects of chlorination; its potential was acknowledged if the perceived disadvantages with the process were overcome. These related to the lack of information on UV application to low grade waters and the impact of various water quality parameters (particularly suspended solids) on design, the lack of any clearly defined design procedures, and the previous history of system fouling. Subsequent investigations focused on these and other aspects of UV disinfection. A number of full scale plants were installed, encouraged by the support of the USEPA through its Innovative and Alternative Technologies funding, under the Construction Grants program. A total of 14 plants were funded under I/A by the USEPA (6). These and other facilities are providing much needed information on the operation and maintenance requirements of the systems and on refinements necessary for current and future installation designs. Plants funded under the I/A program are listed on Table 7-1.

Table 7-1. Municipalities That Have Received I/A Funds for Designing and/or Constructing UV Disinfection Facilities (October 1978 to June 1981)

Village of Suffern, New York	Northfield, Minnesota
Woodstock, New York	Albert Lea, Minnesota
Crawford, New York	Pella, Iowa
Rhineback, New York	Cassville, Missouri
Smithburg, Maryland	Dexter, Maine
Clear Spring, Maryland	Kennebunk, Maine
Evanston, Wyoming	Heston, Kansas

There are a significant number of plants now installed throughout the United States for the disinfection of treated municipal wastewaters. A list of UV installations which are operational, or in the design, bid, or construct stage is presented in Tables 7-2, 7-3, and 7-4 (5). This information is further reduced in Table 7-5 to reflect the distribution of plants with regard to size; as indicated, the existing plants in operation are predominantly small (less than 3800 m^3/d or 1 mgd), while plants in the planning or construction stage tend to be larger.

7.2 Disinfection of Wastewaters by Ultraviolet Radiation

7.2.1 Ultraviolet Light

As presented in Figure 7-9, the ultraviolet region of the electromagnetic spectrum is generally defined as those radiations with wavelengths greater than the longest X-ray and less than the shortest wavelength visible to man; these wavelengths typically are set between 40 and 400 nanometers (nm). The ultraviolet region itself is divided. Near ultraviolet radiation is between 300 and 400 nanometers; far ultraviolet is between 200 to 300 nm. These two bands of ultraviolet are observed in solar radiation. Extreme ultraviolet radiation describes that energy between wavelengths 40 and 200 nm; energy at these wavelengths is strongly absorbed by air and its observation requires working in a vacuum or in a gas which does not absorb the energy.

Qualitatively, light is almost universally described by its wavelength. For the sake of convenience, a single unit of wavelength will be used throughout this chapter. This is the nanometer (nm), or 10^{-9} m. There are 10 Angstroms per nanometer.

A more fundamental quantity describing electromagnetic radiation is its frequency of vibration. The frequency and wavelength of radiation are related by:

$$c = \nu\lambda \qquad (7\text{-}1)$$

where:

c = the velocity of light (3×10^{10} cm per second in free space)
ν = frequency of vibration (vibrations per second)
λ = wavelength (cm)

A number of terms are used to express the quantity of radiation. These physical units relate to work or energy. The ones most commonly used are the erg, calorie, and watt-second (joule); all are measures of total quantity of energy or work. The time rates at which this energy is delivered in the corresponding

Table 7-2. Summary List of Facilities in the U.S.A. or Canada Utilizing Ultraviolet Light (UV) Disinfection Which are in Design

Facility Name	Design Firm	Size, mgd		Other Treatment Processes	Equipment and/or Comments
		Design	Start-up		
Bristol (CONNECTICUT)	Keys Associates Jim Geremis (401) 861-2900	10.75	7.0	Activated sludge	(Equipment not yet purchased)
Ridgefield (CONNECTICUT)	Albertson, Sharp & Ewing Mike Pastore (203) 846-4356	0.120	0.040	RBC System	N.C.
Salmon (IDAHO)	Ellsworth Engineering, Inc. Gary Marshall (208) 523-1662	Not established	Not established	Aerated lagoon/ *Facultative lagoon* (secondary cell)	N.C.
Ucon (IDAHO)	Forsgren-Perkins Engineering Dick Dyer (208) 356-9201	0.115 (80 gpm)	0.07-0.115 (50-80 gpm)	Facultative lagoon (secondary cell); UV is followed by 4-acre storage pond & land disposal	N.C.
Camp Point (ILLINOIS)	W.H. Klinger & Associates Dan Oliver (217) 223-3670	0.16	0.11	Two-cell stabilization pond (30-day Det. time) w/intermittent sand filter	N.C.
Iowa City (IOWA)	Keenstra & Kimm, Inc. Jim Kimm (515) 225-5000	13.0	9.0	Activated sludge	N.C.
Kennebunk (MAINE)	E.C. Jordan A. Peter Krauss (207) 775-5401	1.3	0.7	RBC system	(Equipment not purchased)
Limestone AFB (MAINE)	Dufresne & Henry Barry Bastian (207) 797-2010	2.5 (ADWF) (peak 5.2)	1.5	RBC system dry weather In wet weather, RBC (2.5 mgd) & up to 4.1 mgd primary effluent	N.C.
Pittsville (MARYLAND)	Harrington & Associates William Harrington (301) 768-5400	0.125	0.06	Oxidation ditch, tertiary filtration	N.C.
Poolesville (MARYLAND)	Kamber Engineers Dennis Kamber (301) 840-1030	0.6	0.35	Sequencing Batch Reactor, pressure filter	Flow through UV unit will be under slight pressure
Morton (MINNESOTA)	Ayres Associates Dale Philstrom (612) 644-0604	0.132	0.067 (dry weather)	Aerated stabilization basin system	N.C.
Calhoun City (MISSISSIPPI)	Willis Engineering Co. Joe Sutherland (601) 226-1081	0.342	0.175	Hydrograph-control release facultative lagoon. Store wastewater during dry weather; discharge 1 to 10 times design flow during rainy season	UV selected due to periods of no flow, not because of effluent requirements.
Deer Lodge (MONTANA)	Christian, Spring, Sielbach & Associates John Connel (406) 656-6000	1.5 (peak 3.3)	1.3	Aerated lagoon system	N.C.
Dillon (MONTANA)	S & A Engineers Robert Scruton (406) 442-1532	0.8 (winter-0.65, summer-1.10)	0.8	Aerated lagoon system, storage basin (180 AF)	N.C.
Lewiston (MONTANA)	HKM & Associates Jim Kaercher (406) 245-6354	2.9	2.3 (current 9.0)	RBC system	N.C.

Table 7-2. (Continued)

Facility Name	Design Firm	Size, mgd Design	Size, mgd Start-up	Other Treatment Processes	Equipment and/or Comments
Stratford #1 (NEW HAMPSHIRE)	Hoyle, Tanner & Associates Gene Forbes (603) 669-5420	0.056	0.053	Slow rate (0.5 gpd/ft^2) sand filters	N.C.
Stratford #2 (NEW HAMPSHIRE)	Hoyle, Tanner & Associates Gene Forbes (603) 669-5420	0.024	0.021	Slow rate sand filters (0.5 gpd/ft^2)	N.C.
Whitefield (NEW HAMPSHIRE)	Phillips & Emberly William Emberly (802) 434-2142	0.185	0.133	Aerated lagoon system	N.C.
Woodstock (NEW YORK)	Phillip J. Clarke David Wright (716) 454-4570	0.230	0.180	Oxidation ditch, sand filters in summer months only	N.C.
Harrimon (NEW YORK)	Phillip J. Clarke John Tarolli (914) 294-8818	4.0	2.0 (existing)	Draft tube oxidation ditch, tertiary sand filters, parallel train: existing activated sludge, aerated polishing lagoon	Lead design firm is: Erickson Schmitt, Al Schmitt (914) 294-8838
Loomis (NEW YORK)	Phillip J. Clarke David Wright (716) 454-4570	0.080	0.054	Overland flow, aerated polishing lagoon	N.C.
Ironton (OHIO)	Brundage, Baker & Stouffer, Ltd George Haggard (614) 888-3100	1.7	1.7	Trickling filter	N.C.
Northridge Subdivision Piedmont (SOUTH DAKOTA)	Hoskins, Western & Sonderegger Al Foster (605) 342-4105	0.023	0.011	Extended aeration	State wants a tertiary filter installed before they will approve design
Beckley (WEST VIRGINIA)	Greenhorne & O'Mara Gary Beech (301) 982-2837	3.5	2.5-3.0	Activated sludge extended aeration (post-aeration follows UV)	N.C.
Buckhannon (WEST VIRGINIA)	Kelley, Gridley, Blair Wolfe Jim Downey (304) 345-0470	2.5	1.5	Oxidation ditch (post-aeration follows UV)	Design around ENERCO
Green Valley-Glenwood PSD (WEST VIRGINIA)	Pentree Will Smith Bob Hazelwood (304) 425-9581	1.5	1.0	Oxidation ditch (post-aeration follows UV)	N.C.
Marshall Co. PSD #1 (WEST VIRGINIA)	Green International Norman Katz (412) 471-5348	0.17	0.10	Extended aeration (package plant)	N.C.
Opequen Hedgesville (WEST VIRGINIA)	HNTB Dave Wright (414) 463-2310	0.80	0.55	Oxidation canal, post-aeration follows UV	N.C.
PAX (WEST VIRGINIA)	G.A. Tice George Tice (304) 255-5400	0.06	0.045	Facultative lagoons	N.C.
Harper Eccles WWTP Raleigh Co. PSD (WEST VIRGINIA)	Greenhorne & O'Mara Turgay Ertugal (301) 982-2800	0.100	0.100	Oxidation ditch (package unit)	N.C.

Table 7-2. (Continued)

Facility Name	Design Firm	Size, mgd Design	Size, mgd Start-up	Other Treatment Processes	Equipment and/or Comments
Salt Rock PSD (WEST VIRGINIA)	Dunn Engineers Dave Schultz (304) 342-3436	0.260	0.246	Oxidation ditch	N.C.
Riverton (WYOMING)	Airex Engineers Harry La Bonde (307) 856-6505	4.95	2.3	Oxidation ditch	U.V. Technologies
Athens (WISCONSIN)	Becker Hoppe Engineering Gerald Bizjak (715) 359-6147	0.225	0.070	Aerated lagoon (3 cells: primary, secondary, storage). Plant has controlled discharge of effluent by using the storage lagoon	Anticipating the use of ENERCO
Little Black (WISCONSIN)	Carl C. Crane, Inc. Victor Marz (608) 238-4761	0.012	0.008 to 0.010	Recirculating sand filters	N.C.
Collingwood (ONTARIO, CANADA)	Ainley & Associates, Ltd. (Owen Sound, Ont.) Colin Kent (705) 445-3451	1.2	0.5 low	Activated sludge, tertiary filtration	Anticipating Trojan Industries

N.C. - No comments.

units are ergs per second, calories per second, and the watt. These units and their cgs equivalents are as follows:

Watt-second (or joule)	w-sec	10^7 ergs
Calorie	cal	4.2×10^7 ergs
Erg	—	—

The intensity or energy density of the radiation is expressed in terms of energy incident upon a unit area. The unit used in the context of this report is the micro-watt per square centimeter (μwatt/cm^2).

Quantum theory states that radiant energy occurs in discrete units, or quanta. The energy of these fundamental units is related to its frequency:

$$E = h\nu = hc/\lambda \qquad (7\text{-}2)$$

where:

E = energy of a single quantum (ergs)
h = Planck's constant (6.62×10^{-27} erg-sec)
c = velocity of light (3×10^{10} cm per sec)
ν = frequency (vibrations per second)
λ = wavelength (cm)

The quantum is a very small energy unit, equivalent to (19.86×10^{-17})/wave length, in cm-ergs. From this expression it is shown that the energy content of a quantum is identical for a given wavelength of light.

7.2.1.1 Source of UV Radiation

Practical application of UV for purposes of disinfection required a high intensity source at the desired wavelengths. This can be traced by the evolution of the mercury vapor lamp. In 1835, Wheatstone described the intense light emitted when mercury is vaporized in an electric arc. The first true mercury vapor lamp was constructed by Downing and Keating in 1896 by passing an electric discharge through mercury in a partially evacuated tube. The problem was that the arcs would eventually go out because of the increase in the pressure of the vapor. Cooper-Hewitt resolved this problem in 1901 by devising a lamp in which the mercury was condensed at the same rate at which it was vaporized. This, along with the development of fused quartz and ultraviolet transmitting glass, initiated the successful commercial development of mercury vapor lamps (7).

The discharge type lamps were relatively inefficient, however, due to their low selectivity in the use of the energy, or electrical input. The generation of heat, excitation of several different spectral lines, and inefficient electrodes resulted in a distribution of energy to many outputs. The key development came in the 1920s, when it was determined that a discharge through a mixture of mercury vapor at a precisely optimum pressure and a rare gas (typically argon) at a somewhat higher pressure was extremely efficient in converting the electrical energy to ultraviolet light. Fully 60 percent of the energy input could be converted to monochromatic radiation at 253.7 nm (8).

The commercial development of the mercury-rare gas discharge lamps was directed to its use as a light source. The development of a suitable fluorescent

Table 7-3. **Summary List of Facilities in the U.S.A. or Canada Utilizing Ultraviolet (UV) Disinfection Which are Under Construction.**

Facility Name	Design Firm	Size, mgd		Other Treatment Processes	Equipment and/or Comments
		Design	Start-up		
Galney Ranch (Arizona)	Greely & Hanson Elizabeth Zureick (602) 992-5000	1.7 (8 mo.) 1.0 (4 mo. winter)	1.7	Extended aeration, sand filters	ENERCO (formerly UV Technologies)
Payson (Arizona)	Moore, Knickerbocker & Assoc. Terry Moore (602) 265-3776	1.7	0.5	Bardenpho (& Clarifiers), backwash filters	UV Technologies (Now ENERCO)
Augusta (Arkansas)	Mehburger, Tanner, Robinson & Assoc. Daryl Laws (501) 375-5331	0.6 (current 0.4)	0.4-0.5	Orbal treatment system, Aeration disks	ENERCO #L1000
Heber Springs (Arizona)	Boulder Engineers Jim Little (501) 362-3118	1.8 (1250 gpm)	1.8 (intermittent pumped flow; ultimately; flow will be continuous)	Three-cell aerated lagoon (facultative) of 19 acres rapid sand filters	UV Purification Systems, Inc. (70 lamps; 9-sec. retention)
Presque Isle (Maine)	Wright, Pierce, Barnes & Wyman Dave Fuller (207) 725-8721	2.3 dry weather (5.4 wet weather)	0.7 (1.9 with I/I)	Oxidation ditch system	Pure Water Systems, Inc. (Two units-1 standby)
Clear Springs (Maryland)	Fellows, Reed & Assoc. Ed Renn (301) 739-5660	0.20	0.12	Oxidation ditch	Not yet purchased
Milford (Massachusetts)	Haley & Ward, Inc. Ben Bugbee (617) 890-3980	4.3	2.0	RBC & tertiary filters	Not yet purchased
Bemidji (Minnesota)	Rieke, Carroll, Muller & Assoc. Warren Kerstan (612) 935-6901	2.7	1.6 (current)	Activated sludge & tertiary filtration	UV Technologies
North Koochiching Area San. District (Minnesota)	Widseth, Smith, Nolting & Assoc. Don Anderson (218) 829-5117	2.3	1.1 (current)	Trickling Filter	UV Purification Systems, Inc.
Bonne Terre (Missouri)	Metropolitan Engineering Robert Vogler (314) 948-3860	0.6	0.4	Oxidation ditch	Not yet purchased
Emminence (Missouri)	Missouri Engineering Co. Corky Stack (314) 364-4003	2.9	0.5	Oxidation ditch, sand filter	UV Purification
Frederick Town (Missouri)	Crane & Fleming Greg Boettener (314) 221-4048	0.85	0.6	Oxidation ditch	UV Purification
Mineral Belt Area WWTP (Flat River, Mo.) (Missouri)	Metropolitan Engineering Robert Vogler (314) 467-3860	2.0	1.6	Oxidation ditch	Not yet purchased
Noel (Missouri)	Allgeier, Martin, & Assoc. Jan Tupper (417) 624-5703	0.2	0.15	Oxidation ditch	UV Technology

Table 7-3. (Continued)

Facility Name	Design Firm	Size, mgd Design	Size, mgd Start-up	Other Treatment Processes	Equipment and/or Comments
Summerset Plant Div., South Jefferson Co. (Missouri)	Horner & Shifrin (314) 531-4321	0.117	0.1	Lagoon facility (30-day detention time)	UV Purification Systems Inc.
Winona (Missouri)	C.B. Simmons C.B. Simmons (417) 732-2092	0.175	0.100	Oxidation ditch	ENERCO
Chinook (Montana)	Robert Peccia & Assoc. Alden Beard (406) 442-8160	0.50 (current 1.10)	0.35	Oxidation ditch	ENERCO
Bennington (Nebraska)	Johnson, Erickson, O'Brien & Assoc. Terry O'Brien (402) 443-4661	0.186	0.065-0.070	Extended-aeration activated sludge	Pure Water Systems, Inc. and U.V. Purification Systems, Inc.
Chatham Township (New Jersey)	Keller, Kirkpatrick Bob Kirkpatrick (201) 377-8500	0.120	very low	RBC, multi-media filter	UV Technologies, Inc.
Rhineback (New York)	Brinnier & Larios Dennis Larios (914) 338-7622	0.130	0.080	Oxidation ditch	ENERCO
Thompson (New York)	Phillip J. Clark & Assoc. David Wright (716) 454-4570	1.0	0.80	Draft tube oxidation ditch, aerated polishing lagoon	ENERCO
Beech Mountain (ski resort) (N. Carolina)	Davis, Martin, Powell Ed Powell (919) 883-0032	0.400 (peak @ 2 1/2X)	0.04-0.20 seasonal	Contact stabilization	ENERCO
Waynesburg (Ohio)	Hammontree & Assoc. Richard Hunsinger (216) 499-8817.	4.0	0.23	Bio-drum	UV Purification Systems, Inc.
Mt. Pleasant (S. Carolina)	E.M. Seabrook, Inc. Louis Couthen or Brian Wright (803) 884-4496	3.2	1.0 (current)	Conventional activated sludge	UV Purification Systems, Inc.
Coalville (Utah)	DMJM Reed Fisher (801) 262-2951	3.0	0.25	Oxidation ditch	Not yet purchased
Baker Heights Berkley Co. PSSD[a] (W. Virginia)	HNTB David Wright (414) 463-2310	0.34	0.34	Trickling filter	ENERCO
Moorefield (W. Virginia)	Kelley, Gidley, Blair, Wolfe Dick Kline (304) 345-0470	0.477	0.400	Aerated lagoon	ENERCO
Evanston (Wyoming)	Eckoff, Watson, Preater John McNeil (801) 486-5621	2.9 (current 2.5)	2.0	Oxidation ditch	UV Purification Systems, Inc.
Worland (Wyoming)	Airex Engineers Harry LaBonde (307) 856-6505	1.12	0.8 (current)	Aerated lagoons	UV Purification Systems, Inc.
Madison Met (Wisconsin)	Consour Townsend Ron Reising (312) 938-0300	50 (peak 115) largest UV facility in world	35-40	Activated sludge w/nitrification	UV Purification Systems, Inc.

[a] PSSD = Public Service Sanitation District.

Table 7-4. Summary List of Facilities in the U.S.A. or Canada Utilizing Ultraviolet Light (UV) Disinfection Which are in Operation

Facility Name	Design Firm	Size, mgd		Other Treatment Processes	Equipment	Comments on Performance
		Design	Current			
Lake Creason, Cow Hide Cove area (AR)	U.S. Army Corps of Engineers "Mac" Montgomery (601) 634-5301	0.015	0.015	Extended aeration-activated sludge, rapid sand filters	U.V. Technologies (Teflon Tubes)	In operation 1-1/2 years. Good performance.
Lake Ouachita, Little Fir area (AR)	U.S. Army Corps of Engineers "Mac" Montgomery (601) 634-5301	0.018	0.018	Extended aeration-activated sludge rapid sand filters	U.V. Technologies (Teflon tubes) Now ENERCO 1 unit, 8-lamps rated for 7500 hrs. use	In operation 1-1/2 years. Good performance. Operates March-Oct.
Lamar (AR)	Burrough, Uerling & Brasuell David Uerling (501) 646-5559	0.106	0.100	Overland flow	U.V. Technologies	Disinfection requirements met. Frequent bulb replacement. Failure to operate automatically. Weir causes UV tanks to be filled with silt. (Identical to Hatfield facility).
Tillonsburg (Ontario, Canada)	Anderson Assoc. Peter Laughton (416) 497-8600	2.4	1.3	Extended aeration	Trojan Industries	In operation for 2 years. Achieving 1/2 100 MPN coliform/100 ml.
Eden (Wisconsin)	Arthur Technology John Masters (414) 922-6973	0.16	0.10	1° clarifier, roughing filter extended activated sludge (for nitrification)	UV Technology	Good performance.
Conifer Center (Colorado)	ADG Engineering, Inc. Roger N. Venables (303) 761-5142	0.015	0.005	Package-activated sludge extended	Ultradynamics (Santa Monica, CA)	No performance problems to date, only 1 month of operation.
Erie (Colorado)	Keith Bell & Assoc. Keith Bell	0.30	0.18	Aerated lagoons (20-day det. time)	ENERCO	Tube fouling. Operational since Dec. 1983.
Cypress-Thompson Creek, Challis (Idaho)	Hamilton & Voeleur (No longer in business) Contact: former employee of H & V Rance Bane Ellsworth Engrg. (208) 523-1662	0.720 (500 gpm)	0	3-stage lagoon	ENERCO	Operates 30 days/year in September. Too early to judge performance; flows too early to judge performance; flows too low. Fecal coliform die-off in ponds is 100%, prior to UV disinfection.
Newdale (Idaho)	Forsgren-Perkins Engrg. Dick Dyer (208) 356-9101 Earl Keemp (801) 364-4735 (Salt Lake City Ofc.)	0.045	0.022	Facultative lagoon (Land disposal)	ENERCO (Model G-30)	Start-up: Nov. 1983-99.9% bacterial die-off. Design engineer is Dick Dyer.
Red Top Meadows (Ketchum) (Idaho)	J.U.B. Engineers James Coleman (208) 733-2414	0.180 (dry 0.06)	0.180	Extended aeration oxidation ditch	U.V.Technologies (Now ENERCO)	Good performance. Fecal coliform count = 0 for 100% fecal coliform kill. Bulbs replaced annually. Operating since 10/82.
Pella (Iowa)	Veenstra, Kimm Engineers Jim Kimm, Mike Foreman (515) 225-8000	3.4	1.5	Activated sludge	Pure Water Systems, Inc. (Quartz jacketed UV lamps)	Mechanical wiper system not functioning. Using chemical cleaning with weak acid solution for lamp jackets (outside) approximated every 2 weeks vs. 6 mos to 1 year. New end seals were provided by the manuf. Operational since April 1982 (Oct. '81 for entire facility).
Heston (Kansas)	Wilson & Co. Engineers & Architects Jim Dowell (913) 827-0433 Contact: City of Heston, Maurice Bowersox or Bill Nitzsche (plant) @ (316) 327-4412 or 327-2535	1.3	0.25-0.30 (wet weather −0.40)	Orbal Activated Sludge (Effluent goes to golf course)	U.V. Purification Systems, Inc.	Ultrasonic cleaning not working up expectations. Consequently, light intensity is not as expected. Must chemical clean the lamps.
Sabbatus (Maine)	Woodward & Curran Frank Woodward (207) 839-6751	0.25	0.1	Imhoff tank & intermittent sand filter system	U.V. Technologies (G-500)	Meeting coliform count requirements. Operating 1 year. Still in shake-down. I/I problems; must bypass due to poor filtration.
Togus VA Hosp. (ME)	Hunter Balleau Barrie Patrie (296) 671-4721 or V.A. Center, Bob White (207) 623-8411 X338	0.2	0.15	Oxidation, ditch system	U.V. Technologies	Operating 1 year. Fecal coliform levels met. Initially, wiring and bulbs were faulty. Currently, ballast that runs UV bulbs weakens, diminishes intensity of bulbs. Must change ballast often; life is less than 2,000 hours.
Old Towne (Maryland)	Allegany County Sanitation Commission Kevin Beachy	0.04	0.03	Extended Aeration activated sludge	Pure Water Systems, Inc.	High maintenance. Initial ballast problem-now corrected. (Auto wiper system shuts down UV unit when tubes get too dirty).
Smithsburgh (MD)	Fellows, Reed & Assoc. Ed Renn (301) 739-5660	0.20	0.12	Extended aeration	U.V. Purification Systems, Inc.	System in start-up.

Table 7-4. (Continued)

Facility Name	Design Firm	Size, mgd		Other Treatment Processes	Equipment	Comments on Performance
		Design	Current			
Thurmont (MD)	Harrington & Assoc. William Harrington (301) 768-5400	1.0 (4.0 weather)	0.3 (2.9 weather)	Oxidation ditch with tertiary filtration	ENERCO	Too soon to tell.
Albert Lea (Minnesota)	Tolz, King, Duvall & Anderson Dave Kirkwold (612) 292-4400	12.53	3.91	2-Stage activated sludge, tertiary filters	Pure Water Systems, Inc.	Good performance. Meeting coliform kill requirements. Coliform count is less than 1 MPN/100 ml. Some minor mechanical problems, which have been corrected. Operating since June 1983.
Northfield (MN)	Bonestroo, Rosene, Anderlick & Assoc. Dick Turner (612) 636-4600	2.5	2.1	2 Stage secondary system (trickling filters & RBC)	Pure Water Systems, Inc.	Poor operation. Have not achieved contract specification for operation. Still putting in corrective measures. Mechanical difficulties; electrical components burned.
Cossville (Missouri)	Allgeier, Martin & Associates Jan Tupper (417) 513-5703	0.5	0.7 (exceeding design capacity)	Oxidation ditch	Aquafine (No longer in market)	Excess heat in UV Bldg. - uncomfortably warm for operator. Fans alleviated this problem. Plant is meeting discharge requirements. In operation 2-3 years.
Clinton (MO)	Bucher, Willis J. Ratliffe Jim Swanson (913) 827-3603	2.0	1.3	Oxidation ditch	U.V. Purification Systems, Inc.	Summer use only; in use for 1 yr. Some ballast problems. Some chemical cleaning problems.
Ozark (MO)	Anderson Engineering Steven Brady (417) 866-2741	0.72	0.20	Oxidation ditch	U.V. Purification Systems, Inc.	In operation 1 year. Initial problem with ultrasonic cleaning system achieving desired bacterial kill.
Briarwood (MO)	Sanders, Stewart, Gaston Paul Kinshella (406) 245-6366	0.180 (peak 0.72) (Health Dept. approved 0.123 to date)	0.0005 (500 gpm)	Oxidation ditch	ENERCO	Just starting up (as of 10 am 3/27/84) Using clear water.
Yellowtail Power Plant (Montana)	Bureau of Reclamation Craig Peterson (406) 657-6141 or Mr. Hergenreider (406) 666-2443	0.0006	0.001	Extended aeration package plant, tertiary filtration	U.V. Technologies, Inc.	Good performance.
Environmental Disposal Corp. (Pluckerman) (New Jersey)	Environmental Design Inc. (out of business) Contact: Ray Ferrara Princeton University (609) 452-4653 or Neil Callahan (Operator) (201) 234-0667	0.85 (590 gpm)	0.43-0.57 (300-400 gpm)	Bardenptho and multi-media filter	U.V. Technologies, Inc. (Now ENERCO)	Good performance. Fecal coliform @ 10-50 MPN/100 ml. (Permit 200 MPN per 100 ml) Feed to UV is "clean" 4-5 mg/l SS. 4-5 mg/l SS. Initially seals leaked - now replaced. Once clay got in stream, colored water, and reduce effectiveness of UV system. 2-750 gpm units run intermittently batch.
Educational Testing (N.J.)	CUHZA Manny Dios (609) 452-1212	0.080	0.030-0.038	Extended aeration, filtration (package, multi-media, high rate)	U.V. Purification Systems, Inc.	Operational since 9/81. Faulty photo cell, replaced in 1981.
Crawford (NY)	Phillip J. Clarke & Associates John Tarolli (914) 294-8818	0.15	0.08-0.085	Oxidation ditch (septic tank effluent)	U.V. Purification Systems, Inc.	Not yet in operation—summer requirements only (Plant on-line since 10/83).
Pennyann (NY)	Hershuy, Malone, & Associates Greg Barbour (716) 381-9250	1.8	0.6	RBC	U.V. Purification Systems, Inc.	In operation since 11/83. High flows blew out UV Tubes; cause uncertain: freezing or obstacles in flow.
Suffern (NY)	Thomas Riddick Norman Lindsay (914) 365-0446	1.9 (peak 4.0)	1.2	"Lighting Complete Mix" aeration (similar to activated sludge.)	U.V. Purification Systems, Inc.	Not yet in operation. Awaiting stabilization of activated sludge system.
McPherson (Kansas)	Wilson & Co., & Architects Jim Dowell (913) 827-0433 Contact: Plant operator, Walt Hundley (316) 241-3940	0.29 (200 gpm) (entire plant @ 2.0 mgd)	0-0.14 (0-100 gpm) (entire plant @ 1.7 mgd)	Trickling filter and Contact Stabilization basin in parallel. Combined discharge bypasses UV to creek or a portion (100 gpm) is discharged to lake, when lake level is low.	Aqua-fine	Poor performance. Coliform count is high. Difficult to keep quartz sleeve over lamps clean. Harness (mineral deposits won't wipe off. Sometimes operators use soap or chlorox. However, it's frequently necessary to chemically clean lamp sleeves.
Marietta (Oklahoma)	Bob McCoy (Now retired) Contact: Mark Daniels State Environmental Agency (405) 276-5493	0.231	0.19	Oxidation ditch	U.V. Purification Systems, Inc.	Maintenance problems i.e. burned out lamps.

Table 7-4 (Continued)

Facility Name	Design Firm	Size, mgd		Other Treatment Processes	Equipment	Comments on Performance
		Design	Current			
Berkeley County (South Carolina)	E.M. Seabrook, Inc. Ryan Wright (803) 884-4496	5.0 (peak 8.0)	3.0	Oxidation ditch	Pure Water Systems, Inc.	In operation for past 9 months. Good performance. Discharge requirements met. Some mechanical difficulties.
Civilian Conservation Center, Nemo	Case, Colter, Inc. (Denver) Ralph Olson (303) 288-1511 Other contacts: U.S.F.S: Carl Erikson (605) 348-3636 Terry Ambraster (303) 234-5223	0.024	0.017	Extended aeration, tertiary filtration.	U.V. Purification Systems, Inc.	Good performance.
Danville (Vermont)	Dufresne & Henry Bobbi Trudell (802) 886-2261	0.070	0.040	Aerated lagoon system (30-day det. time)	U.V. Purification Systems, Inc.	Good performance. Meeting discharge requirements.
Jacksonville (VT)	Dufresne & Henry Bobbi Trudell (802) 886-2261	0.050	0.025	RBC system, which treats septic tank effluent	Ultra Dynamics, Inc.	Good performance. Meeting discharge requirements.
Pawlet (VT)	Dufresne & Henry	0.040	0.023	RBC system; which treats septic tank effluent.	U.V. Purification Systems, Inc.	Good performance. Meeting discharge requirements. Operating since 12/83.
Whitinghamn (VT)	Dufresne & Henry	0.013	0.006	RBC system, which treats septic tank effluent.	Ultra Dynamics, Inc.	Good performance. Meeting discharge requirements. Operating since 12/83.
Cumberland Hospital	Gresham, Smith & Partners David Shood (803) 572-1300	0.030	0.002 to 0.003	Extended aeration	U.V. Purification Systems, Inc.	Discharge requirements not met in 6 of 9 months. In compliance for 2/84.
Tangier Island (VA)	Shore Engineering Emmett Ranson (804) 787-2773	0.1	0.032	RBC	Aquafine	Difficulty meeting dicharge requirements.
Lander (Wyoming)	Western Design Consultants Mr. Chen (801) 486-5621	1.82	1.3	Aerated lagoon	ENERCO	Burning one end of the UV lamps due to float switch problem.
Rock Springs (WY)	Johnson, Fermelia, and Crank Dale Crank (307) 877-9093	2.0	2.0	Oxidation ditch	UV Technologies	Test runs of system resulted in overheating of bulbs. Fans were used to correct this.
Brooklyn (Wisconsin)	Carl C. Crance, Inc, Dennis Truttman (600) 238-4761	0.160	0.025-0.030 (low)	Oxidation ditch	U.V. Purification Systems, Inc.	150-200 fecal coliform count. (Currently there is no state requirement.)
Cross Plains (WI)	Mead & Hunt, Inc. Bill Buth (608) 233-9706	0.450 (avg.)	0.180 to 0.200	Extended aeration oxidation ditch	U.V. Purification Systems, Inc.	Good performance. Requirements met.
Ettrick (WI)	Davy Engineering Mike Davy or Arnie Finski (608) 782-3130	0.064	0.040	RBC	U.V. Purification Systems, Inc.	Problems-w/turbidity & coli killo.
Holomon (WI)	Davy Engineering Mike Davy (608) 782-3130	0.8	0.4	Extended aeration	U.V. Purification Systems, Inc.	Problems-start-up only Operational since 10/83.
Deerfield (WI)	Carl C. Crane, Inc. Dennis Truttman (608) 238-4761	0.195	0.100 flowmeter not working.	Extended aeration oxidation ditch	U.V. Purification Systems, Inc.	In start-up for 2 months. Still in shake-down.
Lodi (WI)	Mid State Assoc. Jim Owen (608) 356-8344	0.620	0.230 to 0.288	RBC	U.V. Purification Systems, Inc.	Operational since 7/83. Poor performance. Requirements not met.
Lyons (WI)	Robbers & Boyd Larry Boyd (414) 763-2652	0.100	0.020 (low)	Oxidation ditch	Aqua fine Corp.	Meeting discharge requirements. Operational since late 1981.
Poynette (WI)	Lakeland Engineers Mark Koletzke (608) 274-3898	0.190 (0.470 wet weather)	0.300	Oxidation ditch	U.V. Technologies (now ENERCO)	Operational since 8/83. Meeting coliform count requirements.
Spring Valley (WI)	Davy Engineering Arnie Pinski or Mike Davy (608) 782-3130	0.189	0.100	RBC	U.V. Purification Systems, Inc.	UV is not to start-up until 4.1.84

Table 7-4. (continued)

Facility Name	Design Firm	Size, mgd Design	 Current	Other Treatment Processes	Equipment	Comments on Performance
Otay (California)	Lowry & Assoc. Matt Tebbetts (619) 283-7145	0.65	0.3	Activated sludge, filtration, pH reduction prior to UV, the R.O. unit.	UV Technology	
Jehovah Witness Church, Tujuas (Puerto Rico)	Radmes Torres (809) 725-5878 (in San Juan, P.R.)	0.22	0.006	Septic tank, sand filter, equalization tank.	Unknown	

Table 7-5. Summary of UV Installations in U.S. in Operation, Construct, or Design Phase

Size (Design)	In Operation	In Construction	In Design
<380 m^3/d (<0.1 mgd)	15	–	7
380-1900 (0.1-0.5)	17	10	10
1900-3800 (0.5-1.0)	7	5	4
3800-19000 (1-5)	11	14	11
19000-38000 (5-10)	–	–	–
3800-190000 (10-50)	1	–	2
>190000 (>50)	–	1	–
	53	30	34

Note: List compiled Spring of 1984.

phosphor for application to the walls of the tube (to convert UV light to visible light) and efficient long-lived electrodes was accomplished in the 1930s. By the 1940s the fluorescent lamp was a commercial reality. Although there was no significant demand for the UV lamp ("germicidal lamps") per se, the successful commercial development of the fluorescent lamp technology resulted in the immediate availability of a relatively inexpensive, efficient, UV source.

7.2.2 Mechanism of UV Disinfection

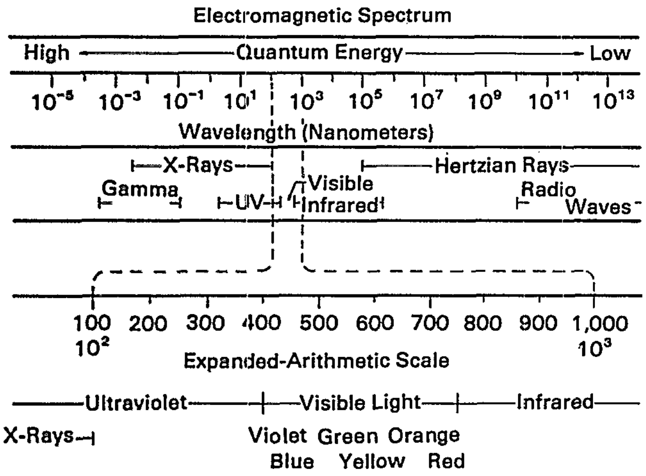
One of the earliest reports relating to the germicidal effects of UV was by Downes and Blount (9). They described the lethal effects of solar radiation on a mixed microbial population and assigned the cause of these effects to shortwave UV radiation.

The early interest in the application of UV for disinfection centered first on potable water. The equipment was not reliable, however, and the lamps were highly inefficient, as discussed earlier. Chlorine was becoming readily available by the early 1900s and was inexpensive. Chlorine also exhibited the very real benefit for potable water applications; this was the ability to maintain a residual. Interest in the application of UV subsequently faded, but not in the effects of UV or its mechanism. Research continued, at first centering on the effects of UV on different organisms and the optimum conditions for germicidal effectiveness. The more recent research, conducted primarily since the early fifties, was directed to the actual mechanisms by which radiant energy affects an organism. These research efforts have been well documented in the literature and it is not the intent of this discussion to give a detailed accounting. Rather a brief description of the basic mechanisms is provided. The reader should refer elsewhere for greater detail (10-17).

The basic premise to understand is that radiation must be absorbed before it can have an effect. Visible light is absorbed by molecules called pigments; color is observed by reflectance or transmittance. Radiation outside the visible spectrum can also be absorbed. Proteins and nucleic acids are basically colorless, but strongly absorb invisible shortwave UV light.

Recall from the earlier discussion of quantum energy that it is constant for a given wavelength, and will change as a function of the wavelength. The longer the wavelength the lower the energy (see Equation 7-2); conversely, the shorter the wavelength, the higher the energy. The effect of a quantum when it interacts with matter is a function of its energy content. Referring to the electromagnetic spectrum on Figure 7-9, infrared radiation at wavelengths greater than 1200 nm has relatively little energy and is unable to effect any chemical change. The energy is immediately converted to heat (hence the infrared heat lamps). At wavelengths from 1200 nm (near infrared) to about 200 nm (far UV) the energy content is sufficient to produce photochemical changes. Radiations with wavelengths less than 200 nm (extreme UV, X-rays, gamma rays, and cosmic rays) have energy contents so high that molecules in their path become ionized.

Figure 7-9. **Electromagnetic spectrum.**



Living organisms can use parts of solar radiation advantageously. The obvious examples are photosynthesis, phototaxis, and vision. The lethal effects are related primarily to the photochemical changes induced by molecular absorption of radiation. Cellular proteins and nucleic acids are strongly absorptive of far UV radiation; the photochemical changes caused by this absorption are very injurious to living cells, hence the bactericidal properties of UV. The most effective spectral region lies around 260 nm, which is the region of maximal absorption by nucleic acids. Cell death following UV radiation is almost entirely attributable to the photochemical damage of these compounds.

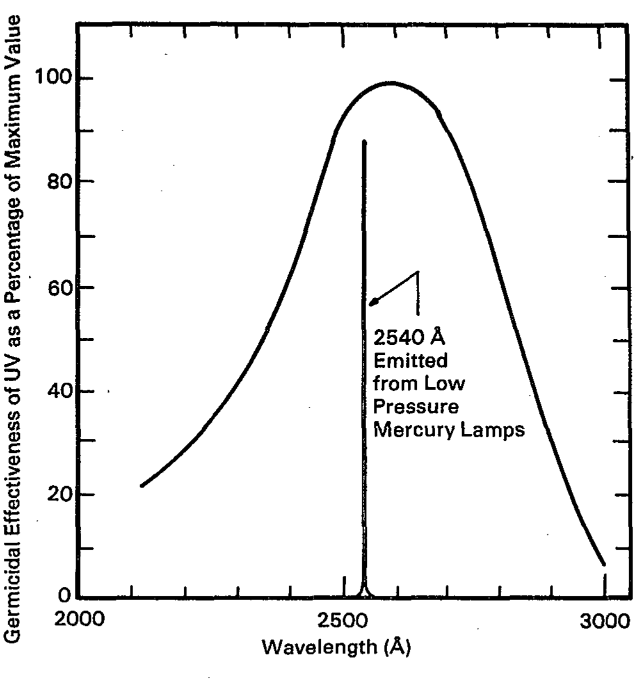
7.2.2.1 Photochemical Damage of the DNA Molecule

Deoxyribonucleic acid (DNA) and ribonucleic acid (RNA) are chain-like macromolecules that function in the storage and transfer of a cell's genetic information. These compounds generally comprise 5 to 15 percent of a cell's dry weight, and effectively define the operations of a cell, particularly the type and quantity of enzyme production. The DNA molecule is considered to be the principal target of UV photons, and the primary component where significant biological effect, or damage, is incurred.

The monomeric units of the DNA (and RNA) are nucleotides. These all have three characteristic components: each has a nitrogenous heterocyclic base which can be either a purine or pyrimidine derivative; each contains a pentose sugar; and each has a molecule of phosphoric acid. There are four different deoxyribonucleotides which comprise the major components of DNA, differing only in their base components. Two are the purine derivatives adenine and guanine; the other two are pyrimidine derivatives cytosine and thymine. Similarly, four different ribonucleotides comprise the major components of RNA. As with the DNA, they contain the purine bases guanine and adenine; the pyrimidine bases are cytosine and uracil. Thus, thymine is characteristically present only in DNA, while uracil is normally present only in RNA.

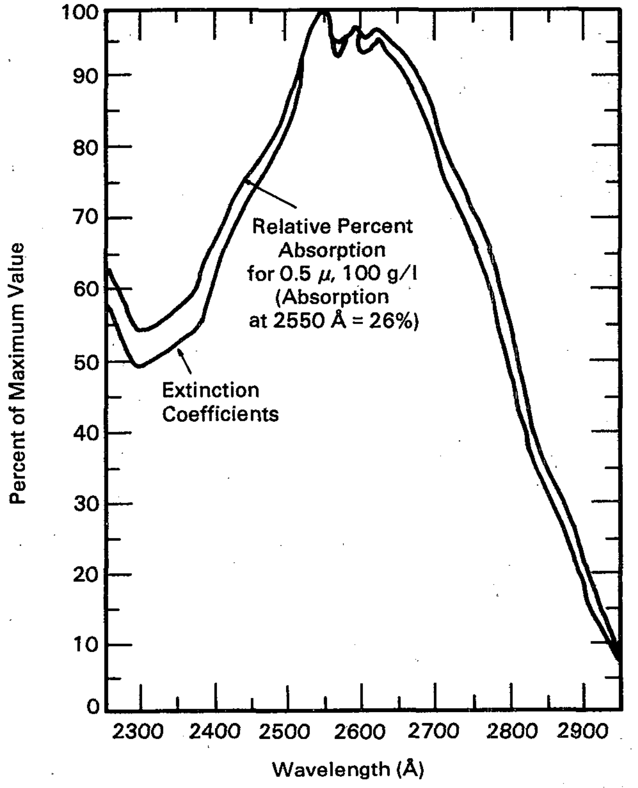
As had been mentioned earlier, the most effective spectral region for germicidal activity lies about the 260 nm wavelength. This is demonstrated on Figure 7-10 which presents relative germicidal effectiveness as a function of wavelength (17). The action spectrum of nucleic acids is very similar to this, as shown by Figure 7-11. On a relative scale, the extinction coefficients (a measure of the inhibiting effect on bacterial colony formation) are plotted as a function of wavelength. Maximal effect is shown to occur between the wavelengths of 250 nm and 265 nm. Overlaying this is the relative percent absorption for a solution of RNA. The similarities are striking, supporting the premise that the lethal effects of UV radiation are induced by the photochemical damage to the cell's nucleic acids.

