



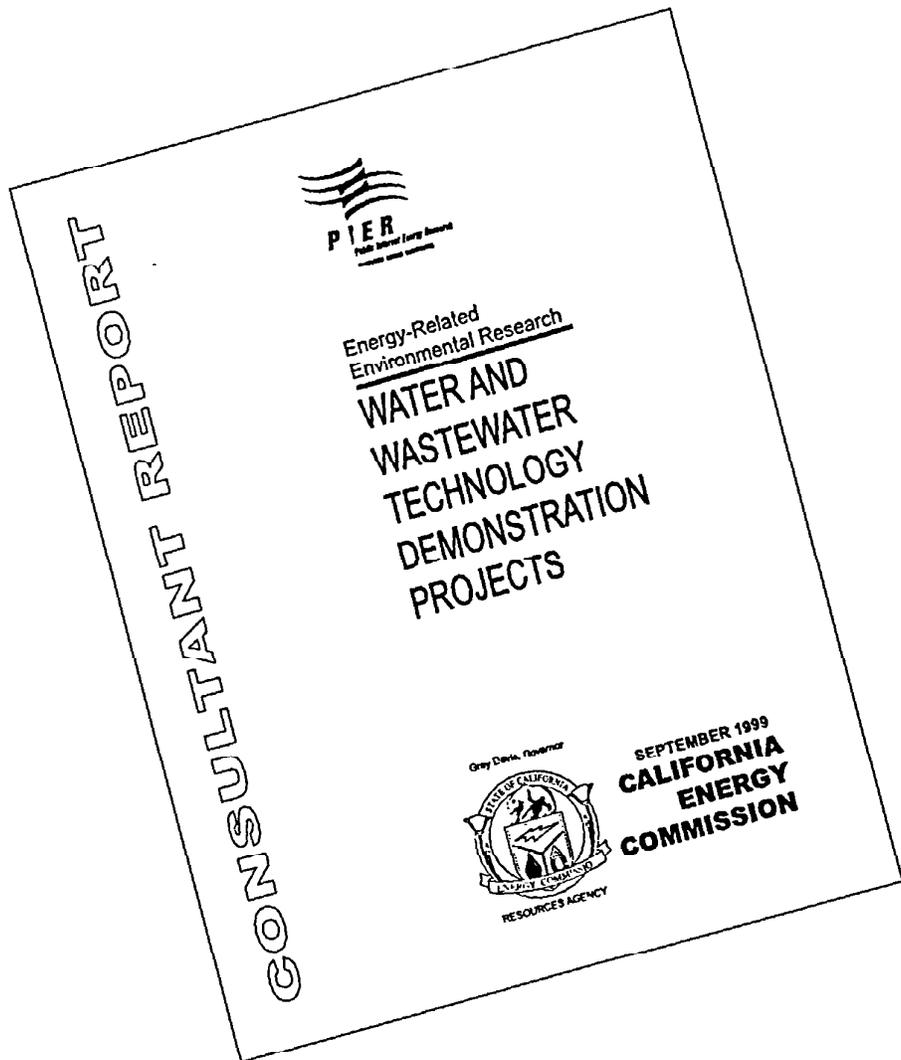
Energy-Related
Environmental Research

WATER AND
WASTEWATER
TECHNOLOGY
DEMONSTRATION
PROJECTS

Gray Davis, Governor



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Preface

The Public Interest Energy Research (PIER) Program supports public interest energy research and development that will help improve the quality of life in California by bringing environmentally safe, affordable, and reliable energy services and products to the marketplace.

The PIER Program, managed by the California Energy Commission (Commission), annually awards up to \$62 million through the Year 2001 to conduct the most promising public interest energy research by partnering with Research, Development, and Demonstration (RD&D) organizations, including individuals, businesses, utilities, and public or private research institutions.

PIER funding efforts are focused on the following six RD&D program areas:

- Buildings End-Use Energy Efficiency
- Industrial/Agricultural/Water End-Use Energy Efficiency
- Renewable Energy
- Environmentally-Preferred Advanced Generation
- Energy-Related Environmental Research
- Strategic Energy Research.

In 1998, the Commission awarded approximately \$17 million to 39 separate transition RD&D projects covering the five PIER subject areas. These projects were selected to preserve the benefits of the most promising ongoing public interest RD&D efforts conducted by investor-owned utilities prior to the onset of electricity restructuring.

Edison Technology Solutions (ETS) is an unregulated subsidiary of Edison International and an affiliate of Southern California Edison Company (SCE). As a result of a corporate restructuring, ETS ceased active operations on September 30, 1999. ETS' remaining rights and obligations were subsequently transferred to SCE.

What follows is the final report for the Water and Wastewater project, 1 of 10 projects conducted by Edison Technology Solutions. This project contributes to the Industrial/Agricultural/Water End-Use Energy Efficiency program.

For more information on the PIER Program, please visit the Commission's Web site at: <http://www.energy.ca.gov/research/index.html> or contact the Commission's Publications Unit at 916-654-5200.

Executive Summary

The Water and Wastewater project was funded by the California Energy Commission with Public Interest Energy Research (PIER) program transition funds, the Electric Power Research Institute (EPRI), and Southern California Edison (SCE). Edison Technology Solutions (ETS) managed the project.

Water supply and wastewater disposal are vital issues for Southern California, which relies on a variety of sources for its water: the Colorado River, the California State Water Project (SWP), local surface water, groundwater, and reclaimed water. Two of California's largest water and wastewater suppliers, the Metropolitan Water District (MWD) and Orange County Water District (OCWD) conducted this six-task research project.

The goal of this project is to identify and develop technologies that can help California municipalities substantially reduce the cost of water and wastewater treatment and improve the operation of treatment facilities. SCE alone provides over 2.0 billion kilowatts hours (kWh) of electricity per year to over 5,000 water and wastewater customers.

Based on the needs of the municipalities, the research consortium developed a six-task project addressing the main issues facing the municipal water and wastewater industry in Southern California. MWD's tasks focused on evaluating research needs for desalination, disinfection of the Colorado River Water (CRW), and bromate control of CRW and California SWP water. OCWD's tasks were directed at wastewater reclamation, focusing on disinfection, microporous membrane applications, and trace organic contaminants removal.

Innovative technologies under the consortium research effort included: low pressure reverse osmosis (RO) system, the carbon-aerogel capacitive deionization (CDI) process developed by the Lawrence Livermore National Laboratory (LLNL), microporous membranes, pulsed ultraviolet (UV), median pressure UV, and electron beam (E-beam). Major study findings include the following:

Colorado River Water Salinity Reduction (MWD)

Objective

- Demonstrate salinity reduction by RO and capacitive deionization with carbon aerogel CDI.

Outcomes

- Conventional treatment both with and without ozone and biofiltration produced an effluent water quality suitable for use with RO.
- Carbon aerogel CDI technology has great potential; the technology is in its infancy and requires more development.
- Larger diameter RO elements are required for membrane technology to be implemented on a large scale.
- Significant energy savings can be realized with the use of ultra-low-pressure RO membranes over the previous generation of low-pressure RO membranes.

- Future CDI evaluations should be conducted at bench scale using deionized water supplemented with artificial ionic matrices.

Recommendation

- Additional research is needed to optimize the use of chemical coagulants and membrane antiscalants to decrease the cleaning frequency of membrane processes.

Pulsed UV for Disinfection and Membrane Fouling Control (MWD & OCWD)

Objective

- Demonstrate that pulsed UV is effective in disinfecting water and reducing membrane fouling.

Outcomes

- Pulsed UV disinfected *Cryptosporidium Parvum* in waters with low UV absorbance, and provided a 99.99 percent *Cryptosporidium* inactivation at UV doses <20 millijoule per square centimeter (mJ/cm²).
- Pulsed UV disinfected Heterotrophic plate count (HPC) bacteria in waters with low UV absorbance results in 99.9 percent bacterial inactivation at UV doses >12 mJ/cm².
- Pulsed UV disinfected MS-2 coliphage (virus) in waters with low UV absorbance gives 99.9999 percent virus inactivation at UV doses >50 (mJ/cm²).
- This technology would be significantly less costly than ozone or membrane *Cryptosporidium* reduction technologies.
- If successful, this technology could be an additional barrier to pathogenic organisms in drinking water treatment.

Recommendation

- Further research is needed to validate preliminary results and to produce full-scale pulsed UV reactors that produce drinking water from a variety of sources.

Bromate Formation and Control

Objective

- Reduce bromate generation in water disinfection processes that utilize ozone.

Outcomes

- Lowering pH prior to ozonation was effective in reducing bromate formation, even at *Cryptosporidium*-level disinfection conditions.
- This technology is expensive compared to chlorination because of the large chemical dosages of acid required for pH control.
- Ammonia is a promising bromate control strategy for the Colorado River water at ozone doses required for enhanced *Cryptosporidium* disinfection.
- Hydrogen peroxide is not an effective bromate control strategy at the ozone doses required for 90 percent inactivation of *Cryptosporidium*.

Recommendations

- Additional studies are needed in this area before ammonia can be implemented as a bromate control strategy.

Disinfection of Reclaimed Water

Objective

- Demonstrate that UV disinfection is a viable alternative to chloramination for disinfecting reclaimed wastewater for non-potable use.

Outcomes

- Ultraviolet disinfection is a viable alternative to chloramination, adding chlorine to wastewater in the presence of ammonia, for disinfection of reclaimed wastewater for non-potable reuse.
- UV is much more effective than chloramination in destroying bacteria.
- A 99.99 percent inactivation of indigenous (native to the environment) coliphage was achieved at an UV dose of about 90 milliwatts seconds per square centimeter (mWs/cm²).
- The pilot plant achieved 99.99 percent and 99.9 percent virus inactivation with an UV dose of about 180 and 100 mWs/cm², respectively.
- Disinfection byproducts are insignificant; formaldehyde (a biodegradable DBP) increased from 1 microgram per liter (µg/L, or □g/L) to between 2 to 10 µg/L.

Recommendation

- A standard method for evaluation of UV disinfection systems is needed in order to make comparisons between vastly different UV systems in existence.

Low Pressure Membrane for Reclaimed Water

Objective

- Identify the optimum features of low pressure, microporous membranes that are commercially available.

Outcomes

- Ease of reclaiming water decreased as the quality of the feedwater decreased.
- In general, microporous membrane permeability decreased as the concentration of suspended solids and biochemical oxygen demand (BOD) in the feed water increased.
- Feed pressure to the RO system increased even though most of the precursors for RO fouling, e.g., suspended solids and microorganisms, were removed by the microporous membrane.
- The best option for treating residuals generated by low-pressure microporous membranes is the use of a second stage membrane to concentrate the backwash, thus reducing the volume of backwash by 85 percent.

- Microporous membranes can be incorporated into membrane bioreactors that operate under aerobic conditions that function similarly to activate sludge, clarification, and filtration systems in which air is used in the digestion of organics.

Recommendations

Further research is needed in the following areas:

- Test oxidant tolerant RO membranes and develop more aggressive cleaning regimens to remove organic foulants from RO membranes.
- Investigate the effect membrane geometry and module potting techniques on the incidence of fiber failure in microporous membrane systems.
- Evaluate the performance of a low-pressure microporous pretreatment system that incorporates the optimum design features identified in this task.
- Evaluate the feasibility of using a microporous membrane to separate suspended and soluble BOD, and a semipermeable RO membrane to concentrate the soluble BOD prior to stabilization in an anaerobic reactor.
- Quantify true operating costs of microporous membrane process used as pretreatment for RO based on the scale up of a promising system identified from this task.

E-Beam for Destruction of Trace Organics

Objective

- Demonstrate that high energy injection (E-Beam) is an effective alternative for water treatment.

Outcomes

- OCWD demonstrated at Water Factory 21 that high energy electron beam injection is an effective alternate water treatment method by meeting or approaching Maximum Contaminant Levels (MCLs) on a variety of priority contaminants applied simultaneously for treatment.
- Costs to install and operate the system are also competitive and potentially much lower than conventional and other emerging alternates.
- Treatment on NDMA, an organic carcinogenic contaminant formed as a chloramination byproduct of nitrate and dimethylamine commonly found in wastewater, showed effective removal rates at costs lower than existing chemical destruction technologies.

Recommendation

- Tests at OCWD Water Factory 21 suggest the better application for the electron beam process will be in high flow rate, single, or multiple constituent treatment scenarios for which destruction of contaminants is a desired endpoint.

Before these technologies can be commercially deployed, additional research is needed to complete current ongoing and newly identified research activities. Specific recommendations for each task are shown in Section 3.0 of this Water and Wastewater Technology Demonstration Projects Report.

This research study represents a significant advancement in science and technology for the California water and wastewater industry. Although not quite ready for commercial

deployment in some of these technologies, California municipalities should be informed of these innovative technologies that may be available in the near future and that have the potential to resolve most of the energy and water issues facing California. Additional research and demonstration funding, however, is needed to complete existing work and to accelerate the commercial deployment of these and other innovative technologies.

Abstract

The Water and Wastewater projects were funded jointly by the California Energy Commission, the Electric Power Research Institute (EPRI), and Southern California Edison (SCE) with project management performed by Edison Technology Solutions (ETS). The primary objective was to identify and develop technologies that could help the California water and wastewater industry reduce the cost of water and wastewater treatment, and improve the overall operation at treatment facilities. Metropolitan Water District (MWD) and Orange County Water District (OCWD) were commissioned to perform a six-task research project. Tasks included: desalination of Colorado River water, pulsed ultraviolet (UV) research on disinfection and biofouling control of surface water, reduction of bromate formation in an ozonation process, disinfection of reclaimed water, low pressure membrane for reclaimed water, and electron beam for trace organics removal. The research was successful in demonstrating microporous membrane and low-pressure reverse osmosis (RO) membrane applications in both surface and reclaimed water. Pulsed UV was found to have significant advantages over conventional disinfection chemicals, and can reduce but not entirely replace these chemicals. Electron beam (E-beam) technology also showed promise in trace organics destruction and can be cost competitive in the near future. When commercially implemented, these technologies can greatly reduce current energy consumption in water and wastewater treatment while improving the quality and performance of the product water.

These projects represent significant advancement in science and technology for the California municipal water and wastewater industry. Additional research and demonstration funding, however, is needed to expand on existing promising technologies and to accelerate the commercial deployment of these technologies.

1.0 Introduction

1.1 Background

Southern California relies on a variety of sources for its water: the Colorado River, the State Water Project (SWP) from Northern California, local surface water and groundwater, and reclaimed water. Currently, approximately 60 percent of the water is imported by the Metropolitan Water District (MWD) and the remaining 40 percent is produced locally. Although there are over 5,000 water/wastewater suppliers in southern California, two of the largest suppliers are the Metropolitan Water District (MWD) and Orange County Water District (OCWD). In addition to being two of the largest water and wastewater suppliers in southern California, these water/wastewater purveyors have a reputation for their accomplished research activities. The following are the major issues facing these municipal water/wastewater suppliers.

MWD, the largest water supplier in the United States, believes that salinity is the most important issue facing southern California. According to MWD, salinity in southern California groundwater basins and in some surface water supplies has been slowly increasing in recent years. One reason for this increase is the high salinity of imported Colorado River Water (CRW), with a total dissolved solids (TDS) concentration of 600 to 800 mg/L. High TDS concentrations cause problems for agriculture, industrial processes, and homeowners through corrosion and scaling of plumbing fixtures and appliances. MWD was interested to evaluate breakthrough technologies for desalinating the CRW.

The Orange County Water District (OCWD), since 1976, has been reclaiming 10 million gallons per day (mgd) of secondary effluent with reverse osmosis (RO) at Water Factory 21 for the replenishment of the local groundwater basin. The existing pretreatment for the RO membranes consists of a five-stage high pH-lime clarification/filtration process coupled with a three stage solid waste handling process. The operating costs of this pretreatment are high due to the problems associated with handling the lime. Approximately one ton of lime is used to treat one million gallons of water. Seventy-five percent of the spent lime is recalcined in a multiple hearth furnace for reuse and the remainder is disposed through landfill. The high pH lime process occupies a large plant footprint so that the expansion of the plant capacity will be land intensive. OCWD showed interest in evaluating innovative electrotechnologies to replace their high lime process and simultaneously meet disinfection discharge requirements.

1.2 Purpose of Report

The goal of these projects is to identify and develop technologies that could help local municipalities substantially reduce the cost of water and wastewater treatment. This customer segment represents over 5,000 Southern California Edison water and wastewater customers that cumulatively consume 2 billion kWh of electricity per year. In addition to being one of the most energy intensive industries in Southern California, this customer segment is inundated with restrictive environmental and health related regulations. Cost for water treatment and wastewater disposal has more than doubled in recent years because of the impending regulatory compliance needs and increased cost of electricity.

The projects will evaluate innovative breakthrough technologies that could significantly reduce energy cost, improve water quality and minimize wastewater disposal problems.

1.3 Technical Description

Based on the needs of the municipalities, the research consortium developed a six-task project addressing the main issues facing the municipal water and wastewater industry in southern California. MWD and OCWD equally shared these tasks. MWD's tasks focused on the Colorado River Water, surface supply water, desalination, disinfection, and bromate control. OCWD's tasks were directed at wastewater reclamation, disinfection, microporous membrane applications, and trace organic contaminant destruction. The following is a summary description of each task. Detailed discussion of the task findings can be found in Section 2.0 of this report. A comprehensive report for each task is presented in the Appendixes.

1.3.1 Colorado River Water Salinity Reduction

Capacitive deionization with carbon aerogel (CDI) electrodes is a new and promising technology for removing TDS and impurities from domestic water supplies. Carbon aerogels are unique, porous materials consisting of interconnected uniform carbonaceous particles 3 to 30 nanometers (nm) with small interstitial pores of <50 nm. This structure leads to the very high specific surface area of 2 to 5.4 times 10 to the sixth power times square foot per pound (2.0-5.4 x 10⁶ ft²/lb), low hydraulic resistance, and exceptional electrical conductivity. The aerogel chemical composition, microstructure, and physical properties can be controlled at the nanometer scale, giving rise to unique optical, thermal, acoustic, mechanical, and electrical properties. For detailed discussion of carbon aerogels, consult the following: Farmer, J. C.; J. H. Richardson, and D. V. Fix., 1996, *Desalination with Carbon Aerogel Electrodes – DRAFT*; and Report UCRL-ID-125298, Rev. 1. Lawrence Livermore National Laboratory, Livermore, Calif. (Dec. 4, 1996).

Ultra-low-pressure reverse osmosis membranes have recently been developed which operate at significantly lower pressures (75 to 150 pounds per square inch [psi]) than conventional low-pressure membranes (175 to 450 psi). The reduced pressure results in significant energy savings. The primary objective of this task was to evaluate new and innovative technologies to substantially reduce the cost of desalinating CRW. The task was conducted in two phases. Phase 1 evaluated carbon aerogel CDI and RO for salinity reduction using conventional treatment as pretreatment. And Phase 2 evaluated carbon aerogel CDI and RO for salinity reduction using conventional treatment with ozone and biofiltration as pretreatment.

During Phases 1 and 2, carbon aerogel CDI and RO were evaluated in parallel, first using conventional treatment as pretreatment, and then conventional treatment with ozone/biofiltration. The quality of the pretreated water was characterized, and the performance of carbon aerogel CDI and ultra-low-pressure RO were evaluated as a unit process for demineralization in terms of the following criteria:

- Process throughput, or permeate flux, expressed as gallons/ft²/day
- Energy efficiency, expressed as kWh per 1000 gallons of product water
- Process recovery, defined as the ratio of product water to feed water, expressed as a percentage
- Aerogel recovery after reverse polarity regeneration

- TDS rejection, defined as one minus the ratio of the concentration in the product water to the feed water, expressed as a percentage
- Permeate feed water quality consisting of general inorganics, organics, microbiological quality, and particles
- Operational reliability, defined as the ability to operate at the same flux, product water quality, and process recovery as a function of time during normal operation.

1.3.2 Pulsed Ultraviolet (UV) for Disinfection and Membrane Fouling Control

The purpose of this task was to:

- Determine the disinfection effectiveness of pulsed UV radiation using microbial indicators for Heterotrophic plate count (HPC) bacteria and virus (MS-2).
- Determine the effectiveness of pulsed UV radiation for *Cryptosporidium* (*C.*) inactivation.
- Evaluate disinfection byproducts (DBPs) formed by pulsed UV irradiation.
- Evaluate the potential for pulsed UV to control biological fouling on membrane surfaces.

HPC bacteria and MS-2 virus were selected due to their previous use in UV evaluations and because of their relative ease in quantification. This task also examined the ability of pulsed UV to disinfect *C. Parvum* oocysts as measured by a human cell-culture infectivity assay. Collectively, these measurements were used to establish the disinfection effectiveness of pulsed UV compared to other UV disinfection systems.

MS-2, HPC, and *C. Parvum* batch samples were irradiated at varying distances from a flash bulb. The bench-top treatment chamber was designed to examine the effects of pulsed UV at distances up to 5 inches (in.) away for plug-flow tests and 6 in. away for batch tests. All experiments used effluent water from Metropolitan's 6-gallon per minute (gpm) pilot plant which treated California State project water using pre-ozonation, coagulation/flocculation, sedimentation and biological filtration. Batch tests treated 4-mL samples that were isolated from the process water in the treatment chamber by a quartz sleeve. In all experiments, 2.4 gpm of the pilot-plant effluent was passed through the inner treatment chamber to simulate a full-scale treatment chamber environment.

1.3.3 Bromate Formation and Control

This study focuses on the formation, control, and removal of bromate formed as a result of using high doses of ozone for purposes of enhanced disinfection. Experiments utilize both bench-scale and demonstration-scale research platforms. Specific objectives of this study are to:

- Determine, on a demonstration scale, bromate production in Colorado River Water, California State Project Water (SPW), and 1:1 blends of the two source waters under conditions of enhanced disinfection for *Cryptosporidium*.
- Assess the effectiveness, on a demonstration scale, of the following bromate control strategies: pH reduction, hydrogen peroxide addition, and ammonia addition.
- Assess the effectiveness of staging ozone doses over two or more contact chambers for bromate control.

- Determine the effectiveness of ferrous iron addition, subsequent to ozonation, for removal of bromate.
- Evaluate hydrogen peroxide, sodium bisulfite, and calcium thiosulfate as quenching agents for the purpose of eliminating ozone residuals in the contactor effluent.
- Using a bench-scale ozone contactor, determine the effects of seasonally varying parameters such as temperature, total organic carbon (TOC), and alkalinity on the production of bromate in CRW and SPW, at bench-scale.

1.3.4 Disinfection of Reclaimed Water

Many emerging UV technologies are now available for disinfection of reclaimed wastewater. These new technologies use various reactor configurations and lamp types including low pressure-high intensity, medium pressure-high intensity, and pulsed power lamps. These UV technologies must be evaluated equally and compared to proven disinfection methods. A proper evaluation method will allow the Orange County Water District (OCWD) to make the best decision on what type of UV disinfection to use for its current and future wastewater reclamation.

The goals of this task were to compare disinfection performance of chlorine to ultraviolet radiation (UV) and to investigate new alternatives for wastewater reclamation. In addition, OCWD wanted to evaluate if UV can be a cost-effective disinfection process. Towards this end, OCWD developed a test protocol to:

- Compare ultraviolet disinfection to chlorine disinfection for municipal wastewater reclamation.
- Develop microbiological methods and capabilities necessary to evaluate disinfection alternatives.
- Implement UV disinfection testing protocol.
- Develop testing methods to compare UV disinfection protocol with the collimated beam apparatus regarding UV disinfection efficacy and dose response curves.
- Select and test a range of UV systems on three classes of reclaimed water and perform surrogate virus seeding studies.
- Evaluate effectiveness of pulsed UV for disinfection of reclaimed wastewater.

1.3.5 Low Pressure Membrane for Reclaimed Water

Low-pressure microporous membrane processes, such as ultrafiltration (UF) and microfiltration (MF), are very effective for the removal of suspended solids and large colloids from secondary effluent prior to demineralization using reverse osmosis. However, unlike the more mature semipermeable membrane processes, reverse osmosis (RO) and nanofiltration (NF), microporous membrane systems do not conform to a standard design or operating regimen. Every commercially available microporous membrane system is unique. Proprietary microporous membranes with distinct symmetry, pore size, porosity, and mechanical properties are assembled into modules with different packing densities by a variety of manufacturing techniques. These modules are integrated into systems that operate under different process flow profiles, loading rates, backwash mechanisms, and cleaning regimens.

This non-uniformity complicates the evaluation of microporous systems during the preliminary design and equipment selection process. The research was initiated to identify optimum features of low pressure microporous membranes for wastewater reclamation application with the intention of steering the industry towards more efficient separation processes.

OCWD's objective was to validate this technology on a pilot scale. After validation, future construction of an efficient commercial water reclamation facility utilizing this technology would become feasible. Specifically, OCWD decided to use a test protocol that it previously developed for data gathering:

- Identify the optimum features of low pressure, microporous membrane process through comparative testing of commercially available UF and MF systems on clarified secondary effluent. Assess the impact of membrane properties, module design, process flow profile backwash mechanism and cleaning regimen on filtrate water quality and fouling of the reverse osmosis membrane.
- Investigate the use of both microfiltration and conventional settling to reduce the volume of the residual waste stream, backwash, generated by microporous membrane processes.
- Investigate the feasibility of integrating a low pressure, microporous membrane system into the wastewater treatment process for the following purposes:
 - Management of peak flows through secondary system.
 - Evaluate performance of UF and MF at elevated loading rates to simulate the short term increased flow conditions that exist during peak storm events.

1.3.6 Electron Beam Destruction of Trace Organics

The Orange County Water District considered a pilot evaluation of the performance of an Electron Beam (E-Beam) for the destruction of trace organics in their treated effluent. This project was funded as an "add-on" by the Commission with recommendations from ETS to provide a preliminary assessment of the effectiveness of E-Beam for trace organics destruction. By the time the contract was authorized and approved by the Commission, OCWD only had approximately 1 month to perform the research and prepare a simplified report to satisfy Commission contractual requirements. The following is a summary of the preliminary research findings by OCWD.

The electron beam technology is an innovative treatment process for contaminated water and other media. The process uses high energy electron injection on aqueous streams and provides instantaneous contaminant degradation, and ultimate destruction.

The purpose of this task is to evaluate the effects of high energy electron injection on a variety of constituents on various water sources utilizing the design developed by OCWD's Water Factory 21 pilot test protocols. The test results were evaluated to determine actual destruction efficiency, understand the effects of local water chemistry on system performance, and develop economics for comparison against other treatment processes.

2.0 Detailed Task Discussions

The following is a summary of the discussions on each task findings. A complete report for each task prepared by individual researchers is included in the Appendixes.

2.1 Colorado River Water (CRW) Salinity Reduction

2.1.1 Salinity Reduction Using Conventional Treatment

Conventional treatment as pretreatment for the Reverse Osmosis (RO) and Capacitative De-Ionization (CDI) was conducted for 2,100 hours. The chemical feeds prior to flocculation included: 2.0 mg/L ferric chloride; 0.5 mg/L cationic polymer; and, 1.0 mg/L sodium hypochlorite. The influent pH was adjusted from approximately 8.0 to 6.8 using sulfuric acid to control calcium carbonate scaling in the RO system. Initially, sulfuric acid was added to the conventional treatment influent; however, the acid injection point was moved to the RO influent for subsequent phases to simulate projected full-scale operation.

2.1.1.1 Pretreatment

Table 2 shows the inorganic and microbial analysis of the conventional treatment plant influent and effluent. Conventional treatment did not significantly alter the effluent concentrations for most inorganic constituents. The exceptions were alkalinity (131 mg/L as CaCO₃ to 82 mg/L as CaCO₃) and sulfate (233 mg/L to 280 mg/L). Both of these changes were caused by the addition of sulfuric acid at the plant influent to control calcium carbonate scaling in the RO unit. Conventional treatment reduced heterotrophic plate count (HPC) bacteria by 64 percent and coliform bacteria to below the method detection limit.

Figure 5 shows the conventional treatment performance as measured by effluent turbidity, particle counts, and silt density index (SDI). Influent particle count and turbidity data were 7300 particles/mL (the average particle counts for 100 percent Colorado River water from February 1 through February 7, 1999) and 2-3 nephelometric turbidity units (NTU) at Typical Weymouth Filtration Plant influent turbidity; average data not available. The box-and-whisker plots show median, 75th percentile, 25th percentile, maximum and minimum values. The conventional treatment plant daily median turbidity and particle-count data were (for samples taken once per minute) 0.08 NTU (range, 0.05 to 2.7 NTU) and 33 particles/mL (range, 0.05 to 5000 particles/mL), respectively. The silt density index (SDI) ranged from 2.4 to 5.2, with a median value of 3.1. The high effluent turbidity occurred immediately after backwash or during air binding episodes at the end of filter runs. Both of these conditions generated false-high turbidities because the turbidimeter was mounted on a section of pipe influenced by the backwash scouring and by bubble entrainment during air binding episodes. These episodes only lasted for 30 minutes. The turbidimeter has been relocated so that the false-high readings were minimized.

2.1.1.2 Reverse Osmosis

Pilot testing with the RO unit was conducted for 2,100 hours using conventional treatment as pretreatment. Antiscalant (1.4 mg/L Pretreat Plus 2000; King Lee Technology, San Diego, Calif.) and pH control with sulfuric acid (up to 45 mg/L) were used to minimize scaling.

Figure 6 shows how the water permeation coefficient (WPC) was affected by treatment conditions. The WPC represents the amount of water that passes through a square foot of membrane each minute (permeate flux) for each psi of net driving pressure at 77°F (25°C). The WPC allows direct comparison of operating data regardless of feedwater composition, operating pressure, water flux, or system recovery. In normal operation, the WPC will decrease over time due to particulate, precipitative, or biological fouling. After 500 hours of operation, the WPC slowly decreased to 14-16 gpm/ ft²-min-psi and remained in that range until 1,700 hours. At this point, the WPC exhibited a sharp, 10 percent decline over the final 400 hours of operation. Physical examination of the lead and final elements revealed that there was a significant build-up of foulant on the membrane in the lead element that reduced water flux through the membrane.

Figure 6 also shows how the salt permeation coefficient (SPC) was affected by treatment conditions. The SPC is a measure of the salt diffusion rate through the membrane normalized to 77°F (25°C). The SPC measures the efficiency of a membrane independent of recovery rate, permeate flux, and operating pressure. Normally, the SPC slowly increases over time, and the salt concentration in the permeate doubles within 3 to 5 years. A significant, short-term increase in the SPC may indicate damaged membranes and is usually accompanied by an increase in the WPC. The SPC may decrease in new or freshly cleaned membranes. In addition, a low SPC value may result from changes in pH or increases in the multivalent ions in the feed water.

The salt permeation coefficient ranged between 0.5 and 1.5x10⁻⁵ cm/sec throughout this phase of the study. Figure 6 shows a gradual decline in the SPC which plateaued at 800 hours. The gradual decline may have been caused by a change in the feed water chemistry. Prior to this study, feed water was switched from blended water (75 percent Colorado River Water and 25 percent State Project water) to 100 percent Colorado River Water, which was higher in multivalent ions. In addition, recirculating a portion of the RO concentrate also increased the total ion concentration in the feed stream. Both modifications increased the conductivity of the feed water into the RO system.

Table 3 shows a summary of selected inorganic and microbial water quality analyses taken under current operating conditions: 100 percent CRW; 85 percent water recovery; and 1.0 gpm concentrate recirculation (which increased the cross-flow velocity by 40 percent). Based on concentration factors, ratio of brine concentration to RO influent concentration, the RO unit performed as expected. However, both aluminum and iron exhibited low concentration factors, 3.1 and 3.4, respectively. These low concentration factors indicate that some of the aluminum and iron were unaccounted for, suggesting that aluminum and iron were retained by the membrane. Aluminum or iron silicate fouling was not suspected here because rapid flux decline was not observed in the last array.

The microbial data (Table 3), indicated that neither fecal nor total coliforms were present in the RO permeate. However, HPC bacteria was found in both RO permeate and brine. The HPC in the permeate was likely due to regrowth, since bacteria are too large to permeate through the membrane. The RO brine HPC counts were 116 times higher than the RO influent water, indicating possible biological growth on the membrane surface or downstream piping.