Figure 7-10. **Relative germicidal effectiveness as a function of wavelength (17).**



The photochemical changes induced by UV radiation on the DNA of an organism have been thoroughly studied. Although several mechanisms exist, the most dominant is the dimerization of two pyrimidine molecules. To visualize this effect, consider the

Figure 7-11. **Relative abiotic effect of UV on *E. coli* compared to relative absorption of Ribose nucleic acid (16).**

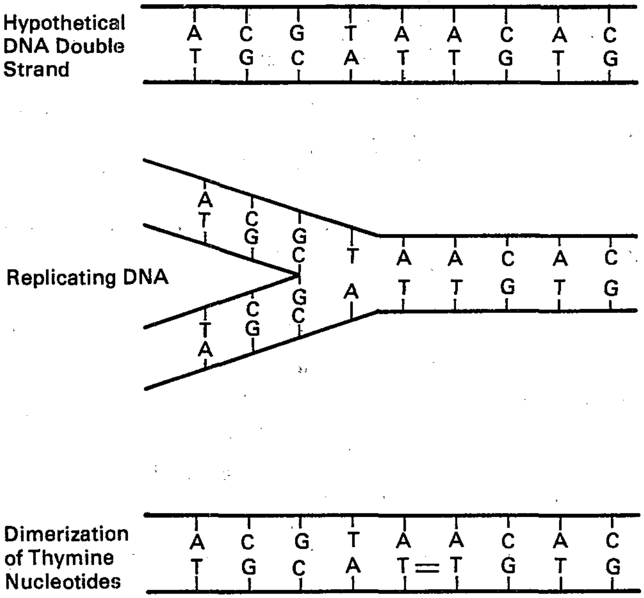


schematic representation of the DNA molecule on Figure 7-12. Recall that the DNA is a long polymer comprised of a double helix chain of simple monomeric units called nucleotides. The order of these nucleotides constitutes the genetic information of the cell. These are represented on the Figure by the letters A (adenine), G (guanine), C (cytosine) and T (thymine).

In the two strands, G is always opposite C and T is opposite A; if damage occurs in one strand the information still remains in the second strand. Thus, to repair the damage, a C is inserted opposite a G and a T opposite an A, and so on. As long as the information is retained on one strand, the second strand damage can be rebuilt. These are enzymatic processes. Before cell division occurs, a duplicate of the DNA is prepared by building a complementary strand to each of the parental strands.

The UV induced dimer between two adjacent pyrimidines in a polynucleotide strand has been demonstrated for all combinations of the pyrimidines (thymine, cytosine, and uracil). The thymine dimer is formed with the greatest efficiency, however. This is shown on Figure 7-12. There are two adjacent

Figure 7-12. **Example of DNA and UV damage to DNA (18).**



thymine monomers on one of the strands; during exposure to UV light new bonds are formed between the two such that a double thymine molecule, or dimer, is formed. Formation of many dimers along a DNA strand makes replication very difficult.

No comparable interactions of the purines have been demonstrated. The effect of the pyrimidine dimerization is a blocking of normal replication. Total and permanent inhibition of DNA replication would in itself be a lethal event.

Alternatively, replication may bypass such a distortion, producing an error in the copy and a subsequent mutant daughter cell which is unable to replicate.

7.2.2.2 Recovery from Photochemical Damage

Just as a cell can be lethally affected by photochemical damage, there is a widespread prevalence in the world of living organisms to repair and reverse the lethal effects of UV. The mechanism is typically a photoenzymatic repair, requiring longer wavelength light in the near UV and visible spectrum. This phenomenon, unique to UV, has been broadly termed photoreactivation. Jagger and Stafford (19) suggested an explicit definition: "the reduction in response to far-ultraviolet irradiation of a biological system resulting from concomitant or post-treatment with non-ionizing radiation."

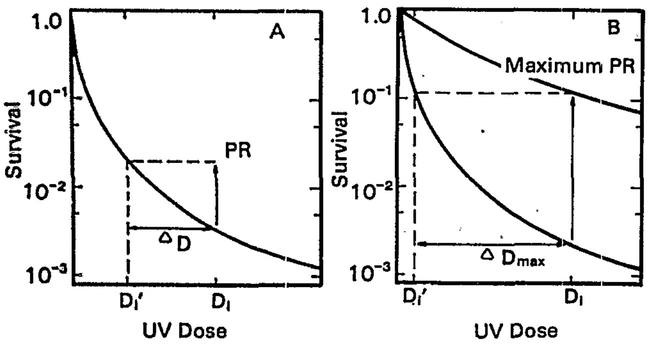
Although not explicitly characterized as such, photoreactivation effects were noted during the first half of this century. These observations related to the counteracting effects of visible and UV light. The discovery of the phenomenon is generally attributed to Kelner (20) and Dulbecco (21), working independently in the late forties. The reader is referred to Harm et. al. (22) and Harm (23) for detailed discussions and reviews of research of this phenomenon.

The repair mechanism is not universal and there is no clearly defined delineation of characteristics which suggest which species would have the ability to repair and which would not. Organisms which have been shown not to have the repair mechanism include *Haemophilus influenzae, Diplococcus pneumoniae, Bacillus subtilis,* and *Micrococcus radiodurans*. Viruses generally do not have the repair ability except when in a host cell which can repair. Organisms shown to photorepair include *Streptomyces, Escherichia coli, Saccharomyces, Aerobactor, Micrococcus, Erwinia, Proteus, Penicillium,* and *Nuerospora*.

The catalyzing, non-ionizing radiation wavelength is not the same for all. It generally falls between 310 nm and 490 nm. In some cases photorepair has been induced by radiations between 230 nm and 240 nm (although these wavelengths are absorbed in the atmospheric ozone layer and would not naturally occur at the earth's surface). It is important to note that photoreactivating light is present in sunlight and, as such, is universally available. The effects are quick, occuring within minutes after exposure to the necessary reactivating light.

Observation of the effect has been accomplished by comparing, after UV radiation, the "dark survival" of cells with "photoreactivated survival." Quantitatively, this is described as the dose decrement. Referring to Figure 7-13, consider a dark survival curve as a function of UV dose. After a UV dose D_i, the culture is exposed to photoreactivating light and the survival increased to the level marked by PR; the resulting survival can be considered equivalent to the dark survival accomplished by dose D_i'. The difference, D_i-D_i', called the dose decrement, can be used as a measure for the extent of photoreactivation. Since the number of lethal photoproducts (thymine dimers) is directly proportional to the UV dose, the dose decrement can be considered a measure of the number of repair events.

Figure 7-13. **Schematic representation of the effects of photoreactivation (23).**



In the case of maximal photoreactivation, the dark and PR survival curves differ by a constant displacement factor. The two curves would coincide if the dose scale for one of the curves was changed by an appropriate factor. This was described as the "principle of constant dose reduction" and the displacement factor was called the dose reduction factor (24):

$$D_i'/D_i$$

this suggests the repair of a constant fraction of lesions. This fraction,

$$1 - (D_i'/D_i)$$

is called the photoreactivable sector, or PRS_{max}.

There are several mechanisms by which these repairs can be made. Rupert (67) established that the most dominant repair mechanism was by a photoreactivating enzyme (PRE). It is similar to other cellular enzymes, except that it requires light energy to initiate its activity. The reaction scheme suggested for this enzymatic repair can be expressed as:

$$E + S \underset{k_2}{\overset{k_1}{\rightleftharpoons}} ES \xrightarrow{k_3} E + P \qquad (7\text{-}3)$$

This is a conventional Michaelis-Menton expression for enzymatic reactions, except that the rate k_3 is absolutely dependent on light energy. E is the photoreactivating enzyme, S is the substrate (the photorepairable lesion), ES is the enzyme-substrate complex, and P is the repaired UV lesion. The enzyme binds the pyrimidine dimers, and upon exposure to the appropriate light energy monomerizes the dimers. The complex formation and dissociation can occur in the dark; the value of k_3, however, is zero in the dark.

A second mechanism has been demonstrated to occur without the light requirement, called dark repair. It is a multi-enzymatic mechanism (termed excision repair) in which the dimer is recognized by an enzyme; this enzyme nicks the dimer from the DNA strand on one side. An exonuclease then releases the dimer completely from the DNA strand and a replicating DNA enzyme then repairs the gap.

Environmental conditions which tend to inhibit active cell metabolism and cell division for a time after exposure will tend to decrease the effects of UV radiation. These inhibitory effects allow time for the cell to repair its DNA before it is erroneously (and lethally) replicated. Such conditions include low temperature or low nutrient levels. Conversely, the repair mechanism will be attenuated by conditions which encourage high growth rates. A population in its lag or stationary growth stage will have a greater chance of recovery because, by definition, it is not replicating its DNA as quickly as an exponentially growing population. The recovery of irradiated phage is also dependent on the physiological condition of its host cell.

Photoreactivation is a phenomenon which can impact the performance and design of a UV system in certain situations. The conditions which exist in a treated effluent are conducive to the occurrence of photoreactivation; the nutrient levels are low and the population of organisms would likely be in the stationary growth stage at the point of the disinfection process. There are several variables involved in predicting the recovery effect in systems such as a wastewater treatment plant. Certainly sunlight, the source of the photoreactivating light, will differ in intensity and spectral distribution according to the season, time of day, and cloud cover. Effluent characteristics will affect the penetration of the photoreactivating wavelengths; this will in fact extend to the receiving water conditions. Shallow, clear receiving streams will be more conducive to repair than discharge to deeper, slow-moving, and turbid receiving waters.

Given the environmental factors which influence the degree and effect of photoreactivation, it is likely far more practicable to control (i.e., account for) the mechanism by increasing the applied UV dose. Manipulating the growth stage of the microorganisms would not be a practical operation in typical wastewater treatment operations. By designing for an increased UV dose, however, the effect of photoreactivation can be concurrently reduced (it cannot be eliminated). This was demonstrated by the discussion of Figure 7-13.

7.2.3 Recent Application of UV to Wastewater Disinfection

The following discussions present a summary of the major studies which have been conducted to date with regard to the application of UV for wastewater disinfection. Several of these studies will be referenced again in subsequent sections. The intent at this point is to present an overview of the evolution of the technology and to identify those efforts which were and are important to the current state-of-the-art. The review is limited primarily to direct applications to wastewater disinfection and, as will be noted, to work reported after 1970. Before this, three studies are presented which, although not directed to wastewater treatment, were important in their evaluation of the applicability of UV on a large scale, its effectiveness, and the factors which would influence its design.

Kelly (25) reported on a study which evaluated the ability of UV to disinfect seawater used for the depuration of oysters. Tests had shown that the activity of the oysters was adversely affected when the water was treated by chlorination/dechlorination. Two different designs were set up to apply the UV; one was at Pensacola, Florida and the other was in Purdy, Washington. Both were tray designs in which the lamps were suspended over a shallow tray which received continuously flowing water. The Pensacola unit was the deeper of the two—6.4 cm (2.5 in); the slow moving system tended to accumulate particulates. The lamps and reflectors were cleaned on a daily basis and the troughs were flushed regularly.

The unit at Purdy was found to be a far more efficient system operationally, most likely because of the higher velocities, the thinner film thickness of the water, the greater agitation, and the improved flow distribution. The liquid level was kept at 1.9 cm (0.75 in) by the placement of a downstream weir. A perforated pipe provided for equal lateral distribution at the inlet to the unit. Six internal baffles were then installed to provide a rolling motion to the liquid as it traveled down the length of the unit. Forced draft ventilation of the unit was also provided in an attempt to optimize the operating temperature of the lamps.

The studies demonstrated the effectiveness of the systems in reducing the coliform levels by greater than three logs from an initial density between 1,000 and 10,000 MPN/100 ml and at turbidities up to 20 JTU. Kelly stated that a dose of 57,600 μWatt-sec/cm^2 was required to accomplish this. The intensity at the surface of the liquid was computed by the author by simply distributing the rated output of the lamps (unreflected) across the surface of the water. The pilot scale studies suggested that, with proper maintenence, the unit could operate with a high degree of dependability. The maintenance requirements were of a housekeeping nature; flushing the units on a periodic basis and cleaning the lamps and reflectors of accumulated debris.

Huff et al. (26) demonstrated that ultraviolet disinfection would be effective in treating ship-board potable water supplies. Doses varied between 4,000 and 11,000 μWatt-sec/cm^2; the intensity was measured by an intensity meter on the side of the UV unit. Effective kill of *E. coli*, *A. aerogenes*, and *S. faecalis* was accomplished. The study also studied the attenuating effects of turbidity, color, and iron on UV intensity and consequent effects on the performance of the UV unit. The apparatus was shown to effectively inactivate certain enteric viruses when operated at the recommended intensity and flow rates.

Hill et al. (27) followed up, in a sense, on the work of Kelly and others in investigating the application of UV for the disinfection of seawater. Their efforts were directed to the virucidal efficiency of the process, which had become the treatment of choice for disinfecting seawater that is to be used for shellfish depuration systems. The Kelly-Purdy UV Seawater Treatment Unit was used for the study. Static tests first demonstrated the effectiveness and rate of inactivation for eight enteric viruses. The exposure required to obtain effective disinfection (99.9 percent reduction) at an applied intensity of 1160 μWatt/cm^2 was as follows:

Poliovirus 1	28 seconds
Poliovirus 2	31
Poliovirus 3	27
Echovirus 1	28
Echovirus 11	31
Coxsackievirus A-9	31
Coxsackievirus B-1	40
Reovirus 1	40

These static bioassays were conducted in shallow, unstirred petri dishes. The devitalization rate determined from dynamic tests with Poliovirus 1 were found to be significantly different than for the same virus under the conditions of the static test. The rate was, in fact, significantly increased in the flowing seawater system. This difference was attributed primarily to UV dose and the mixing effects provided by the unit. In all, the study concluded that the continuously flowing UV units would be highly effective for the inactivation of viruses in contaminated seawaters.

At approximately the same time, a federally sponsored study was investigating the application of UV to wastewaters which were being discharged to shellfishing waters in St. Michaels, Maryland. Roeber and Hoot (28) used a system similar in design to the Kelly-Purdy shallow tray unit to disinfect the effluent from an activated sludge plant.

First order reductions were observed in total coliforms and bacteriophage densities, with a tailing effect after 99.99 percent kill of the total coliform. The average dose required to reduce coliform densities to 70 MPN/100 ml or less was estimated by the investigators to be 25,000 μWatt-sec/cm^2. The intensity was calculated on the basis of measured estimates of the absorbance coefficient of the wastewater at 253.7 nm and direct measures of the intensity in the liquid. The average intensity was then estimated by integration of the Beer-Lambert equation over the fluid depth:

$$\text{Average Intensity} = I_o\,[(1-e^{-kd})/kd] \qquad (7\text{-}4)$$

where:

I_o = initial intensity (μW/cm^2)
d = depth of the fluid (cm)
k = the absorption coefficient (cm^{-1})

This average intensity would be multiplied by the average detention of the system in order to estimate the applied UV dose. The ultraviolet transmittance of the liquid averaged approximately 65 percent; its value was related more to the organic makeup of the water (as COD) than the turbidity. The study indicated a dependence on the initial coliform density and suggested that higher turbidity levels would affect the unit's performance.

Roeber and Hoot also presented the results of a series of photoreactivation tests. Coliforms and bacteriophage were shown to exhibit significant repair upon exposure to sunlight for one hour. In all, the report concluded that ultraviolet disinfection would be practicable for application to well-controlled activated sludge plant effluents. They suggested that at optimum conditions, an energy consumption of 0.092 kWh/m^3 would be required to accomplish a coliform level less than 70 MPN/ 100 ml.

Singer and Nash (29) reported on a study which evaluated UV disinfection of a secondary step aeration plant effluent. The UV system was one normally used in potable water applications and consisted of a closed vessel containing nine germicidal lamps. The lamps were sheathed in quartz sleeves and were continually submerged. Flow would enter the rectangular reactor at one end perpendicular to the lamps, turn and flow down the length of the reactor and exit through a pipe located on the same side as the inlet pipe. This was one of the first reported applications of the submerged quartz configuration to wastewater disinfection. A water quality meter was attached which measured the intensity at a single point in the reactor. This was found to inversely correlate with the turbidity, suspended solids, and BOD of the wastewater.

Singer and Nash reported a tailing of the first order reduction with increasing dose, such that a base level of coliforms would be present in the waste. This was attributed to a degree of short-circuiting and the suspended solids in the wastewater. Effluent levels less than 200 MPN/100 ml could be consistently achieved as long as suspended solids levels were kept below approximately 22 mg/l. The report concluded that UV would adequately and cost-effectively disinfect a step aeration effluent and recommended further work to investigate closer spaced lamps to counteract the poor quality of the effluent and suggested that greater attention be paid to the hydraulic characteristics of the reactor. The quartz surfaces were found to require a periodic chemical cleaning, although the appropriate frequency was not determined.

Oliver and Carey (30,31) investigated methods to apply UV to conventional wastewater treatment systems. In a sense, the scheme was a modification of the Kelly-Purdy design; lamp units (double lamp units with reflector and ballast) would be suspended above secondary clarifiers, close to the clarifier overflow weirs. The effluent would be exposed to about the same dosage as it rises to the surface and passes over the weir in a thin film. Earlier laboratory studies indicated that relatively low dose levels were required to achieve a 2-log reduction (Log survival = -2) in total and fecal coliforms and fecal streptococcus (32). They also determined that the bacterial inactivation was independent of light intensity; thus, a relatively low intensity arrangement could be implemented as long as sufficient exposure time were provided to achieve the desired dose. A significant result of these tests was the demonstration of the bacterial occlusion by particulates in the water. When sonication was applied as a pretreatment before UV exposure, greater inactivation efficiencies were accomplished. The ultrasonics appeared to disperse the particulate aggregates, making the bacteria contained in the particles more susceptible to the UV radiation.

Pilot studies of the overflow weir arrangement were conducted and provided excellent results. The investigators also noted that the lamps did not foul and maintained a constant output over a six-week period. The absorbance of the effluent ranged between 0.12 and 0.25 a.u./cm with little direct effect on the system efficiency. A dose of 130 Watt-sec/imperial gallon was reported to achieve a reduction of 2 logs. This is relatively inefficient, reflecting the ineffective use of the lamps' output by having them suspended above the surface of the wastewater. Greater energy efficiency would obviously be obtained by submersing the source in the liquid. Nevertheless, the report concluded that UV would be highly effective and would be competitive with the use of chlorine.

A limited study conducted in Syracuse, New York, evaluated the use of UV for the disinfection of waters from combined sewer overflows (CSO) (33). One liter samples were irradiated in a bell-jar vessel in which the walls were equidistant—5.7 cm (2.25 in)—from the UV lamp. The intensity at the lamp surface was computed to be 5,800 μWatts/cm^2. The study suggested that a dose of 500,000 μWatt-sec/cm^2 would be required to achieve a residual coliform level of 2500 MPN/100 ml and concluded that UV disinfection of CSO waters was feasible and particularly attractive because of the absence of a residual. The scheme studied by Oliver and Carey (30), which was suggested for the CSO application (the UV lamps would be suspended over high rate swirl separators) would be impractical, however, because of the high dose requirements.

Petrasek, et al. (34), reported on a study conducted in Dallas, Texas, which investigated the feasibility of ultraviolet disinfection of treated municipal wastewater effluents to achieve fecal coliform levels less than 200 MPN/100 ml. Two system configurations were evaluated: the first system was the Kelly-Purdy shallow tray design; and the second system was a closed vessel design with the lamps (enclosed in quartz sleeves) submerged in a 53.6-liter (14.1-gal) chamber. The flow was parallel to the lamps.

The investigators concluded that the more appropriate system configuration for wastewater disinfection was the submerged quartz system. The Kelly-Purdy design, although effective in disinfection, was less efficient in the utilization of UV energy, would require greater space, and was susceptible to solids deposition in the unit. Hydraulic studies indicated that the unit operated poorly at low flows and at deeper liquid depths. Actual retention times approached theoretical when the unit was operated at shallow depths—2.5 cm (1 in)—and higher velocities. Relatively high dispersion was still observed under these conditions.

The hydraulic analyses of the submerged system also indicated relatively poor flow characteristics when compared to the ideal condition of plug flow. The time distribution curves constructed for a number of flows indicated a high degree of dispersion. The investigators also demonstrated the importance of the absorbance coefficient of the wastewater. This was spectrophotometrically measured at 253.7 nm and effectively quantified the UV "demand" of the liquid. The parameter was not found to be affected significantly by the suspended solids or turbidity of the water.

The authors recognized the lack of any direct measurement capability for estimating the UV intensity within a complex multi-lamp system, such as the submerged quartz unit. An approach suggested by the report was to mathematically calculate the intensity at any point in the reactor; they assumed that the radiation emits perpendicularly from the lamp and that the lamp is an infinite line source. The average intensity was then calculated on an areally-normalized basis. The technique accounted for the attenuation of intensity due to the absorptive characteristics of the liquid. No attempt was made to account for the deterioration of the lamp output or the quartz surfaces. The estimated intensity was found to correlate well with the total coliform reduction of the unit at a constant flow to the unit.

The authors found that UV was effective in the inactivation of Type 1 poliovirus and F2 coliphage. The coliphage was suggested as a good indicator for the inactivation of poliovirus. Photoreactivation was also investigated; static tests indicated that subsequent exposure of UV irradiated coliforms to sunlight for 30 minutes induced a 1.1 log increase in total coliforms, and a 0.6 log increase in fecal coliforms.

Scheible and Bassell (35,36) reported the results of a full-scale prototype demonstration study conducted at the Northwest Bergen County Water Pollution Control Plant, Waldwick, New Jersey. The unit, similar to that shown on Figure 7-5, was a submerged quartz system comprised of 400 lamps. These were arranged in a symmetrical array, axially parallel to one another, and equidistant on horizontal and vertical centerlines. The flow path was perpendicular to the lamps. The spacing between quartz surfaces was only 1.25 cm (0.5 in), imposing a thin film of liquid as the wastewater passed through the lamp battery. The lamp battery was inserted into two bulkhead walls constructed in an existing chlorine contact chamber. The arrangement simulated an open channel, with the lamp battery inserted into the channel and the inlet and outlet planes of the lamp battery wholly exposed to the wastewater.

The time-distribution characteristics of the unit were not measured directly, although it was established that a relatively uniform velocity field existed across the exit plane of the lamp battery. The time of exposure was assumed to be the theoretical detention time of the unit, which ranged between one and four seconds under normal operating conditions. In order to estimate the dose under a given set of sampling conditions, the authors incorporated the use of the radial light model, as had been desribed by Petrasek, et al. (34), to calculate the intensity. In this case, however, the authors calculated the incident intensity at the surface of the lamp (dividing the rated UV output by the quartz surface area) and accounted for some deterioration in UV output with time. The lamps were presumed to be transparent to UV from a neighboring lamp, which is not the case, and likely resulted in an overestimate of the applied dose. The intensity calculations were demonstrated for any given symmetrical system and presented as a function of spacing, UV absorbance coefficient, and lamp rating. Dose was estimated by multiplying the intensity (which varied as a function of the UV absorbance coefficient and the output of the lamps) by the theoretical detention time.

Empirical regressions were developed to describe the performance of the system as a function of the applied dose. These reflected more an attempt to linearize the correlation of the coliform reduction with the dose; this was effective in determining the system performance and design requirements for the specific plant application. The UV absorbance coefficient was suggested to be the key wastewater parameter for the design, control, and monitoring of the UV disinfection process. It correlated well in this case with the wastewater COD.

An extensive series of static and dynamic photoreactivation tests were conducted during the study. The static-bottle technique, in which a light and dark bottle are held in sunlight for a period of one hour, was suggested as an effective and practical procedure for evaluating and/or monitoring the effects of photoreactivation. The results of the tests indicated a degree of temperature dependency, although this may have implicitly included the influences of reduced sunlight intensity during the winter months. At 10°C (50°F), a 0.3 log increase in coliform density was observed, while at 20°C (68°F) the increase was approximately 1.0 log. A detailed economic evaluation showed the cost of the process to be approximately \$0.008/m^3 (\$0.012/1,000 gal) for secondary treatment plants (1979\$). When compared to alternative processes, UV was found to be less costly than ozonation, more expensive than chlorination, and competitive with chlorination/dechlorination.

Severin (37) reported on the application of a commercially available UV system to the disinfection of a variety of treated municipal effluents. In determining dose, he estimated the average intensity by the integrated solution of Beer's law over a fluid depth, as presented by Roeber and Hoot (Equation 7-4) and by Petrasek et al. The UV unit was a closed flow-through vessel containing 10 lamps set longitudinally; flow was directed parallel to the lamps. A nominal average depth of liquid was computed for the reactor to use in the computation of the average intensity. Severin assumed a single source in the estimate of intensity and did not account for the additive affects of a multi-lamp system.

Experiments were conducted at a number of wastewater treatment plants which provided different levels of treatment. The quality of the wastewaters was very good in most cases, except where the effluents were artificially adjusted to yield higher absorbance values. Typically, the effluents had an absorbance coefficient between 0.2 and 0.3 cm^{-1} (base e). The inactivation results of the overall experimental effort were described by a disinfection model which was linear with time to the one-third power. Time was assumed to be the theoretical detention time of the unit. The least squares regression yielded the expression:

$$\mathrm{Log}_{10}(100\ N/N_o) = -1.73\ (P_{avg}/P_o)\ (t^{1/3}) + 2.598 \qquad (7\text{-}5)$$

The expression was explicitly described as an empirical relationship applicable only to similar reactors under the same water quality conditions. P_{avg}/P_o is the ratio of the average intensity to the incident intensity.

Hydraulic tracer analyses of the system indicated that the unit did not behave in a plug flow fashion but exhibited a relatively high degree of dispersion. The author indicated that this deviation from ideal plug flow was not enough to account for the empirical observation that disinfection is a funtion of time to the one-third power. Overall, the conclusion was that UV was highly effective for wastewater disinfection. Further work was recommended in the areas of intensity estimates and the direct analysis of the impact of hydraulics and the effect of channelized flow in a non-uniform intensity field.

Johnson and Qualls (38) reported the results of a substantial pilot and laboratory scale effort which focused on a number of parameters which were key to the understanding and design of the UV process. In the early phases of the work they investigated the performance of two commercial units on contact stabilization effluent. The first unit was based on the close spaced, or thin film concept and contained fourteen 25 Watt submerged lamps at nominal spacings of 1.25 cm (0.5 in). The second unit utilized six 40-Watt, widely spaced lamps submerged in an 11 liter exposure chamber. Significant differences were noted, with the widely spaced unit providing far better performance. This was attributed to short-circuiting occurring in the first unit. The study also reported significant photoreactivation in UV irradiated samples which had been exposed to sunlight for 45 minutes at 25°C (77°F). Total coliforms were found to increase in density by 1.4 logs. The degree of photoreactivation was found to decrease with decreasing temperature.

A point source summation model was applied by the authors to estimate the intensity in a multi-lamp system. This calculation method treats a finite line source (the tubular germicidal lamp) as a series of point sources radiating in all directions. The attenuation of the radiation was inversely proportional to the square of the distance from the point source and was absorbed by the liquid medium according to the Beer-Lambert law. The intensity at any given point in a system was assumed to be the sum of the radiation received from all point sources in the system. The authors also demonstrated that direct beam, spectrophotometric measurement of the absorbance of the liquid overestimated the UV absorbance because scattered light would be measured as absorbed light, when in fact it is still available. The true absorbance of the liquid should be estimated by correcting for this scattering effect. The authors also demonstrated a spherical integration method for measuring the actual output of a lamp at any given time.

Qualls et al. (39), reported on the effect of suspended solids on the performance capability of the disinfection process. Aside from the effect on the UV absorbance characteristics of the liquid, the major impact of the suspended solids normally found in a biologically treated domestic wastewater is the harboring or occlusion of the bacteria within the particle. These are protected from the UV radiation and will be measured as viable organisms subsequent to UV disinfection. The authors suggest that these protected coliforms are the major factor in limiting disinfection efficiency at -3 to -4 log survival units. Improved disinfection would be accomplished (if necessary) by prefiltration or improvements in solids-liquid separation steps at a typical plant.

A bioassay approach to estimate the actual dose in a system was proposed by Qualls and Johnson (40). A dose-response relationship would first be developed for a known, pure culture (spores of *Bacillus subtilis* were suggested as an appropriate test organism). This involved exposing the organism to a known, and measurable, intensity of collimated light (at 253.7 nm) over several exposure times. Dose would be the product of this measured intensity and the time of exposure. The calibrated spore would then be injected into a system. The response (log survival ratio) would then be compared to the calibration curve to determine the effective dose delivered by the UV unit. This technique was further applied to a dynamically flowing system concurrent with a conservative tracer to determine the time distribution characteristics of the unit. The spore survival ratio is measured, in this case, at several times after injection. By accounting for the hydraulic distribution in this fashion, the authors demonstrated that it was possible to implicitly solve for the average intensity within the unit. The estimates of intensity by the point source summation calculation method were found to compare favorably to the intensity estimates made by the bioassay technique. The authors proposed the use of this technique to evaluate, and/or compare UV systems, and as a method to separately evaluate the intensity and residence time distribution.

Bellen et al. (41), applied the bioassay technique to seven commercial UV units to determine the dose application performance of each unit and to compare their effectiveness. The application was to be shipboard potable water supplies. They coupled this with a separate analysis of the residence time distribution for each unit. The bioassay method was suggested as the only available technique to directly compare the performance of commercial units.

Haas and Sakellaropoulos (42) presented a series of rational analysis solutions for UV disinfection which incorporated first-order inactivation kinetics with the hydraulic characteristics of the UV reactor. These were allowed to range from a completely mixed reactor to a perfect plug flow reactor. Their analysis supported the premise that the hydraulic characteristics of a reactor can strongly influence disinfection efficiency. The optimum hydraulic regime for disinfection involves turbulent flow with minimal axial mixing. Plug flow would be supported in systems with high aspect ratios (ratio of length to hydraulic radius). In similar fashion, Severin et al. (43) concluded that mixing in the radial direction (perpendicular to the direction of flow) was beneficial to disinfection efficiency. Mixing in the longitudinal direction (direction of flow) was not advantageous, although they suggested that some degree of axial (longitudinal) mixing may have to be accepted in order to ensure adequate radial mixing. Severin et al. (44), suggested the use of a series event inactivation kinetics model to describe the inactivation of coliforms in reactors of differing mixing characteristics. The model effectively predicted efficiency, and suggested the importance of discouraging stratified flow conditions (lack of radial turbulence) in a non-uniform intensity field.

Nehm (45) reported on pilot plant studies conducted at the Nine Springs Water Pollution Control Plant, Madison, Wisconsin, which evaluated commercial units from four different manufacturers. The objectives of these studies were to assess operational requirements; no attempts were made to measure dose or to scale-up to design. The results showed that consistent performance could be accomplished by all systems when they were kept clean. Scale formation occured on both quartz and Teflon surfaces. This was found to be readily removed by the addition of a citric acid solution to the reactor. The report recommended the capability of chemically cleaning on any scale-up, in addition to a UV intensity sensor to monitor the status of the quartz surfaces. As a post-script to these studies, UV disinfection was recommended for the 2,200 L/s (50 mgd) plant, and the system is currently under construction. This was also one of the first major plants to require bioassays of equipment which could be scaled up so as to compare the performance capability of commercial units. This was imposed as a pre-bid qualification specification.

Ho and Bohm (46) and Bohm et al. (47) reported on the pilot scale application of UV to the disinfection of a number of tertiary and secondary effluents at Ontario (Canada) Water Pollution Control plants. The process was demonstrated to consistently meet effluent disinfection goals. Total coliform and fecal coliform densities were targeted at 2000 and 200 MPN/100 ml in the effluent, respectively. Corresponding reductions were demonstrated for *Pseudomonas aeruginosa*, fecal strepococci, and *E. coli*. *Salmonella spp.* were reduced to less than 4/100 ml in 80 percent of the samples. The investigators propose that UV transmission is a good surrogate parameter for correlating effluent water quality to expected UV effectiveness. The authors estimated dose by calculating intensity from the Roeber and Hoot model and the measured mean contact time.

Bohm et al. (47), also assessed the effects of photoreactivation. Total coliforms increased by approximately two logs, fecal coliforms by one to two logs. Little or no photoreactivation was shown for *Pseudomonas aeruginosa* and fecal streptococci. Some increase was observed for the *Salmonella ssp.* Temperature did not affect repair, nor did dilutions with stream water by a factor of 1 to 10.

Whitby et al. (48) reported the results of a full-scale evaluation of a commercial UV disinfection process at the Tillsonburg Water Pollution Control Plant, Tillsonburg, Ontario (Canada). The system design, in this case, allowed the use of the plant's existing secondary effluent channels to retrofit a UV unit (see Figure 7-7). The units were comprised of a series of four-lamp (quartz-sheathed) modules. The modules had the lamps arranged in a vertical row; these were suspended on a support frame which had been inserted into the effluent channel. The number of modules was dependent upon the width of the channels. The wastewater flowed parallel to the lamps. The lamp battery was kept submerged at all times by a downstream control gate.

A direct comparison between UV and chlorination in disinfection efficiency was made in this study. UV was found to outperform the chlorination system; if photoreactivation were allowed to proceed in the UV exposed samples, the performance of the two processes was similar. This applied to the total and fecal coliform analyses. Fecal streptococci do not photoreactivate and substantial reductions were accomplished by the UV systems. A spore forming bacterium, *Clostridium perfringens*, which is known for its resistance to disinfection, was also tested. UV was found to be nearly twice as effective as chlorination in the inactivation of this organism. Similarly, the units accomplished greater than 99.97 percent inactivation of bacteriophages, as compared to the chlorination process, which averaged 95.1 percent.

The studies at Tillsonburg also included fish (rainbow trout yearlings) toxicity studies downstream of the plant's discharge which compared the effects of the chlorinated effluent and the UV irradiated effluent. Complete mortality was observed within 24 hours during the chlorination study; the UV disinfection test was non-lethal for a 48-hour exposure period. The report suggests very consistent operation of the system over an 18 month period. Maintenance was minimal; the lamps had been manually cleaned once and the lamp life had extended for greater than 12,000 hours for the reporting period.

Kirkwold (49) reported on the installation and performance of a UV disinfection system at the Albert Lea Water Pollution Control Plant, Albert Lea, Minnesota. The systems (see Figure 7-6) were reported to be operating well, after some startup problems, and performing very effectively. In fact, due to the high quality of the plant effluent, only a small fraction of the overall system is needed on a continuous basis. The author suggests that the cost of operating the UV system is $5.3/1000 m^3($0.02/1000 gal); this is less than half the operating costs estimated for a comparable chlorination/dechlorination process.

Scheible et al. (50) reported on the early studies conducted for an EPA project at a New York City treatment plant. They discussed the development of software to compute the average intensity of UV in a system of any lamp configuration. They presented the computed intensity as a function of UV absorbance and showed the effect of lamp spacing and lamp output. A procedure was also presented to measure and analyze the retention time distribution of a UV system. A subsequent article reported further results of the New York City project, particularly with regard to the maintenance of UV systems and the effect of quartz fouling and lamp aging (51). The development of a disinfection model was also suggested which incorporated the hydraulics of a system and the inactivation rate of coliforms as a function of the intensity.

7.2.3.1 Current Evaluations of UV Disinfection

A significant amount of the information used to assemble this chapter on the UV process is from studies which were as yet unreported. Only drafts can be cited at this time, although these may be formally reported at the time this manual is printed and released. The following is a short review of these studies. Further reference will be made to them as appropriate in the subsequent discussions.

A major research and demonstration project was completed recently under joint sponsorship by the USEPA and the New York City Department of Environmental Protection (52). Conducted at the Port Richmond Water Pollution Control Plant, Staten Island, New York, it investigated the performance of three large scale UV systems in the disinfection of secondary effluent and high rate settled raw wastewater. The systems included two 100-lamp quartz systems which differed only in the spacing between quartz surfaces—1.25 cm and 5.0 cm (0.5 in and 2.0 in). The third unit used Teflon tubes to carry the wastewater, with the lamps suspended outside the tubes. The effluent characteristics were highly variable; the suspended solids ranged between 5 and 50 mg/l, with a UV absorbance coefficient (base e) between 0.25 cm^{-1} and 0.5 cm^{-1}. Primary effluent was also treated to determine the application of UV to combined sewer overflow wastewaters. These were characterized by high suspended solids levels and UV absorbance coefficients between 0.5 cm^{-1} and 1.0 cm^{-1}.

The UV process was found to be very effective in the disinfection of secondary effluent. Log survival ratios between -3 and -4 could be achieved under practical loading conditions. Similarly, it was shown that a log survival ratio up to -3 could be accomplished with primary effluent. The studies also indicated that the quartz systems were more energy efficient than the Teflon system. An empirical system loading rate was suggested to monitor and compare systems; this was the ratio of the flow (Q) to the actual output of the UV system (W), in watts at 253.7 nm. This output would account for the age of the lamps and the degree of fouling on the surfaces through which the energy must be transmitted.

A major element of the Port Richmond study was the development of a protocol for the design of a UV disinfection process. The resulting model incorporates the retention time distribution of the system, and the inactivation rate of the bacteria described as a function of the calculated intensity in the reactor. This intensity was calculated by the point source summation technique and can be adjusted for the measured (or assumed for design purposes) average lamp output and the losses of energy due to absorption by the liquid and the fouling of the quartz (and Teflon) surfaces. The model was found to correctly respond to the variables associated with the UV process when applied to the Port Richmond experimental data. The study demonstrated the importance of suspended solids in the application of UV. Coliforms occluded by suspended particles will not be affected by UV and will, in effect, set the limiting density which can be achieved by UV radiation.

Photoreactivation effects were demonstrated in this study for both total and fecal coliforms. The results showed a constant fraction increase over densities measured immediately after UV exposure, regardless of the initial UV dose. The study also provided suggestions with regard to the maintenance and monitoring of the system which can enhance the efficiency and cost-effectiveness of the process. A cost analysis of the process shows it to be competitive with a comparable chlorination system.

As a follow-up to the funding of UV installations under the I/A program, the USEPA contracted for a number of post-construction evaluations (PCEs) to assess the status of these plants. These were summarized by White et al. (53). Six plants were visited: Pella, Iowa; Suffern, New York; Northfield, Minnesota; Togus, Maine; Eden, Wisconsin; and Lodi, Wisconsin. Overall, the report was favorable to the application of the UV process. It cited the problems relating to equipment fabrication (ballasts, wiring), and O&M requirements which were more than originally anticipated. These will be discussed further in later sections. An earlier survey of a number of UV facilities in 1983 was limited due to the lack of operating experience at the plants.

A series of special studies were conducted at four operating UV plants to assess plant operations, plant performance, and to determine the applicability of the process evaluation techniques developed by the Port Richmond project (4,54). A one-month study was conducted at the Suffern, New York facility. The units (see Figure 7-4) were evaluated for their hydraulic characteristics, and performance in the disinfection of fecal coliforms and fecal strep. Tracer studies indicated some dispersion and a reduced effective volume. Analysis of data showed that the systems were adequately sized and would be able to meet performance requirements at design loads. The tests conducted at Suffern showed that sodium hydrosulfite is a very effective chemical cleaning agent. Side by side testing of the system's ultrasonics cleaning device indicated that it was of limited benefit, and because of its energy costs, not cost-effective.

The Teflon system at Eden, Wisconsin was also evaluated. As expected, the Teflon tube arrangement was good hydraulically. Tracer analyses showed relatively low dispersion. Analyses indicated, however, that the system would not be able to meet requirements under design loading conditions. Although not directly tested at Eden, significantly reduced levels of transmissivity by the Teflon was suggested by the analysis as a major cause of the poor performance. Special studies were conducted to evaluate a method to directly measure the Teflon transmittance, utilizing chemical actinometry techniques. These showed that virgin Teflon transmitted 75 to 85 percent of the UV (the range being a function of thickness). Samples of Teflon from the Port Richmond facility, which were heavily fouled, were found to transmit only 5 to 30 percent of the UV. When the surfaces were thoroughly cleaned, the transmittance increased to 65 to 70 percent. Limited field studies were also conducted at the Vinton, Iowa facility (see Figure 7-3). The hydraulic tracer analyses indicated a large degree of dispersion and a significant reduction in effective volume.

A series of studies were recently completed which used the same pilot plant to test wastewaters at several wastewater treatment plants (55). The tests were directed to investigating the inactivation rate as a function of the intensity, and the level of bacterial occlusion by the suspended solids.

7.3 Process Design of the UV Wastewater Disinfection System

This section presents the design protocol for the UV process. The mathematical expressions which are presented are based on the analysis of the system as a chemical reactor, incorporating the retention time distribution of the system, and the inactivation rate as a function of the intensity of UV radiation within the reactor. The discussions in this section will be in the following format; first, the process design model is presented. This expression is the framework about which the process is designed and the system sized. The data requirements for the system design are then presented in general form. As will be shown, the design information falls into three major categories: hydraulics; UV radiation intensity; and wastewater characteristics (including bacterial sensitivity to UV).

Second, the hydraulic design considerations, as they relate to the UV reactor, will be discussed. These will encompass the flow and dispersive characteristic of the reactor, and the considerations of head loss, turbulence, and effective volume.

Third, the intensity of UV in the reactor will be discussed. The procedure for estimating the average intensity will be presented, including solutions for almost all practical UV lamp configurations.

Fourth, the relevant wastewater quality parameters are reviewed. Aside from the normal water quality parameters we will need to know, such as coliform densities, UV absorbance coefficients, etc., the discussions will also present procedures to determine the coefficients which describe the sensitivity of the coliforms to UV, and the densities associated with particulates normally found in treated municipal wastewaters.

Finally, the design protocol is presented in Section 7.4, incorporating the discussions of this section. The protocol will be demonstrated by using the design example which has been carried throughout this manual.

7.3.1 Process Model to Describe UV Reactor Performance

7.3.1.1 First Order Kinetics for the UV Process

Recalling the discussions presented in Chapter 4, the inactivation of bacteria by UV radiation can be approximated by the first order expression:

$$N = N_o e^{-kIt} \qquad (7\text{-}6)$$

where :

N = bacterial density after exposure to UV (organisms/L^3)

N_o = the initial bacterial density (organisms/L^3)
k = inactivation rate constant (L^2 $Watts^{-1}T^{-1}$)
I = the intensity of the germicidal UV energy (Watts/L^2)
t = time of exposure (T)

The intensity is the rate at which the energy is being delivered to the liquid; in the context of this report, intensity has the unit microwatts per square centimeter (μWatts/cm^2). When multiplied by the time to which an entity is exposed to this rate, the quantity of energy, or dose, is determined:

$$\text{Dose } (\mu\text{Watt-sec/cm}^2) = \text{Intensity } (\mu\text{Watt/cm}^2) \times \text{Time (seconds)} \qquad (7\text{-}7)$$

The rate constant, k, is the slope of the relationship of $\ln(N/N_o)$ as a function of the dose. It is generally held that the intensity and time are reciprocal in their effect on dose.