At 2,100 hours of operation, the conventional treatment as pretreatment phase of the project was concluded. The first and last element of the RO skid were removed and an in-house

autopsy was performed. The autopsy revealed a dark, reddish-brown, gelatinous foulant on both elements. The foulant deposition was significantly more pronounced on the first element of the RO skid. Elemental analysis of the foulant layer revealed greater than 50 percent iron content; (Table 4). It was theorized that residual ferric hydroxide coagulant passed through the conventional treatment process and reacted with the antiscalant to form a precipitate which deposited on the RO elements. The degree of deposition appears to be directly related to the position of the element in the RO system. Ning and Stith showed that low levels of iron and silica can interact with organic polymers e.g., naturally occurring matter (NOM) or antiscalant and precipitate on both the prefilter and RO elements. Similar results were also reported by Filteau at an RO plant in Tustin, California, (Filteau, G.H., Nitrate Removal from Contaminated Groundwater through Reverse Osmosis, AWWA Membrane Technology Conference, Baltimore, Maryland, 1993). Iron fouling was not seen when alum and polyaluminum chloride were used as the pretreatment coagulants. Therefore, it was assumed that the acrylate-based antiscalant was incompatible with ferric chloride and caused the membrane fouling. A non-acrylate-based antiscalant will be used in future studies to assess this possible interaction.

2.1.1.3 Capacitive Deionization

Batch-Mode

Figure 7 shows a typical batch-mode CDI experiment with regeneration. For batch-mode operation, 3.3 gal of conventional pretreated CRW was recirculated at 200 mL/min without voltage applied until the reservoir conductivity and the effluent conductivity approached steady-state. This marked the beginning of the experiment, shown as 0.0 hours on (Figure 7). A 14 percent reduction in conductivity (954 to 796 microsiemen per centimeter [$\mu\text{s}/\text{cm}$]) was observed during the circulation of the initial test water without applied voltage. Thus, physical sorption of the feed water removed some salinity. After reaching steady state, 1.6 volts was applied to the CDI stack continuously for the duration of the sorption phase. The effluent conductivity decreased rapidly as ionic species in solution were removed through electrosorption to the electrodes. After 2.75 hours, no appreciable decrease in conductivity was observed and the experiment was terminated (453 $\mu\text{s}/\text{cm}$).

Table 5 shows the water quality analyses conducted for the above batch-mode experiment. The TDS decreased 51 percent from 618 to 303 mg/L. The capacity of the carbon aerogel material for salinity removal was 6.6 mg TDS/g carbon aerogel. This represents a two-fold increase in carbon aerogel capacity over previous experiments. However, the theoretical sorption capacity for carbon aerogel was 73 mg TDS/g carbon aerogel. The increase in sorption capacity may be due to extensive flushing of the carbon aerogels prior to the experiment. The CDI unit showed little reduction of microbiological species (32 percent reduction of HPC). The regenerant water was not sampled for bacteria.

Removal of monovalent ions such as sodium (62 percent) or chloride (80 percent) was greater than divalent ions such as calcium (49 percent) or sulfate (34 percent). Further evaluation will be needed to confirm this surprising result and explain this phenomena.

During regeneration, the water pump was stopped and the original test water was left in place 3.5 hours, as recorded on Figure 7. The CDI unit was shorted out by connecting both terminals with a shorting cable. After the stack voltage dropped to 0 v, the power supply leads to the CDI

unit were reversed and a reverse polarity voltage equal to the sorption phase voltage was applied for 5 minutes to drive off sorbed constituents. The effect of polarity reversal produced a dramatic decrease in regeneration time for the carbon aerogel unit (Table 6). Without polarity reversal (i.e., electrically grounding the CDI unit and then reversing the current through the unit to desorb ions from the carbon aerogels back into solution) the regeneration time was greater than 4 hours.

After 5 min of reversed current, the unit was again shorted out and the water pump restarted at 200 mL/min. Effluent conductivity rose rapidly as ions left the aerogel surface and re-dissolved (recombined) in solution. About 85 to 90 percent of the ions (as indicated by the conductivity reading) were recovered in approximately 1 hour of regeneration (shown at 4.5 hours on Figure 7). The regeneration cycle was completed after a stable asymptote was again reached (682 $\mu\text{s/cm}$) at 5.75 hours. Approximately 71 to 76 percent of most constituents were desorbed from the CDI stack at this time (Table 5). Notable exceptions were silica (87 percent recovery) and potassium (57 percent recovery). Silica exhibited poor sorption (20 percent removal) onto the carbon aerogels and the high recovery indicates that silica—a poorly charged ion—was not removed to any large degree by CDI. It was unknown why potassium exhibited such low recovery (57 percent) during regeneration.

The unit was then turned off, and stored over night. After 20 hours, the test water was again recirculated through the unit where additional salts were desorbed from the CDI stack (conductivity increased from 682 to 774 $\mu\text{s/cm}$). After the second regeneration period, 19 percent of TDS was unaccounted for. Additional experiments indicated that the regenerant water could be reused at least three times without any observed decrease in CDI performance (Table 7). By reusing the regenerant water three times, a water recovery of at least 75 percent was estimated. Currently, it is unknown how often the regenerant water can be reused before scaling occurs during the regeneration cycle. The ability of antiscalants to increase recoveries is also unknown.

Continuous-Mode

Figure 8 shows a typical continuous-mode CDI experiment with regeneration. Continuous-mode operation was conducted simply by flowing the test water once through a pre-charged (1.6 v) stack. Initial volume and conductivity of 100 percent CRW were 4.8 gallons and 920 $\mu\text{s/cm}$, respectively. The flow rate for continuous mode was 100 mL/min.

Because of mixing within the stack and the unit's inherent low capacity, the total effluent conductivity increased rapidly. Complete breakthrough (779 $\mu\text{s/cm}$) was observed at 3.0 hours (complete breakthrough was defined as 779 $\mu\text{s/cm}$ because of system losses as shown during batch tests). The sorption capacity of the carbon aerogel material under continuous-flow operation was 7.9 mg TDS/g carbon aerogel.

At 3.0 hours, the complete sample volume (4.8 gallons) had passed through the CDI unit, hence the conductivity measurements after this point fluctuated dramatically due to a no-flow condition. After the total volume of test water was sent through the stack, the regeneration cycle was conducted. For regeneration, the original test water was re-pumped into the unit. Once the stack was refilled, the same regeneration procedure for batch-mode was conducted. As with batch-mode operation, the regeneration cycle of approximately 1 hour yielded 80 percent

recovery of salts (based on a final regenerant conductivity of 730 $\mu\text{s}/\text{cm}$). Additional flushing of the CDI unit over subsequent days was needed to completely restore the carbon aerogels.

Table 7 shows a continuous mode experiment conducted using three CDI units to determine the effect of increasing the carbon aerogel surface area. The results indicate that by increasing the number of CDI stacks (carbon aerogel surface area) the salinity removal was increased. However, this experiment did not show a linear relationship between the number of CDI stacks and salinity removal. This may indicate that CDI becomes less efficient in removing salts as the salt concentration decreases.

2.1.2 Phase 2: Salinity Reduction Using Conventional Treatment and Ozone/Biofiltration

2.1.2.1 Pretreatment

Table 8 shows the inorganic and microbial analysis of the conventional treatment plant influent and effluent. Unlike the conventional treatment phase, acid injection was prior to the RO inlet rather than prior to the conventional treatment plant's flocculation basin. Therefore, conventional treatment with ozone/biofiltration did not significantly alter the effluent concentrations for all inorganic constituents.

Conventional treatment with ozone/biofiltration reduced heterotrophic plate count (HPC) bacteria by 92 percent and coliform bacteria to below the method detection limit. However, the sampling location for the pretreatment effluent was post chloramination. Chloramination likely inactivated a large percentage of bacteria leaving the filter; thereby preventing them from being accurately counted. Therefore, the amount of bacteria leaving the pilot plant may be significantly higher than those indicated by the data. To accurately determine HPC count, sampling for bacteria will need to be done prior to chloramination. Using biologically active filters with post chloramination did not significantly alter the HPC counts versus conventional filtration without chloramination (515 colony forming units [CFU]/mL versus 707 CFU/mL). Operating the filters both conventionally and biologically reduced the total coliform and fecal coliform counts to below the method detection limit.

Figure 9 shows the conventional treatment with ozone/biofiltration performance as measured by effluent turbidity, particle counts, and silt density index (SDI). Influent particle count and turbidity data were 3,261 particles/mL (average particle counts for 75/25 percent blend of Colorado River water and State Project water) and 2-3 NTU (Typical Weymouth Filtration Plant influent turbidity; average data not available), respectively. The box-and-whisker plots show median, 75th percentile, 25th percentile, maximum and minimum values.

The conventional treatment plant operated with ozone and biologically active filters generated daily median turbidity and particle-count data of (for samples taken once per minute) 0.07 NTU (range, 0.05 to 0.3 NTU) and 223 particles/mL (range, 2.2 to 11105 particles/mL), respectively. The SDI ranged from 2.1 to 3.7, with a median value of 3.0. Data for turbidity and SDI were comparable to those obtained during the conventional phase of the project. However, particle count data were a full order of magnitude higher for conventional treatment with ozone/biofiltration (mean 223 particles/mL) than conventional treatment alone (mean 23 particles/mL). This change may be due to the ozone/biofiltration process changing the number and size distribution of the particles, the reduced effectiveness of the pretreatment

process resulting from the operation at a higher pH, or from the bacteria sloughing off the dual media filter. Further work is needed to resolve this uncertainty.

2.1.2.2 Pretreatment Reverse Osmosis

Figure 10 shows the water permeation coefficient (WPC) and salt permeation coefficient (SPC) for the complete RO system. The overall WPC ranged between 14 and 19×10^{-5} gal/ ft²-min-psi. No appreciable decline in WPC was observed through out the test run. However, the differential pressure across the first array exceeded 28 psi after 860 hours of operation. The RO unit was taken off-line at this time to clean the elements and to modify the system with two 2.5 inch elements so that recirculation would not be required to maintain high flow velocities through all elements. A review of the data for each array did not reveal any significant decline in WPC. The salt permeation coefficient ranged between 0.6 and 0.9×10^{-5} cm/sec throughout this phase of the study. This finding was within manufacturer's guidelines.

Table 9 shows a summary of selected inorganic and microbial water quality analyses taken under current operating conditions: a blend of 75 percent CRW and 25 percent SPW; 85 percent water recovery; and 1.0 gpm concentrate recirculation. Based on concentration factors (CF) [ratio of brine concentration to RO influent concentration], the RO unit performed as expected. However, a statistical t-test indicated that for all CFs (excluding barium, aluminum, and iron) the average CF for conventional treatment with ozone/biofiltration (CF 5.6) was significantly less than those observed under the same operation conditions with conventional treatment (CF 6.0) [$\alpha = 0.05$]. The lower CFs resulted from the lower rejection or higher salt passage by the membrane during the ozone biofiltration pretreatment study. The additional passage of TDS (2.8 versus 2.2 percent) reduces the degree of concentration. As during previous studies, the concentration factor for aluminum was lower than other ions (CF 4.3). These data indicate that aluminum was retained by the membrane.

The microbial data, as shown in Table 9, indicate that neither fecal nor total coliforms were present in the RO permeate. However, HPC bacteria was found in both RO permeate and brine. Analysis of the number of HPC bacteria entering and leaving the RO, both as permeate and brine, indicate that bacteria were retained by the membranes; possibly fouling the membranes. Membrane fouling of the lead elements was evident by a greater than 30 psi increase in differential pressure across the first array. This finding, however, was not accompanied by a decrease in WPC across either the total RO system or the first array. Element autopsies of the lead and final RO elements were performed to identify the foulant. During the removal of the first element from the vessel, a light brown floc was observed in the residual water in the RO vessel indicating the possible presence of either organic or biological material. The foulant was lighter in color than the foulant observed during the previous autopsy on conventionally treated water. A visual inspection of the lead membrane revealed a thin coating of a reddish-brown foulant. However, the amount of foulant found on the lead element membrane was significantly less than that observed during conventional treatment phase. The amount of foulant on the membrane surface did not appear sufficient to cause a 30 psi increase in pressure across the first array. A nominal coating of a similar foulant was observed on the terminal element as well. However, the magnitude of contamination was significantly less than that found on the lead element.

The Energy Dispersive Spectroscopy (EDS) analysis of the foulant in Table 10, indicated that the amorphous layer was composed primarily of aluminum and iron silicates (silica [23 percent], aluminum [23 percent], and iron [18 percent]). The 26 percent sulfur content is most likely from the polysulfone membrane support layer. The silicates may be colloidal in nature from the source water, or possibly precipitated silicates from the terminal element recirculated to the head of the RO system. The former explanation might be more probable given the previous experience with aluminum silicate fouling of the terminal elements during the previous conventional treatment. Scanning electron micrographs (not shown) showed numerous bacteria and biological material deposited on the membrane surface. A review of the lead element visual, operational, and EDS data suggests that a combination of particulate and biological fouling had occurred. The 2.0 mg/L chloramine residual in the feed water did not completely deter the adhesion of bacteria onto the membrane surface. Prior data collected using conventional treatment as pretreatment did not reveal the presence of bacteria on the membrane surfaces. Therefore, conventional treatment with ozone/biofiltration may be more prone to biological fouling than conventional treatment alone.

The terminal membrane surface was mottled with aluminum silicates, as determined by EDS on Table 10. However, operational data did not reveal any loss of flux or increase in pressure across the last array due to these deposits.

2.1.2.3 Capacitive Deionization

Due to CDI's poor performance during the conventional treatment phase of the project, no tests using conventional treatment with ozone/biofiltration as pretreatment were conducted. The CDI units were placed in storage until Lawrence Livermore National Laboratory (LLNL) could supply new and improved test units. The scope of the CDI portion of the project was altered to include a more fundamental, bench-scale understanding of CDI's salinity removal potential. This additional work will be completed in the near future and may be reported in the Final Report.

2.1.3 Analysis of Results

A true comparison between RO and CDI cannot be made at this time due to incomplete data. However, a preliminary comparison follows for informational purposes only. Table 11 shows a preliminary comparison of RO and CDI evaluation criteria. The RO performance data were based on 100 percent CRW, 85 percent water recovery, and 1.0 gpm concentration recycle. The average operating pressure under these conditions was 125 psi, ranging from 113 to 132 psi. The CDI unit operated in batch mode with 100 percent CRW and used a regeneration cycle which included a 1.6 volt polarity reversal for 5 min. To date, RO demonstrates superior performance over CDI in terms of permeate flux, energy consumption, process recovery, salt rejection or removal, permeate feed water quality, and operational reliability. Process recovery for CDI was based on treating 9 gal (over three consecutive runs) while using 3 gal of regenerant. The maximum process recovery for CDI has yet to be determined. The operational reliability for CDI was deemed poor due to its low regenerative capacity and its inability to operate continuously. It should be noted that CDI technology was still in its infancy, while RO has been under development for more than 30 years.

Table 11 also shows the performance criteria for RO using conventional treatment with ozone/biofiltration as pretreatment. When using ozone/biofiltered water, the RO operated at 113 psi versus 125 psi for conventionally pretreated water. This resulted in 1.66 kWh per 1000 gal and 1.37 kWh per 1000 gal for conventionally treated and conventionally treated with ozone/biofiltration as pretreatment, respectively. However, it should be noted that the RO unit fouled three times faster using conventional treatment with ozone/biofiltration than conventional treatment due to biological fouling. Over time, this higher fouling rate will result in higher operational cost for ozonated/biofiltered water due to increased RO downtime and increased cleaning frequencies.

The difference in average operating pressure between the two evaluation phases may be attributed to different cleaning regimes prior to testing. Cleaning prior to the conventional treatment phase consisted solely of an acidic cleaner, while the conventional treatment with ozone/biofiltration phased used a combination of acidic and caustic cleaners. Results from our laboratory showed that acidic cleaning alone was insufficient in completely removing all deposits from the membranes. Therefore, when treating CRW with conventional treatment and RO, both an acidic and caustic cleaner should be used.

Both conventional treatment with and without ozone/biofiltration produced effluent waters that satisfied membrane manufacturer's guidelines for turbidity (<1 NTU), and SDI (<5 SDI). However, each pretreatment was hampered by inherent operational liabilities. Chemicals added either to the conventional treatment plant (i.e., coagulants) or prior to RO (e.g., antiscalants) can have adverse interactions with various water quality constituents. Alum, a common coagulant, imparts excess aluminum into the RO influent water causing silicate fouling. Ferric chloride residuals may adversely react with acrylate-based antiscalants to form colloidal particulates that impair RO performance.

Additional problems were introduced by recirculating a portion of the RO concentrate. While not presented in this report, difficulties with concentrate recirculation included:

- increased potential for precipitative fouling in the terminal elements
- seeding of the primary elements with condensation nuclei; thereby, increased potential for particulate and biological fouling in the front of the RO system
- hampered data interpretation.