7.3.1.2 Incorporation of Particulate Coliform Densities

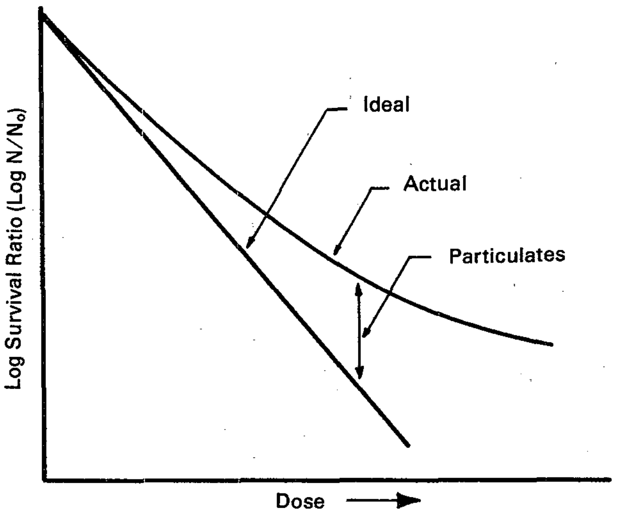
Although the first order expression is generally a good first approximation of the response to a given dose, direct testing on mixed cultures will often show a reduced efficiency with increasing dose. In the disinfection of treated wastewaters by ultraviolet radiation, in particular, this is attributed to the aggregation or the occlusion of bacteria in particulate matter (28,31,33,36). Ultraviolet light is unable to penetrate this material and effect inactivation of the bacteria. Qualls et al. (39) presented data which demonstrated that removal of particles (by filtration) large enough to harbor coliforms exerted dramatic effects on the dose-survival relationships. They concluded that these protected coliforms were the major factor limiting improved disinfection at -3 or -4 log survival units. Thus, as the dispersed or singlet bacterial organisms are inactivated, continued elevation of the dose will show a diminishing response as the residual active bacteria are protected in the particulates. This is schematically presented on Figure 7-14. In light of this, Equation (7-6) can be more accurately written as:

$$N = N_o' \exp(-kIt) + N_p \qquad (7\text{-}8)$$

where N_o' is the initial, non-aggregated density, and N_p is the density associated with the particulates and unaffected by the UV radiation. When considering treated domestic wastewater, $N_o' >> N_p$, such that the total initial density, N_o, can be considered equal to $N_o' + N_p$. The expression can then be written:

$$N = N_o \exp(-kIt) + N_p \qquad (7\text{-}9)$$

Figure 7-14. Effect of particulates on UV disinfection efficiency.



7.3.1.3 UV Process Design Model

Considering Equation 7-9, the use of a single exposure time presumes the ideal case of perfect plug flow in the reactor, with no axial dispersion. Under actual conditions, this ideal plug flow does not exist. Axial dispersion and velocity gradients will cause a distribution of residence times; this will be a function of the dispersion characteristics of the reactor, which can be quantified by defining the spread (or variance) of the time distribution relationship for a specific reactor (see Chapter 4).

A disinfection model was developed and reported by Scheible et al. (56). Reference is made to that report for a detailed development of the model. It presumes the first order expression given as Equation 7-9, but also incorporates the dispersive properties of the reactor, in effect describing the residence time distribution of the reactor under steady-state conditions. This model forms the basis for the design protocol presented herein for the UV process. The general expression is written:

$$N = N_o \exp\left[\frac{ux}{2E}\left\{1 - \left(1 + \frac{4KE}{u^2}\right)^{1/2}\right\}\right] + N_p \qquad (7\text{-}10)$$

where:

N = the bacterial density remaining after exposure to UV (organisms/100 ml).
N_o = the initial bacterial density, measured immediately before entry into the UV reactor (organisms/100 ml)

x = the characteristic length of the reactor, defined as the average distance traveled by an element of water while under direct exposure to UV (centimeters)

u = the velocity of the wastewater as it travels through the reactor (cm/sec). This is calculated as:

$$u = x/(V_v/Q)$$

where V_v is the void, or liquid volume in the reactor (liters) and Q is the total flow (liters/second). In cases where significant "dead volume" is indicated within a reactor, V_v is adjusted to more closely approximate the "effective" liquid volume.

E = the dispersion coefficient (cm^2/second). E quantifies the spread of the residence time distribution of a particular reactor.

K = the rate of bacterial inactivation ($seconds^{-1}$)

N_p = the bacterial density associated with the particulates and unaffected by exposure to UV.

The rate of inactivation, K, is expressed as a function of the UV intensity. Thus, for a given time of exposure (or distribution of exposure times), the rate of inactivation will increase (or decrease) with an increase (or decrease) in the intensity. This is shown graphically on Figure 7-15. In this fashion, an expression can be developed in which K is estimated as a function of intensity:

$$K = f(\text{Intensity}) \qquad (7\text{-}11)$$

As will be discussed in a later subsection, this correlation will be developed by relating the log K to the log Intensity, where the intensity is the average reactor intensity, I_{avg}. This yields the expression (when transformed),

$$K = a\, I_{(avg)}^{b} \qquad (7\text{-}12)$$

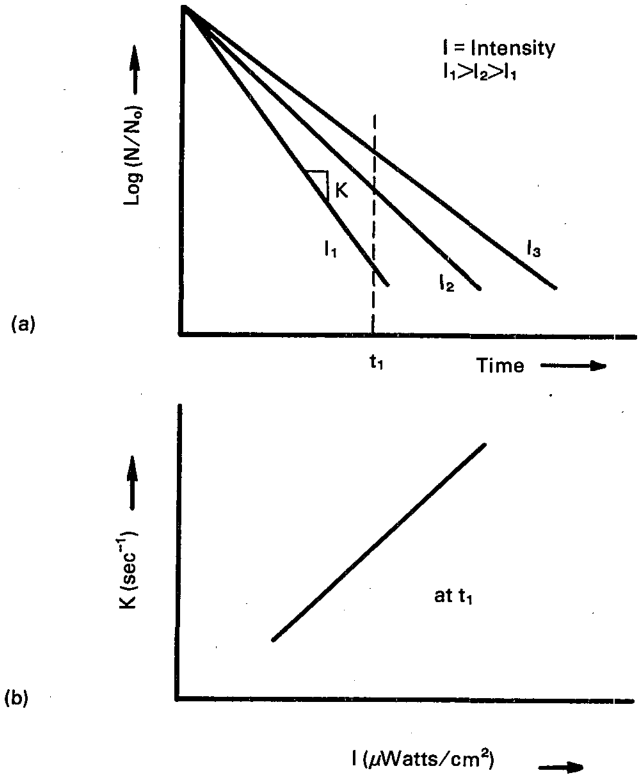
where a and b are the slope and intercept of the linear regression.

With regard to the particulate bacterial density, N_p, this is generally described as a function of some measureable index of particulate density in a wastewater, such as suspended solids or turbidity. Suspended solids is used in the context of this manual since it is the parameter most commonly measured and most relevant to wastewater treatment applications. The value of N_p is described as a function of the suspended solids by the correlation of the log effluent coliform density to the log SS. When transformed, the expression is in the form:

$$N_p = c\, SS^m \qquad (7\text{-}13)$$

where SS is in mg/l. As will be discussed in a later section, the effluent densities for this analysis must be generated under very high dose conditions. In this fashion, it is appropriate to assume that the remaining bacteria are those which were occluded in the particulate matter and were unaffected by the UV radiation.

Figure 7-15. The rate K increases with increasing intensity for a given residence time.



By incorporating Equations (7-12) and (7-13), into Equation (7-10), the UV design model can be expressed as follows:

$$N = N_o \exp\left[\frac{ux}{2E}\left\{1 - \left(1 + \frac{4E\, aI_{avg}^{\,b}}{u^2}\right)^{1/2}\right\}\right] + cSS^m \qquad (7\text{-}14)$$

7.3.1.4 Data Requirements to Use the UV Design Model

The information that would be required to effectively use the UV design model relate to the characteristics of the wastewater to be disinfected, and to the physical characteristics of the reactor itself. Consider first the wastewater application; the data which must be generated either by direct testing, or by estimates based on experience include:

Initial bacterial density, N_o. Explicit in the expression, this should be determined under the average and maximum conditions anticipated for the plant.

Flow, Q. The flows to be handled by the disinfection process. This is implicitly required to determine velocities and loadings to the system.

UV absorbance coefficient, α. This will affect the intensity of radiation in the reactor, and is a direct measure of the energy "demand" of the wastewater.

Suspended solids (SS). This will be defined by the permit limitations the plant is designed to meet. As discussed later, the particulate bacterial density will be related to the suspended solids concentration.

Particulate bacterial density, N_p. This density associated with the particulate forms the minimum density level which can be achieved by the UV process. It is typically determined as a function of the suspended solids concentration.

Coefficients, c and m. These are determined from Equation (7-13) and describe the particulate coliform density associated with the suspended solids.

Rate of inactivation, K. This rate is a measure of the sensitivity of the bacteria to UV radiation, and will be site specific. As discussed, the value of K is estimated as a function of the intensity of UV radiation which a particular reactor can deliver. An estimation of this rate will, therefore, require knowledge of the actual intensity levels within the UV reactor.

Coefficients, a and b. From Equation (7-12), these describe the rate of inactivation as a function of the average intensity.

The remaining parameters which are to be addressed relate to the physical design of the reactor:

Velocity. The velocity of the liquid, as described above, is set by the rate of flow, Q, and by the physical dimensions of the reactor. Specifically, these are the characteristic length, x, and the liquid volume, V_v, of the reactor.

Length. The characteristic length of the reactor is the distance traveled by the liquid while under direct exposure to UV light.

Dispersion coefficient, E. This parameter accounts for the deviation of the reactor's hydraulic behavior from that of perfect plug flow; in effect, the distribution of residence times at steady-state is forced by the dispersion coefficient.

Average intensity, $I_{avg.}$ The average intensity in a reactor is a function of the lamp (i.e., UV energy) density in the reactor and the UV absorbance characteristics of the liquid.

In the situation when one is evaluating an existing system, the physical dimensions are fixed. The task is to properly calibrate the model, which can then be used to assess the system's capacity and to optimize its operations. When a new system is to be designed, the approach is to establish the wastewater parameters, ideally by direct bench or pilot scale testing, and then to determine the optimum hardware configuration and sizing.

7.3.2 Characterization of the Hydraulic Behavior of a UV Reactor

Recall the objectives of an effective hydraulic design, as it would apply to the design of the UV disinfection reactor. (The reader should refer to the discussions in Chapter 4 regarding hydraulic considerations in disinfection reactor design.) First, the unit should be a plug flow reactor (PFR) in which each element of fluid passing through the reactor resides in the reactor for the same period of time. Second, the flow motion should be turbulent radially from the direction of flow. This is to allow for each element to receive the same overall average intensity of radiation in the non-uniform intensity field which exists in the reactor. The tradeoff in this requirement is that some axial dispersion will be introduced, yielding a dispersive or non-ideal flow reactor.

Third, maximum use must be made of the entire volume of the reactor; conversely, dead spaces must be minimized, such that the effective volume is very close to the actual volume available.

When evaluating an existing reactor, the hydraulic evaluation should entail direct testing of the unit to establish the residence time distribution (RTD). Subsequent analysis, as described in Chapter 4, can serve as an excellent diagnostic tool in examining non-performance or to determine system capacity.

New systems design requires the engineer to specify equipment configurations which will be hydraulically efficient. The indices and dispersion characteristics discussed in Chapter 4 can serve as design specifications. Evaluation of commercial reactors can rely on the development and evaluation of the necessary hydraulic information from scaleable pilot units or full scale modules.

The following discussions present the major elements of effective hydraulic design for UV systems:

- residence time distribution (RTD)
- dispersion
- turbulence
- effective volume

7.3.2.1 Residence Time Distribution

The evaluation of a specific reactor relies on the construction of the RTD appropriate for that reactor configuration. This can be accomplished by a number of experimental procedures; subsequent analysis of the residence time distribution curves determines the hydraulic characteristics of the unit. Experimental procedures and analysis techniques were presented in Chapter 4. Particular attention should be paid to the discussions regarding reactors with short residence times.

Consider the analysis of a specific RTD to demonstrate the appropriate calculations and interpretation. The example is taken from the Port Richmond project described by Scheible et al. (55). Unit 2 in this study was a submerged quartz system configured in a fashion similar to that shown on Figure 7-6. It contained 100 lamps, parallel to one another, each held in quartz sleeves with an outer diameter of 2.3 cm. The method by which the RTD was developed was described in Chapter 4 and presented schematically on Figure 4-4(b). The relevant unit characteristics are:

x = distance between tracer input and output, which is approximately equivalent to the lamp battery dimension in the direction of flow.
= 47 cm

V_v = liquid volume of reactor
= 100 liters

T = theoretical mean residence time, V_v/Q
= 7.0 seconds

At a flow of 890 Lpm, the velocity is computed to be 7 cm/sec. Figure 7-16 presents the tracer curve and resultant RTD developed for one run. The upper panel is the F-curve developed by the so-called step input tracer analysis described in Chapter 4 (see Figure 4-3(c)). A salt tracer is continuously injected upstream of the lamp battery until a steady-state concentration is read by the conductivity probe positioned immediately downstream of the battery (all at $t < 0$).

At $t = 0$, the salt injection is discontinued. The trace on the upper panel is a record of the die-away at the downstream probe until a new steady-state condition is reached, which in this case, is the background level. Recalling that the derivative of the F-curve is the C-curve (see Figure 4-3), the C-curve can be constructed by plotting the slopes of tangents (dc/dt) drawn at points along the curve against time. This is shown on the middle panel of Figure 7-16.

The analysis of the RTD curve can be accomplished graphically by breaking the curve into discrete areas at discrete time intervals. The calculations are demonstrated by the following, where c_i is dc/dt and t_i is the corresponding time:

t_i	c_i	$t_i c_i$	$t_i^2 c_i$	Cumulative (%)*
3.8	0	0	0	
4	0.003	0.012	0.048	0.47
4.6	0.011	0.0506	0.233	2.46
5.1	0.02	0.102	0.520	6.44
5.4	0.028	0.1512	0.816	12.3
6.3	0.039	0.246	1.548	21.9
7.0	0.049	0.343	2.401	35.3
7.2	0.061	0.439	3.162	53.5
8.3	0.044	0.409	3.806	68.4
10.0	0.029	0.29	2.9	79.8
11.2	0.017	0.190	2.13	87.0
13.0	0.01	0.13	1.69	92.2
13.9	0.008	0.111	1.55	96.5
14.1	0.005	0.071	0.994	99.2
16.2	0.001	0.016	0.259	100.0
Σ =	0.325	2.561	24.39	

*Cumulative percent tracer at observation j =

$$\frac{\sum_{i=1}^{j} t_i C_i}{\sum_{i=1}^{n} t_i C_i} \times 100$$

where $j \leq n$
n = total number of observations

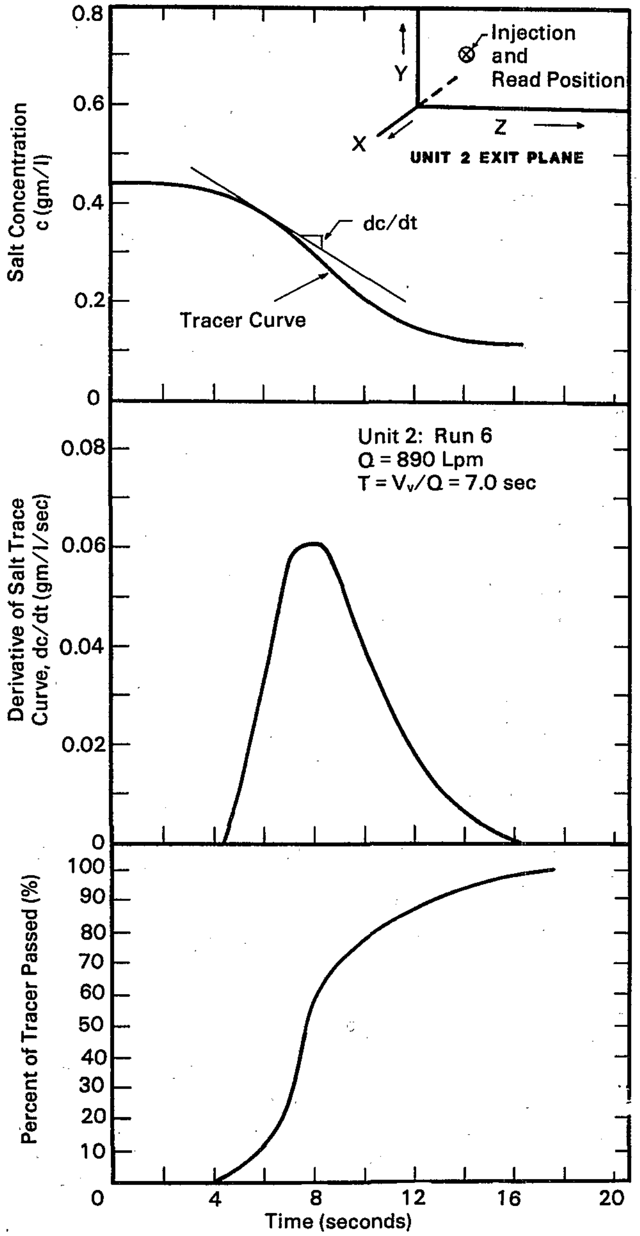
The mean residence time, θ, is the centroid, or first moment of the distribution (from Equation 4-13):

$$\theta = \frac{\Sigma t_i c_i}{\Sigma c_i} = \frac{2.561}{0.325} = 7.9 \text{ seconds}$$

The last column in the above calculation is the cumulative area as a percent of the total. By plotting this against time, as shown on the lower panel of Figure 7-16, one can display the cumulative tracer with time. This then allows one to evaluate any number of the indices defined by Rebhun and Argaman (57) and discussed in Chapter 4. For this particular tracer analysis, the following parameters are determined:

ratio of initial to theoretical time	$t_f/T = 3.9/7 = 0.56$
ratio of peak to theoretical time	$t_p/T = 7.2/7 = 1.03$
Morrill Dispersion Index	$t_{90}/t_{10} = 12.4/5.3 = 2.3$
ratio of mean residence to theoretical time	$\theta/T = 7.9/7.0 = 1.13$
ratio of median to mean residence time	$t_{50}/\theta = 7.2/7.0 = 1.03$

Figure 7-16. **Example of RTD curve developed for Unit 2 at Port Richmond by the step input method (55).**



These analyses suggest that although the reactor is one which behaves in a relatively plug flow mode, there is a degree of dispersion. The ratio of t_{90} to t_{10} (Morrill Dispersion) has a value of 2.3; this is characteristic of plug flow, but is greater than 2.0, a suggested design goal (<2) for disinfection reactors.

The ratio t_f/T is greater than 0.5, suggesting little significant evidence of short-circuiting. This is supported by the t_p/T being approximately equal to 1.0. The ratios of t_{50}/θ and θ/T are also very close to 1.0; this indicates that the entire reactor volume is used effectively.

Consider, as a way to gain a perspective for effective hydraulic design, the analysis of a Teflon tube reactor. The tubular reactor design most closely approximates the ideal hydraulic configuration for a disinfection system. Pipe flow with high length to diameter ratios (this is often referred to as the aspect ratio), are particularly conducive to plug flow, low dispersion conditions.

The step input tracer technique was used to test the Teflon system at Eden, Wisconsin (4). The unit is similar in design to that shown on Figure 7-8. The Teflon tube length is 1.8 m, with a diameter of 6 cm; this yields an aspect ratio of 30. A tracer analysis was conducted at a flow of 65 Lpm. The velocity was 38 cm/s, with a theoretical residence time of 4.7 seconds. The mean residence time was determined to be 4.9 seconds. As presented above, the following indices were computed from the RTD analysis:

$$t_f/t = 0.68$$
$$t_p/T = 0.98$$
$$t_{90}/t_{10} = 1.4$$
$$\theta/T = 1.04$$
$$t_{50}/\theta = 1.02$$

The ratio of t_f/T is 0.68 in this case, which indicates no evidence of short-circuiting. The Morrill index (t_{90}/t_{10}) is significantly less than 2 (1.4), and the values of θ/T and t_{50}/θ are both approximately 1, as is t_p/T. From this analysis, the tubular reactor is a good example of a plug flow design with little dispersion, and maximal use of the reactor volume.

7.3.2.2 Dispersion Characteristics

The design expression (Equation 7-10), incorporates the dispersion characteristics of a reactor. Recall from Chapter 4, the discussion of the dispersion coefficient, E, (cm^2/sec), and the dimensionless dispersion number, d, which is equal to E/ux, (u is the velocity and x is the characteristic length).

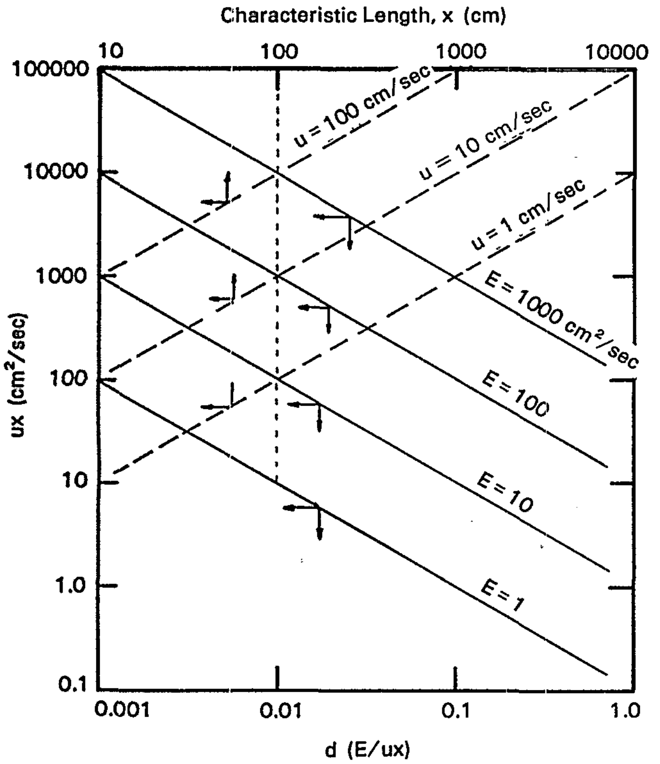
The ideal reactor design for UV disinfection forces the dispersion number very low, preferably less than 0.01. From its definition, this will be forced by a low dispersion coefficient, a high velocity, and/or a long dimension, x. This is schematically presented on Figure 7-17.

Consider the two examples presented earlier. From Chapter 4, we know that the dispersion number can be estimated from the mean and variance. The mean residence time for the Port Richmond unit (Figure 7-16) was computed to be 7.9 seconds. The variance can be determined (Equation 4-14):

$$\sigma^2 = \frac{\Sigma t_i^2 c_i}{\Sigma c_i} - \theta^2 = \frac{24.39}{0.325} - 7.9^2$$

$$\sigma^2 = 12.64 \text{ sec}^2$$

Figure 7-17. Relationships of velocity, length, and dispersion.



The dimensionless variance is computed (Equation 4-19):

$$\sigma_\theta^2 = \frac{\sigma^2}{\theta^2} = \frac{12.64}{7.9^2} = 0.203$$

The value of d can be estimated by (Equation 4-21):

$$\sigma_\theta^2 = \frac{2E}{ux} - 2\left(\frac{E}{ux}\right)^2 (1 - e^{-\frac{ux}{E}})$$

or

$$\theta^2 = 2\,d - 2\,d^2 (1 - e^{-(1/d)})$$

Ignoring the second term on the right, the first approximation of d is 0.102. Adjusting by trial and error for the second term, the value of d becomes 0.104. This suggests that the first Gaussian approximation was adequate, i.e.:

$$d \simeq \frac{\sigma_\theta^2}{2}$$

Note also that d is << 0.5; thus, it is reasonable to consider it as a closed vessel. The value of d (0.104) also confirms the moderate to highly dispersive nature of the reactor.

The tubular reactor variance is 0.624 sec^2. The dimensionless variance is 0.026; from this, the dispersion number, d, is estimated to be 0.013, reflecting the low dispersion, plug flow nature of this particular reactor design.

The dispersion number and the dispersion coefficient are utilized in the design equation for the UV process (Equation 7-10). Correlations can be developed to estimate the dispersion number as a function of reactor characteristics relating to friction losses, hydraulic radius, velocity, etc. Such predictive models exist for pipe systems. Such models have not been developed for UV reactors. There is limited data available to attempt this; certainly further work is needed in this effort.

The dispersion coefficient should be expected to vary with velocity. In current practice, values of E are selected to represent conditions under high flow; a design goal is then set with the dispersion number (e.g., d = 0.02 to 0.05) which will force limits on ux for the given E. This is discussed further with the design example presented in Section 7.4, particularly as it relates to the impact on head loss.

7.3.2.3 Turbulence and Head Loss

An important consideration in the hydraulic design of a UV reactor is the turbulence of the fluid. By having turbulent flow, any particle has an equal probability of being at any point in the cross-section of the conduit, as it travels in the direction of flow. The importance of turbulence lies in the fact that the intensity field in the reactor, regardless of the way the lamps are configured, is non-uniform. Thus, if a particle is forced to move erratically by the turbulent conditions, it will likely see all intensity levels in the non-uniform field. In this case, then, it is acceptable to use the average intensity in the reactor to evaluate dose levels microorganisms receive as they move through the reactor. If true laminar conditions existed, streamlines may move through areas of low-intensity and receive little dose relative to the streamlines moving close to the lamps.

Flow through a reactor can be characterized by eddies, swirls, and irregular movements of large fractions of the fluid; these do not constitute turbulence. They may more correctly be described as "disturbed flow." Thus, a reactor flow may be laminar, but have disturbances; the lamps themselves can be the source of these disturbances. Turbulence is generally induced by high friction losses and high velocities.

Turbulence indicated by head loss. If the log of the head loss for a given length of uniform pipe is plotted against the log of the velocity, it will be found that effectively two regions exist. Where the velocity is low enough to assure that laminar flow exists, the head loss, h_L, due to friction, will be directly proportional to the velocity, u:

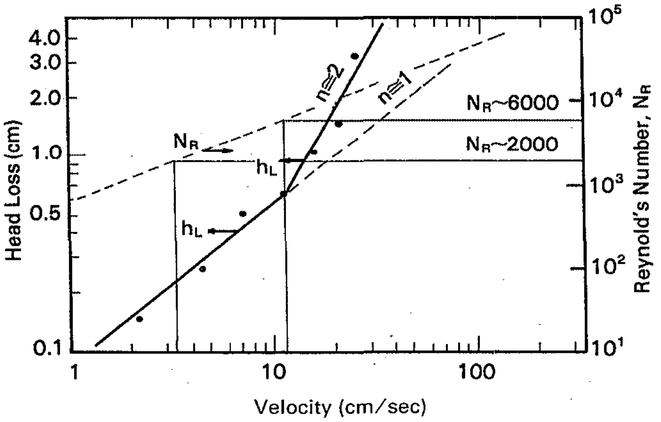
$$h_L \simeq u^n, \text{ where } n = 1 \text{ (laminar)} \qquad (7\text{-}15)$$

As the velocity increases, at some point turbulence is induced and the h_L is found to increase at a higher rate than the increase in velocity. In uniform pipe this value of n is greater than 1.0;

$$h_L = u^n, \text{ where } n = 1.75 \text{ to } 2.0 \text{ (turbulent)}$$

Such measurements were taken during the Port Richmond study for the quartz systems. The data for one of the quartz units are presented on Figure 7-18; these demonstrate the transition from the laminar to the turbulent flow condition. At velocities less than 10 cm/s (log u = 1.0), the value of n is approximately 1.0; at velocities greater than 10 cm/s, the n is estimated to be 2.0.

Figure 7-18. **Log-log plot of head loss against velocity for Unit 2 at Port Richmond indicating transition from laminar to turbulent flow regime (55).**



This type of information can be developed by direct measurement on full scale modules or hydraulically scaleable pilot units. Care should be taken that the head loss due only to the lamp reactor itself is being measured. Losses due to reactor entrance and exit conditions should be separated from the analysis.

Turbulence indicated by Reynolds Number. Velocity is not the only factor that determines if a flow is laminar or turbulent. The criterion is the Reynold's number. This dimensionless number, N_R, is the ratio of inertia forces to friction forces in a completely filled conduit. Thus:

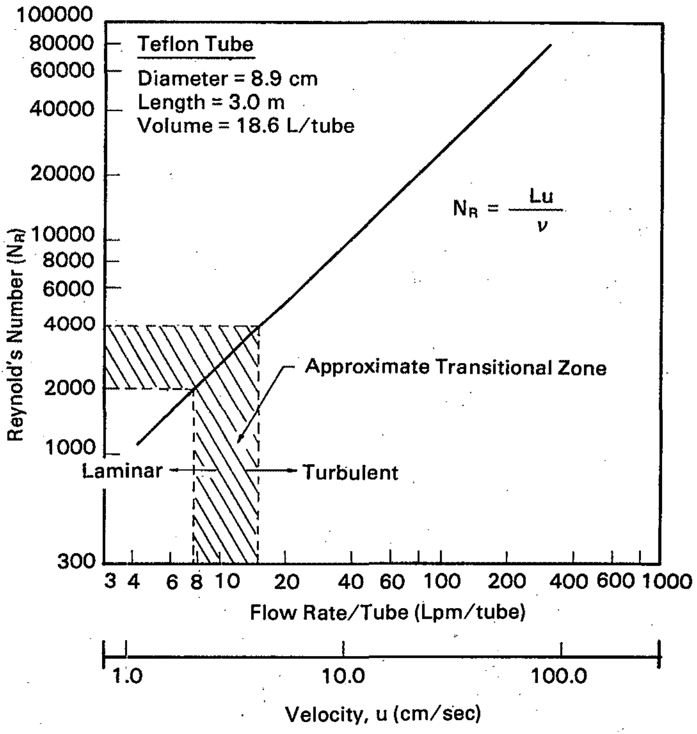
$$N_R = \frac{L^2 u^2 \rho}{L u \mu} = \frac{Lu}{\nu} \qquad (7\text{-}16)$$

where:

u = velocity
ρ = density of fluid
μ = viscosity
L = linear dimension significant to pattern of flow
ν = kinematic viscosity

The linear dimension, L, for pipes is generally taken as the pipe diameter. A straightforward example of the Reynolds number analysis is the tubular flow array. The linear dimension is taken as the tube diameter. The Reynolds Number is plotted on Figure 7-19 against the velocity and the flow rate per 3-meter long Teflon tube.

Figure 7-19. **Estimates of Reynold's Number for 8.9 cm diameter Teflon tubes.**



Typically, the transition from laminar to turbulent flow will occur at a Reynolds number above 4,000; laminar conditions usually exist at values less than 2,000. Turbulent conditions exist in the Teflon tube at flow rates greater than 15 lpm or a velocity greater than 4 cm/sec. This is well below the normal operating range typically encountered for these systems.

The submerged quartz arrays do not allow as straightforward an analysis. The linear dimension is not as clearly defined. It is suggested that an approximation of the hydraulic radius be used in these analyses in order to estimate a Reynolds Number:

$$\text{Hydraulic Radius, } R_H = \frac{V_v}{A_w} \qquad (7\text{-}17)$$

where:

V_v = liquid, or void volume (cm^3)
A_w = total wetted surface area (cm^2)

The void volume is the total reactor volume minus the volume occupied by the lamps and quartz sheaths. The wetted surface area is the sum of the surface areas of the quartz sleeves and the internal wall area of the reactor.

For a circular conduit flowing full, the hydraulic radius is equal to one-fourth the diameter of the conduit. Thus, for Equation 7-16:

$$L = 4R_H \qquad (7\text{-}18)$$

The Reynolds Number can then be estimated for the quartz, submerged reactors:

$$N_R = \frac{4R_H u}{\nu} \qquad (7\text{-}19)$$

Consider Unit 2 from the Port Richmond study as an example of this calculation:

$$R_H = \frac{V_v}{A_W} = \frac{10.2 \times 10^4 \text{ cm}^3}{7.2 \times 10^4 \text{ cm}^2} = 1.4 \text{ cm}$$

At a kinematic viscosity of 0.0098 cm^2/sec (water at 20°C), the Reynolds number is estimated:

$$N_R = \frac{4\,(1.4)\,u}{0.0098} = 580(u) \qquad (7\text{-}20)$$

where the velocity, u, is in cm/s. This estimate of N_R is plotted on Figure 7-18 for Unit 2. From this analysis, the breakpoint appears at an N_R slightly less than 6,000. This is somewhat higher than the N_R of 4,000 normally considered as minimal for turbulent flow. The linear dimension estimate ($4R_H$) may be a factor in this. The fact is, that as long as the method of estimating the Reynolds number is kept consistent, it is possible to qualitatively evaluate a unit design for turbulence.

Table 7-6 presents a summary of the Reynolds Numbers estimated for several lamp array configurations. These are based on the estimated hydraulic radius and the design flow range for each unit. As shown, all systems would typically operate at velocities high enough to yield turbulent flow. If a Reynolds Number of 6,000 were set as a minimum, the minimum velocity can also be shown. Such a criterion can be used in establishing specifications for new systems designs. It should be understood that although the difference in performance has not been demonstrated, it is implicit in the physical mechanism of UV disinfection that the liquid be in turbulent motion. The Reynold's Number calculation offers a method to qualitatively evaluate this criterion.

Estimating Head Loss. There is little information on head losses caused by the UV lamp batteries. This will vary according to the size of the reactor, the velocity, and the placement of the lamps. The tubular reactor head loss can be estimated from pipe flow equations, assuming smooth wall friction coefficients. With regard to the quartz systems (submerged), one would expect higher losses in the unit with the lamps placed perpendicular to the flowpath, than those configured with the lamps parallel to the flowpath.

Scheible (56) reported an empirical relationship developed for the quartz units at Port Richmond. This would be representative of only that type of configuration: open channel structure, uniform lamp array, and the flowpath perpendicular to the lamps. Direct testing would have to be done on alternative configurations in order to determine head losses, although a similar approach can be taken in doing so.

The head loss expression is based on Darcy's equation for pipe flow:

$$h_L = \frac{fxu^2}{d_r 2g} \qquad (7\text{-}21)$$

where x and u are the length (cm) and velocity (cm/s), respectively, d_r is the diameter (cm), and f is the coefficient of friction. In this case, d_r is set as the approximate hydraulic radius of the system from Equation 7-17. Finally, a new coefficient of friction is defined:

$$c_f\,(\text{cm}^2/\text{sec}^2) = \frac{f}{R_H\,2g} \qquad (7\text{-}22)$$

such that:

$$h_L = c_f\,(x)\,(u)^2 \qquad (7\text{-}23)$$

At Port Richmond, direct h_L measurements of the lamp batteries in the two quartz systems yielded a c_f between 0.000173 and 0.00023 sec^2/cm^2.

The head loss can become a factor in the design of a reactor. Earlier discussions cited the dispersion number as a good design guideline; e.g., designing a reactor to yield a low d value. Recalling that:

$$d = \frac{E}{ux}$$

Table 7-6. Summary of Reynolds Number Estimates for Different Lamp Configurations

System	Configuration[d]	R_H (cm)	Hydraulic[b] Flow Range (lpm)	Equivalent Velocity (cm/sec)	Reynolds Number N_R	Velocity at $N_R = 6000$ (cm/sec)
Port Richmond 1[e]	uniform array	4.7	1,000 - 2,600	3.5 - 9.1	6,720 - 17,500	3.1
Port Richmond 2[e]	uniform array	1.4	1,000 - 2,600	9.3 - 24.0	5,400 - 14,000	10.3
Vinton[f]	concentric array	4.74	2,000 - 4,500	6.8 - 15.4	13,000 - 30,000	3.1
Suffern[f]	staggered array	3.65	3,000 - 10,000	7.3 - 24.3	10,900 - 36,200	4.0
Port Richmond 3[e]	tubular array	2.23[a]	40 - 100[c]	10.8 - 27.0	9,800 - 24,600	6.6

[a] Diameter/4.0.
[b] Design flows.
[c] Liters/min/tube.
[d] See section on UV intensity for definition of array configurations.
[e] Reference 56; also see Figures 7-6 and 7-8.
[f] Reference 4; also see Figures 7-3 and 7.4, respectively.

it can be seen that for a given E, the d is forced by ux. Thus, for an increasing E, the ux must be proportionately increased to maintain a fixed value of d. We see by Equation 7-23, however, that increasing velocity can have dramatic effects on the head loss, since it will increase by the square of the velocity. Increasing x can be done to some degree; this must stay within practical limits, however. One should understand that the head loss we are considering is that incurred through the lamp battery itself; additional head losses will be incurred at the inlet and outlet structures.

In all, d will need to be reconciled with the head loss. Increasing the acceptable d will allow one to reduce and keep the h_L within design limits. Generally, for gravity flow systems, one should design about a d between 0.02 and 0.05. One should understand that these are still excellent dispersion numbers for a disinfection reactor.

7.3.2.4 Effective Volume

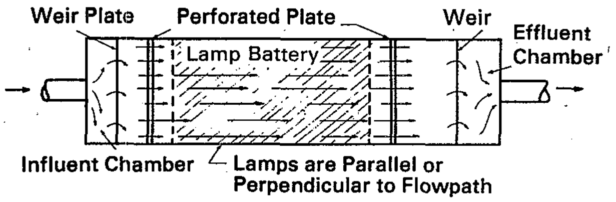
The lamp battery volume is that portion of the total system occupied by the UV lamps. By this fact, it is very important that the reactor is designed such that full use be made of the entire volume. Dead zones or short-circuited areas mean ineffective use of lamps and power, the two components which comprise a major portion of the capital and operating costs. This is primarily a design consideration for the submerged quartz systems. The tubular array configuration should inherently provide for maximal use of the Teflon tube volume.

Maximum use of the reactor volume is directly related to the approach and exit conditions of the reactor. The design intent must be to first enter the front plane of the lamp battery with equal fluid velocity at all points. This same condition must exist at the exit plane. The approaches taken to encourage this include:

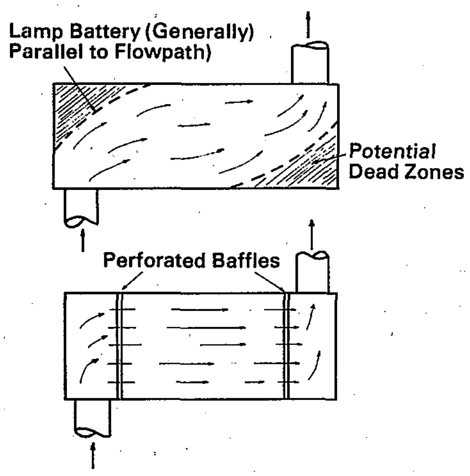
- *open channel flow before and after the lamp battery,*
- overflow weirs placed the full length of the reactor,
- perforated stilling walls before and after the lamp battery, and
- unidirectional flowpath throughout the approach, battery, and exit sectors.

Figure 7-20 schematically displays these considerations. The upper panel shows a plan-view of an open channel type configuration. The inlet and outlet chambers should be independent of the lamp battery. Weirs can be placed across the width of the reactor to distribute the flow evenly across the unit. The perforated baffle plates then serve to distribute the flow evenly both horizontally and vertically. The weir

Figure 7-20. Inlet and outlet considerations for submerged quartz systems.



(a) Open Channel-Type Configuration



(b) Sealed Cylindrical Reactor Configuration plates may not be needed before the perforated baffle if there are no extreme velocity gradients coming into the unit. By having the same arrangement on the exit side, the flow paths are kept stable and unidirectional through the lamp battery.

In the middle panel of Figure 7-20, a sealed cylindrical shell type reactor is shown. The lamps are generally parallel to the flow path and the wastewater enters and exits the reactor perpendicular to the lamps. As shown, there is the tendency to induce flow channelling with this arrangement, causing dead zones in the reactor. A solution is shown on the lower panel, in which perforated baffle plates are installed at both ends of the unit. These serve to distribute the flow over the entire cross-sectional plane of the reactor.

To evaluate a system for effective volume, the indices derived from the RTD curve are useful. The ratio of the mean residence time to the theoretical residence time (θ/T) should be approximately 1.0. The actual fraction is reflective of the actual volume being effectively utilized.

It is suggested that the θ/T ratios be used in specifying the hydraulic design of a UV reactor. A value greater than 0.9 is an appropriate requirement.

7.3.2.5 Summary Considerations for Effective Hydraulic Design

In summary, the key points to address when evaluating or specifying the design of a UV reactor are as follows:

Residence Time Distribution. This should be constructed at a number of flow conditions for an existing system; it should also be required when specifying commercial systems. The RTD provides key information on the actual or anticipated hydraulic behavior of a reactor.

Plug Flow. This can be quantitatively described by indices derived from the RTD analysis. Appropriate guidelines for specifications can be:

$t_f/T > 0.5$
$t_{90}/t_{10} < 1.0$
$t_p/T > 0.9$
$t_{50}/\theta = 0.9 \text{ to } 1.1$

Additionally, the dispersion coefficient, E, should be relatively low (<500 cm^2/s); the dispersion number should be less than 0.1 (preferably less than 0.05 if the head loss is acceptable).

Reactor designs which are conducive to plug flow have high aspect ratios, x/L. Thus, the length, x (i.e., the distance in the direction of flow), should be significantly higher than the appropriate cross-sectional dimension, L. In tubular reactors, such as the Teflon tube units, this is the diameter. In submerged quartz units, L is $4R_H$, where R_H is the hydraulic radius. As a guideline, an aspect ratio greater than 15 should be incorporated into a reactor design.

Maintenance of plug flow within a reactor will be influenced by the approach and exit conditions. The design should have minimal disturbances at the inlet and exit planes of the lamp battery; directional changes in the flowpath would best be made outside of the lamp battery.

Dispersion Number. A key goal is to minimize the dispersion number, d. A design goal should be to have a d between 0.02 and 0.05. Levenspiel (58) suggests that this would be representative of a plug flow reactor with low to moderate dispersion. This can be accomplished by increasing the product of ux, even in a system with a relatively high dispersion coefficient. The designer should be aware, however, that extended lengths and higher velocities will cause higher head losses. In certain situations some adjustment of the dispersion number may be necessary in order to meet specific head loss requirements.

Turbulence. Radial turbulence is necessary due to the non-uniform intensity field. The reactor design should induce an estimated (by the procedure discussed earlier) Reynold's Number greater than 6,000 at minimum flow. If possible, it would be beneficial to confirm the laminar/turbulent flow transition velocity by direct head loss measurements on the lamp battery.

Head Loss. Direct measurements should be required for full-scale modules or scaleable pilot units as part of commercial equipment specifications. These should be determined over a wide velocity range and should exclude entrance and exit losses.

Effective Volume. Maximal use of the reactor lamp battery is essential to keep the process cost-effective. This will be related directly to the reactor's inlet and outlet design. The goal must be to have equivalent velocities at all points upon entering and upon exiting the lamp battery. Stilling walls (perforated baffles), and weirs should be incorporated into reactor designs to assure this. Guidelines for specifying commercial equipment should require the ratio θ/T to be greater than 0.9 and/or the t_p/T to be greater than 0.9.

7.3.3 Estimation of the Average Intensity in a UV Reactor

The second element of dose, after time, is the intensity of energy during the exposure time. Recall that the intensity is the rate, or flux, of delivery of photons to the target. In the UV process design model, Equation (7-10), the rate of bacterial inactivation is described as a function of the intensity. By this fact it becomes important to be able to quantify the intensity in a given system. The intensity in a reactor is a function of the UV source (output), the physical arrangement of the source relative to the wastewater (the arrangement of the lamps and their placement in or out of the liquid), and the energy sinks present which will attenuate the source output before it can be utilized for disinfection purposes.

The UV source, as discussed earlier, is typically the low pressure mercury arc lamp. Table 7-7 presents lamp specifications for a series of germicidal lamp models. The lamps generally used in UV reactor systems are equivalent to the G64 and G36 units. The overall lamp length is approximately 0.9 m for the G36 and 1.6 m for the G64. The arc length defines the active, light emitting portion of the lamp (0.75 m and 1.5 m, respectively.) The diameter of the lamp is small, typically 1.5 to 1.9 cm. The lamp envelope is made of fused quartz or other highly transparent (to the 253.7 nm wavelength) glass, such as Vycor.

In the quartz systems, the individual lamps are sheathed in quartz sleeves only slightly larger in diameter (2.3 cm) than the lamp and the entire lamp/quartz bundle is submerged in the flowing liquid. In systems where the wastewater does not contact the quartz or lamp surface, separate conduits carry the wastewaters. The conduits are translucent to the UV light, with the lamps placed near the outside conduit wall.

Determining the intensity at any point in these complex lamp reactors is not straightforward. At present, there is no commercially available detector which can measure the true intensity in such a system. The problem lies in the fact that the detectors are planar receptors; only energy striking a flat surface will be measured. Such detectors will intercept fractions of light striking the surface at an angle. Only light which is normal to the surface, i.e., collimated light, however, will be wholly measured. "Cosine-corrected" detectors attempt to compensate for this by adjusting for the angular light. These still measure, however, only the planar intensity. Where light is not collimated, as is the case with a multi-lamp UV reactor, the flux of energy is three-dimensional. This same concept is enforced when the target of the radiation is considered. In turbulent motion, all particles can be expected to receive equal exposure.

Several approaches have been proposed to estimate light intensity, including chemical actinometry, biological assays, and direct calculation. The two procedures which have received the greater attention are the bioassay and direct calculation methods. The bioassay procedure has been applied in a limited fashion for a number of design specifications, primarily as a technique for quantifying the dose delivered by a specific piece of UV equipment. It can also be used to implicitly derive the intensity within a system.

The second method which is used and which is generally emphasized within the context of this manual, is the direct calculation of intensity. This is accomplished by the point source summation method. The discussions briefly describe the calculation framework which yields the average intensity as a function of the UV absorbance coefficient of the wastewater. A series of solutions are then presented which the designer can use to estimate intensity in a number of lamp configurations. Finally, discussions are presented regarding the factors of lamp deterioration and enclosure fouling which will absorb the UV energy, and thereby reduce the intensity.

7.3.3.1 Bioassay Procedure to Estimate Intensity and UV Dose

The assay procedure has been proposed as an effective method for estimating delivered dose and system intensity (40). This technique is shown schematically on Figure 7-21.

UV sensitive pure culture is calibrated to the UV dose using the collimated light device shown on Figure 7-21(a). The collimating device allows one to accurately measure the intensity directly with a commercial radiometer. Aliquots of the bacterial suspension are then exposed to this given intensity for a series of fixed time intervals, yielding known doses. The response is then plotted against the dose. This dose-response relationship serves as the calibration for the subsequent reactor assays (Figure 7-21(b)).

The unit to be tested is set to the desired flow and operating conditions and the culture is injected into the influent. The effluent is then sampled with time and assayed for the known bacterium. This same procedure is repeated without the lamps in operation. The resulting densities are as shown on Figure 7-21(c). For each time interval, the log survival rate is determined. The equivalent dose can then be estimated from the dose response calibration curve. When the dose is plotted against time (Figure 7-21(d)), the slope of the correlation is the dose-rate, or intensity.

This method for determining intensity in a system can require a fair sized laboratory effort. A quality analysis requires very frequent sampling and analysis and should be replicated to assure precision. This can be costly and is not cost-effective when compared to the alternative calculation method. It should be noted that Qualls and Johnson (40) used this assay procedure to independently verify the point source summation calculation method.

A possible simplification of the procedure for estimating the intensity involves injecting the bacterium stock at a steady rate. The effluent is assayed and the dose is determined from the log survival ratio. This is

Table 7-7. Examples of Low Pressure Mercury Arc Lamp Specifications (Courtesy Voltarc Tubes, Inc., Fairfield)

	G36T6L G36T6H	G37T6VH	G36T6	G64T5L	G64T6L
Lamp Watts	39	40	36	65	62
Lamp Current, mA	425	425	425	425	425
Ultraviolet Output, watts (at 100 hrs, 253.7 nm)	13.8	14.3	12.7	26.7	25.5
Microwatts/cm^2 @ 1 meter	120	124	110	190	180
Ozone Generation (approximate gm/hr)	H .5 L 0	15	0	0	0
Nominal Length,					
inches	36	37	36	64	64
cm	91.4	94.0	91.4	162.6	162.6
Arc Length					
inches	30	31	30	58	58
cm	76.2	78.7	76.2	147.3	147.3
Tube Diameter, mm	15	15	19	15	19
Tube Material	H: Vycor 7912 L: Vycor 7910	Quartz	Vycor 7910	Vycor 7910	Vycor 7910
Rated Life, hours (average, at 8 hrs/start)	7500	7500	7500	7500	7500

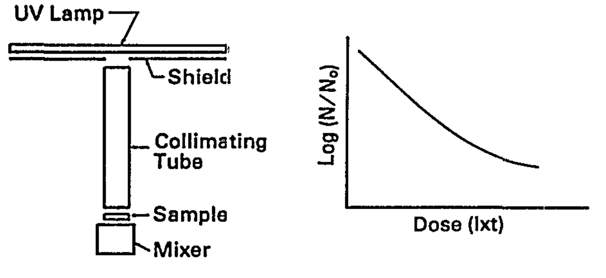
repeated over several flow rates. A separate RTD curve is developed for the reactor using a conservative tracer. The intensity can then be implicitly derived by trial and error calculations to match the RTD and the steady-state dose-response of the unit.

Estimating the Dose by the Bioassay. The steady-state bioassay method, generally using *Bacillus subtilis* spores, has been used as a specification in demonstrating the dose capacity of a given system. The dose is estimated as a function of flow rate through a scaleable pilot module. A minimum dose is then cited for the equipment specification.

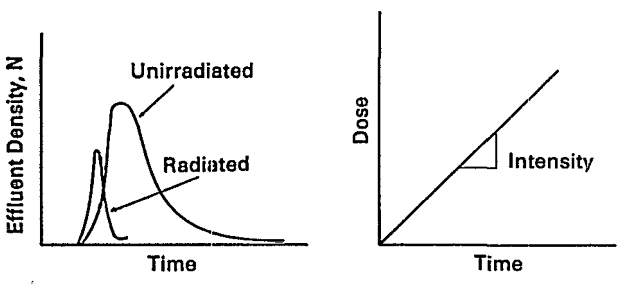
The bioassay procedure, although a valid and unique experimental design, has several disadvantages which, at present, detract from its use as a routine testing and evaluation procedure. It is not standardized, and results will vary from lab to lab, both in the calibration and system assays. It can be costly and cumbersome to accomplish on a routine basis (particularly as a procedure to intermittently monitor a unit for intensity), although this should resolve itself with continuing direct experience. It is also wholly empirical and limited in its use for extrapolation to alternative system configurations.

On the positive side, however, the bioassay can offer several advantages. It is an independent verification of system design, and implicitly of design procedures.

Figure 7-21. Schematic of bioassay procedure for estimating dose and intensity.



(a) Dosing Apparatus (b) Dose-Response Calibration



(c) Equipment Assay Results (d) Estimate of Intensity

As such, it can be used effectively as a post-construction performance test or to compare the performance of competing commercial units during design and/or bid phases of a facility installation.

The following is an outline of the procedure which can be used to develop the dose performance curve for a commercial unit. Example results are given from an actual test series conducted in response to specifications for the Bristol, Connecticut plant expansion; these demonstrate the data generated from the bioassay analysis.

1. Selection and Culturing of Bacterial Culture

   The species selected for the assay should be one which is relatively easy to culture, identify, and harvest, and which has a dose-response which is reproducible and consistent. *Bacillus subtilis* spores, which are used in the example on Figure 7-22, have been used on several recent equipment assays. Originally used by Johnson and Qualls (38), the spores are relatively resistant to UV, and have been shown to be very consistent and reproducible within a specific harvest. *M. lutea* has also been used successfully and shows a response similar to that of the coliform group.

   The culture should be harvested in sufficient quantity such that all necessary dose calibration and assay work can be accomplished with a single harvest. The *B. subtilis* are particularly suited to this since they can be stored for long periods and have been shown to retain their dose-response behavior through this period.

   In situations where units are being compared (such as in a pre-qualification procedure for several manufacturers), a single organism should be specified. Additionally, the use of a single laboratory should be encouraged. It has not been established that reactors will yield the same effective dose for organisms with different dose-response relationships.

   It has also been shown that different dose-response curves are developed lab to lab for the same organism. The better approach, then, in order to assure consistency and the ability to validly compare different units, is to use one organism generated from a single mixed batch and to have one laboratory conduct the calibration and equipment assays.

2. Dose-Response Calibration

   A dose-response calibration curve can be developed using the laboratory collimated beam apparatus shown schematically on Figure 7-21(a). As an example, the apparatus used for the example bioassay had a G8T5 lamp as the UV source. All but two inches of the lamp were shielded. The exposed portion of the lamp was suspended above a 25 cm long, 5 cm diameter non-reflective tube. The sample to be exposed was placed in a petri dish below the tube. The sample size (20 ml) was sufficient to give a liquid depth of 1 cm.

   The purpose of the tube is to collimate the light, such that the light reaching the liquid is perpendicular to the surface. In this manner, the light can be accurately measured by a radiometer detector.

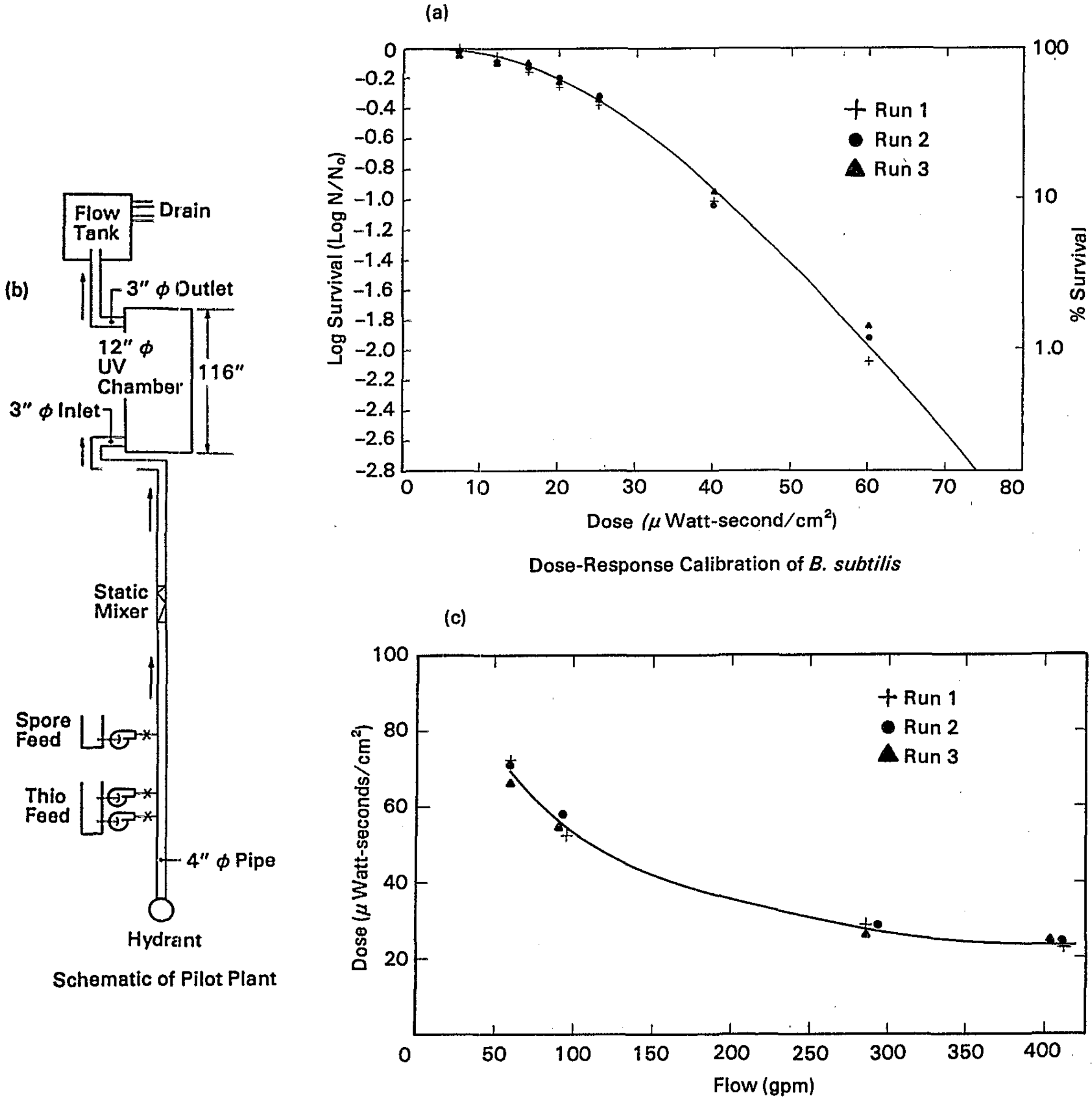
   Prior to exposing the sample, the intensity of the ultraviolet radiation is adjusted (by movement of the lamp position) to 100 $\mu W/cm^2$ at the surface of the sample. A narrow band, calibrated detector should be used for these measurements.

   During exposure, the sample should be gently stirred continuously, using an insulated magnetic mixer with a micro spinbar. The organism density should be adjusted to approximately 10^5; a buffered water should be used for dilution. After exposure, samples should be plated immediately, using a culturing medium appropriate to the organism being assayed. The same medium should be used for both the dose calibration and the equipment assay tasks. In situations where a clean, potable water can be used for the carrying liquid in the equipment assay, a non-selective nutrient medium can generally be used for microorganism enumeration. This was the case for the example bioassay; total plate count agar (pour plates) were used to grow the *B. subtilis* spores.

   Five to seven exposure times should be run to develop the dose-response curve. Thus, at a set intensity of 100 $\mu W/cm^2$, running exposure times between 50 and 400 seconds would yield a dose (I x t) range between 500 and 4000 $\mu W\text{-}sec/cm^2$.

   The dose runs (series of exposure times) should be conducted in triplicate, each from a separate dilution of the stock suspension. In all three dose runs the controls and exposed samples should be sampled in triplicate; three dilutions should then be plated in triplicate. A minimum of two controls (unexposed) should be run with each dose run, representing time zero and the longest exposure period.

Figure 7-22. **Example of bioassay analysis of commercial UV system to determine dose (By permission of Ultraviolet Purification Systems, Bedford Hills, New York).**



Dose-Response Calibration of *B. subtilis*

Schematic of Pilot Plant

Assay Results of Test Unit

A dose-response calibration curve developed for the example assay is given on Figure 7-22(a). The log of the survival ratio (Log N/N_o) for the *B. subtilis* is plotted against the delivered dose.

3. Test Unit and Experimental Setup

The test unit to be evaluated by the bioassay should closely simulate the design of the full-scale system proposed for the treatment facility. Since the UV equipment is generally comprised of modules, the test unit need simulate only one module. In some cases, it may be practical to test the full scale module itself.

Particular attention should be paid to scaling the hydraulic design of the full-scale unit. The parameters of dispersion and the indices developed from the RTD of a unit can be considered in specifying the test unit. These were discussed in Section 7.3.2. Additionally, the test unit should have a similar aspect ratio (ratio of length to diameter, or cross sectional dimension), and inlet and outlet designs. In particular, the inlet and outlet velocities should be equivalent. Although not always specified, it is recommended that residence time distribution curves be developed for the test module.

Generally, specifications should require that the test unit, once its similitude is established, be tested at the hydraulic loads to be encountered for the full scale system. This is determined on the basis of flow per unit lamp, e.g., lpm/lamp. The range of flows to be tested should encompass the peak design flow anticipated for the plant. The performance requirement generally specifies that the system sizing would meet a desired dose level under peak design conditions. Other requirements imposed on the test are that the lamp output be reduced to simulate end of life conditions; this is generally considered at 70 percent of the lamps' nominal output. The lamp output can be reduced by using a rheostat to adjust the voltage, or by using lamps which have reached 70 percent (this should be confirmed by direct measurement, see Section 7.5.1) of their original output.

The transmittance of the carrying water should also be adjusted to yield an absorbance coefficient, or percent transmittance, anticipated under design conditions. This is accomplished by adding a chemical which will absorb energy at the 253.7 nm wavelength, but will not interfere with the test. An appropriate compound is sodium thiosulfate.

Figure 7-22(b) presents, schematically, the experimental setup to test the UV module for the example bioassay. The water source was potable water from a hydrant at a wastewater treatment plant (with appropriate backflow protection devices). Sodium thiosulfate is injected at a rate needed to yield (by direct measurement) the desired transmittance.

The microorganism suspension is injected in similar fashion to yield a desired density level. Both the thiosulfate and spore suspension are injected upstream of an in-line static mixer to assure a homogenous solution before entering the UV chamber. Flow rates are set and measured by determining the rate of fill in a large (1000 L) tank. The tank drains to the primary clarifiers in this particular case.

4. Experimental Field Test Procedure

Three to four flow rates should be tested in triplicate; these should, at minimum, bracket the peak design flow. Once the appropriate water flow rate is set through the unit, a near saturated solution of sodium thiosulfate is metered directly into the water line. The feed rate is adjusted until the desired transmittance level is reached in the effluent.

Once the water is adjusted to the desired transmittance with sodium thiosulfate, the *B. subtilis* spore suspension (or other test organism) which is continuously mixed, is metered into the line with a second metering pump. The feed rate is adjusted, in this case, at each flow setting to yield an influent density of approximately 10^4 spores/ml.

The flow with thiosulfate and spore suspension is continued long enough to allow a minimum of seven volume changes in the unit before sampling. The influent and effluent are then sampled in triplicate using sterile sampling containers. The influent and effluent sampling lines should be kept flowing continuously to assure that the samples taken are representative of the run being done. An additional sample is taken of the influent for percent transmittance analysis.

Samples should undergo immediate (within four hours) analysis. A minimum of three dilutions should be plated in triplicate, using the appropriate medium. The percent ultraviolet transmittance at 253.7 nm should be measured by standard spectrophotometric procedures.

The results of the example bioassay are presented on Figure 7-22(c). The log of the survival ratio (Log N/N_o) are first determined from the experimental data. These are then used to determine the effective dose delivered by the test unit by reference to the dose-response calibration curve (Figure 7-22(a)). This effective dose is then plotted against the corresponding flow rate. The example unit was operated at 70 percent lamp output (set by adjusting the voltage); the water transmittance was 70 percent (at 253.7 nm).

The relationship presented on Figure 7-22(c) allows one to then determine the flow (or flow per unit lamp) which corresponds to the minimum desired dose. As discussed earlier, this has often been used in equipment specification as a pre-bid or bid qualification requirement. The bioassay is also used to set specific performance requirements for equipment supplied to a facility.

7.3.3.2 Calculation of the Average Intensity by the Point Source Summation Method

The calculation approach is suggested as the method of choice because of its versatility and flexible application to varying configurations. The technique used to calculate intensity is the point source summation method. A brief description is presented herein; the reader is referred elsewhere for a detailed discussion of the calculation framework (52).

The point source summation technique was evaluated by Jacob and Dranoff (66) for light intensity profiles in a perfectly mixed photoreactor and was first applied to UV disinfection reactors by Johnson and Qualls (38). It presumes that the lamp is a finite series of point sources that emit energy radially in all directions. The intensity at a given point in a reactor would be the sum of intensities from each of these point sources.

Intensity Attenuation. UV intensity will attenuate as the distance from the source increases. This occurs by two basic mechanisms: dissipation and absorption. Dissipation is simply the dilution of the energy as it moves away from the source. The area upon which the energy is being projected is increasing; thus the energy per unit area is decreasing. This dissipation can be calculated by surrounding the point source by a sphere of radius R:

$$I = S/(4\pi R^2) \qquad (7\text{-}24)$$

where I is the intensity at a distance R in μWatts/cm^2, R is the distance in centimeters, and S is power available from the UV source in μWatts. Thus, dissipation is seen to attenuate the intensity as the inverse of the radius squared.

The second attenuation mechanism relates to the absorptive properties of the medium through which the energy is transmitted. This is best described by Beer's Law:

$$I = I_o \exp[-\alpha R] \qquad (7\text{-}25)$$

where I_o is the intensity at a given surface on the source (μWatts/cm^2), α is the absorbance coefficient of the medium through which the energy is passing (cm^{-1}), and R is the distance at which I is measured relative to the point represented by I_o. The absorbance coefficient reflects the absorbance at the specific wavelength being emitted; in the case of the low pressure mercury arc lamps, the wavelength is 253.7 nm.

Combining Equations 7-24 and 7-25 yields an expression which describes the intensity at a given distance from a single point source of energy:

$$I = [S/(4\pi R^2)] \exp(-\alpha R) \qquad (7\text{-}26)$$

This equation serves as the basis for the point source summation calculation technique. A basic assumption is that a receiver (i.e., a microorganism) passing through the reactor is infinitely small and is spherical; by this it can then be presumed that the energy emitted from any point source element of the lamp will strike the receiver normal to its surface.

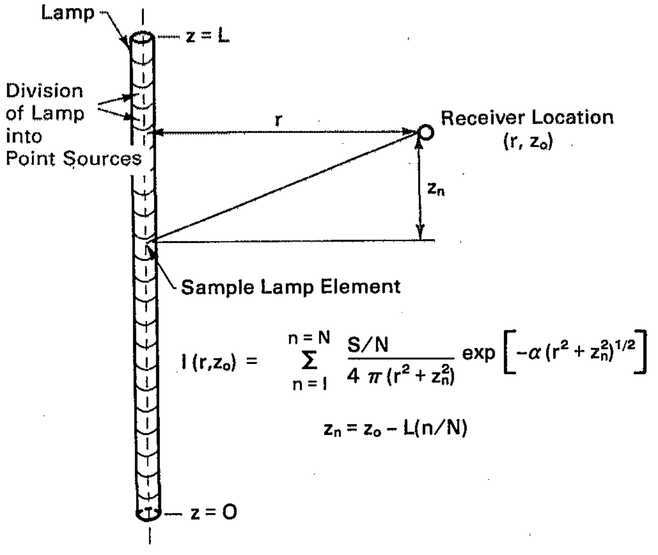
The model analysis also neglects the phenomena of reflection, refraction, diffusion, and diffraction of light and assumes that the absorptive properties of the liquid are independent of the light intensity. The intensity at a receiver is then the summation of the intensities from each of the point source elements of a lamp, (or lamps in a multilamp system). Figure 7-23 is a schematic representation of this calculation. As shown, the intensity at the receiver location (r, z_o) is the summation of the intensities from each of the lamp elements:

$$I(r,z_o) = \sum_{n=1}^{n=N} \frac{S/N}{4\pi(r^2+z_n^2)} \exp[-\alpha(r^2+z_n^2)^{1/2}] \qquad (7\text{-}27)$$

where N is the number of point source elements in the lamp. The value of z_n is:

$$z_n = z_o - L(n/N) \qquad (7\text{-}28)$$

Figure 7-23. Lamp geometry for point source summation approximation of intensity.



A practical analysis of intensity in a submerged or Teflon tube lamp battery system requires that the calculations be made at numerous receiver locations within the lamp battery. This is accomplished by dividing the cross-sectional space between lamps into an equal-area grid system. The average of the receivers located in the center of equal area grid elements would then be equivalent to the average intensity within the total grid area. Within a system, this grid is expanded to encompass all or a section of the unit and can be moved about to evaluate boundary effects and other configurations which may affect the overall unit average intensity.

The model takes into account the geometry of a system, the characteristics of the lamps and enclosures (e.g., quartz or Teflon), and the given UV absorption properties of the fluid. Since the low pressure mercury arc lamps are excellent absorbers of light at the 253.7 nm wavelength, the model calculations presume that any energy at this wavelength entering a lamp from a neighboring lamp will be completely absorbed by that lamp.

Figure 7-24 is presented to illustrate the intensity field calculated by the point source summation method. This example shows four lamps, spaced 5 cm apart. The important note to this is the non-uniformity of the intensity field and thus, the need to have turbulent flow, as discussed earlier, such that a particle will have the opportunity to be exposed to all intensity levels. In this fashion, it is appropriate to use the average intensity computed for the reactor.

Computationally, the point source summation method is not convenient and is best handled by computer with the appropriate software. In lieu of this, a series of solutions have been developed, and are presented in this manual, which describe the average intensity for almost any practical lamp configuration which would be considered by the designer.

7.3.3.3 Nominal Average Intensity Estimates for Alternative Lamp Configurations

Different lamp configurations will yield different nominal intensities in the reactor. Calculations have been performed for a number of designs, and subsequently reduced to show the nominal intensity as a function of the UV density of the reactor, and the wastewater absorbance coefficient.

UV Density. The UV density, D, is defined as total nominal UV power (at 253.7 nm) available within a reactor divided by the liquid volume of the reactor:

$$D = \text{total UV output/liquid volume} \qquad (7\text{-}29)$$

$$= \text{UV watts/liter}$$

As an example, Unit 1 in the Port Richmond study contained a total of 100 lamps. The liquid volume (internal reactor volume minus the volume occupied by the quartz sleeves) was approximately 375 liters. The lamps were the G37T6VH lamps described in Table 7-7. The nominal output is shown as 14.3 W/lamp; thus, the total UV output is 100 x 14.3 = 1430 W. The UV density of the reactor is then 1430 W/375 liters, or 3.8 W/L.

Obviously, the density will be directly related to the spacing of the lamps. The closer the spacing, the higher the UV density of the reactor.

Lamp Array Configurations. Four lamp "arrays" are considered; these are in common use today and, in effect, cover almost all practical configurations one would consider. The only assumptions which are made are that the lamps are always parallel to one another and that the single array pattern is continuous and symmetrical throughout the reactor. Both are appropriate and would be expected from a practical design. The four arrays are: (1) uniform array; (2) uniform staggered array; (3) concentric array; and (4) tubular array.

Uniform Array—
A cross-section of a uniform array is given on Figure 7-25(a). The lamps (with quartz sleeves) are arranged in even horizontal and vertical rows, with the centerline spacings equal in both directions.

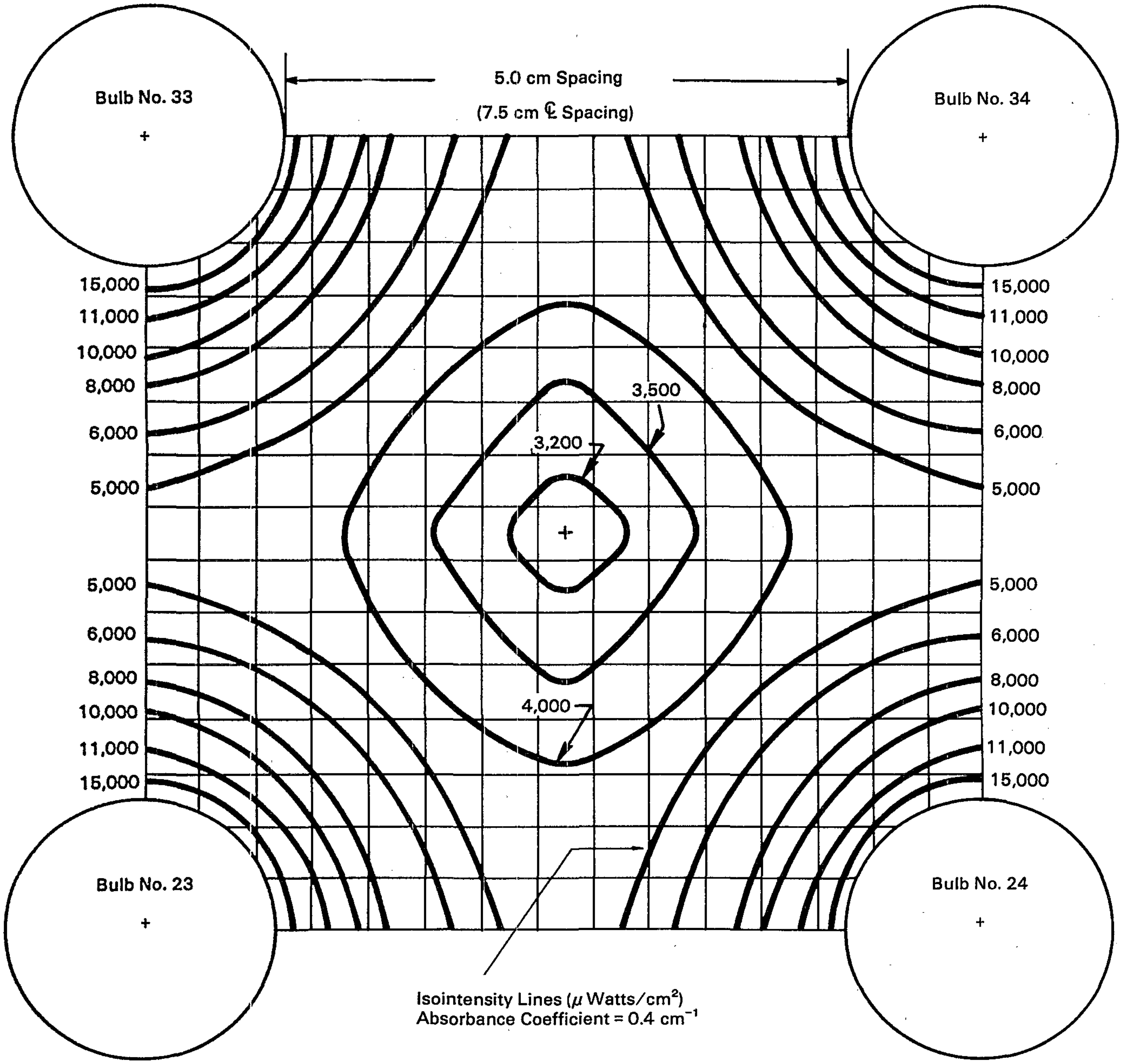
Staggered Uniform Array—
This is similar to the uniform array, except that the alternating vertical rows are offset by one-half the vertical spacing, S_v, as shown on Figure 7-25(b). The flowpath is typically perpendicular to the lamps; the staggered effect is designed to influence turbulence.

Concentric Array—
In this configuration, the lamps are arranged in concentric circles. This is illustrated on Figure 7-26(a), which is the cross-section of the unit at the Vinton, Iowa wastewater treatment plant. Typically, the array is designed to shut off banks of lamps. The banks are distributed throughout the reactor; this in effect, alters the UV density of the reactor as lamp banks are turned on or off. As more and more lamps are turned off, this can possibly cause very non-uniform intensity fields. This is not the case with the uniform arrays.

Tubular Array—
The tubular array describes the Teflon tube systems in which the lamps are suspended outside and parallel to a Teflon conduit. This is illustrated on Figure 7-26(b). The lamps and tubes are stacked vertically in alternating rows, with equivalent vertical and horizontal centerline spacing.

When considering the tubular array configuration, it is important to understand the effect which the number of vertical rows of tubes in a system have on the lamp requirement (and the consequent UV density). The meters of arc required for each meter of Teflon tube will vary with the number of vertical tube rows in the system. This is shown on Figure 7-27

Figure 7-24. Illustration of the intensity field calculated by the point source summation method.

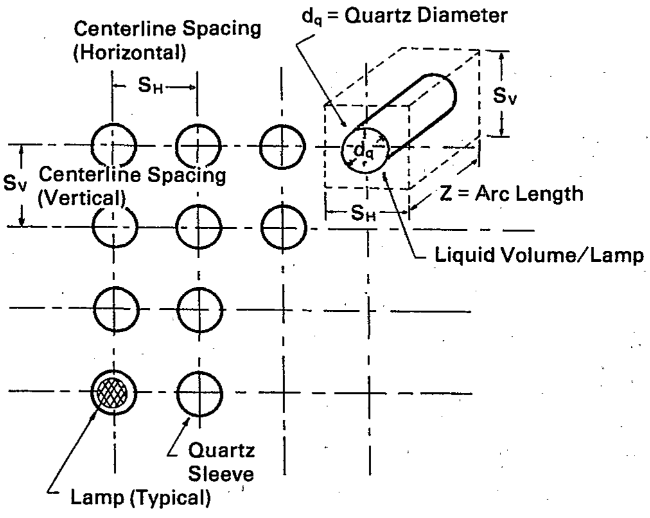


where n is the number of vertical rows of Teflon tubes in the unit. The number of lamp rows would be n+1. Thus, if the system has only one row of tubes, two rows of lamps would be required, yielding the ratio of 2 meters arc/meter tube. As n becomes greater, the ratio approaches 1.0. From Figure 7-27, the greater efficiency is achieved in systems with greater than 10 Teflon tube rows. In computing the UV density for the UV intensity solutions in this manual (Figures 7-31 and 7-32), the ratio of meters arc to meters tube is assumed to be 1.1.

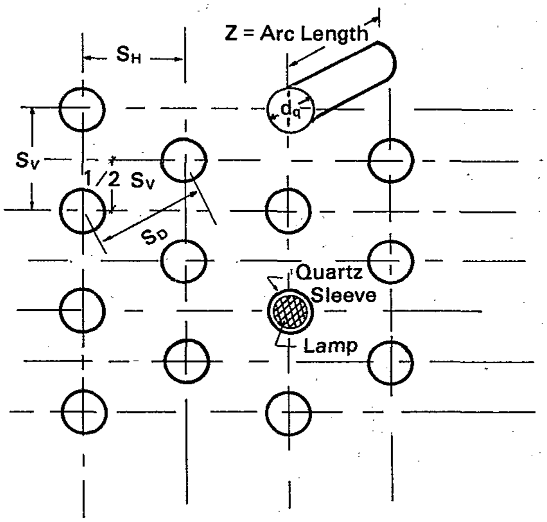
Nominal UV Intensity Estimates. Figure 7-28 presents the estimated nominal UV intensity as a function of the UV density for the uniform lamp array configuration. These are shown for a range of UV absorbance coefficient values between 0.2 and 0.9 cm^{-1}. In similar fashion, solutions are presented in Figure 7-29 and 7-30 for the staggered uniform array, and the concentric array, respectively.

Figure 7-31 presents the solutions for the tubular array, in which the lamp/tube/lamp centerlines are

Figure 7-26. **Schematic of uniform and staggered uniform lamp arrays.**



(a) Uniform Array



(b) Staggered Uniform Array

15 cm, as shown on Figure 7-26(b). The density for the system is adjusted by changing the diameter of the Teflon tube. Current designs use the 15 cm centerline configurations with a Teflon tube diameter of 8.9 cm.

An additional analysis is presented on Figure 7-32 for the tubular array configuration. This presents a comparison of the solutions for the 15 cm centerline spacing to the more compact (but equivalent density), 10 cm centerline spacing configuration. The tube diameter is 6 cm (D = 1.95 W/L); the nominal UV intensity is shown as a function of the UV absorbance coefficient. By drawing the lamps in closer to the Teflon tube, greater energy efficiency is achieved. The compactness of the tubular array, however, will be influenced and limited by practical fabrication considerations.

As shown on Figure 7-33, the greater efficiency is demonstrated by the quartz arrays. The tubular arrays are shown to be less efficient in accomplishing an intensity for a given UV density. This is an artifact of the physical constraints on the unit fabrication. In order to arrange the lamps and tubes and still maintain the system for easy assembly/disassembly and for access to these components, there will be limits as to how closely spaced the lamps and tubes can be.

7.3.3.4 Energy Loss Factors to Adjust the Nominal Average Intensity Estimate

An important note applies to the solutions presented in Figures 7-28 through 7-33; the intensity is calculated at the nominal output of the lamp and assumes that the quartz sheath or the Teflon tube will transmit 100 percent of the energy emitted by the lamp. Thus, the term "nominal" average intensity. Under actual operation, and for design purposes, this nominal average intensity must be adjusted to account for the aging of the lamps, and the consequent reduction in UV output, and for the losses of energy as it passes through the quartz sleeve or Teflon tube wall. These losses are due to the quartz or Teflon wall itself and to fouling of the inside and outside surfaces. Thus, in order to estimate the actual intensity under a given set of conditions, it is necessary to adjust the nominal intensity:

$$I_{avg} = (\text{Nominal } I_{avg}) \times (F_p) \times (F_t) \qquad (7\text{-}30)$$

where:

F_p = the ratio of the actual output of the lamps to the nominal output of the lamps

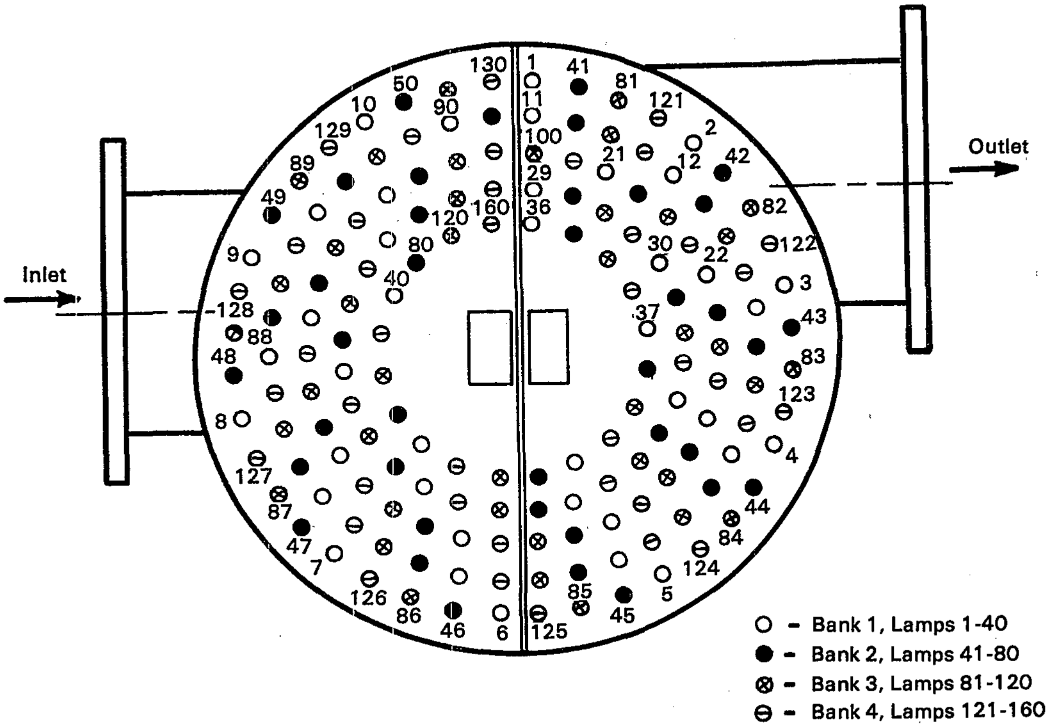
F_t = the ratio of the actual transmittance of the quartz sleeves or Teflon tubes to the nominal transmittance of the enclosures; the nominal transmittance is presumed to be 100 percent in the intensity calculation

Procedures are described in a later section to directly monitor the average values of F_p and F_t for a given system, and are strongly recommended as control procedures in a plant's O&M program.

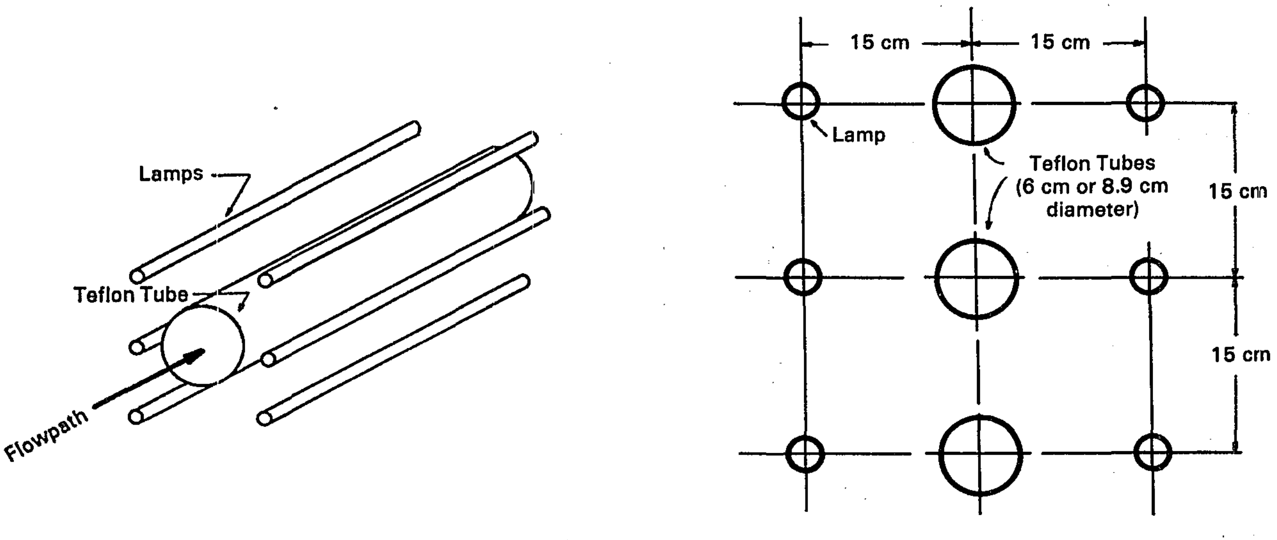
When designing a new unit, it is suggested that the system be designed at an average F_p of 0.7, which is representative of a lamp inventory output at approximately one half its operating life. In a sense, economics come into play here. The UV lamps are expensive. Their rated life of 7,500 hours can sometimes be greatly exceeded, but at a cost of reduced output. The inefficiency may be balanced by not having to buy new lamps.

The F_t should reflect the anticipated maintenance input; if the system will be well attended, a reasonable

Figure 7-26. Schematic of concentric and tubular lamp arrays.