Concentrate recirculation was instituted to increase the cross-flow velocity in the terminal elements to prevent aluminum silicate fouling without resorting to significant RO design modifications. Future testing should incorporate RO design considerations to increase the cross-flow velocity across all RO elements to avoid silicate fouling in the tail end of the plant without the use of concentrate recirculation.

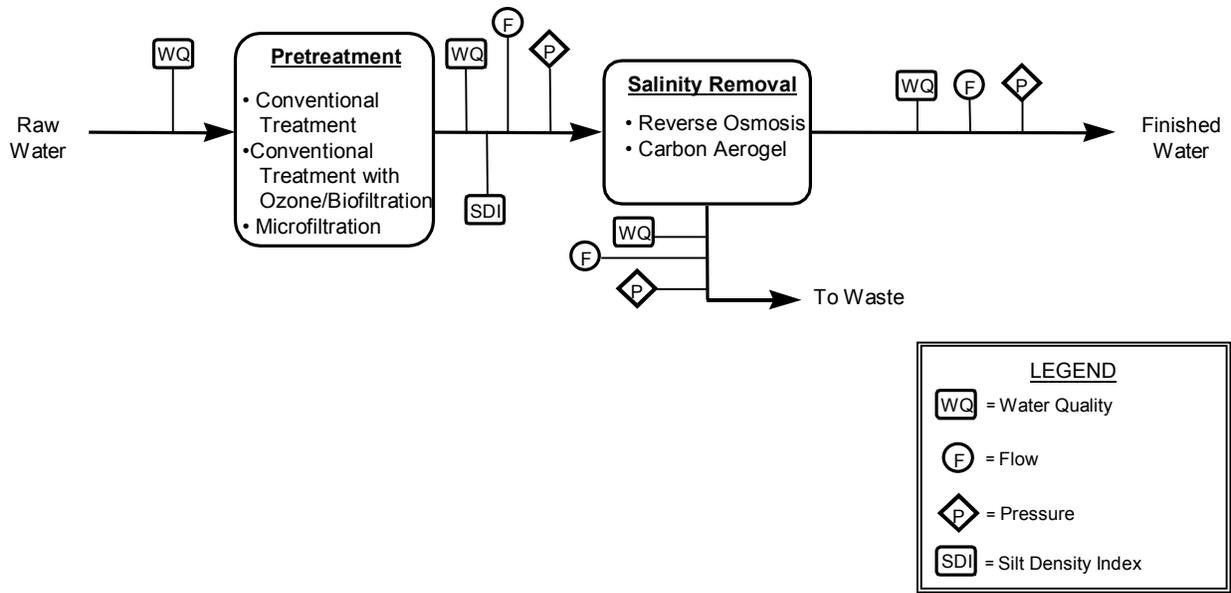


Figure 1. Research Approach

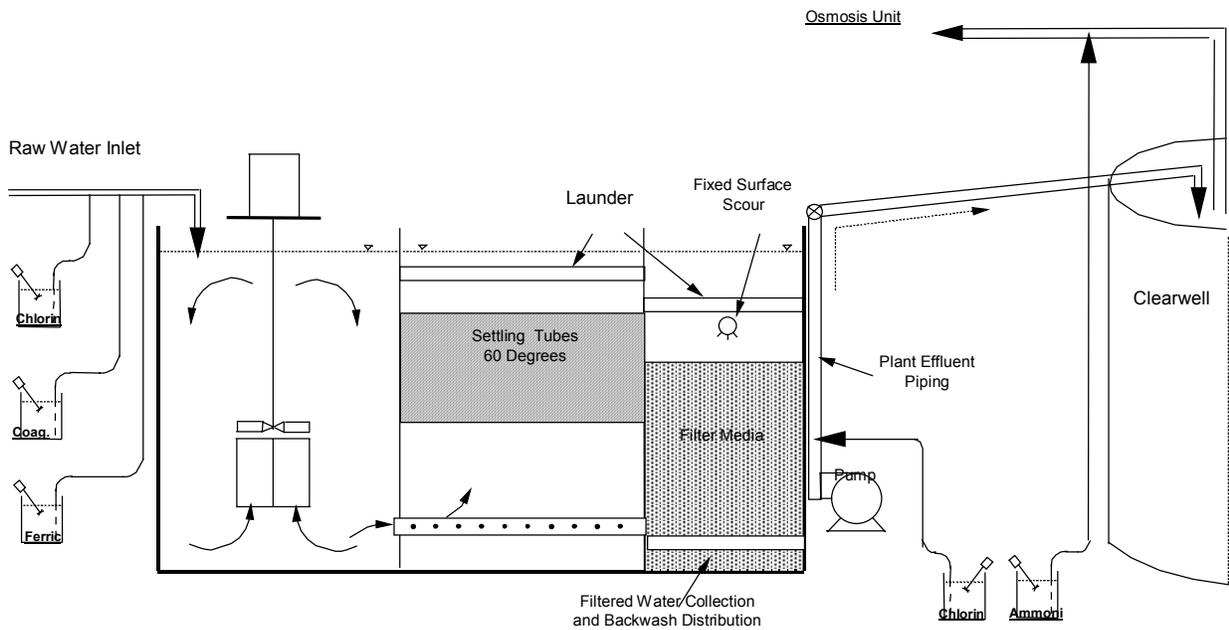


Figure 2. Schematic Diagram of Conventional Treatment Plant

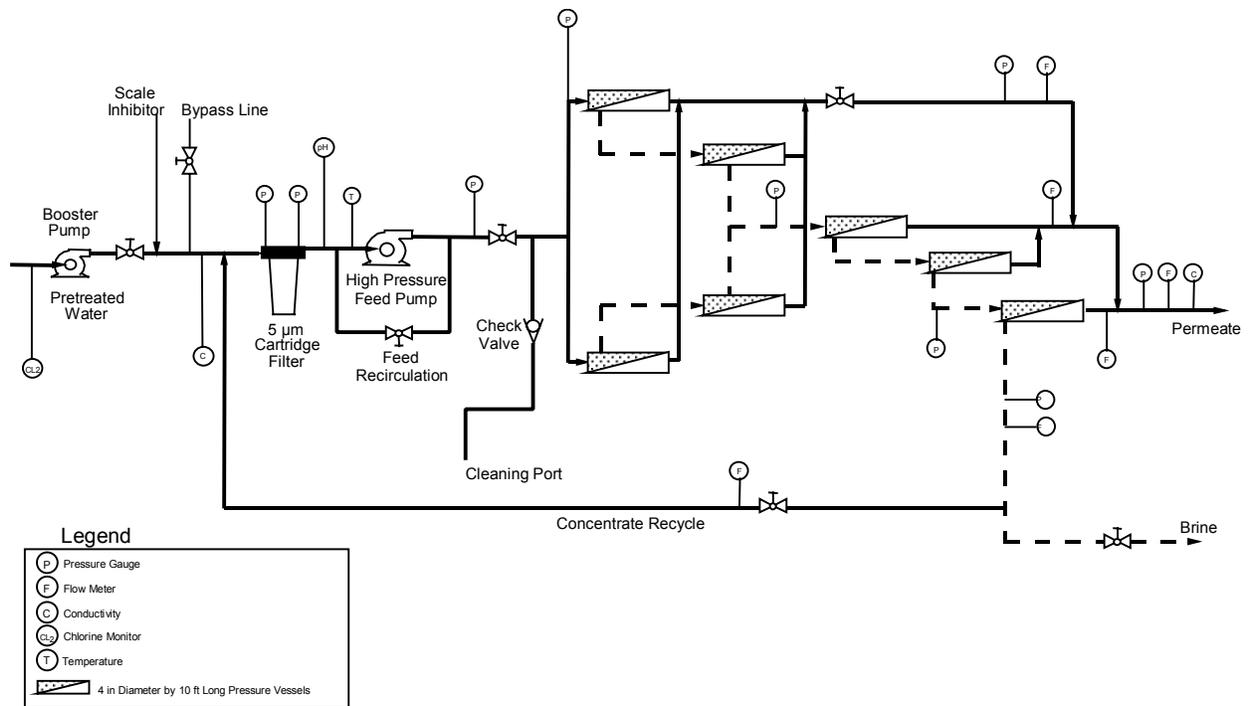


Figure 3. Schematic Diagram of Reverse Osmosis Pilot Unit

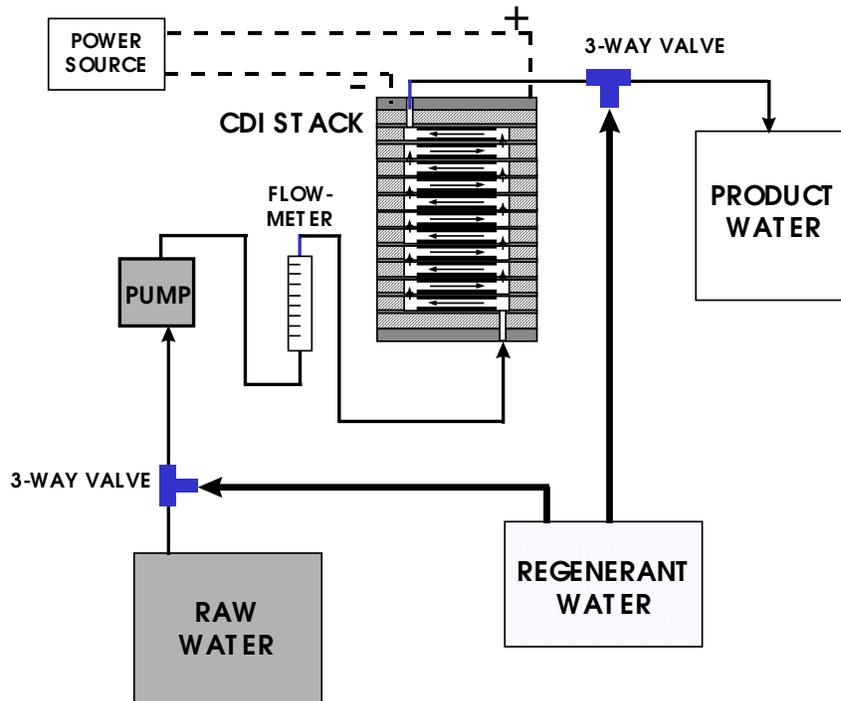


Figure 4. Schematic of Capacitive Deionization with Carbon Aerogel Process

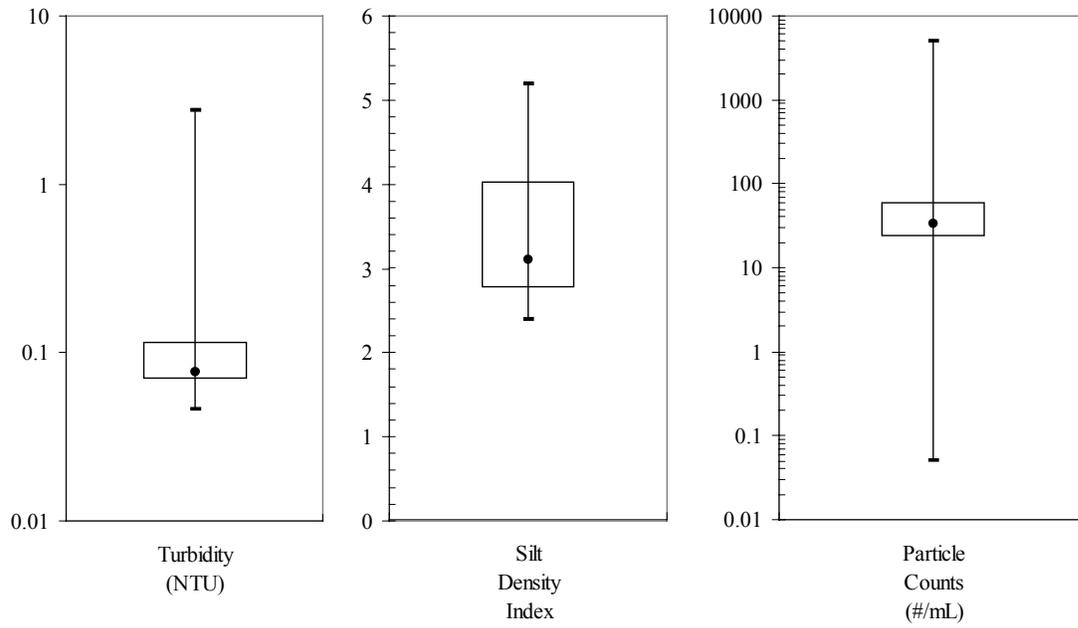


Figure 5. Pretreatment Performance for Phase 1: Conventional Treatment

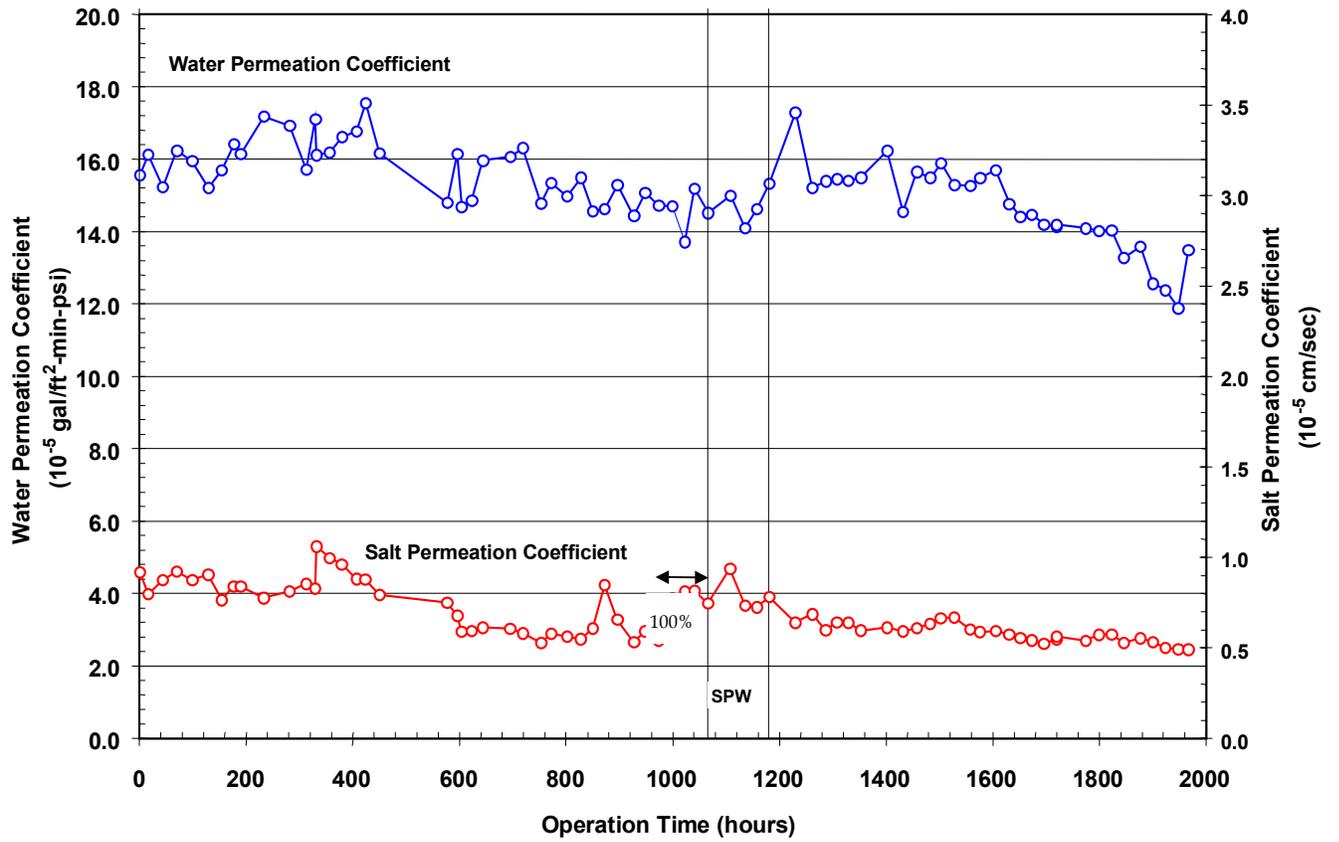


Figure 6. Water Permeation Coefficient and Salt Permeation Coefficient During Phase 1: Reverse Osmosis Using Conventional Treatment