(a) Concentric Array



(b) Tubular Array

Figure 7-27. **Effect of Teflon system sizing on the power requirement efficiency.**

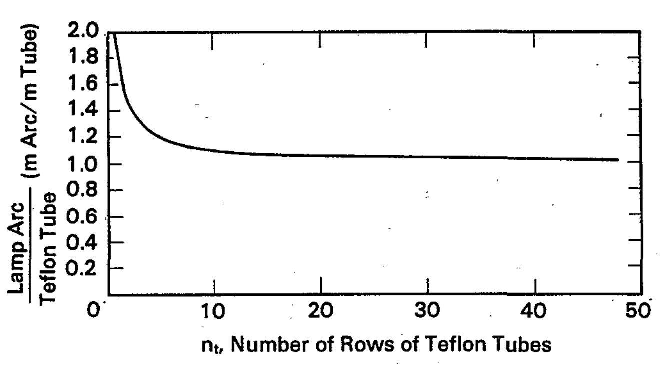


Figure 7-28. **Uniform lamp array intensity as a function of the reactor UV density and UV absorbance coefficient.**

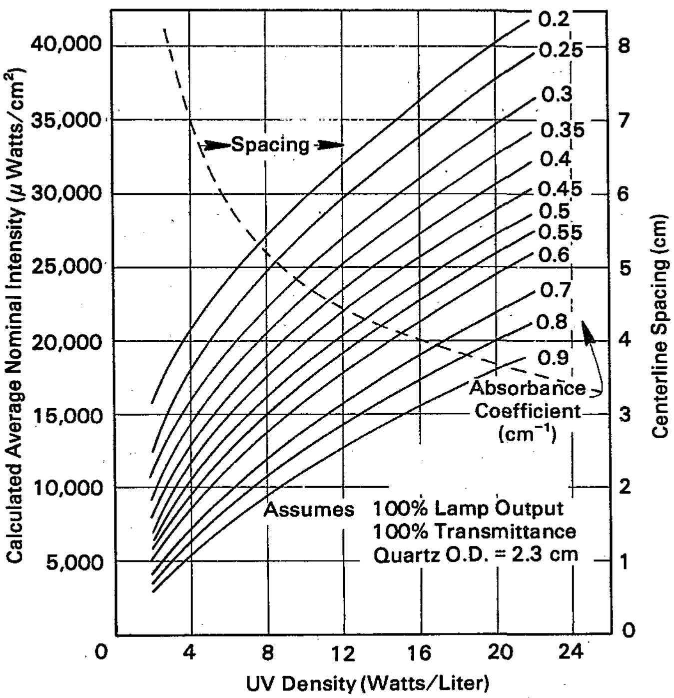


Figure 7-29. **Staggered uniform array intensity as a function of UV density and UV absorbance coefficient.**

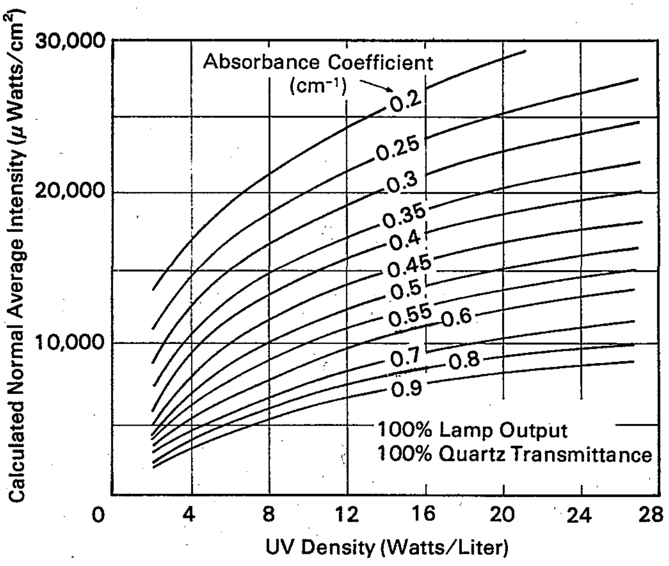
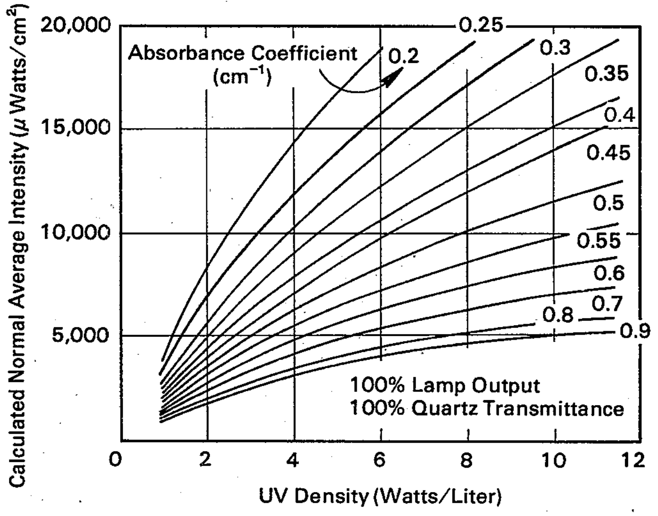


Figure 7-30. **Concentric lamp array intensity as a function of UV density and absorbance coefficient.**



Concentric Array
(see Figure 7-25(a))
100% Lamp Output
10% Quartz Transmittance

Figure 7-31. **Tubular array (15 cm ℄) intensity as a function of UV intensity and UV absorbance coefficient.**

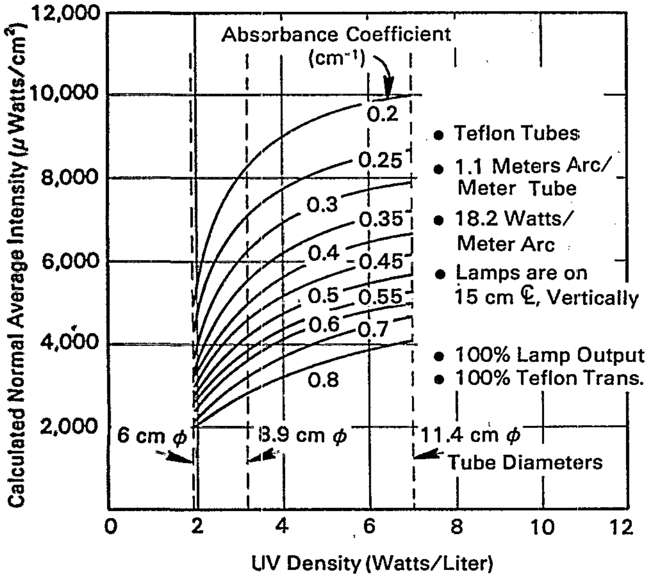
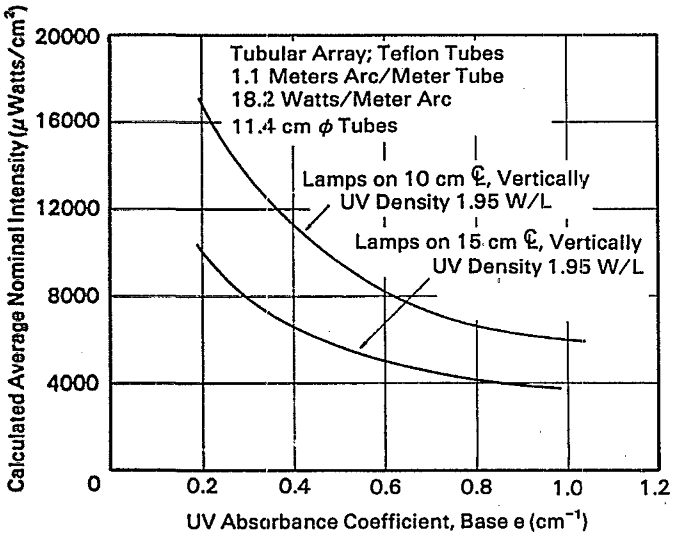
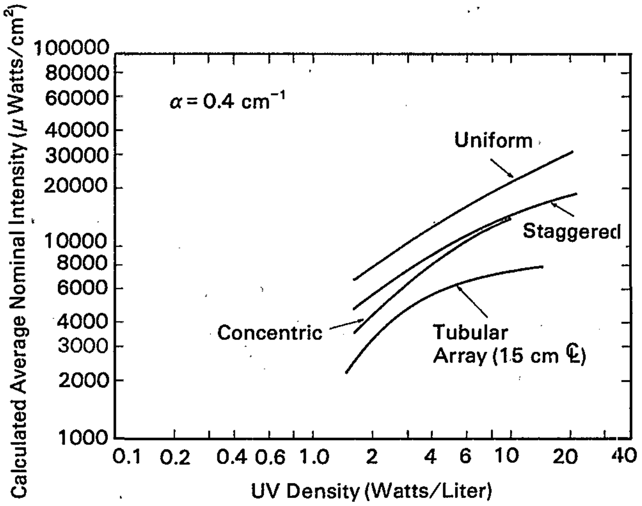


Figure 7-32. **Effect of centerline spacing on intensity for tubular arrays of the same UV density.**



minimum F_t would be 0.7 for a quartz system and 0.6 for a Teflon system. If minimal operator attention is anticipated, or if the enclosures are especially prone to fouling due to wastewater conditions, an F_t of 0.5 to 0.6 may be more appropriate for the quartz units and 0.4 to 0.5 for the Teflon unit.

Figure 7-33. **Calculated intensity as a function of UV density for different lamp array configurations.**



7.3.3.5 Summary—Estimation of the Average Reactor Intensity

The arrays which have been presented in these discussions encompass almost all configurations which are utilized in current systems, and which are practicable for future applications. Thus, the relationships on Figures 7-28 through 7-32 represent, in effect, solutions to the nominal average intensity calculations for any given practical lamp configuration. As long as the arrangement being considered can be represented closely by one of these solutions it is not necessary to independently apply the point source summation calculations. This would apply to units which will not be strongly affected by boundary conditions, which holds for systems with greater than 40 lamps.

Finally, the nominal intensity of the given configuration must be adjusted by the factors F_p and F_t to account for the actual (or planned) average lamp output and average transmittance of the quartz or Teflon enclosure.

7.3.4 Wastewater Quality Considerations in the Design of a UV System

Certainly, a major element in the effective design of a UV disinfection system is a clear and concise understanding of the characteristics of the wastewater to be treated. These are directly related to degree of pretreatment the material will receive before the disinfection step and certainly affect the sizing and performance of the disinfection system. Pretreatment ranges from very minimal, as in the case of combined sewer overflows and primary effluents, to very high quality, as in the case of advanced tertiary plants. Typically, with regard to UV disinfection application, the wastewater will result from secondary and advanced treatment facilities.

The following discussions will present the key elements of the wastewater application as they apply to the design (and subsequent operation) of a UV disinfection system:

- wastewater quality parameters. These are the flow (Q), initial bacterial density (N_o), suspended solids (SS), and UV absorbance coefficient (α);
- estimated bacterial density associated with the suspended solids (N_p). This includes a determination of the coefficients c and m in Equation (7-13);
- estimated inactivation rate (K). Recall that this is set as a function of the average intensity, requiring a determination of the coefficients a and b in Equation (7-12).

Relevant wastewater characterization data (including estimates of the coefficients, a, b, c, and m) from existing wastewater treatment plants are summarized to demonstrate the range of values to be expected under typical wastewater applications.

Photoreactivation is then discussed, with a presentation of field data to demonstrate levels of repair one can expect to occur. Finally, some discussion is given to sampling considerations and to suggested monitoring programs.

7.3.4.1 Key Wastewater Quality Parameters

The four wastewater parameters which most affect the design or performance of a UV system are the flow, initial bacterial density, suspended solids (or some measure of the particulates in the wastewater), and the UV absorbance of the wastewater.

Flow Rate. The flow rate is set by design of the main plant and projections of the hydraulic load to the plant. In evaluating the design requirements for the disinfection process, some consideration should be given to the equalization effects of the treatment processes before disinfection. This can have an effect on the sizing of the UV system.

Flow estimates should be for the design year of the plant. There should also be some knowledge of the progressive increase in the flows through the design life of the plant in order to determine if the system can be phased in by the addition of modules as the demand increases. Some consideration should also be given to the hydraulic load to the unit. The flow rates important to the design and evaluation of the system are those typically considered for wastewater treatment systems:

Annual average daily flow
Maximum 7-day average flow
Maximum 30-day average flow
Peak daily flow
Peak hourly flow

For disinfection, average flows are not critical to the design sizing; rather they are important to estimating average utilization of the system for operation and maintenance needs. Peak flows should be used for sizing, particularly reflecting diurnal variations.

Initial Coliform Density. The performance of a UV disinfection system is directly related to the initial density of the indicator organisms. This is not a parameter which is generally monitored at a treatment plant, particularly where the disinfection is accomplished by chlorination. In the case of disinfection by UV, however, it is critical. Performance is given by the log of the survival ratio, N/N_o, or by the number of "logs" the density is reduced.

Expected initial densities cannot be predicted solely from the type of treatment process preceding the disinfection process. Order of magnitude levels are given in Chapter 2 as guidelines. Examples drawn from several plants (presented in a later discussion) vary widely and do not correlate well with the types of systems or plant residence time. It is recommended that these data be generated before design; effluents can be analyzed from similar plants in the area, or at the existing facility if an upgrade or retrofit is being considered.

Suspended Solids. From the development of the disinfection model, it is clear that the occlusion of bacteria in the particulates will have a significant effect on the design of a UV system. It is recommended that the suspended solids measurement be used as the primary indicator to quantify these particulates.

The level of suspended solids in the effluent of a wastewater treatment facility is, in effect, set by the design of the plant. This then will limit the range of suspended solids concentrations to be considered in the design of the UV process. A further consideration is to understand the variability associated with the effluent suspended solids. As an example, if a plant is designed not to exceed 30 mg/l on average for any consecutive 30-day period, the suspended solids levels it must meet on an annualized basis will likely be between 10 and 20 mg/l. This can affect the sizing of a UV facility and determination of its average operational requirement.

UV Absorbance. The one parameter which is solely in the venue of UV disinfection is the UV "demand" of the wastewater. Specific organic and inorganic compounds in the wastewater will absorb energy at the 253.7 nm wavelength. This absorbance will affect the intensity of the radiation within the reactor; in specific design situations, the level of absorbance will affect the sizing of a system and possibly the configuration (spacing) of the lamps. Recall from the discussions of intensity, and its calculation in a complex reactor, that the final product of these calculations is the average nominal intensity as a function of the UV absorbance coefficient.

There are a number of ways to express the absorbance of a wastewater. First, consider the manner in which it is measured. The wastewater sample is placed in a quartz cell (transparent to the 253.7 nm wavelength) of a given width. A spectrophotometric measurement of the absorbance is made of a direct beam of light (at 253.7 nm) which is passed through the quartz cell containing the liquid. A detector determines the amount of light which passes through, and by inference, the amount of light "absorbed" by the liquid sample can be determined. The output of this measurement is absorbance units per centimeter, or a.u./cm. The pathlength is set by the quartz cell; typically this is 1 centimeter.

The transmittance of the wastewater is a common parameter used to describe the "demand" of the wastewater. This can be determined from the absorbance measurement, and is most often expressed on a percent basis:

$$\%\,\text{Transmittance} = 100 \times 10^{-(\text{a.u./cm})} \quad (7\text{-}31)$$

Conversely, the percent absorbed is simply 100 percent minus the percent transmittance. The parameter which is most often used for design purposes, and is the parameter used within the context of this manual, is the UV absorbance coefficient, expressed in base e:

$$\text{UV absorbance coefficient, } \alpha = 2.3(\text{a.u./cm}) \quad (7\text{-}32)$$

The unit of the UV absorbance coefficient is cm^{-1}. The reader is referred to Section 7.1.2 which presents typical absorbance levels for varying degrees of treatment.

The single beam, spectrophotometric method for measuring the UV absorbance of the liquid is the simplest procedure, requiring minimal effort and instrumentation. It is important to note however, that this "direct" UV absorbance measurement assumes that light which does not pass through the cell and is not seen by the detector has been absorbed by the liquid. This is not necessarily the case, especially in samples which have suspended or colloidal particles in the liquid. These will cause a portion of the light to be scattered; the light is still available, but it will not be seen by the detector since it has been deflected from its direct path through the quartz cell. Thus, the direct method tends to overestimate the "true" absorbance of the liquid.

Johnson and Qualls (38) and Scheible et al. (52), demonstrated that suspended or colloidal particles will not absorb any significant amount of light energy and will in fact scatter the light back to the liquid. It becomes important, therefore, that the absorbance measurement must in some fashion account for the scattering effect and give a value representative of the true absorbance of the liquid. Note that the procedures to calculate the intensity in a reactor inherently presume that the UV absorbance coefficient reflects the true absorbance of the liquid.

The Port Richmond study incorporated the use of a standard accessory to the UV/Visible spectrophotometer which would correct the absorbance measurement for the effect of scattering. A sphere, in effect, surrounds the quartz cell; any scattered light is absorbed on the surface of the sphere, which integrates the quantity of light collected and corrects the absorbance measured by the direct beam detector. This absorbance, referred to by the Port Richmond report as the "spherical" absorbance coefficient, is felt to more closely represent the true absorbance of the liquid, and is more appropriate for use in estimating the intensity in a reactor.

In the case where the capability to measure the corrected UV absorbance coefficient is not available, the UV absorbance coefficient should, at minimum, be determined on filtered samples by the direct method. In most cases this will give an approximation of the true absorbance. The results would be further improved if this is accomplished by membrane filtration to remove particles greater than 1 micron in size. Care should be taken to prewash the filters; in some instances the filter material itself can contribute UV absorbing materials.

At Port Richmond, limited testing was conducted on an in-line continuous monitor of the UV absorbance. This was a prototype instrument which would continuously sample the influent of a UV system and determine the UV absorbance at 253.7 nm; the instrument would also correct, to some degree, for scattering. The monitor was found to respond well to the absorbance of the wastewater. Such a direct monitor would be useful, in conjunction with the flow rate meter, in controlling the operations of a UV system.

The daily average, maximum 7-day, and the maximum 30-day average UV absorbance coefficient would be important to design. Unlike the suspended solids, which is limited by permit and by the treatment process, the UV absorbance coefficient is not a parameter describing treatment goals. It is an artifact of the wastewater and the treatment of that wastewater. Thus, an estimate will have to be made of the UV absorbance coefficient, and its variability, either by direct measure of the treated wastewater (as in the case of an existing plant) or a similar wastewater undergoing the same degree of treatment.

In summary, the parameters of primary concern regarding the characteristics of the wastewater are the flow rate, UV absorbance coefficient, suspended solids, and initial coliform density. The peak design condition for a plant should not necessarily consider the concurrent occurrence of these parameters as the worst case condition. If sufficient data are available, running averages of combinations of these parameters should be constructed to determine the maximum 7-day and 30-day average of the combined parameter set.

The maximum 7-day coliform density may occur under average flow conditions, as an example. Analysis of the data in this fashion may allow a more realistic system sizing to achieve the desired performance under the anticipated worst case condition.

7.3.4.2 Estimating N_p, the Bacterial Density Associated with the Suspended Solids

The estimate of N_p requires generating data under high dose levels. Recalling earlier discussions, the premise is that by determining the residual density after high doses, the residual can be attributed to those bacteria which were occluded from the radiation. These are then correlated to the suspended solids concentration which were present in the same sample.

In effect, one is generating information which is well out on the "tail" of a typical dose-response curve (see Figure 7-14). An example is given on Figure 7-34, which shows the log effluent fecal coliform (after exposure) plotted against the log effluent suspended solids. These data are from the Port Richmond study. A linear regression analysis yields the expression (when transformed):

$$N_p = 0.26\ SS^{1.96} \qquad (7\text{-}33)$$

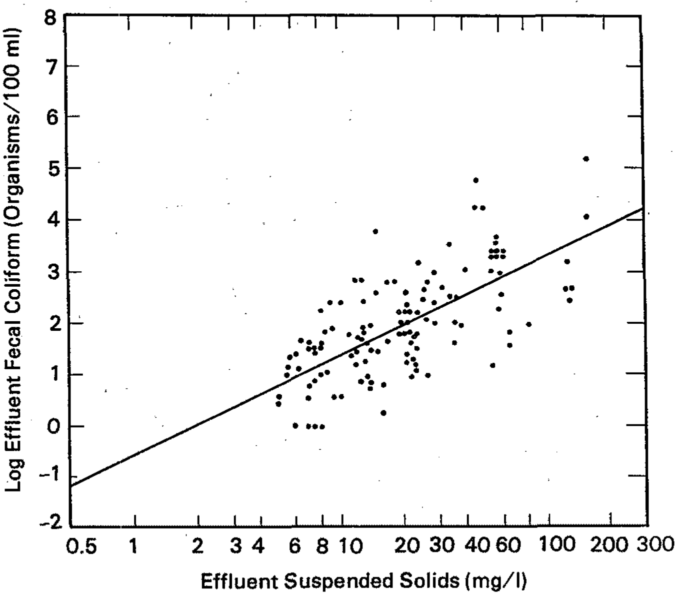
where N_p is in the units colonies/100 ml and SS is in mg/l. The coefficients 0.26 and 1.96 are the values of c and m, respectively, from Equation 7-13. The intercept of the regression on Figure 7-34 is c; m is the slope.

It is best to determine c and m by direct testing. Values determined at several plants are presented in a later section. These had been developed from flow-through tests. Although batch tests have not been conducted, one should expect that these would provide acceptable data under the proper conditions.

7.3.4.3 Estimating the Inactivation Rate, K

It is necessary, again, to generate a specific set of data to estimate the inactivation rate as a function of the average intensity. The required samplings are those in which the operating conditions would not allow for maximum kill at some point in the reactor. In other words, the apparent dose is low enough such that a significant coliform density would still be evident in the exposed effluent. This allows a valid estimate of the rate of inactivation in a specific sampling utilizing the initial and final coliform densities. One is, in effect, operating in the portion of the dose-response curve (see Figure 7-14), where the relationship is linear, with a constant slope.

Figure 7-34. **Example of deriving an estimate of the residual fecal coliform density associated with particulates as a function of suspended solids (55).**



The data are generated by piloting a system on the subject wastewater. For proper analysis of the data, the hydraulic characteristics of the unit need to be determined; these are the dispersion number and the dispersion coefficient. Once the data subset is developed, the rate coefficient can be estimated for each sampling by manipulation of the model equation (Equation 7-10) to solve for K. Estimates are first made of N_p from the suspended solids data and Equation 7-13, as described above. This is then subtracted from the densities measured after exposure:

$$N' = N - N_p \qquad (7\text{-}34)$$

Solving Equation (7-10) for K yields:

$$K = [u^2(P^2 - 1)]/4E \qquad (7\text{-}35)$$

where:

$$P = 1 - [2E \ln(N'/N_o)]/ux \qquad (7\text{-}36)$$

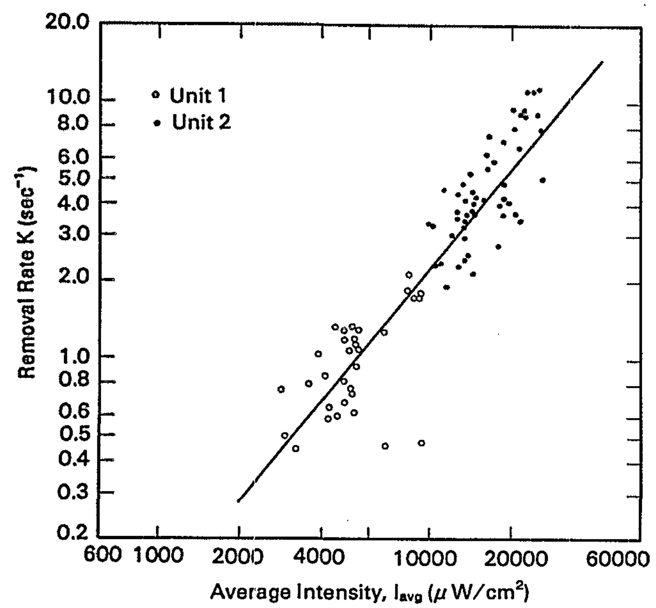
The rate K has the units second^{-1}. Equation (7-35) can be used to solve for K for each sampling; the inputs are the observed initial and final coliform densities, the velocity based on the observed flow rate, the x dimension based on the operating condition of the unit, and the dispersion coefficient determined for the reactor. These values of K are then correlated to the estimated I_{avg} in the reactor corresponding to the conditions for each sampling.

The correlation of log K as a function of log I_{avg} for fecal coliform data generated at Port Richmond are presented on Figure 7-35 as an example of the rate analysis. Linear regression analysis yielded the expression (when transformed):

$$K = 0.0000145\,(I_{avg})^{1.3} \qquad (7\text{-}37)$$

where I_{avg} is the average intensity in μWatts/cm^2. The coefficients 0.0000145 and 1.3 are the values of the intercept a and the slope b, respectively, of Equation 7-12. Values of a and b derived at a number of plants are presented in a later section.

Figure 7-35. An example for deriving an estimate of the inactivation rate for fecal coliforms as a function of the calculated average intensity (56).



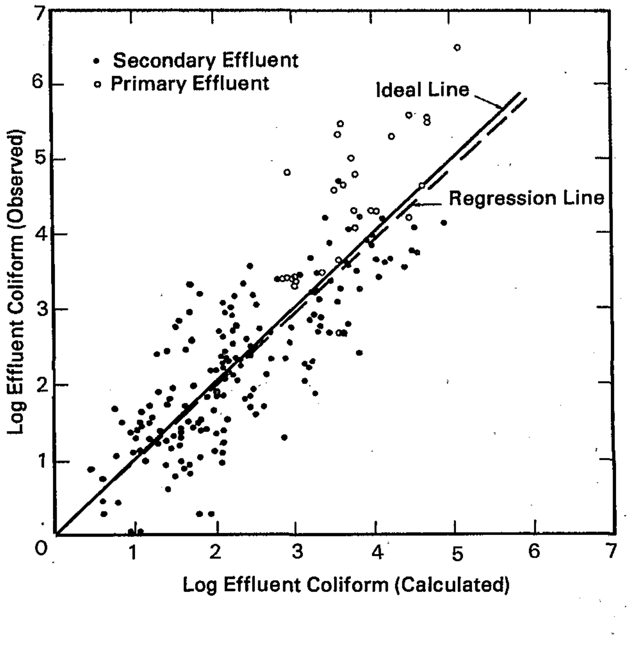
Batch testing, although not applied or demonstrated at the time of this publication, can also be used to derive the coefficients a and b. A clear advantage to the batch technique would be its independence of hydraulic considerations.

7.3.4.4 Checking the Coefficients Determined for Use in the Model

The coefficients a, b, c, and m are specific to a given wastewater application and reflect the site-specific sensitivity of the microorganisms to UV (a and b) and the level to which microorganisms are occluded in the effluent suspended solids. When these are determined by direct piloting, it is appropriate to verify their values by checking against data generated independent of the data set used to derive the coefficients.

Influent and effluent data should be collected over a range of conditions. Using Equation 7-14, calibrated to the proper coefficients and operating conditions, the performance (Log N/N_o) can be predicted for a given sample. This can then be compared to the observed performance. Again, an example of this type of analysis is presented on Figure 7-36, which shows the observed versus predicted fecal coliform densities from the Port Richmond study. An analysis of the data indicated that the regression line for the observed versus predicted correlation was not significantly different from the ideal line in which the slope is 1.0 and the intercept is 0.0. In all, the analysis suggests that the model correctly responds to the varying operating parameters of the UV system, and when properly calibrated, will successfully predict performance under any matrix of operating and hardware configurations.

Figure 7-36. An example of the comparison of disinfection model estimates to observed effluent fecal coliform densities (56).



7.3.4.5 Summary of Wastewater Data from Existing Plants and Recent Field Studies

Wastewater quality data were compiled from several wastewater treatment plants. This was done to give the reader a perspective on water quality characteristics, particularly those parameters relevant to the

UV disinfection process. The list of plants is summarized on Table 7-8, including a brief description of the treatment process at the plant and the level of treatment the plant is designed to achieve. The plants were selected because full or pilot scale UV evaluations had been conducted at each yielding a consistent set of data relevant to UV.

The Port Richmond plant is listed first; data are given for both the primary and secondary treatment levels (52). The Suffern, Vinton, and Eden data are from special one-month studies (54). The Northfield plant (59) is an existing full scale facility and is one of the few facilities that routinely monitors on a frequent basis. Data from the 1984 disinfection season are presented. The Tillsonburg plant is a full-scale UV facility in Ontario; data are presented from a long-term demonstration study (48). The Northwest Bergen study entailed a one-year full-scale pilot evaluation (36).

Five plants are listed from Ontario, Canada: Toronto-Main, Georgetown, Milton (secondary and tertiary), Hamilton, and Toronto-Lakeview. They were all sites of UV pilot plant studies (47). Finally, the last four plant sites, New Windsor, Newburgh, Suffern, and Monticello (secondary and tertiary), are from special pilot studies conducted in 1985 (52). These studies also investigated the inactivation rate, K, and the particulate coliform density, N_p, as will be presented in the following discussions.

Initial Bacterial Density. Average bacterial densities are summarized on Table 7-9. These are all geometric means. The fecal coliforms are listed in all cases; this remains the indicator of choice in almost all permitting activities. Other indicators are listed when available.

As can be seen, there is little consistency in the density levels, nor is there any obvious correlation to the level of treatment. The fecal coliforms range from 10^4 to 10^6; for the purpose of preliminary sizing and/or system designs, a reasonably conservative initial fecal coliform density, N_o, would be 2 to 5×10^5 for secondary treatment levels, and 1 to 2×10^5 for tertiary levels.

Treated Effluent Quality. Table 7-10 presents average UV absorbance coefficient data and suspended solids information for each of the plant sites. Where available, turbidity, COD, and TKN data are also provided. Of particular interest is the UV absorbance coefficient; in all cases, the direct measurement (unfiltered) is provided. In several cases, both the direct and spherical measurements (unfiltered and filtered) are presented. Recalling the discussions in Section 7.3.4.1, the preferred method is the spherical (unfiltered) since it corrects for scattering and is the most representative of the actual UV absorbance characteristics of the wastewater. An alternative is to use the direct method on a filtered sample; except for a few cases, this is shown to give a reasonable approximation of the spherical UV absorbance coefficient.

The direct analysis of an unfiltered sample is certainly the easier procedure, requiring no sample preparation or special accessory to the UV spectrophotometer. Figure 7-37 indicates an excellent correlation between the direct (unfiltered) and spherical (unfiltered) data from nine plants. Thus, for preliminary design purposes, it is reasonable to estimate the spherical absorbance coefficient from the direct (unfiltered) analysis:

$$\alpha_S = 0.6\,(\alpha_D)^{0.64} \qquad (7\text{-}38)$$

where α_S and α_D are the spherical and direct (unfiltered) absorbance coefficients (base e), respectively, with the unit (cm^{-1}).

Inactivation Rate, K. Several plant sites were evaluated to directly determine the inactivation rate, K, as a function of the UV intensity (52). These used 2 pilot units; each had 12 lamps and differed only in spacing, and, therefore, in intensity. The results are presented on Figure 7-38, including the coefficients a and b (Equation 7-12). The Port Richmond regression (from Figure 7-34) is also presented (this also used two units, each differing significantly in intensity).

There is significant variability among the regression coefficients presented on Figure 7-38, although the actual values of K are not as varied. At the lower intensities (~3000) the average K ranges between 0.2 and 0.6 sec^{-1}, a factor of 3. The K ranges between 1.6 and 3.8 at an I_{avg} of 10,000, a factor of 2.4.

The data from the four plants were combined; the resulting regression line is also given on Figure 7-38. The Port Richmond plant is shown to be nearly equivalent to this combined regression.

At this point, sufficient data are not available to clearly demonstrate a uniformity in the K rate as a function of the I_{avg}. For this reason, it is strongly suggested that these data be generated by direct testing for specific plant applications. Preliminary design calculations can use an $a = 1.4 \times 10^{-5}$ and a $b = 1.3$, based on the results of the Port Richmond and combined plants regression analysis. The reader should understand that the I_{avg} is estimated on the basis of the spherical absorbance coefficient.

Particulate Denisty, N_p. The same plant studies which had evaluated the inactivation rate, also evaluated, on a limited basis, the coliform density associated with the suspended solids. These data are shown on Figure 7-39. As would be expected, there is

Table 7-8. Wastewater Treatment Plants Which are Sources of Wastewater Characterization Data

Treatment Plant Location	Treatment Level	Description
Port Richmond WPCP Staten Island, NY	Secondary	Step aeration activated sludge, secondary clarification
Port Richmond Staten Island, NY	Primary	High rate primary clarification
Suffern, NY	Advanced	Trickling filter; single stage (nitrification) activated sludge, clarification
Vinton, IA	Secondary	Extended aeration activated sludge, final clarification
Eden, WI	Secondary	Activated sludge, secondary clarification
Northfield, MN	Secondary	RBC and secondary clarification
Tillsonburg, Ontario	Secondary	Activated sludge; clarification
NW Bergen, NJ	Advanced	Single stage (nitrification) activated sludge; clarification
Toronto-Main, Ontario	Secondary	Conventional activated sludge
Hamilton, Ontario	Secondary	Conventional activated sludge
Georgetown, Ontario	Secondary	Conventional activated sludge
Milton, Ontario	Secondary	Conventional activated sludge
Milton, Ontario	Tertiary	Conventional activated sludge; rapid sand filtration
Toronto-Lakeview Ontario	Secondary	Conventional activated sludge
New Windsor, NY	Secondary	Trickling filters; clarification
Newburgh, NY	Secondary	Activated sludge; clarification
Suffern, NY	Advanced	Trickling filters; single stage (nitrification) activated sludge; clarification
Monticello, NY	Secondary	Oxidation ditch; secondary clarification
Monticello, NY	Tertiary	Oxidation ditch; secondary clarification; sand filtration

considerable scatter, not unlike the variability shown on Figure 7-34 for the Port Richmond study. The data tend to fall about the regression line developed for Port Richmond. This poor correlation can be due to several factors, including analytical precision at low levels; site differences; differing particle size distributions; etc. For preliminary design purposes, it is suggested that the coefficients c and m be set to 0.25 and 2.0, respectively.

7.3.4.6 Photoreactivation

The phenomenon of photoreactivation had been described in Chapter 7.2.2.2. Unique to ultraviolet radiation, the mechanism involves the repair of damage caused by exposure to UV, allowing for subsequent replication of the organism. The enzymatic mechanism generally involved requires subsequent (or concurrent) exposure to light at wavelengths between 300 and 500 nm; such light is available in sunlight and in most incandescent and fluorescent light sources.

The procedure to quantify the effects of photoreactivation is by the so-called static bottle technique. The method involves splitting an exposed sample to three aliquots: the first is set immediately for coliform enumeration; the second is placed in a bottle opaque to visible light; and the third is placed in a bottle

Table 7-9. Initial Bacterial Density Before Disinfection[a] (organisms/100 ml)

Plant	Treatment Level	Total Coliforms	Fecal Coliforms	Fecal Streptococci	*Escherichia coli*	*Pseudomonas aeruginosa*
Port Richmond	Secondary	1,020,000	361,000	–	–	–
Port Richmond	Primary	31,700,000	12,500,000	–	–	–
Suffern	Advanced	–	95,500	25,120	–	–
Vinton	Secondary	–	89,100	–	–	–
Eden	Secondary	–	44,700	21,400	–	–
Northfield	Secondary	–	75,500	–	–	–
Tillsonburg	Secondary	53,000	15,600	1,240	–	–
NW Bergen	Advanced	190,000	48,000	–	–	–
Toronto-Main	Secondary	1,300,000	120,000	33,000	110,000	2,500
Hamilton	Secondary	1,300,000	180,000	15,000	30,000	1,300
Georgetown	Secondary	77,000	9,200	3,500	8,300	29
Milton	Secondary	140,000	23,000	1,900	22,000	140
Milton	Tertiary	77,000	13,000	820	6,600	40
Toronto-Lakeview	Secondary	160,000	14,000	2,500	12,000	80
New Windsor	Secondary	–	1,900,000	–	–	–
Newburgh	Secondary	–	77,000	–	–	–
Suffern	Advanced	–	60,800	–	–	–
Monticello	Secondary	–	1,124,000	–	–	–
Monticello	Tertiary	–	736,000	–	–	–

[a] Geometric Means; data from analysis of grab samples.

Table 7-10. Treated Effluent Characteristics From Several Wastewater Treatment Plants[a]

		Average UV Absorbance Coefficient (base e) (cm^{-1} at 253.7 nm)							
		Direct		Spherical					
Plant	Treatment Level	(Total)	(Filtered)	(Total)	(Filtered)	Suspended Solids (mg/l)	Turbidity (NTU)	COD (mg/l)	TKN (mg/l)
Port Richmond	Secondary	0.466	0.404	0.372	0.358	14.3	4.0	44.5	6.9
Port Richmond	Primary	0.865	0.747	0.593	0.533	80.9	25.7	134.0	–
Suffern	Advanced	0.290	0.273	0.282	0.271	8.3	4.4	34.4	–
Vinton	Secondary	0.331	0.275	0.296	0.260	8.3	–	28.9	–
Eden	Secondary	0.391	0.354	–	–	33.2	8.6	39.0	–
Northfield	Secondary	0.378	–	–	–	12.2	–	–	–
Tillsonburg	Secondary	0.250	–	–	–	6.1	1.9	14.6	1.2
NW Bergen	Advanced	0.390	–	–	–	6.4	3.7	28.0	14.0
Toronto-Main	Secondary	1.07	–	–	–	26.8	6.3	92.0	28.6
Hamilton	Secondary	0.565	–	–	–	12.2	3.2	39.1	26.5
Georgetown	Secondary	0.348	–	–	–	5.8	2.0	34.8	14.5
Milton	Secondary	0.366	–	–	–	9.1	2.0	38.8	9.2
Milton	Tertiary	0.271	–	–	–	2.6	1.2	23.2	4.3
Toronto-Lakeview	Secondary	0.657	–	–	–	6.4	3.2	49.1	3.9
New Windsor	Secondary	0.894	0.705	0.578	0.500	31.6	–	–	–
Newburgh	Secondary	0.739	0.495	0.518	0.444	26.7	–	–	–
Suffern	Advanced	0.454	0.379	0.374	0.362	7.3	–	–	–
Monticello	Secondary	1.627	0.589	0.766	0.352	72.4	–	–	–
Monticello	Tertiary	0.687	0.440	0.416	0.328	13.3	–	–	–

[a] Arithmetic means; data from analysis of grab samples.

Figure 7-37. **Correlation to estimate the spherical absorbance coefficient from direct unfiltered absorbance coefficient.**

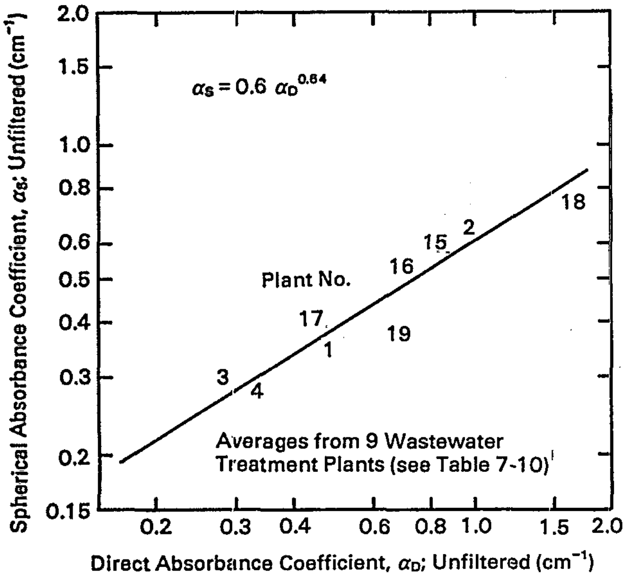
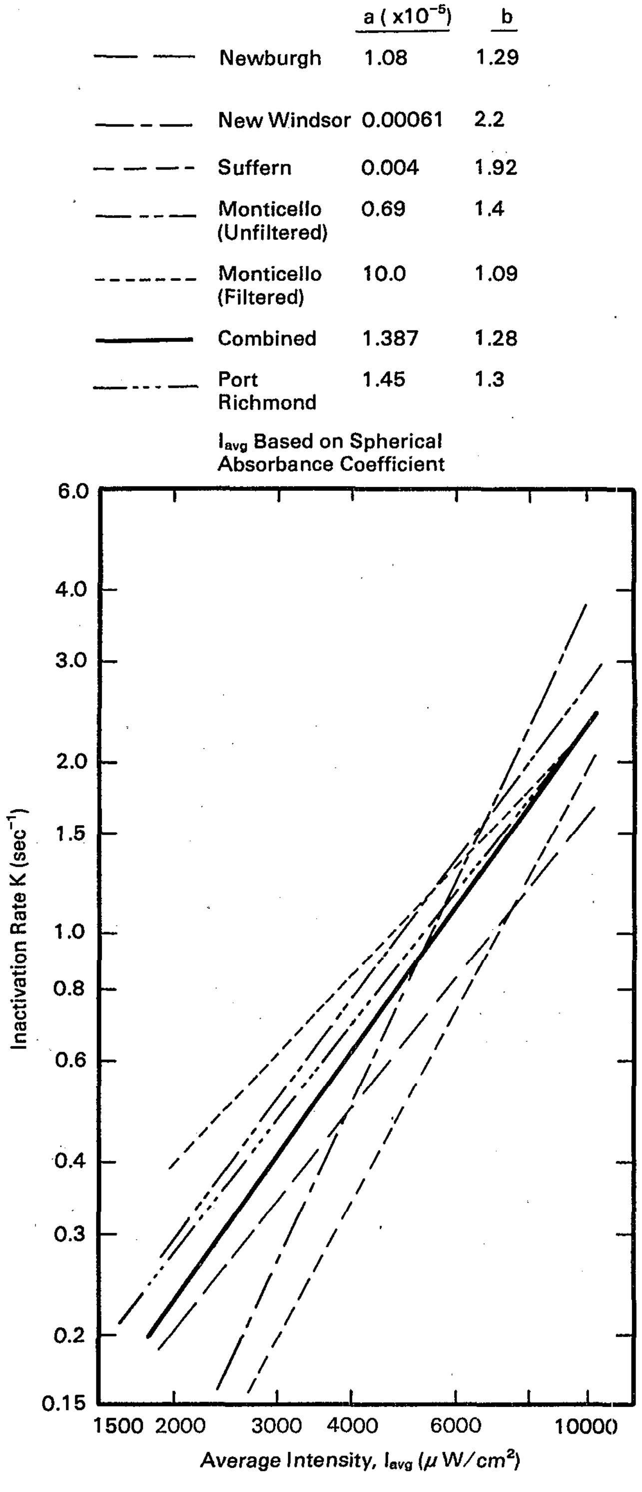


Figure 7-38. **Comparison of inactivation rate estimates from several wastewater treatment plants.**



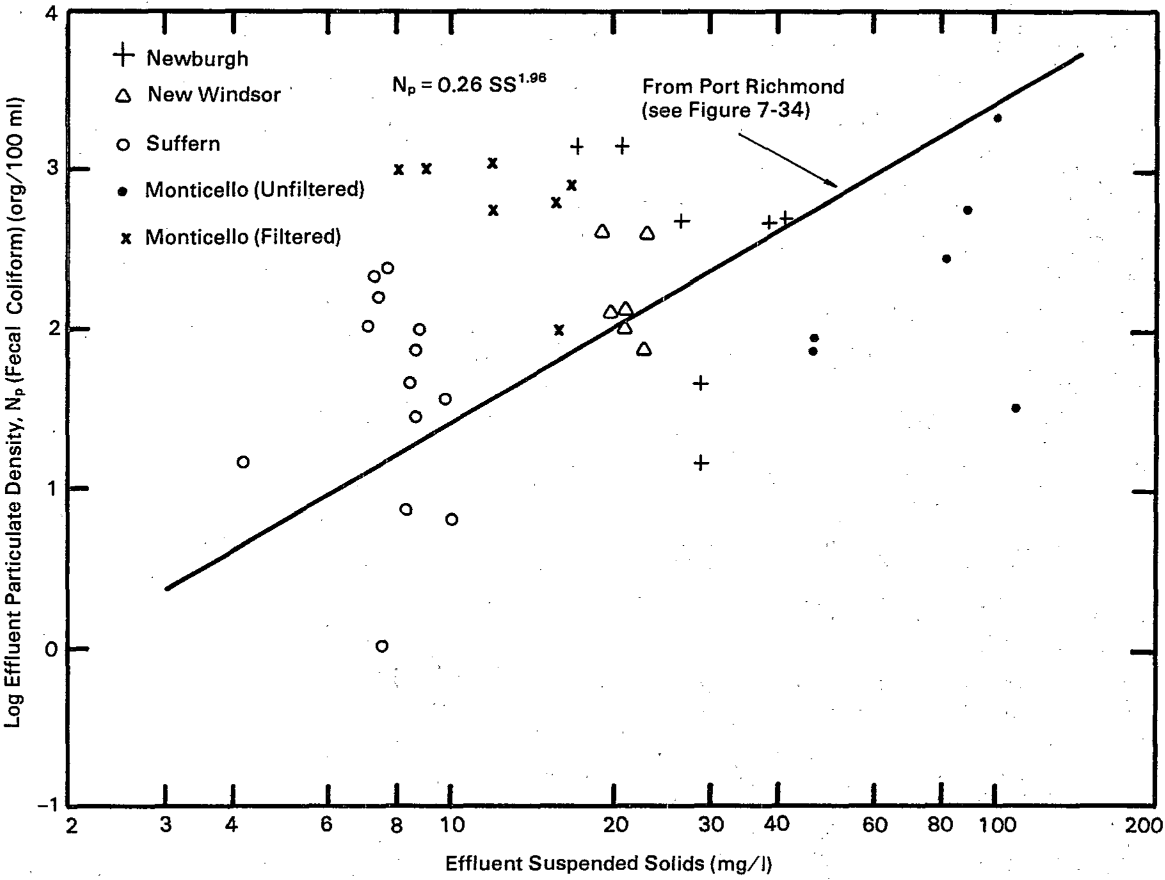
		a (x10^{-5})	b
— —	Newburgh	1.08	1.29
——— —	New Windsor	0.00061	2.2
— — — —	Suffern	0.004	1.92
—— - - -	Monticello (Unfiltered)	0.69	1.4
- - - - - - -	Monticello (Filtered)	10.0	1.09
———	Combined	1.387	1.28
—— - - - —	Port Richmond	1.45	1.3

transparent to visible light. The two bottles are then held for sixty minutes (this time is not standard, it can vary from one half hour to three hours), at 20°C, exposed to sunlight. The samples would then be set for coliform enumeration; the opaque bottle is the "dark" sample, and the transparent bottle the "light" sample. Holding the visible light exposure at constant temperature (20°C) is not critical. In situations where it is desired to monitor photoreactivation seasonally, it is more appropriate to suspend the two bottles just below the surface of the plant effluent or the receiving water. In this fashion, the degree of photoreactivation is being monitored under current temperature conditions.

The repair mechanism has a dependency on temperature (36). The analysis suggests that at 10°C, a two-fold increase in fecal coliform density will be caused by photoreactivation. If the temperature is approximately 20°C, a ten-fold (one log) increase in the effluent density is observed.

A series of tests were also conducted as part of the Port Richmond study to evaluate the impact of photoreactivation. Again, the tests centered on the indicators total and fecal coliforms, and utilized the static bottle technique. The resultant data are presented on Figure 7-40, which is taken from the Port Richmond report. Total and fecal coliforms at time 0 and 60 minutes (light) are plotted as a function of the loading parameter, Q/W, which is the ratio of flow (Lpm) to the UV output of the system (Watts at 253.7 nm). The suspended solids averaged 11.8 mg/l during

Figure 7-39. Estimation of N_p from several plants.



this period and the average spherical absorbance coefficient was 0.448 cm^{-1}. The temperature was approximately 23°C during the test period. As shown on the figure, the photoreactivation mechanism causes an increase of approximately 1.3 logs in either the total or fecal coliform. The increase is relatively constant, regardless of the applied dose. This is expected, since the degree of photoreactivation is independent of dose.

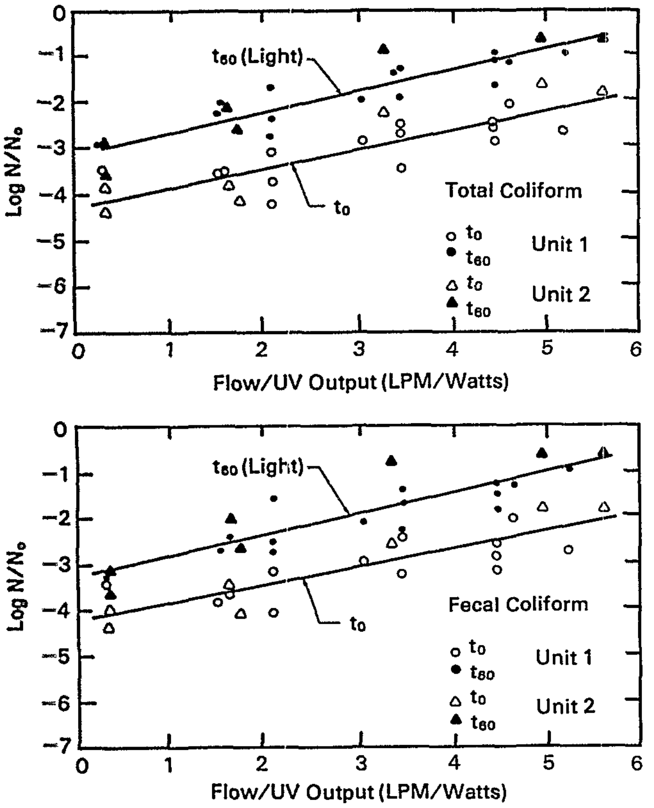
At present, the effects of photoreactivation are not directly addressed in most state permitted activities. Thus, it is appropriate to minimize the effect in the sampling and analysis of the exposed effluent. The degree to which this phenomenon exists among the pathogenic organisms is not fully understood; as an example streptococci do not photorepair, while *Shigella* do exhibit the ability. Viruses generally cannot photorepair, except in cases where the host cell can photoreactivate.

If the permitting agencies require that the photorepair phenomenon must be addressed in assessing the performance of a UV disinfection process, the design of the process can accommodate such a requirement. Based on the results of previous studies, it is also appropriate to assume that total and/or fecal coliforms would still serve as adequate indicators of pathogen activity. The design of the system must now address the inactivation of an additional fraction of the initial coliforms in order to meet guidelines after photorepair takes place. The critical design period is during the warmer temperature, higher sunlight intensity summer months. The system should generally be designed to accomplish approximately one additional log reduction. Thus, if it is determined that a 3-log reduction will be required to meet the permit levels, the system should be designed to accomplish a 4-log reduction to account for the effect of photoreactivation.

7.3.4.7 Wastewater Sampling Considerations for Design

The design and performance monitoring of a disinfection process is based on the measure of bacterial density. These are typically coliforms or fecal strep. Sampling for these can be accomplished only on a grab basis. Furthermore, the sampling for coliforms must typically be accomplished during daylight hours, often set between 10 a.m. and 3 p.m. This time period

Figure 7-40. **Photoreactivation effects for total and fecal coliform at Port Richmond (55).**



is considered the maximum loading period for a treatment plant, and reflects the maximum density levels with regard to the disinfection process. With this in mind, it is appropriate that the data that are generated to characterize the effluent for suspended solids, UV absorbance, initial coliform density, and flow should also be collected as grab (or short-term) analyses, and should correspond to the time of day in which the system is to be monitored for disinfection performance.

Twenty-four hour composites, which are typically collected at a plant, will not directly reflect conditions under which the system should be designed. At minimum, sufficient data should be generated to understand the variability of these parameters during a diurnal period. There is an additional benefit which can be gained from this information. The data will likely indicate significant improvement in absorbance and lower initial densities and suspended solids during the off-peak, early morning hours. The system can be adjusted to account for this, potentially resulting in energy savings.

The data base developed for design should reflect analyses of grab samples taken during the peak load hours. The parameters which must be monitored include suspended solids, coliform density, and the UV absorbance coefficient. Secondary parameters useful to characterizing a wastewater for design purposes are grease/oil, iron, and hardness. These will be important in considering the cleaning requirements of a system.

7.4 UV Disinfection System Design Example

The preceding sections presented the various elements important to the design of the UV process. These considered the hydraulic design of a reactor, the intensity of radiation, and the wastewater quality for the specific application. All were related to a design protocol based on a disinfection model (Equation 7-10).

This section presents a design example to demonstrate the use of the design protocol. It does not attempt to provide comprehensive solutions, but rather the procedures and calculations which the designer can use for specific applications. This can encompass several situations: to design a new reactor; to determine the adequacy of a proposed reactor design (e.g., by equipment manufacturer); and/or to evaluate the capacity and design adequacy of an existing system.

7.4.1 UV Disinfection Design Example

The example presented in this section of the manual is intended to illustrate the design considerations involved with development of a UV disinfection system. The wastewater treatment design information for the example design problem is shown in Table 7-11. The plant is an air-activated sludge plant located at an elevation of 1,070 m (3,500 ft) above sea level. The influent to the UV disinfection system is the effluent from the secondary clarifiers.

For this example, we are assuming that the wastewater data (fecal coliform and UV absorbance) represent grabs taken during the peak diurnal period (see discussion in Section 7.5.4). The disinfection portion of the plant will operate under gravity flow. There are no area constraints and the plant treats domestic wastewaters only. The plant's permit calls for year-round disinfection.

7.4.2 Setting Design Conditions and Parameters for Equipment Sizing

From the disinfection model (Equation 7-10), the information needs, aside from wastewater characteristics described earlier, are an estimate of the dispersion properties of the proposed reactor configuration (i.e., dispersion coefficient and the dispersion number), the inactivation rate, K, and the coliform density associated with the particulates, N_p.

Table 7-11. Example UV Disinfection System Design

Average Daily Wastewater Flow	28.4 mL/d (7.5 mgd)
Peak Daily Wastewater Flow	56.8 mL/d (15.0 mgd)

NOTE: The daily peak flow rate will not exceed 15.0 mgd because of storm flow equalization facilities.

Start-up Daily Average Wastewater Flow	13.2 mL/d (3.5 mgd)
Start-up Peak Daily Wastewater Flow	28.4 mL/d (7.5 mgd)
Average Effluent BOD/TSS	15/15 mg/l
Maximum Daily Effluent BOD/TSS	30/30 mg/l
Design Required Effluent Fecal Coliform	
Weekly Maximum Limitation (Geometric Mean)	400 per 100 ml
Average Monthly Limitation (Geometric Mean)	200 per 100 ml
Disinfection System Influent Fecal Coliform	
Geometric Mean Concentration (Daily)	500,000/100 ml
Maximum Concentration (7-day Mean)	2,000,000/100 ml
Maximum Concentration (30-day Mean)	1,000,000/100 ml
UV Transmittance (% at 253.7 nm)	
Daily Average	70%
Minimum 30-day Average	65%
Minimum 7-day Average	60%

7.4.2.1 Model Coefficients

The hydraulic characteristics are a direct function of the reactor configuration, particularly the lamp placement and spacing, and the reactor's inlet and outlet design. This can best be characterized by direct testing of full-scale modules, or hydraulically scaleable pilot modules. Alternatively, these data can be required from the equipment manufacturer as a bid and/or warranty specification. The procedures for these analyses have been described earlier. Specifically, the information should encompass the following:

a. Residence time distribution curves developed at several flow rates. These should encompass, at least, the minimum, average, and maximum design flows for the system.

b. Head loss measurements, again over a range of velocities (i.e., flow rates).

Calibration of the model requires direct determination of the inactivation rate as a function of the intensity and an estimate of the residual coliform density associated with the suspended solids. These are the coefficients a, b, c, and m in Equation 7-14. They will be site specific and will need to be determined experimentally. As more field experience is gained with the application of UV, these coefficients may be found to cluster about certain levels relative to the type of wastewater and the degree of treatment. Estimates of these coefficients for several plants were summarized in Section 7.3.4.5. These may be used as approximations for first-cut design estimates; it is recommended at this point, however, that direct testing be conducted to verify and/or refine these estimates.

7.4.2.2 Testing Requirements

Testing need be directed only at the data which are necessary for the model calibration. Demonstration of long term performance at a given loading or an evaluation of O&M needs over time are not issues which can be effectively resolved by limited piloting or lab tests. These elements of a system evaluation are best answered by observation of existing full scale facilities and the experiences of engineers and operators directly involved in their design and operation.

The tests should incorporate the ability to obtain data at two significantly different intensity levels. If direct piloting is conducted, this may best be accomplished by using two units which differ significantly in lamp spacing, such that the UV density, and consequently the intensity, differ significantly. The sizing of the systems does not need to be large, although it is recommended that the units should have 10 lamps (0.75 arc lamps would be sufficient) at a minimum. There are no restrictions on the configuration of the lamps relative to flow (e.g., parallel or perpendicular), or in the arrangement of the lamp array (the lamps should be parallel to one another, however).

The wastewater effluent should be piped to the unit and there should be the capability to vary the flow rate. Flow rates must be accurately measured. The sampling and analyses will center on measurement of the influent and effluent bacterial density (typically, this will be total and/or fecal coliforms), the UV absorbance coefficient, and the suspended solids.

The pilot unit lamp configuration would first require evaluation by the point source summation method to calculate the intensity in the unit as a function of the UV absorbance coefficient. This would be similar to the relationships presented in Figures 7-28 through 7-32, depending on the array configuration. During the term of the pilot study, direct measurements should be made to determine the output of the units' lamps and the transmittancy of the quartz and/or Teflon enclosures. These tasks can be accomplished by methods which are described in Section 7.5.1. The information on lamp output and quartz transmittance would be used to adjust the calculated intensities. The adjusted intensity would also be affected by the UV absorbance coefficient measured at the time of sampling.

The hydraulic characteristics of the pilot units would then have to be defined directly by running tracer analyses. The procedures were described in Section 7.3.2, including the analysis of the resulting data. The required information would be an estimate of the dispersion coefficient, E, and an estimate of the actual detention time (mean) versus the theoretical detention time. If the mean detention time is significantly different from the theoretical detention time, then this measured value should be accounted for in the analysis of the system by adjusting the effective volume.

The procedure for determining N_p was discussed earlier. Samplings should be conducted under very high dose levels in the pilot units (high exposure times); the premise is that the residual density measured after such a high apparent dose is attributable to the bacteria being protected by occlusion in the suspended solids. Linear regression analyses of the log effluent residual coliform density as a function of the log of the effluent suspended solids will yield the coefficients c (intercept) and m (slope). It is important to evaluate the units in this fashion over a significant range of suspended solids concentrations. It may be necessary to artificially adjust the wastewater to accommodate this requirement.

Influent and effluent data should be collected at flow rates which are high, yielding low "apparent" doses in order to determine the rate of inactivation; this requires that the exposure time be sufficiently low to allow significant bacterial density levels in the effluent. The rate, K, is then determined by solution of the model expression. A linear regression analysis of the log K as a function of the log intensity will then yield the intercept and slope, which are the coefficients a and b, respectively.

Testing can also be conducted on a bench-scale basis to determine the UV coefficients. Although procedures have not been reported in the literature, simple batch test methods can be used. This, in turn, would greatly simplify the testing requirements described above. Care should still be taken, however, in devising the experimental apparatus. In order to simulate the high intensity levels experienced in the multi-lamps full-scale units, the bench-scale batch units should use several lamps to attain higher intensities. The collimated beam apparatus in Figure 7-21 would not be adequate as shown; intensities generally do not exceed 1,000 $\mu W/cm^2$, even with modifications of the arrangement.

7.4.3 Assumptions for the Design Example

Given the protocol presented in the earlier discussions, the disinfection model can be used to determine the optimum design for a given application. The model approach allows the testing of several design scenarios and any number of unit configurations. At minimum, it allows the designer to evaluate directly the systems proposed by manufacturers.

The primary design objective (and operating goal) is to maximize the loading to the system while still meeting performance goals. For the UV disinfection process, this UV loading is defined as the ratio of the flow, Q, to the nominal UV wattage (at 253.7 nm) of the reactor, W_n:

$$\text{UV Loading} = Q/W_n = \text{Lpm/UV Watt, nominal}$$

Additionally, we will define the performance of a reactor as the log of the survival ratio, Log N/N_o. Thus, our goal is to design a system which can handle the maximum loading, Q/W_n, and meet the desired Log N/N_o.

At the start, let us make the following assumptions. The model coefficients, based on direct testing, are:

$a = 1.45 \times 10^{-5}$
$b = 1.3$
$c = 0.25$
$m = 2.0$

Further, the assumptions we will make regarding the reactor are:

a. Quartz system with a uniform lamp array. This was described in section 7.3.5.3 and schematically presented on Figure 7-25(a).

b. The centerline spacing will be 6.0 cm. The average nominal intensity is presented on Figure 7-28 as a function of the UV absorbance coefficient.

c. The lamps will be G64T5, or equivalent (see Table 7-7). The lamps will have insignificant transmission at 185 nm, in order to minimize the production of ozone. This ozone is generated in the air gap between the lamp and the quartz sleeve; the ozone absorbs energy at the 253.7 nm wavelength, resulting in attenuation of the UV energy before it can reach the liquid.

d. The lamps will be 1.6-m long with an effective arc length of 1.47 m; the nominal UV output is approximately 18.2 W/m arc.

e. Each lamp is sheathed in a quartz enclosure with an outer diameter of 2.3 cm.

f. The lamps will be configured axially parallel to one another; the flow path will be perpendicular to the lamps.

g. The values of the energy loss factors, F_p and F_t (see Sections 7.3.3.4 and 7.5.1) are set at 0.8 and 0.7, respectively.

h. The maximum allowable headloss through the lamp battery is set at 40 cm. This is exclusive of the entrance and exit losses for the reactor.

Regarding the wastewater characteristics, several adjustments are made to the Table 7-11 parameters to reflect diurnal and maximum average design conditions.

For convenience, the problem is restated. The treatment plant is a conventional activated sludge facility with the following effluent:

Maximum 30-day average BOD_5 = 15 mg/l
Maximum 30-day average SS = 15 mg/l
Maximum 7-day average BOD_5 = 30 mg/l
Maximum 7-day average SS = 30 mg/l
Maximum 30-day average
fecal coliform = 200 org/100 ml (GM)
Maximum 7-day average
fecal coliform = 400 org/100 ml (GM)

The anticipated design hydraulic capacity of the plant, will be:

Average daily flow (dry weather) =
7.5 mgd (28,000 Lpm)
Peak daily flow = 15.0 mgd (56,000 Lpm)

These flows are anticipated at 5 years; the average daily flow is expected to be 3.5 mgd at startup. The relevant wastewater characteristics are:

Average daily fecal coliform = 5×10^5 org/100 ml
UV transmittance (% at 253.7 nm)
Daily average = 70 percent
Minimum 30-day average = 65 percent
Minimum 7-day average = 60 percent

Other wastewater quality characteristics relate to the variability of the parameters; these can be established by the analysis of data collected over an extended period of time:

Ratio of maximum 7-day average flow/average daily flow = 1.25
Ratio of maximum 30-day average flow/average daily flow = 1.1

Ratio of maximum hourly flow/average 24-hour flow = 1.3

7.4.4 Design Sequence

The following steps will comprise the sequence of calculations for the design example:

1. determine UV density, D,
2. establish intensity as function of D; adjusted for loss factors,
3. establish inactivation rate, K,
4. set hydraulic parameters to accommodate dispersion and headloss limitations,
5. establish UV loading-performance relationship,
6. establish performance goals, and
7. reactor sizing.

7.4.5 Design Example

The following calculations demonstrate the procedure for sizing the UV system. The wastewater characteristics have been given on Table 7-11; the design criteria and assumptions are discussed in Section 7.4.3 and 7.4.4.

Step 1—Reactor UV Density

The liquid volume per lamp (see Figure 7.25(a)) is computed:

$$V_v/\text{Lamp} = (S^2z) - [(\pi d_q^2/4)z] \qquad (7\text{-}38)$$

where S is the centerline spacing (cm), z is the lamp arc length (cm), and d_q is the diameter of the quartz sleeve (cm). For the uniform array, with

S = 6.0 cm
z = 147 cm
d_q = 2.3 cm

$$V_v/\text{lamp} = [(6.0)^2(147)]-[\pi(2.3)^2\,147/4]$$

$$V_v/\text{lamp} = 4700\ \text{cm}^3\ (4.7\ \text{liters})$$

The UV density, D, is calculated from Equation (7-29):

$$D = (1.47\ \text{m arc} \times 18.2\ \text{W/m arc})/4.7\ \text{liters}$$

$$D = 5.7\ \text{W/liter}$$

Step 2—Intensity

The nominal average intensity can then be estimated from Figure 7-28 (uniform array) for this density and the anticipated wastewater conditions, i.e., absorbance coefficients. These are summarized on Table 7-12. The percent transmittance is first converted to the UV absorbance coefficient, α(base e). The nominal average intensity is then determined from Figure 7-28. Note that the absorbance coefficient information has been assumed to be derived from measurements corrected for scattering (α_s-Total).

This nominal average intensity must then be adjusted to account for the anticipated average lamp output in the reactor and the minimum average transmittance of the quartz sleeves. Recalling $F_p = 0.8$ and $F_t = 0.7$:

$$I_{avg} = \text{Nominal } I_{avg} \times 0.8 \times 0.7$$

The adjusted I_{avg} values are given in Table 7-12 to reflect these conditions.

Step 3—Inactivation Rates

Once the adjusted intensity values are determined, the inactivation rates can be estimated from Equation 7-12, with the coefficients a and b determined earlier. The rates are given on Table 7-12 for each of the design conditions.

Table 7-12. Estimate of Intensity and Rate, K for Design Example

Design Condition	UV Transmittance at 253.7 nm (%)	UV Absorbance Coefficient (cm^{-1})	Nominal[a] Average Intensity ($\mu W/cm^2$)	Adjusted[b] Average Intensity ($\mu W/cm^2$)	Inactivation[c] Rate (sec^{-1}) K
Average Daily	70	0.35	17300	9700	2.21
Maximum 30-Day Average	65	0.43	15100	8450	1.85
Maximum 7-day Average	60	0.51	13000	7300	1.53

[a] From Figure 7-28 for a density of 5.7 W/L
[b] Assumes an F_p = 0.8 and an F_t = 0.7
[c] a = 0.0000145; b = 1.3, Equation 7-12.

Step 4—Set Hydraulic Rates

Recalling the discussions on hydraulics (Section 7.3.2), the maximum performance would ideally be accomplished in a perfect plug flow reactor. Since there will always be the non-ideal case, the goal is to design a plug flow reactor with a low dispersion number, d. As was also discussed, this dispersion number must be reconciled with the headloss incurred by forcing the d to be low.

Scheible et al. (54) had estimated the c_f in Equation 7-23 for headloss to be between 0.00017 and 0.00023 sec^2/cm^2, based on direct head loss measurements for quartz units with a flowpath perpendicular to the lamps. This estimate would apply only to this type of configuration. There is little direct data available to determine c_f for a variety of configurations. It is very important that these data be generated; the designer should at least specify this in the equipment specs.

It is suggested that a conservative value of c_f = 0.00025 sec^2/cm^2 can be used in estimating headloss in most quartz reactors where the flow path is perpendicular to the lamps. One should understand that this applies only to the lamp battery itself and does not take into account losses from pipe inlets, stilling walls, etc.

The example plant is to operate by gravity with the maximum allowable headloss due to the lamp battery set at 40 cm. We can use this to set a practical design goal with regard to the dispersion number. From Equation 7-23:

$$h_L = c_f (x) (u)^2$$

$$40 \text{ cm} = 0.00025 (x) (u)^2$$

this implies that:

$$x \cong 160{,}000 (u)^{-2} \qquad (7\text{-}42)$$

We have defined d as:

$$d = \frac{E}{ux}$$

Consider the product, ux. We can estimate u and x over a range of ux values which will keep the h_L below 40 cm. As an example, at:

$$ux = 10{,}000 \text{ cm}^2/\text{sec}$$

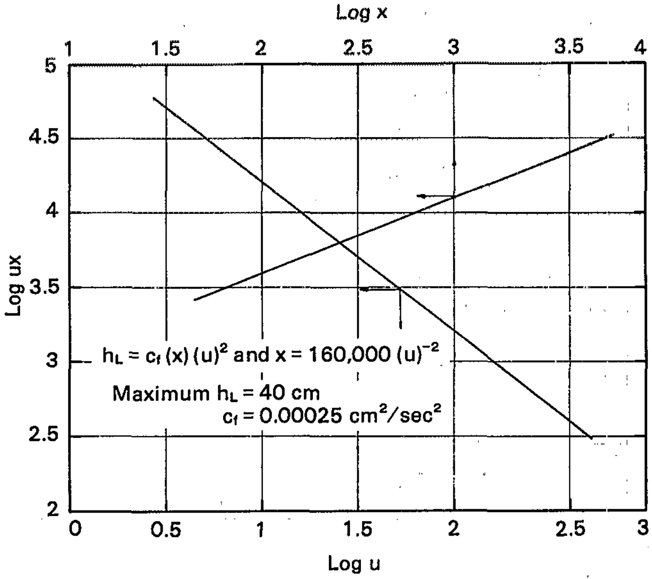
substitute Equation 7-42:

$$(u)^{-1} = \frac{10000}{160000} = 0.0625$$

$$\text{or } u \cong 16 \text{ cm/sec}$$
$$\text{and } x \cong 10{,}000/16 = 625 \text{ cm}$$

Figure 7-41 presents similar solutions; these are all to yield an $h_L < 40$ cm for the example system.

Figure 7-41. Example of calculating the limiting U and X on the basis of head loss (flowpath perpendicular to the lamps).



A dispersion number should then be approximated for design purposes. From earlier discussions, a practical range of d is 0.03 to 0.05. If we consider d = 0.03, this implies a plug flow reactor with low to moderate dispersion. The components of d are the dispersion coefficient E, the velocity u, and the characteristic length, x.

Little direct testing information is available to select a dispersion coefficient as a function of the selected reactor configuration. Procedures have been given to develop the RTD curve for a given unit; the d and E can be estimated from these data. For this reason, it is strongly recommended that it be developed directly, or that the equipment manufacturer supply direct, certified, test data from hydraulically scaleable units.

In the long term, as more information is developed, it may be possible to develop empirical relationships which will give reasonable approximations of E. To date, direct testing on quartz units has yielded an estimated E between 10 and 500, with values typically between 50 and 200 cm^2/sec. Little data is available on the Teflon tube configuration; these indicate a value between 10 and 50 cm^2/sec. Although E will likely vary with flow rate (i.e., velocity), it is sufficient to consider one E; this should be representative of the maximum flow conditions. For purposes of design, one should check that the sizing, based on a selected, d, u and x, implies an E less than 300 cm^2/sec.

Let us select a characteristic length, x, of 200 cm for the UV unit (this can be varied in evaluating alternative design configurations). From Figure 7-41, this implies a ux of 5,600 cm^2/sec and a u of 5,600/200 = 28 cm/sec. Maintaining u below 28 cm/sec will assure an h_L below 40 cm at peak flow.

Recall that we set a d of 0.03 as a design guideline:

$$d = 0.03 = \frac{E}{ux} = \frac{E}{5600}$$

This implies an E of approximately 170 cm^2/sec, which falls within acceptable limits.

In summary, as the first approximation of the system design, we are setting:

d = 0.03
x = 200 cm
u = 28 cm/sec
E = 170 cm^2/sec

Step 5—UV Loading

Recall the disinfection model (Equation 7-10):

$$\frac{N}{N_o} = \exp\left[\frac{ux}{2E}\left\{1 - \left(1 + \frac{4EK}{u^2}\right)^{1/2}\right\}\right] + N_p$$

The next step in the design procedure is to develop the relationship of the reactor performance (Log N/N_o) as a function of the maximum UV loading (Q/W_n), based on the above relationship, and the assumptions discussed earlier.

Before this, one should understand that for a given loading, there is an equivalent nominal exposure time, t_n:

$$t_n = \left(\frac{V_v}{W_n}\right) / \left(\frac{Q}{W_n}\right)$$

At the 6-cm spacing for the uniform array the liquid volume, V_v, associated with the lamp and quartz sleeve is 4.7 liters. Thus, for the reactor configuration we are considering for this example:

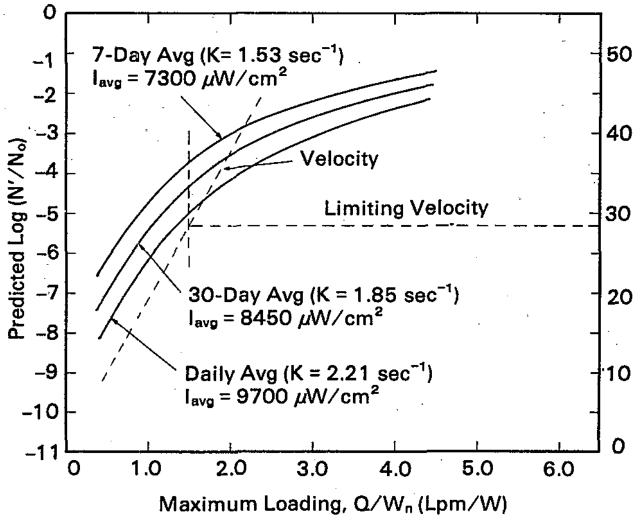
$$\frac{V_v}{W_n} = \frac{4.7\ L}{1.47\ m\ arc \times 18.2\ W_n/m\ arc} = 0.176$$

and

$$t_n = \frac{(0.176)}{(Q/W_n)}$$

Calculations are summarized on Table 7-13 for the design conditions. The log (N'/N_o) values are plotted on Figure 7-42 as a function of the Q/W_n. Understand that the performance is based on the non-particulate effluent fecal coliform density, N'. The particulate fecal coliform density, N_p, is additive. The velocity is also plotted as a function of the loading; the limiting velocity is based on the maximum headloss.

Figure 7-42. **Predicted performance as a function of loading for design example.**



Uniform Array, 6.0 cm ℄ Spacing

Solutions at:

d = 0.3
x = 200 cm
E = 170 cm^2/sec
D = 5.7 W/L
$F_p = 0.8$
$F_t = 0.7$

Table 7-13. Calculations of Performance on the Basis of Loading for the Design Example

UV Loading, Q/W_n (1 pm/W_n)	Nominal[a] Exposure Time, t_n (seconds)	Characteristic Length, x (cm)	Velocity[b] u (cm/sec)	Log N'/N_o[c] Daily Average $k = 2.21\ sec^{-1}$	Maximum 30-Day Average $K = 1.85\ sec^{-1}$	Maximum 7-Day Average $K = 1.53\ sec^{-1}$
0.5	21.1	200	9.5	-7.8	-7.0	-6.2
1.0	10.6	200	18.9	-6.2	-5.4	-4.8
1.5	7.04	200	28.4	-5.0	-4.4	-3.7
2.0	5.38	200	37.2	-4.2	-3.6	-3.1
3.0	3.52	200	56.8	-3.1	-2.6	-2.2
4.0	2.64	200	75.8	-2.4	-2.0	-1.7

[a] $t_n = [(0.176)/(Q/W_n)] \times 60$ sec/min

[b] $u = x/t_n$

[c] $\frac{N'}{N_o} = \exp[\frac{ux}{2E}\{1 - (1 + \frac{4EK}{u^2})^{1/2}\}]$ where $E = 170\ cm^2/sec$, $x = 200$ cm, and $N' = N - N_p$ (see Equation 7-34)

Step 6—Establish Performance Goals

Returning to the plant design, it is necessary to determine the performance goal for the system design conditions. These calculations are summarized on Table 7-14. The design values for N_o are set to reflect the variability of the data base, as given on Table 7-11. Regarding the suspended solids, in order to meet the maximum 30-day permit level, it is necessary to achieve an overall average no more than 50 to 70 percent of the maximum 30-day average. For this example, then, the daily average SS is set at 10 mg/l.

Table 7-14. Estimation of Reactor Performance Requirements for the Design Examples

	Daily	Maximum 7-Day	Maximum 30-Day
Initial Fecal Coliforms Density, N_o (org/100 ml)	500,000	2,000,000	1,000,000
Suspended Solids (mg/l)	10[a]	30	15
Particulate Coliforms Density, N_p (org/100 ml)	25	225	56
Permit Requirement, N (org/100 ml)	200	400	200
Performance Goal, N' (org/100 ml)	175	175	144
Log (N'/N_o)	-3.45	-4.05	-3.84

[a] Assumed value; acknowledges that the verage daily is generally 50 to 70 percent of the maximum 30-day average

The values of N_p are calculated on the basis of Equation 7-13, with the coefficients c and m equal to 0.25 and 2.0, respectively. These are then subtracted from the permitted effluent fecal coliform densities to yield the design performance goal. This can then be used to compute the design Log (N'/N_o).

Step 7—Reactor Sizing

Table 7-15 presents a summary of the reactor sizing calculations for the design examples. The I_{avg} and performance goals are restated. The maximum allowable loadings are then determined from the performance curves on Figure 7-42.

The lamp requirement is estimated on the basis of using 1.47 m arc lamps with a UV output of 26.7 W/Lamp:

$$\text{Number of Lamps} = [(Q)/(Q/W_n)]/26.7\ W_n/\text{Lamp}$$

Table 7-15. Sizing Calculations for the Design Example

	Daily Average	*Maximum 7-Day*	*Maximum 30-Day*
Adjusted I_{avg} ($\mu W/cm^2$)	9700	7300	8450
Performance, Log (N'/N_o)	-3.45	4.05	-3.84
Maxmium Q/W_n (lpm/W)[a]	2.63	1.35	1.90
Flow, (lpm)	28000	35000	30800
Adjusted Design Flow (lpm)[b]	36400	45500	40000
Peak Dry Weather Flow (lpm)	57000		
Peak Wet Weather Flow (lpm)	76000		
Nominal Exposure Time, t_n (seconds)[c]	4.0	7.8	5.6
Characteristic Length, x^n (cm)	200	200	200
Nominal Velocity, u (cm/sec)[d]	50.0	25.6	35.7
Lamp Requirement	518	1262	788
at peak dry weather	811		

[a] *From Figure 7-42*

[b] Set flows to peak diurnal conditions; adjusted daily = 1.3 x daily; adjusted 7-day = 1.25 x adjusted daily; adjusted 30-day = 1.1 x adjusted daily.

[c] $t_n = (0.176)/(Q/W_n) \times 60$ sec/min

[d] $u = x/t_n$

From Table 7-15, the maximum design requirements are the 7-day maximum and the peak wet weather conditions. The minimum number of lamps required is appproximately 1,300.

Table 7-16. Reactor Sizing Requirement for the Design Example

	Daily Average	7-Day Maximum	30-Day Maximum
Required Number of Lamps (see Table 7-15)	518	1262	788
	Reactor Sizing Requirement— Perpendicular Flow Path		
Length (x) cm	200 (34 lamps)	200 (34 lamps)	200 (34 lamps)
Height (y) cm	96 (16 lamps)	222 (37 lamps)	138 (23 lamps)
Width (z) cm	150 (1 lamp)	150 (1 lamp)	150 (1 lamp)
Total Lamps	544	1258	782

Once the number of lamps is determined, the actual modular configuration of the UV system can be considered. Table 7-16 summarizes the combined reactor dimensions which will satisfy the critical design elements of the Q/W_n (i.e., number of lamps) and the characteristic length, x. This is to form a uniform array reactor in which the flowpath is perpendicular to the lamps.

Recall that, the x dimension is set by dividing the required x by the centerline spacing S, to determine the number of lamps in the x direction. Thus, for the design example, we can determine the number of lamps in x:

$$200\ \text{cm}/6\ \text{cm} = 33$$

As shown on Table 7-16, this is set at 34 for a total x dimension of 204 cm. The width of the reactor is set by the length (i.e., effective arc length) of the lamp. This is approximately 1.5 m. The height then, is set to satisfy the total lamp requirement: For the average daily condition:

No. of Lamps in y = 518/34 = 15.2 or 16 lamps

This yields a total y dimension of 96 cm.

The maximum requirement is set by the 7-day condition. A possible arrangement of UV modules would be as follows:

4 modules; 408 lamps/module (Total 1500 lamps)
Each 34 lamps long (2.0 m)
12 lamps high (0.72 m)
1 lamp length wide (1.5 m)

The average condition can be met by 2 units; three modules would be required under the peak condition. The fourth unit would be solely for standby.

The reactor modules can be installed in any number of configurations. The critical consideration is the design of the approach and exit portions of the system. This was discussed in Section 7.3.2.4. In both parallel and perpendicular flow path cases, the key is to simulate open channel flow with a constant velocity profile across the cross-sectional plane on both the inlet and outlet sector of the lamp battery (see Figure 7-20). Weirs and stilling walls should be used on both ends of the reactor to effect this even flow distribution. Other design considerations regarding system layout and facilities requirements are discussed in Section 7.5.

7.4.4 Summary

The preceding calculations are given to demonstrate the design protocol. The numbers used should not be used for an actual design application. The procedure, once the model is calibrated, can be very effective. Several alternative configurations can be evaluated and, most importantly, the sensitivity to the key design parameters can be assessed. This applies particularly to the hydraulic parameters of velocity, dispersion, and head loss.

The calculations also demonstrate the importance of having direct test information on several aspects of the design:

Wastewater:	flow and flow variability UV absorbance initial coliforms suspended solids coefficients a, b, c, and m
Reactor Characteristics:	RTD curve (range of flows) head loss (lamp battery only) dispersion coefficient

The reader is referred to Section 7.5 for discussions of other design aspects relating to O&M and to facilities requirements.

7.5 System Design and Operational and Maintenance Considerations for the UV Process

Sections 7.3 and 7.4 present the protocols by which a UV disinfection system can be designed and evaluated. The discussions centered on the design basis and the process elements which are key to the design. These were the hydraulic behavior of the unit, the calculation of the intensity in the reactor, and the generation of the appropriate wastewater characterization data. Finally, an example was presented to illustrate the design procedure.

This section presents other peripheral topics which the designer (and operator) must consider. These address system design elements which will affect the operation and maintenance of the system, and the overall economics. Specifically, the following items are presented:

First, the factors that affect the reactor intensity will be addressed. These relate primarily to the lamp output and lamp aging, and the attenuation of intensity due to fouling of surfaces in the reactor.

Second, methods are presented to monitor a system directly for lamp aging and unit fouling.

Third, design considerations are presented which will encourage effective maintenance, and assure the disinfection performance of the unit.

Fourth, a brief discussion will be given in regard to the major reactor components and system controls.

Fifth, the safety aspects of the system will be discussed.

Sixth, and finally, design considerations will be presented as they relate to ancillary facilities, materials, labor requirements, and system layout.

7.5.1 Factors Affecting UV Intensity in a Reactor

A critical element in the evaluation of a system's performance or in the design of a system is the actual energy available in the germicidal range. The key is to understand how efficiently the 253.7 nm energy is being utilized and, conversely, how it is being lost. This information is necessary to make the required adjustments to the calculated nominal average intensity solutions presented in Figures 7-28 through 7-32, and quantified in Equation 7-30 as the factors F_p and F_t.

The reduction in available energy can be divided into two major areas: the loss of lamp output (F_p) and the change in transmittance of the enclosures separating the lamp from the liquid (F_t). These enclosures are typically the quartz sleeves or the Teflon tubes.

7.5.1.1 UV Lamp Output

Electrical discharge lamps generate light by transforming electrical energy into the kinetic energy of moving electrons, which is then converted to radiation by some kind of collision process. Mercury vapor kept at an optimum pressure in the presence of a rare-gas (generally this is argon—which is the reason for the blue-green glow seen with germicidal lamps), is a remarkably efficient emitter of light at 253.7 nm, when an electrical potential difference is applied across the device. The basic process as it takes place in the discharge lamp can be described in three steps:

1. Free electrons are accelerated by a potential difference applied across the lamp. This voltage drop is maintained by an external source of power, the current being the motion of the electrons.
2. The kinetic energy of the accelerated electrons is released as they collide with atoms in the plasma.
3. The energy of the atoms is dissipated as radiation as the atoms relax back to their lowest energy levels.

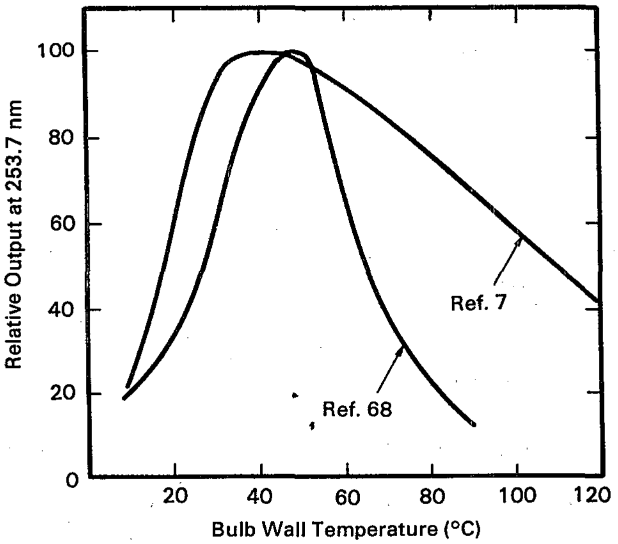
The lower the vapor pressure of mercury in an electric discharge, the greater the intensity of the mercury resonance line at 253.7 nm. Exploiting this fact, construction of the low-pressure mercury arc lamp yields a lamp which is nearly monochromatic in its radiation. The greatest output occurs under conditions which favor having the highest numbers of "excited" atoms close to the wall of the lamp. In this way, the radiation they emit as they relax will have a higher probability of passing through the wall, and not being re-absorbed by atoms in the low energy state. This is a reason for the thin tubular design of the lamps, the smaller the pathlength (diameter of the tube) the less likely the radiant energy will be re-absorbed. In the same fashion, the lower the pressure, the greater the chance of the radiation reaching the wall before making a collision. Under higher pressures the concentration of the atoms in the excited resonance state is high, but because of this higher density, the chances of the radiation reaching the bulb wall is small. Conversely, at extremely low pressures the number of available atoms becomes too small.

The resonance output of the lamp depends to a large extent on maintaining the optimum conditions of the discharge requirements in the vapor, and on the operating conditions during the life of the lamp. In large part, many of the factors which establish the output are independent of the designer or operator and will simply be set by the environmental conditions the unit must operate under. Still it is important to understand the major factors which come in to play.

Output at any given time will be influenced by lamp temperature and by the voltage potential applied across the lamp. Additionally, output at the resonant frequency will always degrade with time of operation due to any number of "aging" factors.

Temperature. As was discussed, there is a maximum resonance output which is dependent on the vapor pressure in the lamp. For a given lamp, this pressure will be influenced by the temperature of the lamp wall. The optimum wall temperature for maximum efficiency is generally between 35 and 50°C (95 and 122°F). Figure 7-43 presents data relating the relative UV output (at 253.7 nm) to the bulb wall temperature. In the quartz systems which are submerged in the flowing liquid, the lamps are insulated by inserting them in the quartz sheaths. The air layer between the quartz and the lamp wall serves as a buffer and prevents the lamps from being cooled by the wastewater. There is little information available regarding actual lamp skin temperatures during normal wastewater disinfection operations. Scheible and Bassell (36) reported that cold water temperatures had little effect on the measured bulb wall temperature. In the study at Northwest Bergen, the lamp temperature averaged 43°C (110°F) at an average water temperature of 21.3°C (70.3°F). At water temperatures averaging 10.5°C (50.9°F), the lamp wall temperature decreased to an average of 40°C.