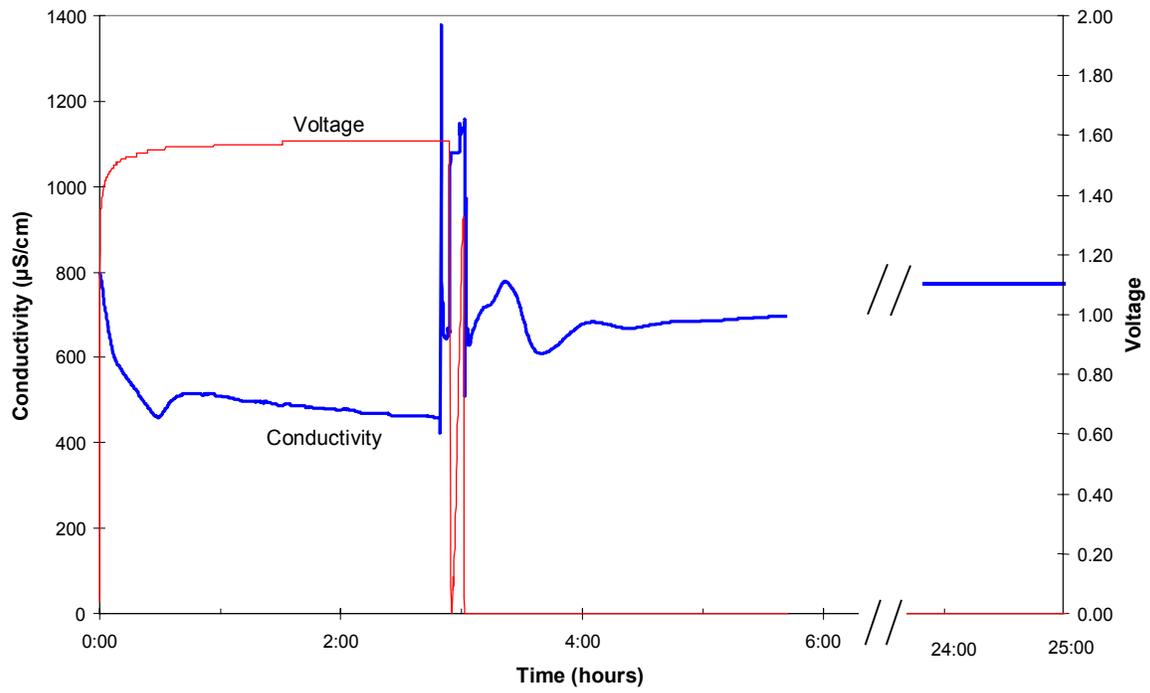


Figure 7. Typical Batch-Mode CDI Experiment with Regeneration

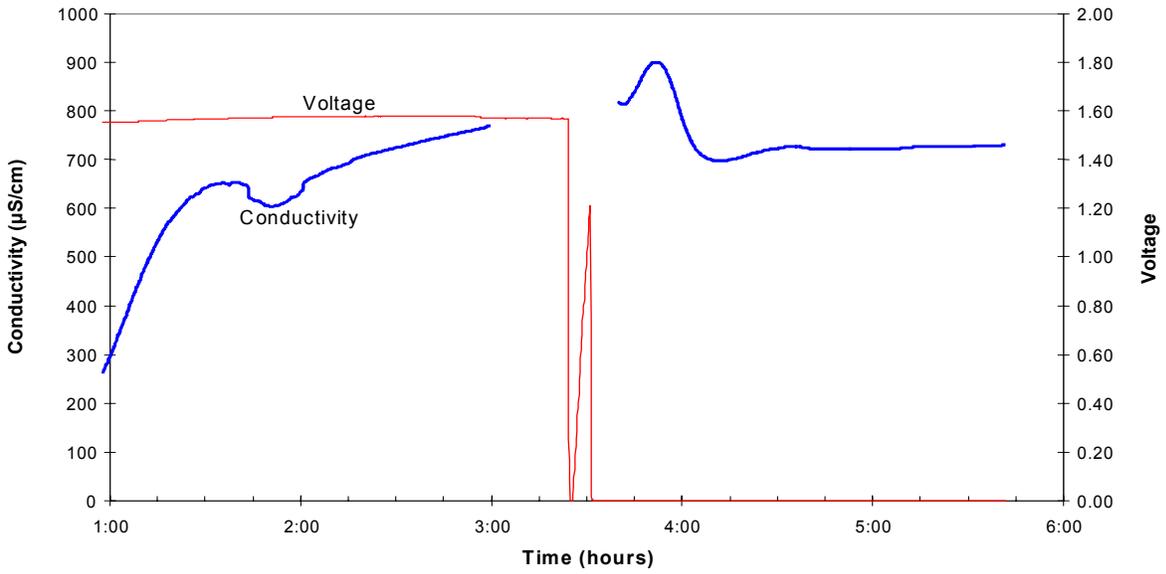


Figure 8. Typical Continuous-Mode CDI Experiment with Regeneration

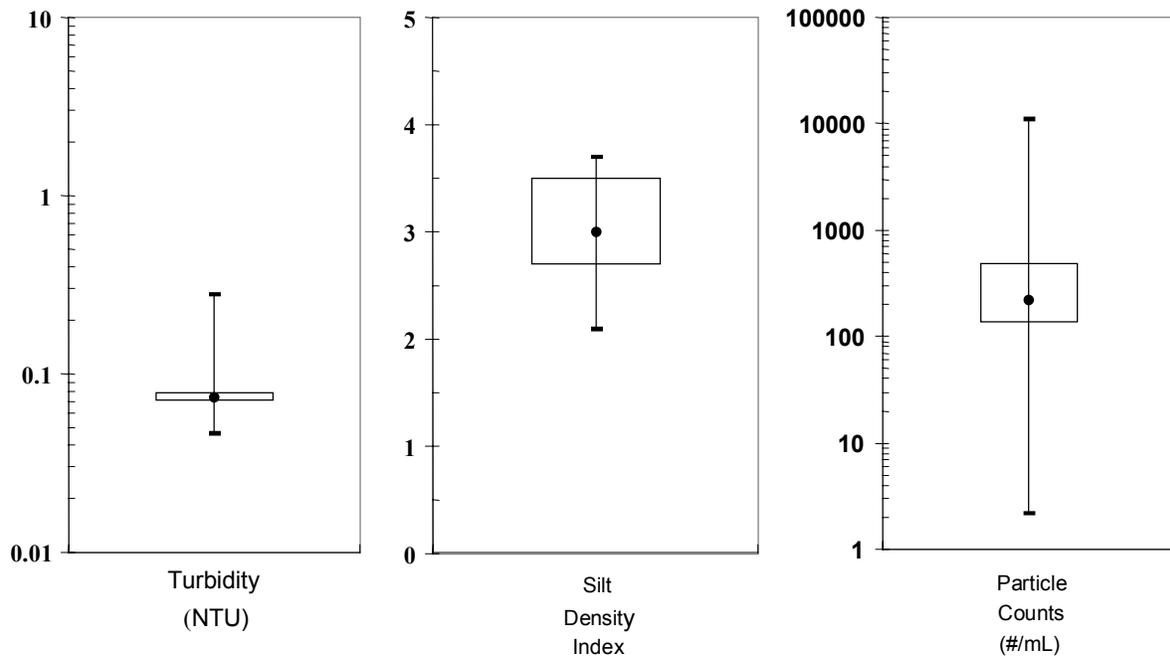


Figure 9. Pretreatment Performance during Phase 2: Conventional Treatment with Ozone/Biofiltration

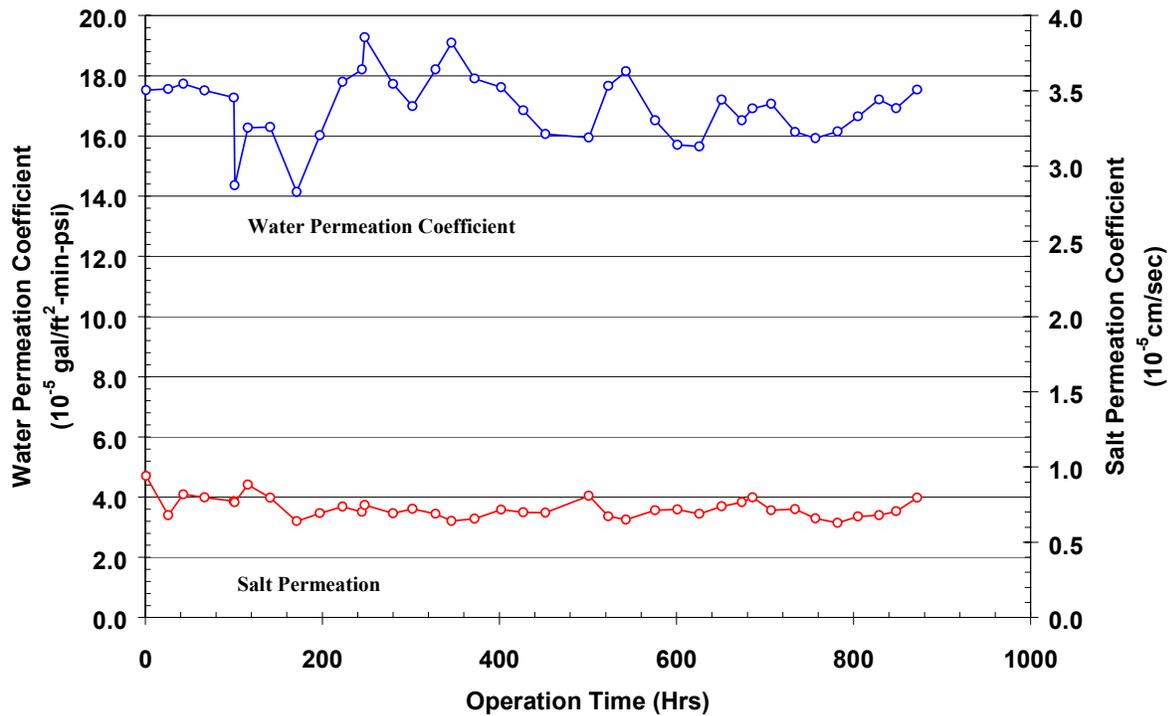


Figure 10. Water Permeation and Salt Permeation Coefficients for Phase 2: Reverse Osmosis Using Conventional Treatment with Ozone/Biofiltration

Table 1. Sampling Schedule for Inorganic and Microbial Analyses

Parameter	Frequency
Cations	2/Month
Anions	Weekly
Trace Metals	Monthly
Total Dissolved Solids	Biweekly
Alkalinity/Hardness	Weekly
TOC/DOC/UV	Weekly
Carboxylic Acids	Biweekly
Heterotrophic Plate Count Bacteria	2/Week
Total/Fecal Coliform	Weekly

Table 2. Water Quality Analyses for Phase 1: Conventional Treatment*

Parameter	Influent	Effluent
Inorganics		
Alkalinity (mg/L as CaCO ₃)	131	79.5
Total Hardness (mg/L as CaCO ₃)	289	289
Total Dissolved Solids (mg/L)	599	619
TOC (mg/L)	2.8	2.2
Calcium (mg/L)	73	73
Magnesium (mg/L)	27.3	27.2
Potassium (mg/L)	4.1	4.2
Sodium (mg/L)	86.3	87.0
Nitrate (mg/L)	1.0	1.2
Silica (mg/L)	9.1	8.9
Chloride (mg/L)	75.2	78.6
Sulfate (mg/L)	234	283
Fluoride (mg/L)	0.30	0.31
Aluminum (□g/L)	46.6	4.5
Iron (□g/L)	53.5	12.8
Microbial		
HPC (CFU/mL)	1,744	705
Total Coliform (P/A)	NS	ND
Fecal Coliform (CFU/100 mL)	26.1	ND

*All data given in average values

ND = Non-detectable

NS = Not sampled

Table 3. Water Quality Analyses for Phase 1: Reverse Osmosis*†

Parameter	Influent	Permeate	Brine	Concentration Factor (Based on average flows)	Rejection (%) (Based on average flows)
Inorganics					
Alkalinity (mg/L as CaCO ₃)	79.5	3.2	471	5.9	96.0
Total Hardness (mg/L as CaCO ₃)	289	2.2	1809	6.3	99.2
Total Dissolved Solids (mg/L)	619	13.8	3837	6.2	97.8
TOC (mg/L)	2.2	0.1	13.8	6.3	95.5
Calcium (mg/L)	73	0.2	417	5.7	99.7
Magnesium (mg/L)	27	0.1	167	6.2	99.6
Potassium (mg/L)	4.2	0.1	24.7	5.9	97.6
Sodium (mg/L)	87	4.3	508	5.8	95.1
Nitrate (mg/L)	1.2	0.2	6.6	5.5	83.3
Silica (mg/L)	8.9	0.3	54.4	6.1	96.6
Chloride (mg/L)	78.6	3.4	484	6.2	95.7
Sulfate (mg/L)	283	4.2	1771	6.3	98.5
Fluoride (mg/L)	0.31	0.02	1.73	5.6	93.5
Barium (□g/L)	106.7	0.31	533	5.0	99.7
Aluminum (□g/L)	4.5	3.2	14.0	3.1	28.9
Iron (□g/L)	12.8	11.1	44.1	3.4	13.3
Microbial					
HPC (CFU/mL)	681	45.7	79105	116	93.3
Total Coliform (presence/absence)	ND	ND	NS	NA	NA
Fecal Coliform (CFU/100 mL)	ND	ND	2.0	NA	NA

* Operating conditions: 85 percent water recovery with 1 gpm concentrate recycle

† All data given in average values

ND = Non-detectable

NA = Not available

NS = Not sampled

Table 4. Energy Dispersive Spectroscopy Analysis of Membrane from Fouled Element for Phase 1: Conventional Treatment

Ion	Lead Element (percent w/w)	Terminal Element (percent w/w)
Iron	48	12
Silica	17	20
Sulfur	14	33
Calcium	10	6
Aluminum	9	6
Potassium	4	ND
Barium	ND	22

ND = Not detected

Table 5. Selected Water Quality Analyses for Phase 1: Capacitive Deionization*

Parameter	Influent	Effluent	Rejection (%)	Regenerant Water	Recovered (%)
Inorganics					
Alkalinity (mg/L as CaCO ₃)	85	8.0	91	56	66
Total Hardness (mg/L as CaCO ₃)	290	144	50	215	74
Total Dissolved Solids (mg/L)	618	303	51	456	74
TOC (mg/L)	2.4	2.1	16	2.7	100
Calcium (mg/L)	72	37	49	51	71
Magnesium (mg/L)	27	13	51	22	81
Potassium (mg/L)	4.2	0.7	83	2.4	57
Sodium (mg/L)	86	33	62	62	72
Nitrate (mg/L)	1.2	0.1	92	0.9	73
Silica (mg/L)	9.1	7.3	20	7.9	87
Chloride (mg/L)	79	16	80	4.5	73
Sulfate (mg/L)	278	183	34	206	74
Fluoride (mg/L)	0.3	0.1	82	0.2	76
Aluminum (□g/L)	ND	ND	NA	ND	NA
Iron (□g/L)	ND	ND	NA	ND	NA
Microbial					
	92,33				
HPC (CFU/mL)	3	62,667	32	ND	NA
Fecal Coliform (MPN/100 mL)	2	NA	NA	NA	NA

* 3.3 gallons of 100 percent CRW, batch-mode

ND = Non-detectable

NA = Not available

NS = Not sampled

Table 6. Capacitive Deionization Polarity Reversal Experiments

Sample Volume (Gallons)	# of Stacks	Mode	Initial TDS (mg/L)	Final TDS (mg/L)	Polarity Reversal (minutes)	Regeneration Time (hours)
5	1	B	564	337	0	4.0
5	1	B	570	355	10	2.5
5	1	B	569	355	5	1.0
5	1	C	574	397	5	1.0

B = Batch-mode

C = Continuous-mode

Table 7. Capacitive Deionization Water Recovery Experiments

Source Water	Sample Volume (gallons)	Mode	Number of Stacks	Initial TDS (mg/L)	Final TDS (mg/L)	Salt Rejection (Percent)	Water Recovery (Percent)
Blend	3	B	1	473	211	55	--
Blend	3	C	1	489	209	57	75
CRW	2.6	B	1	558	259	54	75
CRW	5	C	3	541	200	63	--