Figure 7-43. **Effect of bulb wall temperature on the UV output of a low pressure mercury arc lamp (7,68).**



In the submerged systems it is not practical under most design conditions to control the lamp temperature. In the non-contact systems, such as the tubular arrays, it is possible to maintain the lamps at their optimum wall temperature by controlling the temperature of the ambient air surrounding the lamps. This is currently being practiced in commercial applications. Heat given off by the lamp ballasts is circulated into the lamp reactor in cases where heat is required; otherwise fans vent the reactor with cooler outside air. These operations are thermostatically controlled.

Voltage. A characteristic of electric discharge arcs is that they have a negative volt-ampere relationship. This means that the voltage decreases with an increase in the current. Such devices are inherently unstable. This instability is counteracted by putting the arc in series with a resistance. This resistance (or reactance in the case of a-c circuits), is called the ballast. This ohmic resistance has a positive volt-ampere characteristic.

Radiance will be a function of the arc current. This fact can be exploited by adjusting the voltage, in order to vary the output of the lamp. Decreasing the voltage will result in a decrease in the current. Such a control mechanism has been installed at full scale facilities as an energy conservation measure. During periods of low UV demand, the lamps are "dimmed" by turning the lamp supply voltage down. This results in a reduction in the power draw of the lamp. Generally, the lamp intensity can be reduced to levels no less than 50 percent before the lamp current becomes too low and the lamp will begin to flicker and eventually turn off.

Lamp Aging. A number of factors combine to effectively age a lamp and limit its useful life. These include failure of the electrodes, plating of the mercury to the interior lamp wall (blackening), and solarization of the lamp enclosure material (reducing its transmissibility). These all cause a steady deterioration in the lamp's output at the 253.7 nm wavelength, such that its output at the end of the lamp's life can be 40 to 60 percent of its nominal output.

The output of the lamp through its life is affected primarily by the extent of blackening and solarization of the glass tube; the actual life of the lamp is governed by the condition of the electrodes. The germicidal lamps are typically of the hot cathode type. These will progressively deteriorate with increasing number of starts. Thus, the lamp life expectancy is generally rated according to the number of times the lamp is started, or the burning cycle. The lamp life normally cited by most manufacturers is 7,500 hours, based on a burning cycle of eight hours. The average UV output at this point is estimated to be 70 percent of the lamps output at 100 hours (note that the nominal output of the low-pressure mercury arc lamps represents its output after a 100 hour "burn-in" period).

An unusually long lamp life has been demonstrated by the units at Tillsonburg, Ontario. Lamps have been in operation for a documented period of greater than 13,000 hours. This may be due to the fact that the entire lamp, including the electrode connections, are submerged and thus cooled by the flowing water (see Figure 7-7). The cooling of the electrodes may be a factor in extending the life of the electrode. Conversely, excessive blackening and deterioration of the end of the lamps has been noted for systems in which the lamps are inserted through a metallic bulkhead. There is no cooling in this case and, in fact, this may cause a buildup of heat.

Measurement of the UV Output of the Lamp. Table 7-7 presented the specifications for a number of slimline type germicidal lamps. Each is specified with a rated UV output at 253.7 nm, expressed in Watts, and the arc length, which is the length of the radiating portion of the lamp column. Figure 7-44 presents the output of the 1.5-cm diameter lamps as a function of the arc length. The slope of this is 18.2 UV W/m of arc. This will be slightly lower for lamps with a diameter of 1.9 cm. One should understand that this is the nominal output of the lamp; as discussed, this will decrease with time.

Figure 7-44. **Nominal lamp output as a function of arc length.**

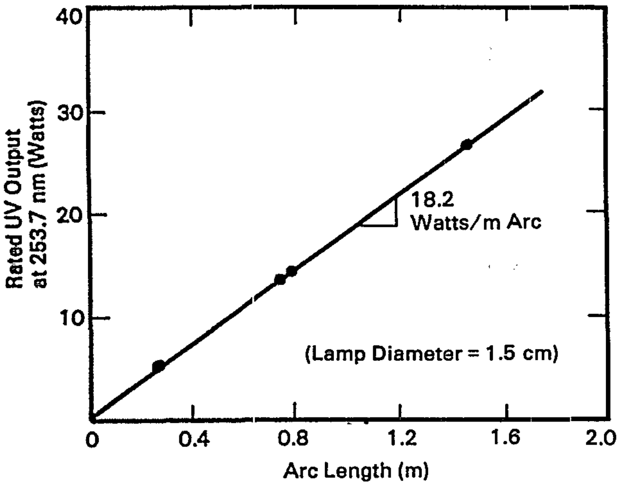
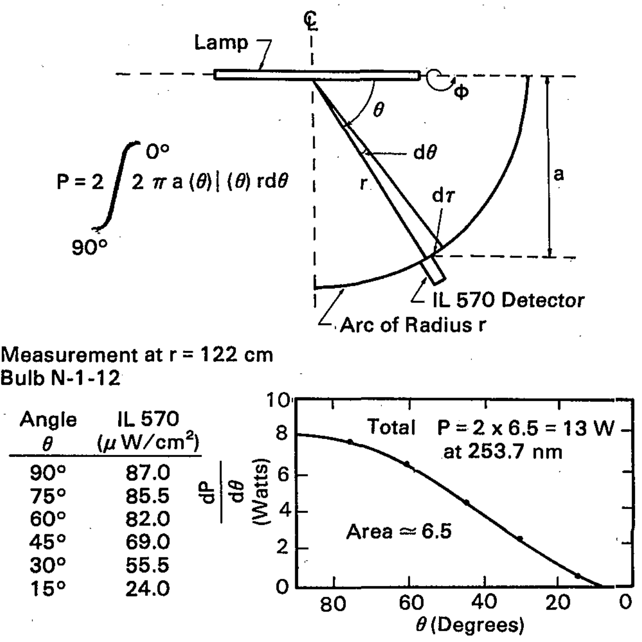


Figure 7-45. **Measurement and analysis technique for estimating the total UV output of a lamp.**



$$P = 2\int_{90°}^{0°} 2\pi a(\theta) I(\theta) r d\theta$$

Measurement at r = 122 cm
Bulb N-1-12

Angle θ	IL 570 (μW/cm^2)
90°	87.0
75°	85.5
60°	82.0
45°	69.0
30°	55.5
15°	24.0

A test procedure for determining a lamp's UV output was described by Johnson and Qualls (38). The procedure, in effect, treats the lamp as a single point source of light. A UV detector measures the intensity along an arc scribed at a fixed radius from the center of the lamp. These intensity readings are integrated over the surface of a sphere with the same radius, resulting in an estimate of the total output from the lamp.

Figure 7-45 is a schematic representation of the experimental set-up to measure the total output and a sample analysis of a set of data. The expression to compute the power, P, at any point at distance r from the center of the lamp, at angle θ, is written:

$$P = 2 \int_{90°}^{0°} 2\pi a(\theta) I(\theta) r d\theta \qquad (7\text{-}46)$$

where:

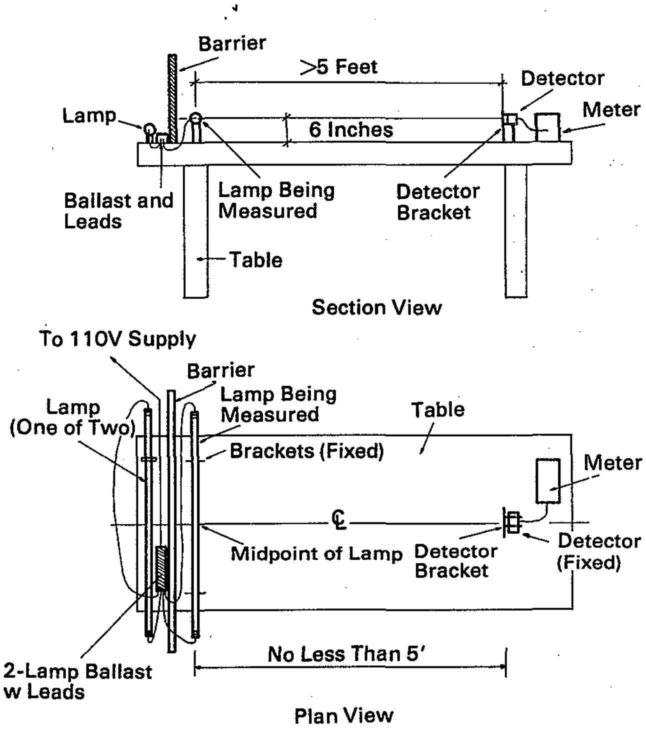
P = the total power at 253.7 nm (Watts)
θ = angle of the detector from the longitudinal centerline of the lamp (0 to 90 degrees)
I = intensity reading at radius r at a point (T) along the arc
r = radius of arc (distance from center of lamp to detectors) (cm)
a = distance from detector to axial line of lamp (cm)

This can be solved graphically by plotting $dP/d\theta$ against θ, as shown by the example on Figure 7-45. The area under the curve is power output at the wavelength 253.7 nm, and represents one-half of the sphere about the lamp. Doubling this power estimate yields the total output of the lamp.

In the Port Richmond study, a total of 9 lamps were measured by this method after 100 to 300 hours operation. The measured output ranged from 12.2 to 15.9 W, with an average output of 13.6 W; this is close to the rated output of 14.3 W given by the manufacturer.

A far simpler technique can be used to implicitly monitor the average lamp output with time. The setup with the lamp and the detector is the same as shown on Figure 7-44. In this case, however, a single intensity reading can be taken at θ equal to 90 degrees and compared to the same intensity reading for a new lamp. Figure 7-46 is a sketch of a simple table-mounted unit which can be used at a full-scale facility.

Figure 7-46. Sketch of lamp monitoring set-up.



Two lamps are operated off a single ballast. The lamp to be measured would be placed on brackets which are in a fixed position from the detector. The other lamp should be operating but placed behind a barrier to prevent it from interfering with the lamp being measured. Between three and five minutes warm-up time should be allowed before a measurement is taken. The leads from the ballast should be wired to end caps which can then be quick fitted onto the lamps. The lamps should not be repositioned by disconnecting them; they should be moved with their leads in-place.

As shown, the setup can be placed on a table-top; the lamp brackets should be thin and should not shield light. The lamp which is not being measured can be shielded by fitting a length of cardboard tubing (or thin opaque plastic tubing) over it. The tube can be slit lengthwise and should fit loosely (to prevent overheating the lamp). Alternatively, as shown on the sketch, a barrier can be set up between the lamps.

The detector is set up at the other end of the table. The mounting bracket should be rigid and fixed. The bracket should allow for removing and inserting the probe without changing the position of the detector in any direction. The objective is to always have the lamp and detector in a fixed position; these positions must also be reproducible from day to day. The detector should be on the same horizontal plane of the lamp centerline and perpendicular to the midpoint of the lamp. The distance between the lamp and the detector should be no less than five feet.

The procedure for monitoring the lamp intensity, once the setup is in place, is rather simple. The idea is to measure a lamp's intensity relative to that of a new lamp. The first step is to measure the intensity, at the fixed distance, of 3 to 5 new lamps which had been burned for approximately 100 hours. The average of the five then becomes the benchmark to determine the relative output of the lamps in use with the system (percent of new lamp average). Each lamp should be tagged and given an I.D. number; this allows direct monitoring of individual lamps and allows the operator to keep an appropriate mix of lamps in a system and to know when to discard a lamp.

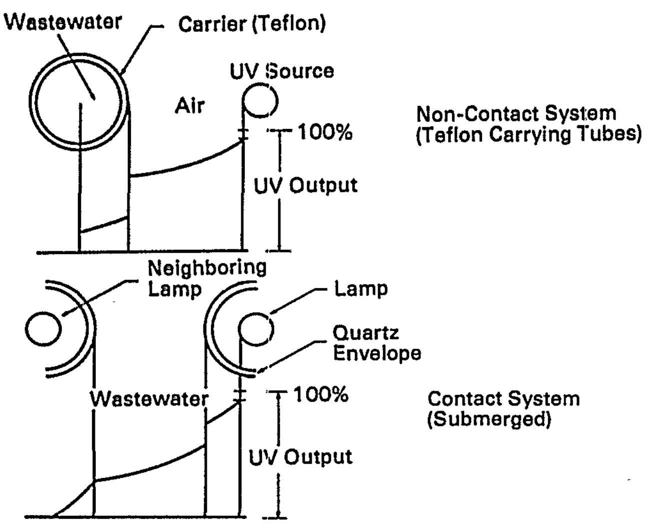
The same procedure is used to monitor the transmittance of a quartz sleeve. In this case a single lamp is used; first the intensity is measured with and without a new, clean quartz sleeve in place over the lamp. Similar measurements are then taken with the unit's quartz sleeves and compared to the transmittance of the new quartz. This can be done before and after the quartz is cleaned.

7.5.1.2 Losses of UV Energy through the Quartz and Teflon Enclosures

The UV output of the lamps themselves can be monitored with time, as discussed above. This is generally a non-controllable parameter, although optimum conditions can be maintained (such as voltage and temperature) to keep the output at its maximum. The lamps will age, however, and lose output. A key consideration regarding the UV source in both system design and subsequent operation and maintenance relate to maximizing the utilization of the source output and understanding the known energy sinks within a given system.

Figure 7-47 schematically presents two typical UV lamp configurations and the known energy sinks. The first is the tubular array; the second is the submerged quartz array. In either system there are several ways the UV energy is lost before it reaches the liquid and can be utilized for its primary germicidal role. First, the lamp wall itself can become dirty. In the non-contact tubular array systems the lamps are in an open air environment within the reactor. The air is often circulated to keep the ballast cool and/or to control the lamp temperature. This can introduce dust

Figure 7-47. **Energy sinks in UV reactor.**



which settles on the lamp surface and becomes an energy absorber. This same problem will also cause the outer surface of the Teflon tubes to become dusty and reduce the Teflon's transmittance. Filters are now installed in such units to minimize these problems.

In the quartz systems, some units are installed which either seal the quartz ends or leave them open. In the open arrangement, convective air currents can carry air (often humid) through the quartz sleeve, causing some deposition on the lamp surface. Additionally, the same air convection will cause the inner surface of the quartz sleeve to become dirty. This may also occur to some degree in sealed systems due to condensation effects, although there is no current information regarding these effects.

Passage through the quartz sleeve or the Teflon tube wall will itself cause a loss of energy. The fused quartz sleeves are highly transmissible of UV at 253.7 nm. The transmittance of the Teflon will vary with the thickness of the tube wall and is typically less then than that of the quartz. Lastly, the surfaces of the quartz sleeves or the Teflon tubes which contact the wastewater will foul and cause the transmissibility of either to be reduced. The O&M tasks will naturally be directed to keeping these surfaces clean and maintaining the maximum transmittance of the quartz or Teflon.

Quartz Systems. The quartz sleeves are typically high quality fused quartz, with a transmittance greater than 90 percent when in a new, clean condition. In the Port Richmond study, relative output readings of a lamp were taken with and without the quartz sleeve in place by the procedure discussed earlier. The average (of 12 measurements) reduction in intensity measured when the quartz was placed over the lamp was approximately 25 percent, with a range of 15 to 35 percent. Obviously, this is significantly different from the 90 percent for a new and clean quartz; the reasons relate to the dirtiness of the surfaces and the loss due to ozone absorption.

The lamps used at Port Richmond were made of quartz, which is transparent to energy at the 185 nm wavelength, a spectral line characteristic of the low pressure mercury arc. Energy at this wavelength will ionize free oxygen to ozone which, in turn, is an excellent absorber of energy at the 253.7 nm wavelength. Thus, with any production of ozone in the gap between the lamp and the quartz sleeve, it is likely that there would be a consequent absorption of UV energy. Direct testing in the Port Richmond project confirmed this effect.

Not all lamps will transmit this energy. In fact, the majority of lamp designs utilize a lamp envelope which has a low transmittance at the 185 nm wavelength. This is shown on Table 7-7. It is recommended that these types of lamps be used for UV disinfection systems.

Table 7-17 is excerpted from the Port Richmond report (54), and shows the effects of fouling on the quartz transmittance. Intensity readings taken immediately after the quartz were removed from the unit averaged 61.8 $\mu W/cm^2$, which was 81.6 percent of the reading obtained with the new quartz sheath. Cleaning the inside surface of the quartz improved the output to an average 70.1 $\mu W/cm^2$, or 92.8 percent of the new quartz reading. Finally, the output increased to 75.3 $\mu W/cm^2$ when the outer surfaces were cleaned, which is essentially equivalent to the new quartz reading.

The inside of the quartz had last been cleaned five months before these tests; the report then ascribed a 15 percent output decrease per five months due to the inside surface fouling. The outside surfaces were cleaned with an acid/detergent solution on a frequent basis, and immediately before these readings were taken. The results indicated that although this procedure was effective, there is apparently a film layer which stays on the surface; this was presumed to cause a loss of approximately eight percent of the UV output at any time.

The Port Richmond report presented a summary analysis of the lamp output and quartz transmittance monitoring conducted during the term of the project. This is repeated herein as Figure 7-48, and provides an excellent example of the importance of both monitoring these conditions and accounting for them in the design of a system. The procedures are relatively straightforward and require little expense in terms of monitoring equipment. The primary commitment is the labor requirement for taking the necessary readings.

The manufacturer's rating for these lamps was confirmed by direct measurement of their UV output; this was shown to degrade to approximately 60 percent of this output after 8300 hours. A 25 percent reduction is taken to account for the quartz sheath absorbance, and losses attributable to ozone absorbance within the air gap between the lamp and the quartz sheath. If low or zero ozone producing lamps are used, this reduction will be approximately 10 percent. A constant eight percent loss is taken to account for the film layer on the outer surface of the quartz sheaths. This was considered the base line loss as discussed earlier; with increasing time between chemical cleaning cycles, this loss will increase. If not attended to (cleaned) this surface fouling can cause significant deterioration in the system performance. The internal fouling, although not as significant as the outside surface, will still have an effect. It should be a requirement, particularly in systems where the quartz ends are open, to clean these surfaces at least once to twice per year.

The lower line on Figure 7-48, therefore, is an estimate of the actual average UV output for the lamps at any time during the operating period for the Port Richmond study. It is significant to note that although a system may start with a nominal output from a lamp source, this is immediately reduced to approximately 70 percent of nominal simply by its placement in the quartz sheath and by the development of a film layer with time. Over the operating age of the lamp, this output, even with good maintenance, will deteriorate to approximately 35 to 45 percent of this nominal output after one year. Thus, it is important to maintain the system and keep this output at its maximum.

Teflon Tubular Systems. A number of tests were also conducted during the Port Richmond study on a sample of the Teflon tubes used in Unit 3 of the project. Lamp intensity readings were taken with and without clean Teflon in place; these readings also evaluated the transmittance with the lamp on both the convex and concave side of the Teflon. The results are presented on Figure 7-49, excerpted from the Port Richmond report. They are not wholly conclusive; there appears to be some apparent effect due to the curvature of the Teflon, with the higher transmittance measured when the lamp is on the inside of the Teflon arc.

All of the used samples were vigorously cleaned in the lab with hot soapy water and a soft brush after the transmittance measurements were made. The UV transmittances after this cleaning are also shown in Table 7-17. As can be seen, considerable improvement was obtained. However, none of the cleaned samples demonstrated UV transmittances at the level one would expect for a virgin Teflon sample of the same wall thickness, suggesting that Teflon may undergo a transformation and lose some of its ability to transmit UV light over time. This may be caused by continued exposure to the UV lamps.

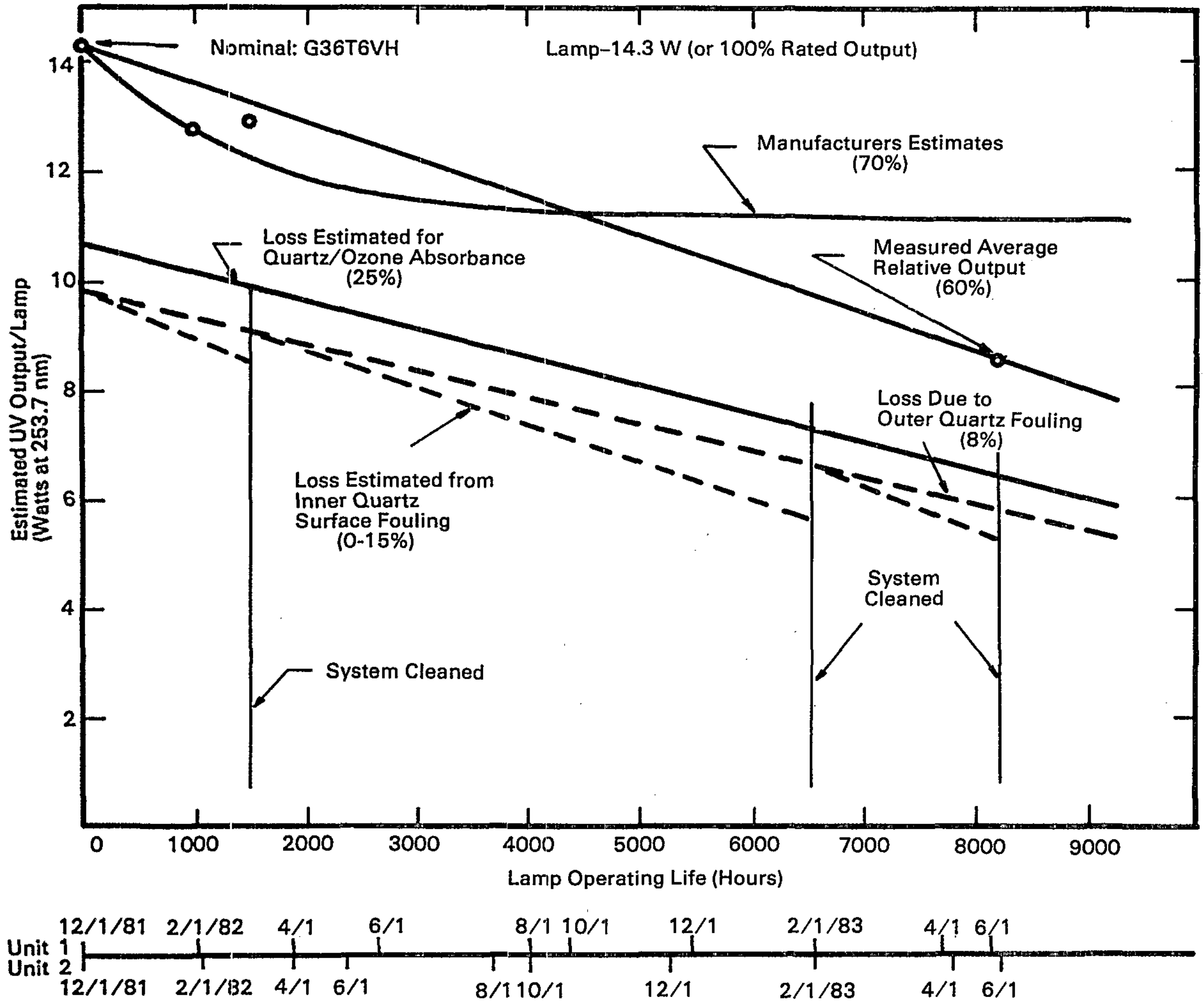
The Teflon used in the Port Richmond tests had been in use for a period of time. The inside surface was observed to be dirty, but was cleaned before the readings were taken. These results compare poorly with the 75 percent transmittance cited by the manufacturer as characteristic of Teflon. In their report, the authors suggested a transmittance level for the Teflon to range between a maximum of 75 percent when new to as low as 30 percent under significant fouling conditions.

A special series of tests were conducted using an alternative and possibly more accurate method to determine the transmittance of the Teflon. These tests (4) involve the exposure of a chemical actinometer to UV light. The compound was 0.006 M potassium ferrioxalate; upon exposure, the ferric ions will be reduced to ferrous ions in proportion to the amount of UV light received by the actinometer. The ferrous concentration is measured spectrophotometrically at 510 nm, using phenanthroline as the color reagent. The chemistry procedures follow those by Baxendale and Bridges (60) and Parker (61).

A sketch of the bench-scale setup is given in Figure 7-50. Note that, although these tests addressed Teflon, the same procedures are applicable also to quartz. Work should be conducted in a darkened room. The ferrioxalate is placed in a 1.3-cm diameter by 8-cm long fused quartz test tube (volume = 6 ml) capped with a non-reactive stopper, and covered with an aluminum foil sheath. The UV lamp is allowed to stabilize for a minimum of five minutes before any test. The test tube containing the actinometer would be clamped in a fixed position from the lamp; exposure would be accomplished by slipping the aluminum foil off the test tube for a preset period of time. The ferrous concentration would then be measured. Exposures are conducted with and without the Teflon tube in place.

The tests are conducted to encompass a range of reaction times. The reaction rate is determined by plotting the ferrous concentration against exposure time. The transmittance of the Teflon is estimated by determining the reduction in the reaction rate of the actinometer. Examples of these results are given in Figure 7-51. Results are given without the Teflon and

Figure 7-48. Approximation of average lamp UV output at 253.7 nm with time for quartz systems, accounting for lamp aging and surface fouling (54).



with different Teflon wall thicknesses. As an example of the calculation, the slope of the line in Figure 7-51 without Teflon is 3.86×10^{-6} M/sec; the slope of the line for Teflon with a wall thickness of 0.81 mm is 3.29×10^{-6} M/sec. The ratio is 0.85, suggesting a transmittance of 85 percent.

Kreft et al. (4) reported results of tests on both new and used Teflon by the actinometry method. Representative virgin Teflon samples of varying diameters and wall thicknesses were obtained from a manufacturer of the ultraviolet equipment. Used Teflon samples were obtained from secondary wastewater treatment plants in Port Richmond, New York; Chinook, Montana; and Beech Mountain, North Carolina. Table 7-18 shows the results of virgin and used Teflon UV transmittance analyses by chemical actinometry. It can be seen that the virgin Teflon transmittance varied with the wall thickness, but not significantly with the diameter of the tube. Figure 7-52 shows the relationship developed from these analyses. Current designs typically have thin walls with a thickness less than 0.9 mm; thus, the virgin Teflon transmissivity for these units can be estimated to be in the range of 80 to 90 percent.

The Teflon samples showed varied degrees of internal fouling. The Port Richmond samples were obtained after a fairly continuous operating period of five months while treating secondary effluent. The Chinook samples were obtained after an operational period of approximately two months. Table 7-18 presents results of the UV transmittance measurements from these used Teflon samples. The Port Richmond samples were heavily fouled with a rust-colored material. UV transmittance for these samples

Table 7-17. **Effects of Fouling on the UV Transmittance of Quartz (54)**

	Intensity Readings					
Tube Number	After Removal from Unit[b] (μWatts/cm^2)	Relative to New Quartz	Cleaned Inside of Quartz (μWatts/cm^2)	Relative to New Quartz	Cleaned Outside of Quartz (μWatts/cm^2)	Relative to New Quartz
4	50.5	66.9%	66.5	88.1%	77.5	102.6%
34	60.5	80.1%	67.5	89.4%	73.5	97.3%
64	63.5	84.1%	70.5	93.4%	74.0	98.1%
74	68.5	90.7%	73.5	97.3%	73.5	97.3%
84	66.5	88.1%	73.5	97.3%	72.5	96.0%
94	60.0	79.5%	69.0	91.4$	81.0	107.3%
Average	61.5	81.6%	70.1	92.8%	75.3	99.8%

[a] Intensity of bulb without quartz is 90.5 μW/cm^2 at 1.2 m. The intensity with a new quartz in place is 75.5 μW/cm^2. This is used as the reference intensity.

[b] Cleaned by acid/detergent solution while in the unit.

Figure 7-49. **Estimate of Teflon transmittance by use of a UV detector (54).**

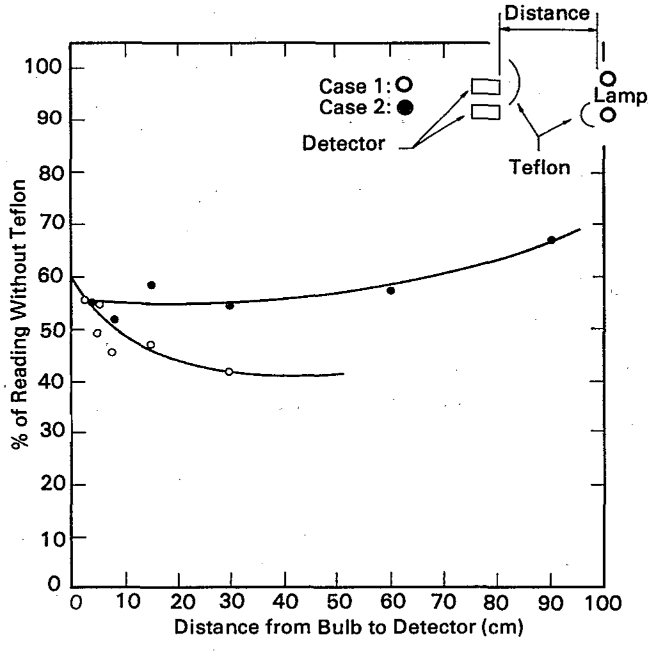


Figure 7-50. **Test set-up to conduct actinometry experiments (4).**

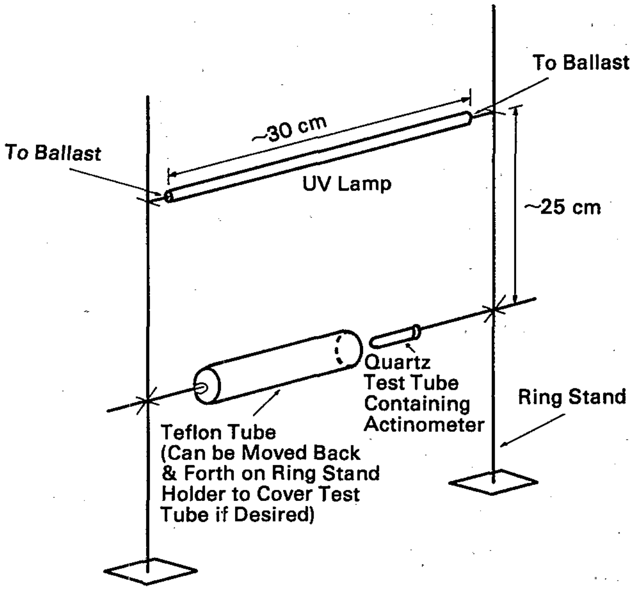


Figure 7-51. Example of chemical actinometry tests to determine Teflon UV transmission (4).

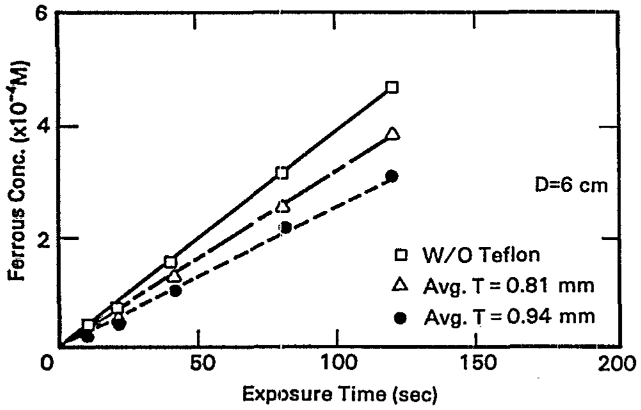


Table 7-18. UV Transmittances of New and Used Teflon as Determined by Chemical Actionmetry

Teflon Tube I.D.	% UV Transmittance As Received	% UV Transmittance After Lab Cleaning
New, D=6 cm, T=0.81 mm	85%	--
New, D=6 cm, T=0.94 mm	72%	--
New, D=8.9 cm, T=0.46 mm	92%	--
New, D=8.9 cm, T=0.84 mm	85%	--
Port Richmond #1[a]	7	68
Port Richmond #2	9	71
Port Richmond #3	5	72
Port Richmond #4	18	64
Port Richmond #5	30	76
Chinook—3800 hours[b]	30	75
Chinook—3800 hours (Cleaned in field with water and vinegar)	66	78
Chinook—100 hours	55	80
Chinook—100 hours (cleaned in field with water and vinegar)	70	85
Beech Mountain #1[c]	33	74
Beech Mountain #2	47	78

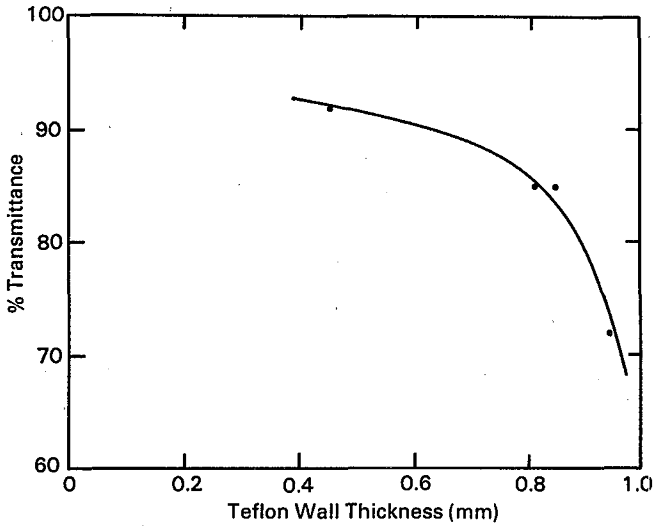
[a] Port Richmond tubes had D=3.5" and T=0.030".

[b] Chinook tubes had D=3.5" and T=0.032". One half of the tube had been exposed to UV light for 3800 hours, the other half for 100 hours. Each of those halves was cut into sections; one section was cleaned with a high pressure nozzle washer with water and vinegar, the other not.

[c] Beech Mountain tubes had D=2.375" and T=0.033".

Note: D = inside diameters, T = wall thickness

Figure 7-52. Effect of wall thickness on Teflon transmittance, as determined by chemical actinometry (4).



ranged from 5 to 30 percent. The Chinook samples were not as obviously fouled, but they did have noticeable whitish precipitate deposits and some scum and grease attached to the Teflon. Their UV transmittance were in the range of 50 to 60 percent. Beech Mountain samples also demonstrated UV transmittances in the 50 to 60 percent range.

7.5.1.3 Summary—Adjustments to the Estimated UV Intensity

The lamp output will decrease with operating time. It is recommended that the system be designed (and subsequently operated) on the basis that the average output is approximately 80 percent of the nominal output. This is equivalent to an F_p of 0.8 in Equation 7-30. With time, it will be necessary to mix newer lamps ($F_p > 0.85$) with older lamps ($F_p < 0.75$) in order to maintain the desired output level. This will require monitoring by the procedures suggested earlier.

The quartz sleeves and Teflon tubes will require effective maintenance to keep their transmittances at reasonable levels. For design purposes, one needs to consider the minimum transmittance to be expected. With the quartz systems this is suggested to be 60 to 70 percent of nominal. This is equivalent to an F_t of 0.6 to 0.7 in Equation 7-30. For the Teflon systems, an F_t between 0.5 and 0.6 is recommended. In situations where the disinfection units are not to be frequently attended (remote, smaller plants), the values of F_p and F_t should be reduced further. Values of 0.7 and 0.4 are suggested, respectively.

7.5.2 System Design Considerations for Effective Maintenance

An overriding concern in the proper maintenance of the UV reactor for effective performance is to keep all surfaces through which the radiation must pass as clean as possible. The effects of surface fouling on energy utilization efficiency were discussed in detail earlier in this section. It is critical, and can very often be pointed to as the primary reason for the non-performance of a particular system. Other concerns relate primarily to the accessibility to UV reactors and to keeping adequate records to control replacement cycles and maintenance schedules.

7.5.2.1 Reactor Maintenance

The most reliable method to determine if a reactor is becoming dirty and requires cleaning is by visual inspection. The unit should be drained and the surfaces observed for fouling. In open systems this can be done rather conveniently and quickly. Reactors which are sealed vessels can be difficult; these designs should accommodate such visual inspections by incorporating large portholes or manways in the reactor shell.

Generally the surfaces of submerged quartz systems contacted by the wastewater will become coated by inorganic scale, very much like boiler scale. This will especially be the case in areas where there is hard water. Additionally, the inside surface of the quartz and the outer surfaces of the Teflon tubes will eventually develop a grimy dust layer, primarily from airborne dirt and water vapor.

Fouling of the reactor's internal surfaces will be signaled by reduced performance efficiency, or by reductions in the intensity measured by in-line probes. While these may provide some signal of fouling, it is still necessary to be able to physically inspect the surfaces.

Subsequent discussions will present procedures and equipment for routine cleaning of the reactor surfaces which contact the wastewater. First, however, it is appropriate to discuss a maintenance task which should be conducted at least once per year, or once per disinfection season. This is to completely overhaul the reactor, cleaning all interior surfaces, and determining the lamp outputs and quartz (or Teflon) transmittances.

These procedures were demonstrated for quartz systems at Port Richmond (54). Each lamp is removed from the reactor and washed with a mild soap solution, rinsed, and swabbed with an alcohol (isopropyl) soaked rag (cheesecloth). Then the interior surfaces of the quartz sleeves are cleaned by the same procedure by using a gun-barrel type cleaning rod to swab the interior surfaces. At the same time, each lamp, which is tagged with an ID number, is measured for relative output. Those which are below a specified level are discarded and replaced with new lamps. These new lamps are also tagged with a number. In this manner each lamp can be traced on the basis of operating time and output. A reactor lamp inventory can then be mixed and controlled to maintain a minimum average output level.

In similar fashion, the quartz should be monitored for transmittance. It may be cumbersome, however, to remove all the quartz from a system. It is recommended, instead, that a representative fraction of the quartz sleeves be monitored. Ten to fifteen percent of the quartz inventory would be sufficient. The same quartz should always be monitored; these would be considered as representative of all quartz in the system. If the tagged quartz begin to show marked deterioration due to aging and wear, it may then be appropriate to broaden the monitoring and to begin replacing the quartz sleeves. This replacement can be accomplished gradually. As with the lamps, there will eventually be a mix of old and new quartz in the system. There is little experience in determining the effective life cycle of the quartz sleeves; certainly it will vary by site, but generally should be between four and seven years.

In Teflon systems, the lamps are removeable on racks; they should be cleaned and monitored in the same manner as the quartz systems. The Teflon tubes should be cleaned on occasion; this can be done by swabbing the tubes with soapy water/alcohol. A non-abrasive material should be used. Each tube should also be monitored for transmittance, just as with the quartz sleeves. This may not be as straightforward, however, because of their limited accessibility and problems in getting direct measurements with a UV radiometer/detector. The actinometry procedure described earlier may be the more appropriate method for this application.

This system overhaul, as mentioned, should be accomplished at least once per year. In the case of plants with seasonal disinfection requirements, the most appropriate time would be before the start of the disinfection season.

7.5.2.2 Routine In-Place Cleaning

Regardless of a particular system's accessory cleaning equipment, it is likely that periodic chemical and/or detergent cleaning will be required to maintain the outer quartz, or inner Teflon surfaces. This is particularly the case where the wastewater is relatively dirty (secondary or primary effluents), has a relatively high grease and oil content, or has a high hardness content. A major cause in fouling the quartz surfaces (and to a lesser extent, the Teflon surfaces) has been found to be inorganic magnesium and calcium carbonates. The inorganic deposition is also easiest to control; simple acidification of the reactor water will generally dissolve the material and restore the surface. In the case of organic fouling (usually from a high grease content) it is necessary to use a detergent or some combination of cleaning agents. This is typically determined by trial and error for a particular situation. The frequency with which this cleaning task has to be accomplished will also be site specific, and will be determined with experience.

A number of systems commercially available offer accessory equipment which are purported to maintain the surfaces of the Teflon or quartz. Currently, these are the mechanical wiper, ultrasonic transducer, and a high pressure spray nozzle. The mechanical wiper and ultrasonic devices are applicable to the quartz systems, while the high pressure spray wash is applicable to both the Teflon and quartz systems. Although these components may be effective in cleaning the appropriate surfaces, intermittent cleaning with chemicals is generally required, and it is strongly recommended that provision be made in the overall system design to allow for chemically cleaning the system. In certain cases, the accessory cleaning devices can be used to assist and increase the efficiency of the chemical cleaning task.

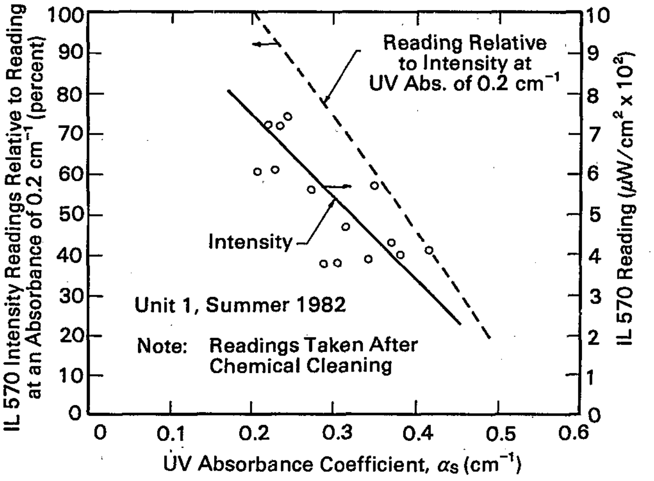
Chemical Cleaning. The task of cleaning the UV reactor on a routine basis is generally a very straightforward and simple task. The procedures and equipment requirements will generally be a function of the type of reactor and the system size. There is no standard procedure or equipment design, nor should there be. A system and procedure should be developed which best suits the application. This will be influenced by the size of the plant, the level of operator attention, the type and characteristics of the wastewater, and the UV system configuration.

The simplist procedure is applicable to the exposed open channel units. Examples are the systems at Pella, Iowa; Northfield, Minnesota; and Tillsonburg, Ontario. These are all open channel systems where the lamp batter can be drained, is accessible, and can be easily inspected visually. The unit at Port Richmond were also similar to these types of systems.

At Port Richmond, a combination of acid and an industrial detergent was used. The unit to be cleaned was isolated by diverting flow to the second unit; acid was then added to adjust the reactor water pH to approximately three. The acid was sulfuric in this case; a pH meter would be used as the entire reactor liquid volume was acidified by direct addition. Once acidified, the detergent would be added in similar fashion. The wiper stroke would usually be increased to assist the cleaning operation. This procedure was found to be very effective and typically required no more than one hour of a single operator's time.

An operational concern regarding this aspect of system maintenance is determining the frequency and/or need to chemically clean. Visual inspection has already been mentioned as the most effective procedure. An in-place intensity monitor would be effective, although there are concerns for this type of fixed place detector. The window for the probe must itself be kept clean in some fashion; additionally, the probe sees only the surfaces in the near vicinity, and the probe cannot directly account for the absorbance characteristics of the wastewater itself. At Port Richmond, a portable radiometer was found to be more effective, when used in conjunction with the UV absorbance measurements taken on each sampling. Intensity measurements were taken by placing the detector at selected (and reproducible) positions along the influent and effluent plane of the lamp battery. Figure 7-53, taken from the Port Richmond report, shows detector intensity readings versus the absorbance readings of the wastewater. These were taken after the unit was chemically cleaned. The relationship is important, not from the absolute intensity readings as a function of absorbance, since these will change with lamp age, but from the relative change as a function of absorbance. In this particular example, the intensity is shown relative to the intensity at an absorbance coefficient of 0.2 cm^{-1}. Thus, by knowing the relative effects of the wastewater absorbance, the operator can make reasonable judgements from the radiometer readings as to the condition of the quartz surfaces. This same procedure can be accomplished with the Teflon system by inserting the detector (always to the same fixed position) into the Teflon tubes.