B = Batch-mode

C = Continuous-mode

Table 8. Water Quality Analyses for Phase 2: Conventional Treatment with Ozone and Biofiltration*

Parameter	Influent	Effluent
Inorganics		
Alkalinity (mg/L as CaCO ₃)	116	114
Total Hardness (mg/L as CaCO ₃)	240	243
Total Dissolved Solids (mg/L)	499	509
TOC (mg/L)	2.7	2.3
Calcium (mg/L)	61	61
Magnesium (mg/L)	23	23
Potassium (mg/L)	3.7	3.7
Sodium (mg/L)	74	75
Nitrate (mg/L)	1.4	1.7
Silica (mg/L)	9.0	8.9
Chloride (mg/L)	64	68
Sulfate (mg/L)	183	188
Fluoride (mg/L)	0.31	0.24
Barium (□g/L)	81	78
Aluminum (□g/L)	18	11
Iron (□g/L)	ND	ND
Microbial		
HPC (CFU/mL)	6675	515
Total Coliform (P/A)	NS	ND
Fecal Coliform (CFU/100 mL)	6.5	ND

*All data given in average values

ND = Non-detectable

NS = Not sampled

Table 9. Water Quality Analyses for Phase 2: Reverse Osmosis*†

Parameter	Influent	Permeate	Brine	Concentration Factor (Based on average flows)	Rejection (%) (Based on average flows)
Inorganics					
Alkalinity (mg/L as CaCO ₃)	114	7.3	467	4.1	93.6
Total Hardness (mg/L as CaCO ₃)	243	1.7	1435	5.9	99.3
Total Dissolved Solids (mg/L)	509	14.7	3078	6.0	97.1
TOC (mg/L)	2.3	0.1	14.1	6.3	95.6
Calcium (mg/L)	61	0.3	268	4.4	99.5
Magnesium (mg/L)	23	0.1	131	5.8	99.6
Potassium (mg/L)	3.7	0.3	19.3	5.2	91.9
Sodium (mg/L)	75	5.0	419	5.6	82.3
Nitrate (mg/L)	1.7	0.3	8.6	5.2	82.3
Silica (mg/L)	8.9	0.5	50.3	5.7	94.3
Chloride (mg/L)	68	3.3	394	5.8	95.1
Sulfate (mg/L)	188	3.1	1314	7.0	98.4
Fluoride (mg/L)	0.24	ND	1.4	5.6	NA
Barium (□g/L)	78	0.3	433	5.5	99.6
Aluminum (□g/L)	11	0.9	48	4.3	91.8
Iron (□g/L)	ND	ND	22	NA	NA
Microbial					
HPC (CFU/mL)	515	13	565	1.1	97.5
Total Coliform (presence/absence)	ND	ND	NS	NA	NA
Fecal Coliform (CFU/100 mL)	ND	ND	2.0	NA	NA

* Operating conditions: 85 percent water recovery with 1 gpm concentrate recycle

† All data given in average values

ND = Non-detectable

NA = Not available

NS = Not sampled

Table 10. Energy Dispersive Spectroscopy Analysis for Phase 2: Conventional Treatment With Ozone/Biofiltration

Ion	Lead Element (percent w/w)	Terminal Element (percent w/w)
Iron	18.2	ND
Silica	23.3	17.4
Sulfur	26.1	64.9
Calcium	4.8	ND
Aluminum	22.9	17.7
Chloride	4.8	ND
Barium	ND	ND

ND = Not detected

Table 11. Comparison of Reverse Osmosis and Capacitive Deionization

Evaluation Criteria	Conventional Treatment Phase		Conventional Treatment w. Ozone/ Bio. Phase
	Reverse Osmosis	Capacitive Deionization	Reverse Osmosis
Permeate flux (gallon/ ft ² /day)	16.1	0.6	15.7
Energy usage (kWh/1,000 gal)	1.66	2.25	1.37
Process recovery (percent)	85	75	85
Aerogel regeneration (hrs)	NA	1	NA
Salinity rejection (percent)	98	51	99
Permeate feed water quality (mg/L TDS)	14	303	27
Operational reliability	Good	Poor	Good

NA = not applicable

2.2 Pulsed Ultraviolet (UV) Radiation for Disinfection and Membrane Fouling Control

2.2.1 Experiment Results

2.2.1.1 Heterotrophic Plate Count (HPC) Bacteria

HPC bacteria in the pilot-plant effluent were readily inactivated by pulsed-UV radiation. Figure 12 shows that a UV dose of approximately 12 mJ/cm² achieved ≥ 3 log₁₀ inactivation of bacteria (to a final concentration of <10 CFU/mL). Data for both the batch and the plug flow experiments showed similar reductions in bacteria concentrations, although additional data is required to fully characterize the HPC inactivation in the plug flow experiments.

Repair/regrowth experiments were also conducted with HPC bacteria after exposure to pulsed-UV irradiation. Figure 13 shows data for irradiated samples which were then placed in the dark (i.e., similar to a pipeline distribution system). Dark-repair/regrowth experiments indicated that HPC bacteria exposed to low doses (9 mJ/cm²) of pulsed UV irradiation and followed with the addition of a residual disinfectant (2.5 mg/L chloramines) showed no HPC growth after 3 days of incubation. Samples irradiated at 27 mJ/cm² and placed in dark incubation showed that HPC bacteria could repair or regrow to 3,000 to 156,000 cfu/mL. Samples exposed to 56 mJ/cm² and incubated in the dark did not show significant repair or regrowth ability and levels of HPC bacteria stayed below 10 cfu/mL. Repair/regrowth results from irradiating HPC bacteria and placing them in light (i.e., similar to an uncovered reservoir. Figure 13 illustrates initial results which indicated that samples irradiated at all the doses examined (up to 56 mJ/cm²) had regrowth or repair to over 10,000 cfu/mL after 3 days. However, the sample irradiated at 9 mJ/cm² followed by 2.5 mg/L of chloramines still showed no net increase in bacteria. Initial results indicate that low UV doses provided adequate reduction of bacteria so that chloramines alone (with no free chlorine contact) may be successfully used as a secondary disinfectant in the distribution system.

2.2.1.2 Malt-Specific (MS-2) Coliphage

Figure 15 summarizes batch inactivation of MS-2. Previous studies have shown that dose-response from pulsed UV outperforms low pressure UV irradiation (a higher log inactivation per UV dose value). Dose-response also depends on the distance a sample is from the flash lamp; see: B. R. Wilson; P.F. Roessler; E.V. Dellen; M. Abbaszadegan; C.P. Gerba, "Coliphage MS-2 as a UV Water Disinfection Efficacy Test Surrogate for Bacterial and Viral Pathogens," In Proc. AWWA Water Qual. Tech. Conf., Toronto, Ontario, Canada (1992). Other investigators demonstrated that joulemeter measurements of UV dose are not consistent at varying distances from the flash lamp, see K. G. Linden and A. A. Mofidi, "Disinfection Efficiency and Dose Measurement for Medium Pressure and Pulsed-UV Disinfection Systems: an Unsolicited Proposal to the American Water Works Association Research Foundation" (University of North Carolina, Charlotte, 1998). This work indicated that the joulemeter measured 100 percent of the UV dose at 6 inches from the lamp, but only 50 to 80 percent of the dose at 3 inches from the lamp. This variation in UV dose measurement by the joulemeter indicated that the differences in dose-response at different distances may not truly exist.

2.2.1.3 *Cryptosporidium (C.) Parvum*

Figure 16 shows the preliminary results from the disinfection of *C. parvum* as a function of UV dose. Greater than 4.5 log₁₀ inactivation indicated that the oocysts were completely inactivated (i.e., no infections were detected by the cell culture assay) and the inactivation was limited by number of influent oocysts in suspension (approximately 1 × 10⁶ oocysts).

2.2.1.4 Disinfection By-Products (DBP)

Table 12 and Table 13 summarize the DBP results. All experiments were conducted at 0.5 inches (in.) from the flash lamp. Experiments were designed to evaluate the effect of pulsed UV light on water which passes next to the lamp when doses were applied to achieve disinfection at the reactor wall. The pulse frequency and water flow for these samples corresponded to UV doses of 660, 1,300, and 2,000 mJ/cm² at 0.5 in. from the flash lamp. These conditions in an 8-in. diameter unit would result in UV doses of 15, 30, and 44 mJ/cm² at a distance of 4 in. away. Table 12 shows that formaldehyde (a biodegradable DBP) increased from less than 1 microgram per liter (µg/L, or □g/L) to between 2 to 10 µg/L after UV treatment. Table 13 shows that the THMs and HAAs formed by chloramines (at simulated distribution system conditions) increased before and after pulsed UV treatment, but the magnitude of the increase was insubstantial. In all cases, the THMs and HAAs never exceeded 3 and 11 µg/L, respectively.

2.2.1.5 Estimated Full-Scale Operating Costs

Full-scale operating costs were estimated based on bench-scale inactivation data. Power consumption of the pulsed-UV system was monitored at increasing frequency, as shown in Table 14. Assuming a design objective of 2-log₁₀ reduction in viruses and an 8-in. diameter treatment chamber, 10 pulses per second would need to be applied to a flow of 1.4 acre-feet per day (ac-ft./day). The operating power consumption would therefore be 52 kilowatt hours per acre-foot (kWh/ac-ft.), or 0.16 kWh/1,000 gal.

2.2.2 Analysis Of Results

As shown in the experiment results section, data indicate that pulsed-UV can effectively inactivate HPC bacteria, viruses (as measured by MS-2 coliphage) and *Cryptosporidium parvum* at UV doses low enough such that DBP levels are not significantly increased. In most cases, a secondary disinfectant such as chloramine would be required to inhibit regrowth or repair of bacteria after UV treatment. A more accurate means of measuring UV dose will be required to effectively scale-up this technology.

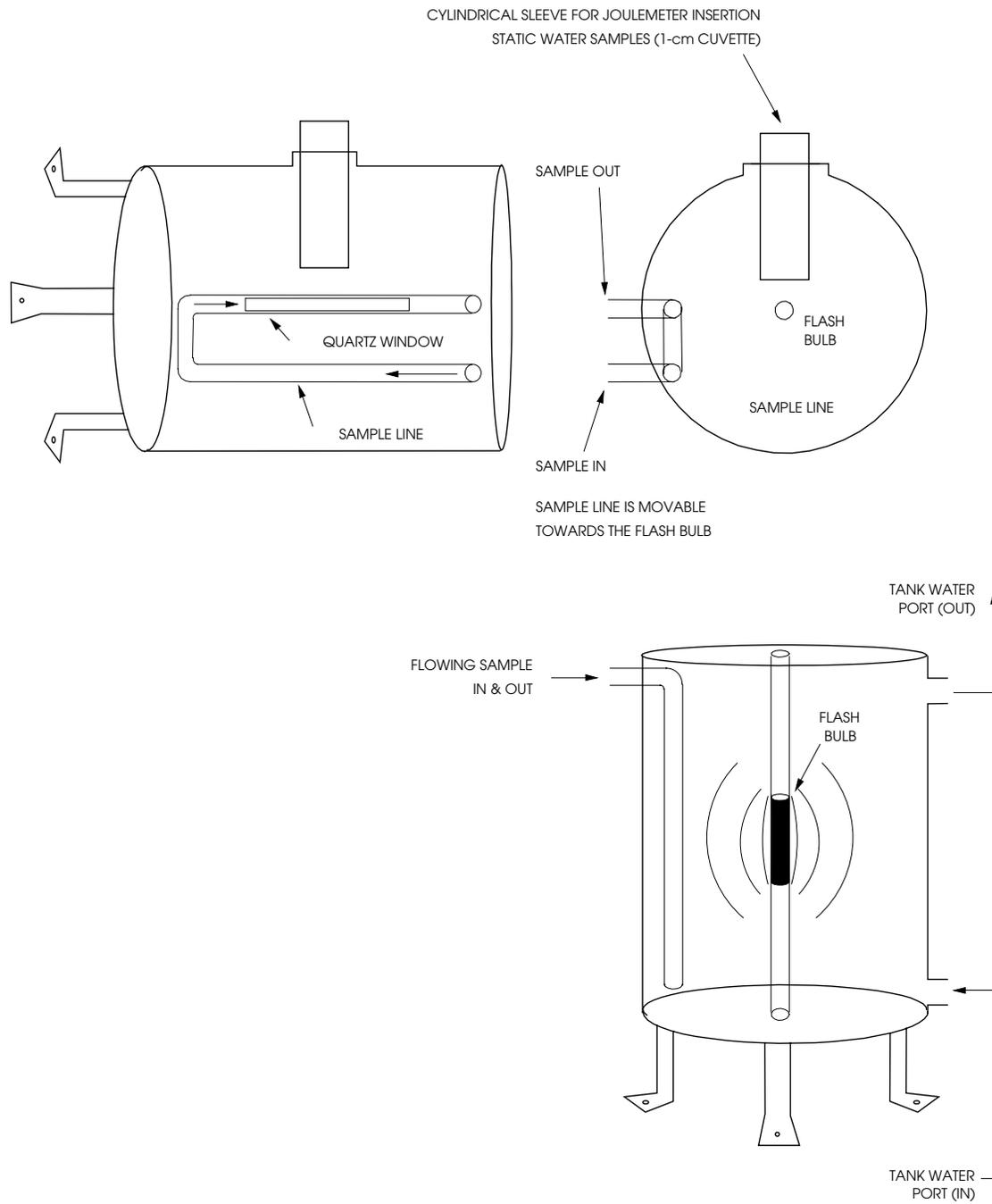


Figure 11. Schematic of Pulsed UV Treatment Chamber

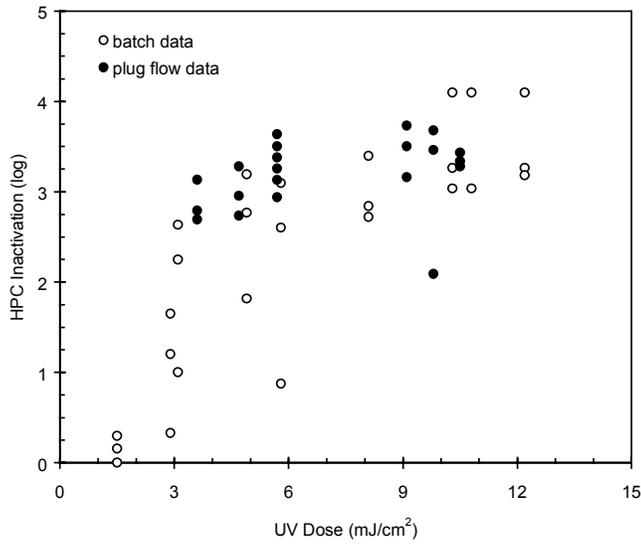


Figure 12. HPC Inactivation Data for Batch and Plug-Flow Experiments

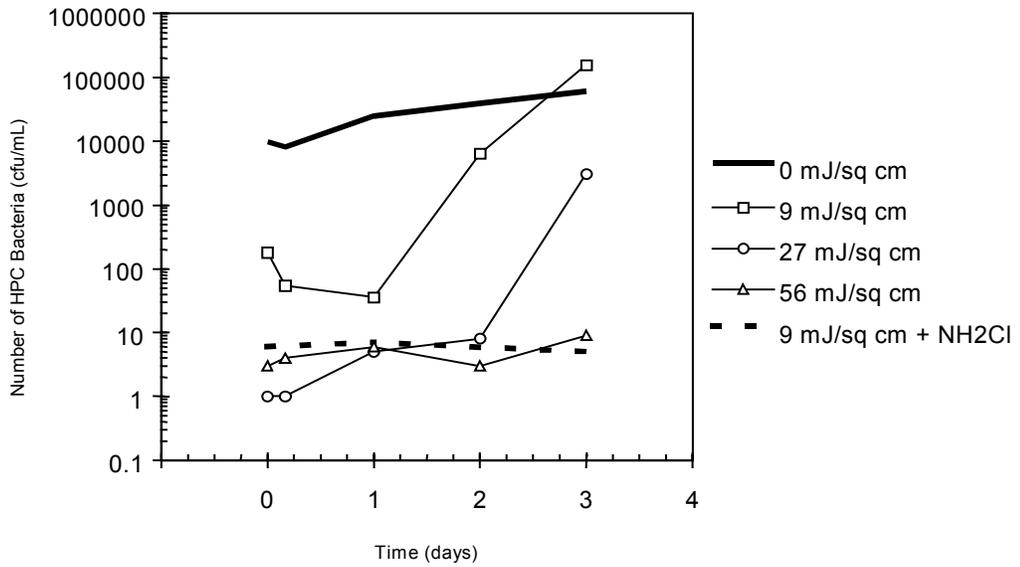


Figure 13. HPC Dark Repair/Regrowth Experiment Results (one experiment)