Figure 7-53. Example of radiometer intensity readings as a function of UV absorbance at Port Richmond (54).



At Pella, Iowa, direct provision was not made for chemical cleaning (62). The system has mechanical wipers which are generally effective, but the units do foul. The operators isolate one of the two units, drain the channel and inspect the lamp battery. If dirty, the unit is rinsed, then sprayed with an acidic detergent. The wiper is stroked across the unit several times and the lamp battery is rinsed once again. The surfaces are inspected and the procedure is repeated, if necessary. The unit is then put back into service. This entire procedure is efficient, requiring one-half to one hour, and is typically done once every one to two weeks.

At Northfield, Minnesota, the quartz units are constructed very much like the units at Port Richmond, with inlet and outlet tanks. The lamp battery is open on both the inlet and outlet planes. Again, this plant had been designed with mechanical wipers and no provision to chemically clean the quartz. The area is marked by very hard water and the quartz were found to foul fairly quickly with an inorganic scale. The operators had cleaned the unit by adding citric acid to the reactor water (having first isolated it from the system) and allowing it to soak for several hours. The wipers would also be kept in operation during this period. This was generally effective, but relatively expensive. Each application involved adding 50 to 100 lb of citric acid, costing over $100 per sequence.

This procedure has subsequently been modified. The citric was replaced by a cheaper mineral acid, (sulfuric acid). Less than a nine pound bottle is needed to reduce the pH to approximately three. It is recommended that the acid input be controlled by a portable pH meter to prevent addition of excess acid. A small metering pump with a small bulk acid drum (stored outside the building), is also recommended to further reduce the cost of the acid (low grade or waste mineral acid is sufficient). A recirculation pump is also added to rapidly mix the tanks and provide agitation. The water level is typically lowered to the top of the lamp battery to further reduce acid requirements. Once acidified, the unit is allowed to stand for a period of time, with the recirculation pump in operation and the wiper moving. The tank is then fully drained (drainage to the head of the plant) and the quartz is inspected. It is rinsed, sprayed with an acidic detergent, and then rinsed again with a high pressure hose. A wand sprayer is used to be sure the internal quartz are reached. The unit is then brought back into service. This procedure is effective and requires only one to two hours to accomplish. The frequency is generally once every week.

The Tillsonburg, Ontario unit is set in the plant secondary effluent channel and does not have any accessory cleaning equipment (63). This too can be easily cleaned because of its accessibility. The unit is isolated, the channel drained, and the quartz cleaned by the procedures described above. Alternatively, this particular design allows for individual lamp modules (each containing four lamps) to be pulled from the unit. The unit does not have to be shut down, and, if necessary, a spare lamp rack can be inserted temporarily. The lamp racks are then inspected and thoroughly washed and rinsed. The unit design also incorporates the application of a proprietary polymeric coating to the quartz surface which is designed to retard fouling.

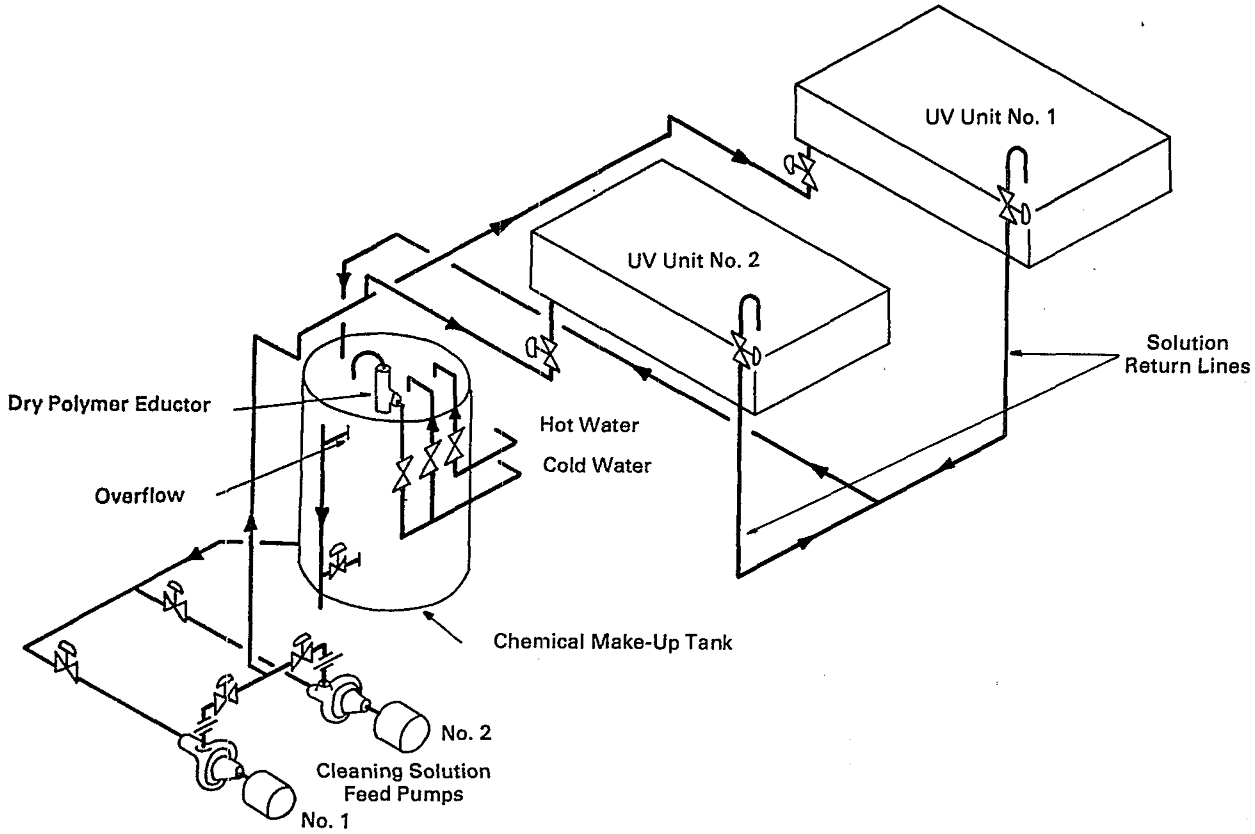
In sealed quartz systems, it is not possible to easily inspect the quartz or to access the internal quartz with sprayers or high pressure wash lines. At Vinton, Iowa the units are designed with ultrasonic devices, and a supplemental ability to chemically clean the reactor (64). The reactor is first drained, then it is filled with clean water and a chemical cleaning agent. These are added through piped inlets to the reactor. The system is also designed with inlet ports to inject either high pressure air or water to agitate the solution inside the unit. The unit is then drained (back to the head end of the plant), and returned to service.

At Suffern, New York, the system is also sealed. Ultrasonics are incorporated as the primary cleaning device, with a supplementary chemical cleaning system (4). The chemical cleaning system includes a solution mix tank, and recirculation pumps. A schematic is provided on Figure 7-54. A solution of the cleaning agent is prepared in the tank with warm water, and then recirculated for a period of time. Food grade citric acid or sodium hydrosulfite is used.

During startup at Suffern, the quartz became heavily fouled (the units had been allowed to sit for a long time filled with wastewater); the citric acid was not effective in this case. The sodium hydrosulfite was very effective; however, it is highly reactive and a strong oxidant. This material would likely be the most effective in sealed systems where the cleaning relies solely on contacting the surface under agitated conditions. It must be handled with great care, however, and special precautions would be required to properly store and handle the material. The chemical manufacturer should be consulted on these aspects.

A special note is also made under this topic of routine chemical cleaning. In several cases, it has been observed that a luxuriant growth (believed to be a fungus) will develop on the wetted metallic surfaces of the UV reactor. Additionally, particularly in reactors which have quiscent zones, sludge accumulations can develop. When the systems are drained and/or cleaned, an effort should be made to remove these accumulations.

Figure 7-54. **Schematic of in-place chemical cleaning system at Suffern, New York (4).**



Mechanical Wipers. A number of full-scale systems incorporate the use of a mechanical wiper. Recall from Figure 7-5 the schematic of the wiper blade on a submerged quartz system. These entail a machined frame in which the wipers are fixed; these then fit over the quartz sleeves. A single frame usually services the entire lamp battery. The wiper is driven by cable (pnuematically pulled) or by a piston. This can then be stroked across the reactor at a preset frequency. Examples of systems which have wipers are Pella, Iowa; Northfield, Minnesota; and Albert Lea, Minnesota.

Generally, the wipers are looked upon favorably by plant operators. At Pella, the wiperblades are a rubber-base ring. These tend to wear and will typically require replacement every one to two years. The wiper does not accomplish its original intent, which was to keep the surfaces clean, precluding the need for chemical cleaning. The wiper is felt to serve a useful purpose, however, by continually removing small debris particles, including grit, plastic fibres, and strings of algae (from the secondary clarifiers and channels). When the unit is taken down for chemical cleaning, the wipers are used to provide a degree of scrubbing.

Similar observations are made at the Northfield plant. The rings, in this case, are made of Teflon. Although they should not wear as quickly, it is found that the Teflon ring becomes distorted. It has no memory; thus, if there is a variation in the surface of the quartz, the Teflon will respond but will not recover its original shape. The wiper is still considered a benefit, however, because of its ability to keep the surface free of debris, and the ability to use it during the chemical cleaning task.

At Albert Lea, the wiper also uses Teflon rings but of a different design (and at present the most current design). These rings are split by cutting the ring on a bias at one point on its circumference. A spring then surrounds the ring, its compression causing the ring to always try to close. In this fashion, if the wiper passes over a section of the quartz which is larger in diameter, the ring will expand. As the quartz diameter decreases, the Teflon will also close down on the surface because of the spring action.

A limited series of tests were performed at the Albert Lea plant to assess the effectiveness of the wipers. These tests involved monitoring quartz clarity from modules which were operated with and without the wiper. The results were not wholly conclusive. Kreft et al. (4), reported that the wipers were effective in reducing the buildup of scale and biological growth on the surfaces of the quartz, when compared to the unit with no wiper in operation. The report also noted problems which must be addressed in the design and operation of system with the wiper devices. The cable drive is sheathed in Teflon; this was found to crack and cause water leaks onto the outside electrical connections (this same problem was reported in the Port Richmond study). This resulted in electrical hazards which on one occasion caused a small electrical fire. The second problem arose due to the misalignment of the wiper frame. This will cause, at minimum, incomplete wiping of the quartz; at worst, the misalignment will cause breakage of the quartz sleeves.

Other considerations which should be taken into account are the time requirements for the disassembly and repair of these wiper mechanisms. Experience up to now has involved the equipment manufacturer accomplishing this task; this may be appropriate since it requires extensive handling of the quartz sleeves and precise alignment of the frame when reassembled. The design of the UV system should address these tasks, particularly their costs, including the costs of replacing the individual wipers (e.g., the spring loaded split Teflon rings).

Ultrasonics. Ultrasonic devices rely on the surface cavitation caused by high frequency sound waves. When properly applied, debris which coats a surface will simply fall off. This concept has been applied at a number of full-scale plants, including the Suffern, New York and Vinton, Iowa installations. The ultrasound transducers are typically inserted across the length or width of the reactor, parallel to the quartz sleeves. They are operated on an on/off cycle, the on-time determined by the specific site requirements.

The units at Vinton each have two transducers which have 1.8 kW ultrasound input; thus, each unit is equipped with a total of 3.6 kW input. The units are rated to have an effective radius of activity of approximately 0.75 m over a 180° arc.

The total lamp power in each of the units at Vinton is approximately 12.8 kW; this yields a ratio of ultrasonic power to lamp power of 0.28. With the ultrasonics operated 25 percent of the time, the energy utilization by the ultrasonics comprises approximately seven percent of the total when all lamps are being operated and 28 percent when one bank of lamps (out of four banks) is in operation. These would increase if the ultrasonics are required for greater periods of time.

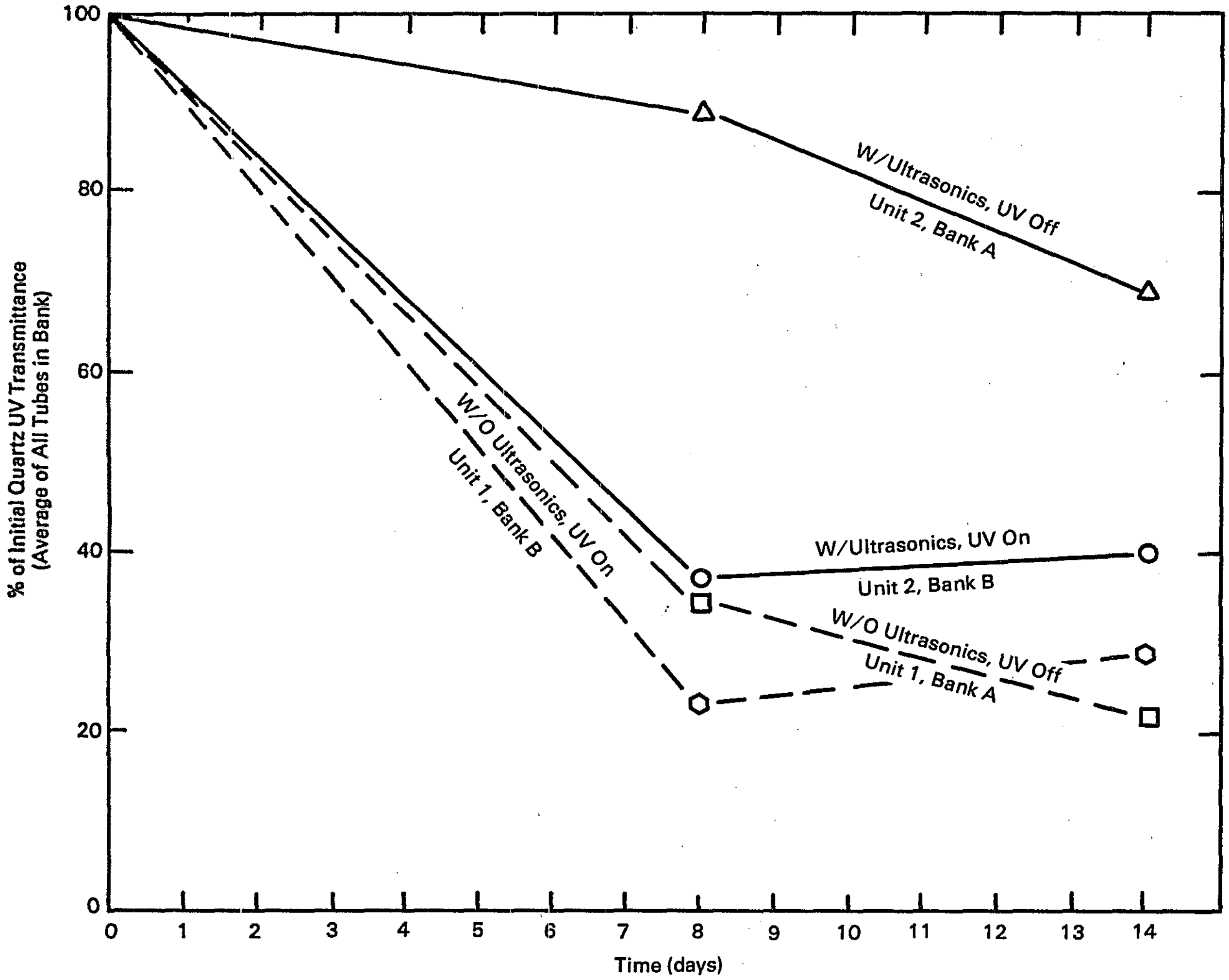
The system at Vinton has been in operation for approximately two years. At first the ultrasonics were operated with approximately 25 percent on-time. This was not felt to be effective and was gradually increased to nearly full time. This is excessive; its cost of operation negates its utility. Hypochlorite had been used to chemically clean the Vinton reactors. This is not effective, and may have contributed to the poor performance of the ultrasonics.

The Suffern units each have one 1.5 kW transducer per two banks of lamps, or two per unit. The ratio of ultrasonic to lamp power is approximately 0.14, or half that at Vinton. A limited series of tests were recently performed at this plant to evaluate the effectiveness of the ultrasonics (4). The tests were conducted over a two week period, using unit 1 without ultrasonics and unit 2 with ultrasonics. The ultrasonics were operated on a cycle of 30 minutes on and 30 minutes off (50 percent on-time). Flow was generally split evenly between the two units. As discussed earlier, the quartz will accumulate buildup differently, depending on whether the lamps remain on or off for extended periods of time. With the lamps on, an inorganic scale tends to develop; with the lamps off, the quartz simply provide a surface for a biological film to develop. The ultrasonics were evaluated as to the ability to retard the buildup of either type of material. In either unit a select number of quartz sleeves were monitored in a bank where the lamps were kept off and in a bank where the lamps were continuously operated.

The results of the evaluation, taken from the study report, are presented on Figure 7-55. The dashed lines are from Unit 1, which operated without the ultrasonics; there appears to be no significant difference between the banks with the lights on or off. It may be said that the bank with the lights on tended to degrade at a faster rate than the bank with no lamps on. By the fourteenth day, however, the transmittance of the quartz sleeves in either bank was between 20 and 30 percent of the initial transmittance.

The results of the test in Unit 2, which operated with the ultrasonics, show a significant difference between the banks with or without the lamps on. With the lamps off, the ultrasonics appear to have been very effective; after 14 days, the quartz transmittance was still approximately 70 percent of the initial transmittance. In the bank which had the lamps on, however, the ultrasonics was ineffective; by day eight, the transmittance was between 35 and 40 percent of the initial transmittance. It appears that the ultrasonics are unable to retard the softening effect in which the inorganic carbonates plate out on the

Figure 7-55. Comparison of ultrasonic cleaning performance at Suffern, New York (4).



quartz surface. This phenomenon will occur only when the lamps are on.

In all, current experience has not confirmed the utility of ultrasonics as a cleaning device. This would especially be the case in areas with hard waters. Additionally, consideration should be given to the cost of the accessory device on the basis of both additional capital cost and added operating cost due to its energy requirements.

High Pressure Wash. This is closely related to the earlier discussions of routine chemical cleaning. Commercially available industrial cleaning units are available which use pressurized water to clean surfaces. The type of hose nozzle or spray wand used to discharge the water will vary and will depend on the application. In the Teflon units, a nozzle is fitted to the end of a flexible hose; spray is directed radially from the nozzle as the hose is snaked down the Teflon tubes. The systems have the capability of educting a detergent or acid solution into the wash stream. The same type of system would be applicable to quartz systems. In this case, however, a wand type sprayer would be more appropriate to reach the insides of the lamp battery. The system is inexpensive, easy to use, and recommended for most UV reactor applications.

Equipment manufacturers for the Teflon tube systems have been supplying a high-pressure nozzle spray cleaning system with several recent installations. Kreft et al. (4) reported on visits to wastewater treatment plants at Chinook, Montana and Rock Springs, Wyoming to observe performance of the cleaning systems. The Rock Springs UV disinfection unit had moderate to heavily-fouled Teflon tubes (black and brown coatings) caused by heavy solids and foam and grease carry-over into the tubes. During cleaning, it was noted that the high-pressure nozzle system was able to remove, in some cases, significant portions of the fouled material from the inside surfaces of the Teflon, but was not consistent for all tubes. In some tubes, very little material was removed and, therefore, the UV transmittance would be assumed to still be minimal. A series of repeated cleanings may be necessary to remove the material from all tubes. This is a critical point; if a few tubes remain dirty, their consequent poor performance can have a dramatic effect on the overall performance of the system.

At Chinook, the amount of fouling on the tubes was not as heavy as at Rock Springs and the high pressure cleaning system appeared to remove most of the material attached to the inside of the Teflon tubes. A sample of a cleaned tube was measured by the chemical actinometry method along with the used samples that were taken from the Chinook plant. It was verified, as shown in Table 7-17, that the in-place cleaning did return the Teflon tube to a UV transmittance in the range of 65 to 70 percent from an uncleaned transmittance of 30 to 55 percent.

The authors concluded that the high-pressure nozzle washing systems have some benefit in helping to clean the interior surfaces of Teflon tubes. Since the Teflon tubes are themselves usually difficult to access, the nozzle cleaning system offers some advantages and provides the operator with a simpler cleaning task. They did indicate, however, that internal swabbing of the Teflon tubes with a soft rag, and possibly with a detergent, will be necessary on an occasional basis to ensure that the Teflon tubes remain in a fairly clean state. It was also noted that many of the plants visited that had Teflon UV units had noticeable amounts of dust on the outside surfaces of the tubes. It is highly recommended that in all installations, a routine maintenance task must be to clean the outside with a rag and, possibly isopropyl alcohol or water; this will improve the transmission of the UV light through the Teflon into the wastewater.

7.5.2.3 Other Design Elements for Effective Maintenance

The following observations are made on the basis of current full-scale and pilot scale operating experiences. These are directed to considerations for design, fabrication, and installation which will ease maintenance tasks or provide for more effective maintenance:

a. The reactors and related tankage should be equipped with drains which will allow for complete and rapid dewatering. Drainage should be to the main plant drainage system.

b. A clean water supply should be permanently available, in addition to all requirements for chemical cleaning.

c. The systems should be designed modularly with the ability to readily isolate a module from the plant flow.

d. A bypass should be constructed around the entire UV disinfection system, particularly in plants which require only seasonal disinfection. This would allow for greater convenience for maintenance tasks during the non-disinfection season.

e. The accessibility to the lamps, quartz sleeves, and Teflon tubes is critical to the ease of maintenance. Manways should be provided on larger scale systems.

f. Strict inventories should be kept of the lamps in use, their relative output, and their estimated cumulative operating life. This should also apply to a more limited extent to the quartz sheaths, Teflon tubes, and ballasts.

g. The reactors and all other accessory equipment should be installed in an area that is adequate to accomplish all the required maintenance tasks. The systems should not be so cramped that it is virtually impossible to work on the units.

h. If reactors are taken out of service, they should always be drained; a clean water rinse would also be appropriate. The units should then be held in a drained, dry condition.

7.5.3 System Components

The major components of the UV systems are the lamps, enclosures, and the ballasts. Discussion of these have been interspersed throughout this text. The following observations are made to highlight the major points to consider when evaluating or designing a new system, or in the operations of an existing system.

a. The low pressure mercury arc lamps are currently the most efficient source of UV radiation. Costwise, the longer arc length lamps are more efficient. Because of the negative effects of the ozone produced by the 185 nm light, lamps with fused quartz envelopes are not recommended. The lamps of vycor or other high transmission glass are appropriate.

b. Care should be taken to minimize temperature effects. In quartz systems, O-ring spacers should be slipped over the lamps to prevent direct contact with the cooler quartz sleeve. These are generally provided in the newer systems.

c. Fittings holding the quartz sleeve should be tight and leakproof. A number of plants have had problems with leaks at these points, causing electrical hazards and corrosion. Additionally, some designs are difficult to disassemble and reassemble, causing excessive labor, quartz breakage, and continuing leaks.

d. The quartz sheaths are fairly standard in composition. Variations have been noted in the wall thickness; attention should be paid to the structural strength of the sleeve. Another variation is the single end quartz in which only one end of the quartz is open; the other end of the quartz is fused close. A plant which will use this type of quartz is the Nine Springs plant in Madison WI. The wiring from the lamp electrode at the fused end of the quartz will be snaked back to the open end of the quartz.

e. Frequent on-off cycles for the lamps will shorten their life. More effective lamp control may be accomplished by voltage dimming, in conjunction with on-off control of banks of lamps.

f. Control panels should be remote from the UV reactor.

g. The ballasts must be properly mated with the lamps being used. It is strongly recommended that both the lamp and ballast manufacturers be consulted on this aspect. It would be appropriate to require certification that the ballast is correct for the UV lamp. The ballasts should be thermally protected; this forces the ballast to shutdown if it overheats.

h. The power panel containing the ballasts must have adequate ventilation to discharge the heat generated by the ballasts. This has been a recurring problem at full-scale installations. The life of the ballast is greatly shortened and in several cases excessive heat build-up caused rapid failure of a number of ballasts. High volume ventilation fans should be installed in the power panels to cool the ballasts. During warm temperature months, this should be vented out of the building.

i. Careful attention should be paid to the electrical wiring of the UV systems, at the points of both fabrication and installation. Improper wiring at several plants resulted in electrical hazards, component failures, and in some cases, small electrical fires. The wiring should be properly sized and the wire covering should be resistant to UV radiation effects. Typically, Teflon coated wiring is specified.

j. The Teflon tubes are generally standard. The variables are the wall thickness and the tube diameter. The transmittance will decrease with increasing wall thickness; very thin walls will limit the structural integrity of the tube, however, causing them to collapse. Greater intensities can be achieved with smaller diameter tubes; this will have to be weighed against increased head losses.

k. Air bleeds should be considered on certain Teflon systems (in particular the pressure units) to minimize air binding in the Teflon tubes.

l. Removeable screens should be placed upstream of the lamp reactor to prevent large debris from entering the system. This is especially important for quartz systems.

7.5.4 Monitoring and Control

System Controls. The sophistication of the monitoring and control systems for the UV process can vary from minimal to fully automated. This is no different from any other unit operation in a treatment facility. It is recommended that the minimum should always be provided; any increased capability should then be considered on a cost-benefit basis. The minimal requirements suggested for the UV disinfection process are flow metering per unit, individual lamp operating monitors, a portable radiometer, power panel temperature (with alarm), and the ability to turn portions of the system on and off on the basis of time. The following observations are made, again as points which should be considered when evaluating or designing a UV system:

a. The units should be arranged such that banks of lamps can be shut off or on. In the simplest mode this can be controlled by timers. Modest adjustments can then be made on a diurnal basis to reflect the normal variation in the plant's flow. This can be further advanced by automatically slaving the lamp bank operations to the plant flow; some systems also will adjust the unit voltage, using bank shutoffs as a gross adjustment.

b. Concurrent with the plant flow, the control of the system can be coupled to the water quality. This is done in some systems by use of an intensity monitor fixed to the reactor; this may cause problems if it cannot be effectively maintained. An alternative method is to utilize a continuous monitor of the wastewater UV absorbance.

c. As had been discussed earlier, it is important that a continuing record be kept of the average output of the lamps, and the transmittance of the quartz and Teflon. Procedures have been given by which to monitor these parameters. It is strongly recommended that these become part of the system's routine O&M, and that the necessary equipment be available to accomplish these tasks.

d. In-line pilot light monitors should be installed in the control panels to indicate each lamp operation. This is normally available on most UV systems. Alarms should be installed to alert the operator if a preset number of lamps fail.

e. The flow to each module should be metered.

f. Elapsed time monitors (non-resetable) should be installed for each bank of lamps. This will allow an accurate accounting of cumulative operating time and will allow the operator to balance the use of the various lamp banks in a system.

g. The temperature in the ballast power panel should be monitored. An alarm system should be available to alert the operator if the panel temperatures exceed an acceptable level.

h. A watt-meter would be useful, particularly in larger systems, to monitor on a continuous basis the power requirements relative to the remainder of the plant.

7.5.4.2 Wastewater Monitoring for an Existing System

The UV disinfection process does not have the monitoring advantage of a measurable residual (as is the case with chlorine). As such, greater care is required in controlling operations efficiently and still maintaining performance. It is strongly recommended that this entail frequent sampling and analysis. A suggested protocol is to sample the system a minimum of three times per week (alternating days) between the hours of 10 a.m. and 3 p.m. The influent to the UV system should be analyzed for suspended solids, UV absorbance, and coliform density (i.e., the bacterial monitor prescribed by the plant's permit). Additionally, the flow rate should be recorded at the time of sampling, as well as the operating conditions of the reactor (number of lamps in operation, etc.). The effluent sample should be analyzed for coliform density.

An important note should be made with regard to the subsequent handling of the exposed UV effluent sample. All precautions should be taken to protect the sample from exposure to visible light (sunlight and normal fluorescent and incandescent light), before the sample has been set and put into incubation. This will prevent the occurrence of photoreactivation. Normal precautions include sampling in an opaque (and covered in foil) sample bottle, and keeping the sample covered during the procedure to set the sample (filtration and plating for the MF procedure, and dilutions and inoculation for the MPN procedure).

Alternatively, if photoreactivation is to be accounted for by the permit requirements, the effluent should be taken with the transparent glass bottle and left in direct sunlight for approximately one hour. Enumeration would then be accomplished by routine procedures. Either the direct Membrane Filter (MF) or the Most Probable Number (MPN) procedures have been demonstrated to yield equivalent recoveries (65).

Collection of such a relatively comprehensive data set is felt to be important in controlling the operations of a UV system. It allows direct evaluation of the system performance under current wastewater conditions and provides a data base from which the disinfection model can be calibrated and/or refined. The model itself then becomes an excellent tool in controlling the system and optimizing operations for maximum use of lamps and minimal use of energy. The continuous collection of the appropriate data also allows a rational approach to troubleshooting the non-performance of a system.

7.5.5 Safety Considerations

Ultraviolet disinfection is basically a safe process; the activity is generated on-site; thus, there are no transport concerns to or from the site, or concerns regarding storage of reactive material. Normal plant safety precautions apply relative to physical layout (railings, etc.) and to electrical hazards. Power supplies are high voltage, requiring the adherence to normal electrical safety codes. Electrical interlocks should be provided to shut off systems when opened (reactor end panels); particular attention should be paid to electrical wiring, groundings, and waterproofing.

The storage, handling, and disposal of the expendable components should also be considered from the standpoint of safety. Storage of lamps, quartz sleeves, and ballasts should be in a separate dry area. Adequate shelving should be designed to store the materials such that they are protected from breakage, and are easily and safely accessed. Used lamps, quartz sleeves, and ballasts which are to be discarded should be repackaged and overpacked for safe disposal.

Personnel safety training should address and require strict adherence to personal protection from excessive UV radiation. A lamp battery would not present a hazard while submerged and operational; the water absorbance will sufficiently attenuate the radiation. These lamp batteries should not be operated while in a dewatered and dry state. Similarly, "dry lamp" systems such as the Teflon unit, should have all covers in-place during operation. Plastic (e.g., plexiglas) will not transmit the 253.7 nm wavelength; this material can be used for the end plates (windows or end plates) or unit shields to protect against exposure but still allow visual inspection of the lamp ends. If it is necessary to engage a system without shields in place (or with the lamp battery exposed) it is absolutely necessary that the proper protective gear be worn by all personnel in the area and that adequate warning signals be active during these operations to warn anyone entering the area. This will also apply during the routine lamp monitoring tasks discussed earlier.

The skin and eyes readily absorb UV radiation and are particularly vulnerable to injury. Sunburn (erythema) is a common example, although this effect is most pronounced with UV light between wavelengths 285 and 300 nm. Absorption by the mucous membranes of the eye and eyelids can cause conjunctivitis (commonly referred to as "welders flash"). The injury becomes apparent 6 to 12 hours after exposure; although painful and incapacitating, the damage is usually temporary.

Personal protection must include plastic goggles (wrap around), or face-shields. These must be rated to absorb the UV spectral lines. Protective clothing should be worn to prevent exposure to the hands, arms, and face.

7.5.6 Facilities Requirements for Full Scale Installations

The facilities needs are divided to two specific areas: equipment and the physical plant. The equipment elements include the hardware requirements directly associated with the installation of the UV process. The installation at a given plant site must then address the hookups and physical plant needs to install the system.

7.5.6.1 Equipment

The equipment generally supplied through the vendor include the UV reactor itself and the ancillary equipment used to control and monitor the system:

a. UV lamp battery (UV reactor)

b. Power supply and power panels (with single point hookup to the plant power)

c. Instrumentation for the control and monitoring of the system; this generally includes a UV intensity monitor per module, pilot monitors for each lamp, controller to direct the number of lamp banks on as a function of flow (and/or water quality), and alarms to signal deficiencies in lamp operation and/or performance.

d. Accessory cleaning equipment; this is generally in the form of chemical cleaning, mechanical wipers, or ultrasonics.

e. Manufacturer's engineering and startup services are generally at the option of the buyer.

f. Replacement parts supplied with purchase; this should include no more than 100 percent lamps, 50 percent ballasts, and 20 percent quartz or Teflon tubes. At minimum, this should be 20 percent, 10 percent and 10 percent, respectively.

When specifying a UV system, the design should analyze the cost of the ancillary monitoring and cleaning devices to be provided with the reactor against this need and benefit. These can account for 15 to 25 percent of the total equipment costs.

7.5.6.2 Physical Plant

The structural and installation requirements are sensitive to the specifics of the site and the equipment to be installed. It is difficult to give a detailed assessment of these needs; by way of guidance, however, a generic installation is considered.

There is a basic space requirement for the UV system at a plant, based on the number of lamp modules to be installed. The smallest plants would generally require only one module; the reactor itself is rather small. A minimum space of 10 m^2 (108 sq ft) should be allowed, however, for the unit. It is further suggested that this space allocation be increased to 25 m^2 (270 sq ft) for large modules which may contain several hundred lamps. Thus, a 500 kW system may contain approximately 6,000 lamps in 10 modules; the total space allocation would be approximately 250 m^2 (2,700 sq ft). In most cases, the system should be housed in a standard building. Certain configurations proposed by engineers/manufacturers (including existing full-scale systems), do not require such housing. The designs generally call for open channel installations of the UV equipment. The power supplies and instrumentation in this case require more stringent specifications with regard to weatherproofing and protection against water/electrical hazards.

Reactors which are housed are generally characterized by piped inlet and outlet structures. The reactors are typically sealed units or have integral influent and effluent tanks attached to the lamp battery. The housing itself may be shared with other unit operations in the plant. Power supplies and control systems are contained in the building, typically remote and elevated from the reactors. Storage for spare parts (lamps, ballasts, etc.) can also be accomplished within the building. The entire area should be adequately ventilated, particularly with regard to humidity control and venting from the power panels.

7.5.7 Estimating O&M Requirements

UV is a capital intensive process, with the equipment requirements directly proportional to the design peak hydraulic and performance needs. The operational and maintenance needs, however, are reflected more by the average utilization of the system. In fact, a key operational consideration is to use only that portion of the system necessary to meet current performance demands. Over-utilization of the system, in an attempt to simplify operations, will have a significant impact on the costs to operate the process. The following discussions focus on the three major elements which comprise the costs associated with the operation and maintenance of a UV system.

7.5.7.1 Labor Requirements

Estimating labor requirements is a subjective task, relying on current experiences and being selective in defining the tasks which should be assigned to the UV process. In estimating labor needs, Scheible et al. (54) assessed the experience in the operation of the Port Richmond plant, as well as previous studies, and the current experience at full scale facilities. The estimates are summarized in graphical form on Figure 7-56. As shown, the labor is divided to three major categories: direct UV operation and maintenance tasks; general maintenance; and system overhaul. It is important to note that the labor estimates are based on the O&M requirements for the entire installed system.

Direct UV Operation and Maintenance. The tasks which are considered in this category may be described as follows:

1. Operations and Monitoring
    - daily systems checks for proper operation;
    - appropriate recording of data (lamps in operation, meter readings, power readings, flow rates, water quality readings, temperatures, etc.);
    - sampling and analysis for SS, bacterial density, UV absorbance;
    - direct manual control of the systems, or the monitoring and control of automatic operational instrumentation.
2. Maintenance
    - checking and maintaining system components (lubrication, etc.);
    - storage and maintenance of appropriate parts inventory;
    - routine systems cleaning, the labor associated with this task will include monitoring of the quartz/Teflon surfaces, switching systems during special cleaning cycles, and maintenance of the chemical feed systems;
    - replacement of worn or broken components in the system.

The labor needs assigned to these direct O&M tasks are estimated to range from 2-3 hr/wk for small systems (less than 100 lamps) to 15-30 hr/wk for larger plants (greater than 1,500 lamps).

General Maintenance. As had been discussed, there are space, building, and ventilation requirements associated with the installation of a UV process. The general maintenance of these physical facilities will be required. For purposes of this discussion, the labor is assigned to the labor requirements for the disinfection process at a plant. It is suggested that approximately one-half the labor required for the direct O&M tasks discussed above be assumed for the general maintenance tasks.

System Overhaul. The reader is referred to the earlier discussions which dealt with the cleaning and direct measurement of the lamps and the quartz/Teflon which comprise the UV reactor. It is strongly recommended that the entire system be broken down on a yearly basis to accomplish the following tasks:

a. clean the outside surface of each lamp;

b. clean the inside surfaces of each quartz sleeve, and the outside surfaces of the Teflon tubes;

c. measure each lamp for relative UV output; replace those which fall below a specified level;

d. measure a representative sampling of the quartz/Teflon enclosures for transmittance, replace those which are worn excessively; and

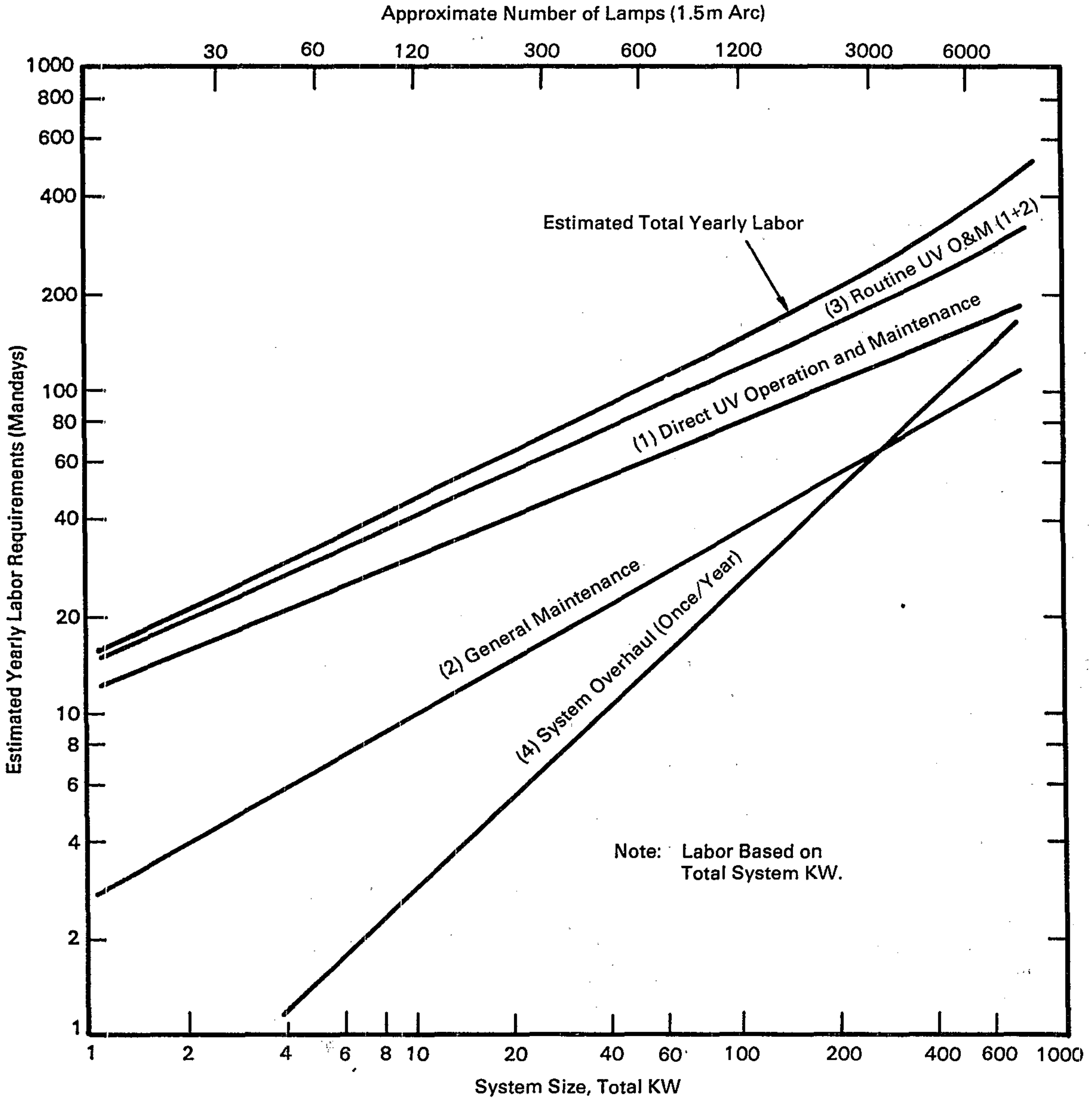
e. check internal components for wear and replace if necessary.

These tasks are suggested to serve as an efficient means to control the system's output and energy efficiency at acceptable levels. By having direct measurements of the unit's average output, the lamps can be utilized to their maximum life. Keeping the surface clean will allow for efficient use of the UV energy.

Based on the experiences of Port Richmond, the labor required to accomplish the system overhaul each year is estimated to be approximately 16 hr/100 lamps. Relative to the total labor requirements for the UV process, the system overhaul is small, but can yield significant overall O&M cost savings.

Total Labor Estimates. The total yearly estimated labor requirement is presented on Figure 7-56. Note that these are based on year-round disinfection. In cases where seasonal disinfection is allowed, the labor estimates for the direct O&M, and the general maintenance tasks would be reduced; the system overhaul is still recommended on a yearly basis. Overall, the labor needs for the UV process are relatively low, ranging from approximately 40 mandays/yr for a small 10 kW (120 lamps) system to approximately 400 mandays/yr for a 400 kW system (5,000 lamps).

Figure 7-56. Estimate of labor requirements for the operation and maintenance of UV systems (54).



7.5.7.2 Materials

The major materials cost associated with the UV system are the lamps, the ballasts, and the quartz or Teflon enclosures. Note that these requirements are considered as a function of the annual average system utilization (kW). Thus, although the system may be sized to meet the peak power demand, the need to replace the major expendable components of the system will depend on their actual use; this can be represented by the estimated annual average utilization of the system. To estimate the annual average requirement for material, the following suggestions are offered:

Lamps. Low pressure mercury arc (1.5 m/arc) are standard. This replacement cycle should be assessed at one year (8,700 hours). This is conservative; there are cases where considerably longer life cycles have been demonstrated.

Ballasts. A single ballast serves two lamps. The average life cycle is five years; this can be considered conservative if the ballast is properly mated and the power panel is properly ventilated to prevent overheating.

Quartz/Teflon Enclosures. The estimate should assume one quartz sleeve per lamp, and one Teflon tube (3 m) per two lamps. An average life cycle of five years is suggested for both, although there is little demonstrated experience in this regard.

Miscellaneous. To account for miscellaneous parts replacement, a cost equivalent to five percent of the annual lamps, ballasts, and enclosures costs are suggested.

7.5.7.3 Power Requirements

The third element, power, should also be estimated on the basis of the annual average system utilization. This requirement can be accounted for by addressing the lamps only. Ancillary power use is relatively insignificant, except in cases where ultrasonic devices are used for cleaning. The total power per lamp is 80 Watts (for 1.5 m arc lamps), including the ballast.

7.5.7.4 Estimating Average Annual Utilization

As discussed, the materials and power requirements (and to a lesser extent, the labor needs) should be based on the annual average utilization. An example of this analysis was provided in Section 7.4 for the design example. It is based on the design needs for average wastewater conditions (flow, UV absorbance coefficient, initial density, and suspended solids). This can be significantly less than the peak system requirement (20 to 30 percent of peak), particularly with plants which are not at capacity and in cases when only seasonal disinfection is required.

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