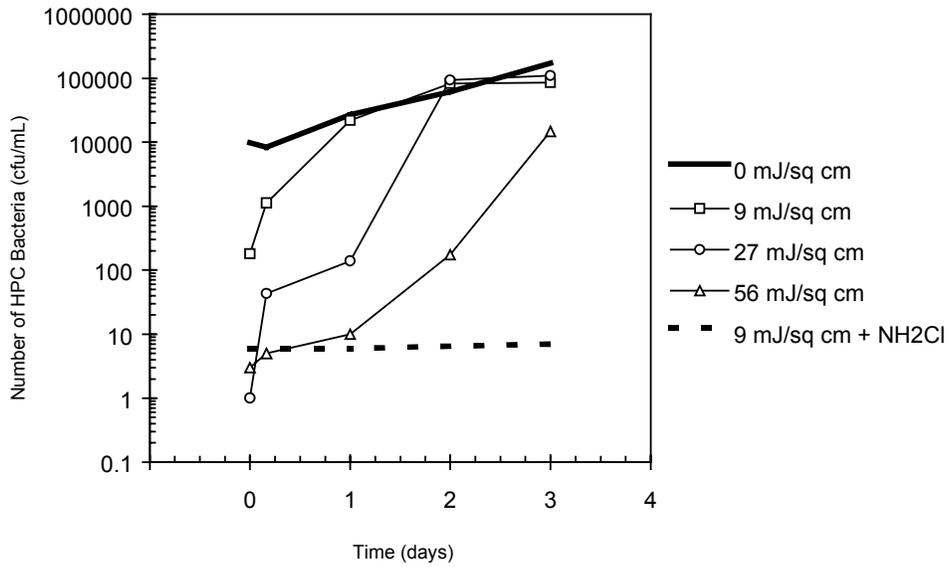


Figure 14. HPC Light Repair/Regrowth Experiments (one experiment)

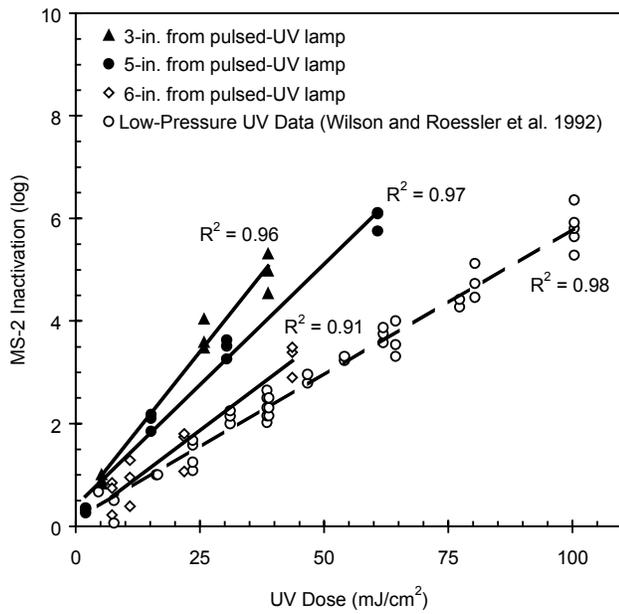


Figure 15. MS-2 Coliphage Inactivation in Batch Mode with Pulsed UV Compared to Low Pressure Inactivation Reported in the Literature

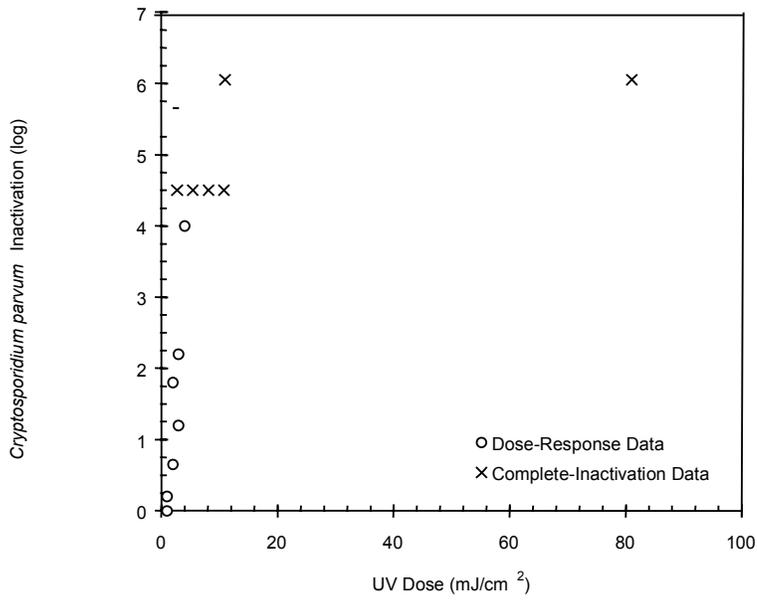


Figure 16. *Cryptosporidium Parvum* Inactivation in Batch Mode

Table 12. Formaldehyde Formation after Pulsed-UV Treatment

UV Dose* (mJ/cm ²)	Formaldehyde (µg/L)			
	Test 1	Test 2	Test 3	Test 4
0 (control)	<1	<1	<1	<1
660	<1	1	<1	6
1,300	2	3	2	7
2,000	6	4	2	10

*Water velocity = 0.5 to 1.5 ft./sec; distance from lamp = 0.5 in.; pulse rate = 29 Coefficient Hz

Table 13. Effect of Pulsed UV Treatment on SDS Chloramine THMs and HAAs

DBP	Condition	Simulated Distribution System DBPs* ($\mu\text{g/L}$)			
		Test 1	Test 2	Test 3	Test 4
THMs	Before UV	1	2	1	2
	After UV†	2	3	2	2
HAAs	Before UV	8	9	8	8
	After UV†	9	11	10	9

*SDS conditions: chloramine residual = 2.5 mg/L, no free chlorine contact, contact time = 24 hour, pH = 8, temperature = 25°C

†Water velocity = 0.5 to 1.5 ft./sec; distance from lamp = 0.5 in.; pulse rate = 29 Hz.

Table 14. Bench-Scale Unit Power Usage and Estimated Full-Scale Treatment Costs

Pulse Rate (Hz)	Average Current (amps)	Voltage (volts)	Average Power (kW)	Daily Energy Use (kWh)	Daily Operating Cost (\$/day)
0	4.8	196	0.9	22	1.78
5	12	194	2.3	55	4.41
10	16	192	3.1	74	5.94
20	24	190	4.6	110	8.80
29	26	190	4.9	117	9.39

Assumptions:

1. Electrical cost = \$0.08/kWh
2. 0 Hz sample describes the unit running in the “simmer” mode
3. Power usage calculated from line current and voltage

2.3 Bromate Formation and Control

2.3.1 Results of pH Control Experiments

Figure 17 shows the formation of bromate in Colorado River Water (CRW), a relatively low-bromide water. At disinfection levels required under the existing Surface Water Treatment Rule (SWTR, which requires less than 5-log *Giardia* inactivation), bromate was not formed at detectable levels (less than 3 µg/L). However, at ambient pH (~8.3) and at a *Cryptosporidium* inactivation credit of ~1 log, significant levels of bromate were formed. Reducing the pH to 7 or below reduced the production of bromate in CRW by at least 50 percent.

Substantially more bromate was formed in SPW or in a blend of State Project Water (SPW) and CRW—even when the bromide level was similar to CRW. The increased alkalinity of CRW may have acted as a scavenger of hydroxyl radicals in this case which would have lessened bromate formation through the radical pathway. Controlling the pH of ozonation in SPW and the blend was also effective in minimizing bromate as seen in Figure 18 and Figure 19. Detectable levels of bromate were still formed at the higher inactivation levels. In SPW from the East Branch of the California Aqueduct, the bromide historically is above 0.10 mg/L most (greater than 85 percent) of the time. By controlling to pH 6.5, approximately 70-80 percent reduction in bromate can be achieved.

Figure 20 shows the effect of the bromide ion concentration on bromate formation. At pH 8, the formation of bromate is highly sensitive to the influent bromide concentration. When ozonating to achieve a 1-log *Cryptosporidium* inactivation, ~48 µg/L of bromate was formed. However, less than 10 µg/L of bromate was formed when the pH of ozonation was below 6.5. Figure 20 and Table 16 note that the effect of influent bromide is largely attenuated at neutral and acidic pH values.

2.3.2 Results of Ammonia Addition Experiments

Figure 21 and Table 17 show the effect on bromate formation in SPW of adding 0.2–2 mg/L ammonia, as well as the corresponding reduction in *Cryptosporidium* inactivation credit. A constant ozone dose resulting in a 1-log *Cryptosporidium* inactivation credit (without ammonia) was used at ambient pH (8.4–8.6) and at an ambient bromide concentration (0.10 mg/L). Addition of 0.2 mg/L (NH₃:Br⁻ = 13:1 mol:mol) and 0.9 mg/L (NH₃:Br⁻ = 54:1 mol:mol) ammonia resulted in reductions of 33 and 61 percent, respectively, in bromate formation, with moderate (~30 percent) decreases in inactivation credit. A higher dose of ammonia (2 mg/L, NH₃:Br⁻ = 117:1 mol:mol) marginally improved bromate reduction (to 72 percent) but substantially reduced inactivation credit (50 percent).

The effects of ammonia concentration on bromate formation (at a constant ozone dose) in CRW are shown in Figure 22 and Table 17, along with corresponding changes in *Cryptosporidium* inactivation credit. Four ammonia doses (0.2, 0.4, 0.7, and 1.4 mg/L) were applied at ambient pH (8.3) and bromide (0.07 mg/L) concentrations. Adding 0.2 mg/L ammonia (NH₃:Br⁻ = 18:1 mol:mol) resulted in a relatively large decrease (67 percent) in bromate formation and essentially no decrease in disinfection credit. Higher ammonia doses (≥ 0.4 mg/L) reduced bromate by >75 percent, to below 3 µg/L mg/L, the Minimum Reporting Level (MRL).

Because CRW has a higher alkalinity and greater pH buffering capacity than SPW (130 mg/L versus 60 mg/L as CaCO₃), adding ammonia had less impact on pH in CRW, as seen in Table 3. As stated by Krasner, S. W.; Glaze, W.H.; Weinberg, H. S.; Daniel, P.A.; and Najm, I. N., Formation and Control of Bromate During Ozonation of Waters Containing Bromide. Jour. AWWA, 85:1:73 (Jan 1993), bromate formation increases with increasing pH:

At elevated pH, bromide is preferentially oxidized to OBr⁻, which can be oxidized by molecular ozone to bromate, whereas HOBr is not oxidized by molecular ozone to bromate.

At elevated pH, there is more hydroxyl radical activity, which can promote bromate formation.

Thus, in SPW tests, the benefits of ammonia addition were offset by higher bromate formation at the elevated pH. In addition, the speciation and stability of bromamines are affected by pH, as is stated by H.Galal-Gorchev and J.C. Morris, "Formation and Stability of Bromamide, Bromimide, and Nitrogen Tribromide in Aqueous Solution," Inorg. Chem. 4(6):899-905 (1965). Thus, an increase in pH may have resulted in a shift to bromamine species that are less stable.

Bromate formation in SPW with and without 1 mg/L ammonia over a range of applied ozone doses (3.1–5.3 mg/L) is shown in Figure 23 and Table 17. The effects of ammonia addition – and subsequent changes in pH – on inactivation credit are shown by the dashed lines of equal inactivation plotted between the two data sets. Note that the addition of ammonia (as ammonium hydroxide) increased the pH and alkalinity of the water. In these tests, adding 1.0 mg/L of ammonia increased the pH by ~0.6 units and the alkalinity by ~2 mg/L as CaCO₃. Comparing these two conditions for an equivalent applied ozone dose showed from a 56 percent reduction in bromate formation at a 3.1 mg/L applied ozone dose to a 70 percent reduction at a ~5.2 mg/L applied ozone dose. Comparing bromate formation on an equal-inactivation basis showed that ammonia may have been somewhat more effective (42- versus 33-percent reductions at 1.1 and 0.4 log, respectively) at a higher *Cryptosporidium* inactivation. Adding 1 mg/L of ammonia resulted in a substantial decrease (e.g., 69 percent at an ozone dose

of 4.6 mg/L) in *Cryptosporidium* inactivation credit as a result of the increase in ozonation pH (8.4 to 9.0).

Figure 24 and Table 17 show a comparison of bromate formation over a range of ozone doses (1.2-2.4 mg/L) in CRW with and without 1 mg/L ammonia. On an applied-dose basis, bromate was reduced by 77 percent at a 2.0 mg/L applied ozone dose and by 81 percent at a 2.4 mg/L applied ozone dose. Comparison on an equivalent-inactivation basis shows a similar trend in which bromate was reduced by 70 percent at 1.0- and 1.5-log *Cryptosporidium* inactivations.

Figure 25 and Figure 26 show bromate formation rates for SPW and CRW, respectively. These data represent samples collected in each of the six contact chambers of the over/under contactor (the samples were quenched with an excess of ethylene diamine) and are plotted against hydraulic detention time. In SPW, bromate formation with ammonia addition occurred almost entirely within the first contact chamber (<4 min detention time). Likewise, without ammonia, bromate formed rapidly in SPW (80 percent formed in chamber 1A). In contrast, bromate formed slowly (without ammonia addition) in CRW (30 percent formed in chamber 1A). These results provides insight into the underlying mechanisms of bromate formation in these two source waters, suggesting that (1) bromate formation in CRW may be predominantly controlled by a direct (molecular ozone) oxidation of bromide; (2) the relatively rapid rate of bromate formation in SPW is probably controlled predominantly by indirect (OH• radical) oxidation; and (3) ammonia is probably more effective in controlling bromate formed through the molecular-ozone-controlled pathway. For both source waters, however, adding ammonia decreased bromate formed within the first contact chamber (e.g., 70 percent for SPW and >25 percent for CRW), suggesting that ammonia is – at least partially – effective in inhibiting hydroxyl radical pathways.

Figure 27 shows changes in ammonia and dissolved ozone concentrations in CRW within the over/under contactor. Changes in ammonia concentration over the 24-min detention time were small (≤ 0.05 mg/L decrease for a ~ 0.4 mg/L dose) and may be attributed to reactions with HOBr/OBr- (forming bromamines) and direct reaction with other oxidants, including dissolved ozone. Based on this data an apparent second order rate constant of $0.5 \text{ M}^{-1} \text{ s}^{-1}$ was estimated, over the 24 min contact time, which is within an order of magnitude of ($\sim 20 \text{ M}^{-1} \text{ s}^{-1}$) for the oxidation of ammonia by ozone. See J. Hoigne and H. Bader, "Rate Constants of Reactions of Ozone with Organic and Inorganic Compounds in Water. III: Inorganic Compounds and Radicals," *Water Res.* 19(8): 993 (1985).

2.3.3 Hydrogen Peroxide Addition

Figure 28 and Figure 29 show the effects of adding hydrogen peroxide on bromate production in SPW and CRW, respectively. Hydrogen peroxide was added upstream of the ozone contactor, and a constant ozone dose (corresponding to 1-log and 1.5-log *Cryptosporidium* inactivations for SPW and CRW with no hydrogen peroxide addition, respectively) was applied. Adding hydrogen peroxide to SPW at 0.05 mg/L ($\text{H}_2\text{O}_2:\text{O}_3 = 0.01:1$ mg:mg) and 0.12 mg/L ($\text{H}_2\text{O}_2:\text{O}_3 = 0.03:1$ mg:mg) had no effect on bromate production and decreased inactivation credit by ~ 30 and ~ 60 percent, respectively. At a constant ozone dose, bromate production in CRW decreased with increasing peroxide dose. Adding 0.14 mg/L of hydrogen peroxide ($\text{H}_2\text{O}_2:\text{O}_3 = 0.07:1$ mg:mg) resulted in a 58-percent decrease in bromate formation. However, this apparent benefit probably resulted from ozone consumption by hydrogen

peroxide, with a corresponding loss of ozone residual. When compared on an equivalent-inactivation basis, only a small (10 to 15 percent) decrease in bromate formation occurred in CRW at lower doses (<0. mg/L) of hydrogen peroxide ($H_2O_2:O_3 <0.05:1$ mg:mg).

Figure 30 shows changes in hydrogen peroxide and ozone residuals within the contactor for the high dose of hydrogen peroxide (0.14 mg/L) in CRW. During ozonation to achieve *Cryptosporidium* disinfection (by maintaining high levels of CT), the application of hydrogen peroxide can hinder the maintenance of an ozone residual (compare Figure 30 and Figure 27) and result in an inadequate CT, Table 16. However, if an ozone residual is not required e.g., for taste-and-odor or micropollutant oxidation, then hydrogen peroxide addition may reduce bromate formation, as well as improve the oxidation objective. See U. von Gunten; A. Bruchet; E. Constantin, "Bromate Formation in Advanced Oxidation Processes," *Jour. AWWA* 88(6): 53-65 (1996).

2.3.4 Cost Comparison of pH Control and Ammonia Addition

Table 18 and Figure 31 compare operating costs for two bromate control strategies: ammonia addition and pH control. Ammonia addition increases pH, and, consequently, higher doses of ozone are required to achieve a given inactivation credit. Adding 1 mg/L ammonia increases pH from 8.4 to 9.0 in SPW and from 8.3 to 8.6 in CRW. This pH increase results in a 26- and 17-percent increases in ozone dose (required for a 1-log *Cryptosporidium* inactivation) for SPW and CRW, respectively. For both source waters, the relative cost of ammonia compared to ozone is small (8 percent). Reducing pH to minimize bromate formation increases the ozone residual half-life and less applied ozone is needed to achieve the same CT. Thus, ozone production costs decrease. However, the moderate decrease in ozone production is offset by the rapid increase in pH control cost. For SPW, operating at pH 6.5 with acid feed is 1.8 times more expensive than operating at ambient pH and ammonia feed. However, comparing the two bromate control strategies when comparable bromate reductions are achieved (Table 17; Figure 31), for SPW, operating at pH 7 with acid feed costs only 20 percent more than adding ammonia. Moreover, with ammonia addition in SPW, acid feed is still needed to reduce the pH for distribution. For CRW, operating at pH 6.5 is six times more expensive than adding ammonia at ambient pH.

2.3.5 Operational Considerations

Several important issues must be considered before ammonia is used in a treatment process. First, adding aqua ammonia, ammonium hydroxide, increases the pH of ozonation. In waters where the influent alkalinity is low e.g., 50 mg/L as $CaCO_3$, the addition of 1.0 mg/L ammonia could raise the pH from 8.0 to 8.8. The increase in pH may require substantially more applied ozone to achieve the same disinfection objective. A 2 mg/L dosage of sulfuric acid could possibly be used to counter the pH increase of the ammonium hydroxide.

Second, the addition of pre-ammoniation would either preclude the ability to use free chlorine downstream or would require breakpoint chlorination to achieve a free-chlorine residual. Though breakpoint chlorination is practiced by a number of water utilities particularly when ammonia is present in the raw water, objectionable tastes and odors may result from the formation of di- and trichloramines.

Third, many ozonation facilities operate biological filters downstream of the ozonation process to remove biodegradable organic matter such as assimilable organic carbon and aldehydes formed by ozone. When fed with a continuous level of ammonia, the biological filters would attain some level of nitrification, which would convert the ammonia to nitrate, with nitrite as an intermediate step. Incomplete nitrification (to nitrite) would further increase the chlorine demand because 1 mg/L nitrite (as $\text{NO}_2\text{-N}$) consumes 5.1 mg/L chlorine (as Cl_2) as it is further oxidized to nitrate. In addition, some nitrifying bacteria may be sloughed off of the biological filters and could seed the distribution system with bacteria, which may increase the occurrence of nitrification in the distribution system.

Finally, because the effluents of biological filters contain high bacterial densities (including total coliforms), a disinfection process is generally desired downstream of the biological filters. If excess ammonia is added to the ozonation process (and is not removed by the biological filters), then chloramines would form when chlorine is added downstream of the biofilters. Because chloramines are a much weaker disinfectant than free chlorine, longer contact times would be required to disinfect the bacteria leaving the biological filters. The additional contact time may not be achievable downstream of filtration.

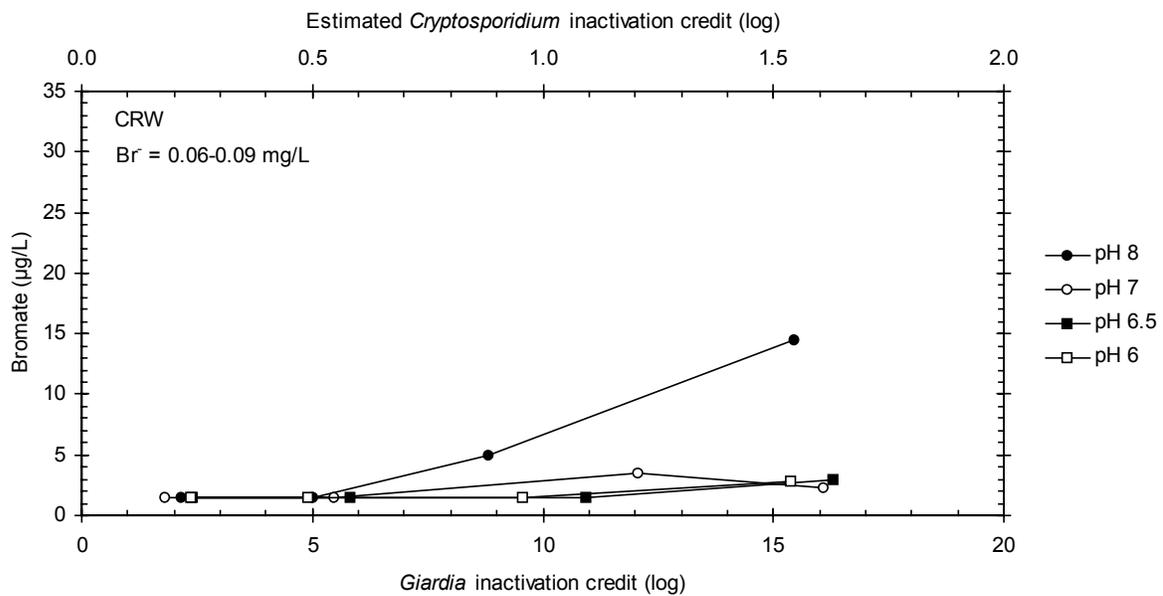


Figure 17. Bromate Formation in CRW (bromide levels of 0.06-0.09 mg/L)

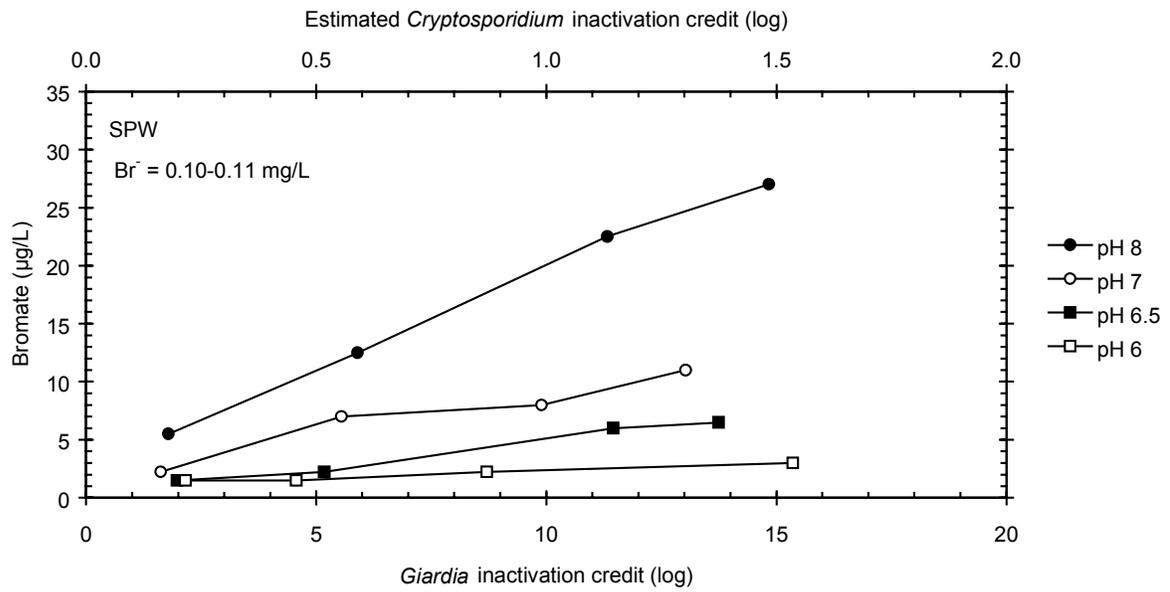


Figure 18. Bromate Formation in SPW with Low Bromide (0.10-0.11 mg/L)

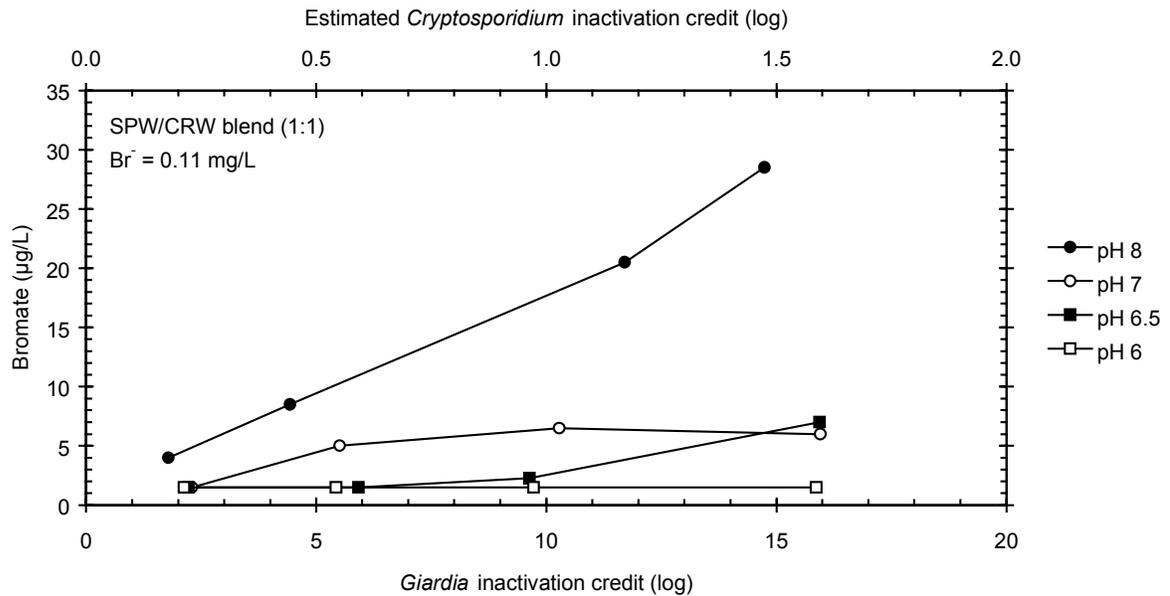


Figure 19. Bromate Formation in a SPW/CRW Blend (bromide = 0.11 mg/L)

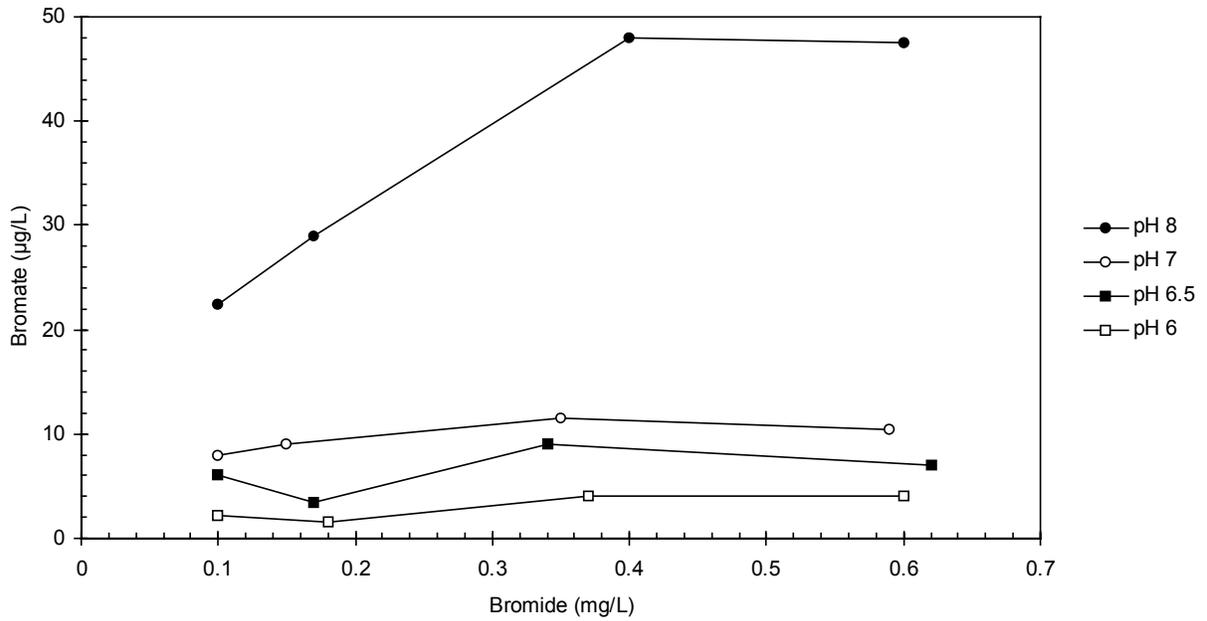


Figure 20. Effect of Influent Bromide in SPW on Bromate Formation at a 1-log (estimated) *Cryptosporidium* Inactivation

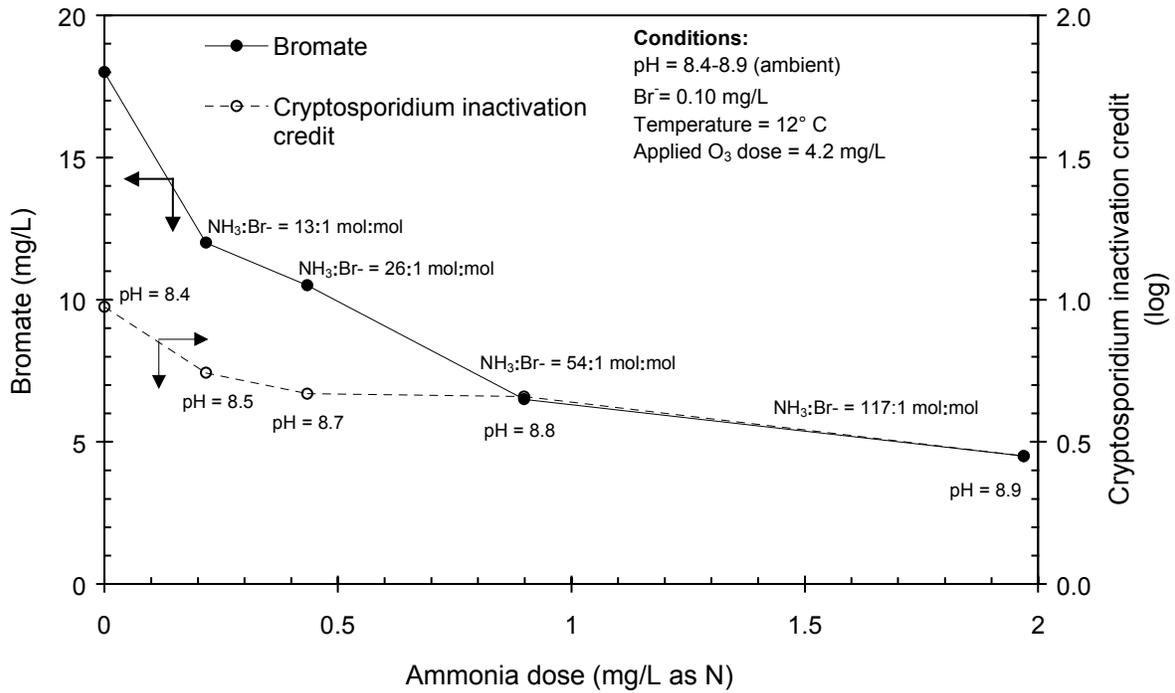


Figure 21. Effect of Ammonia Dose on pH, Bromate Formation, and *Cryptosporidium* Inactivation Credit in SPW

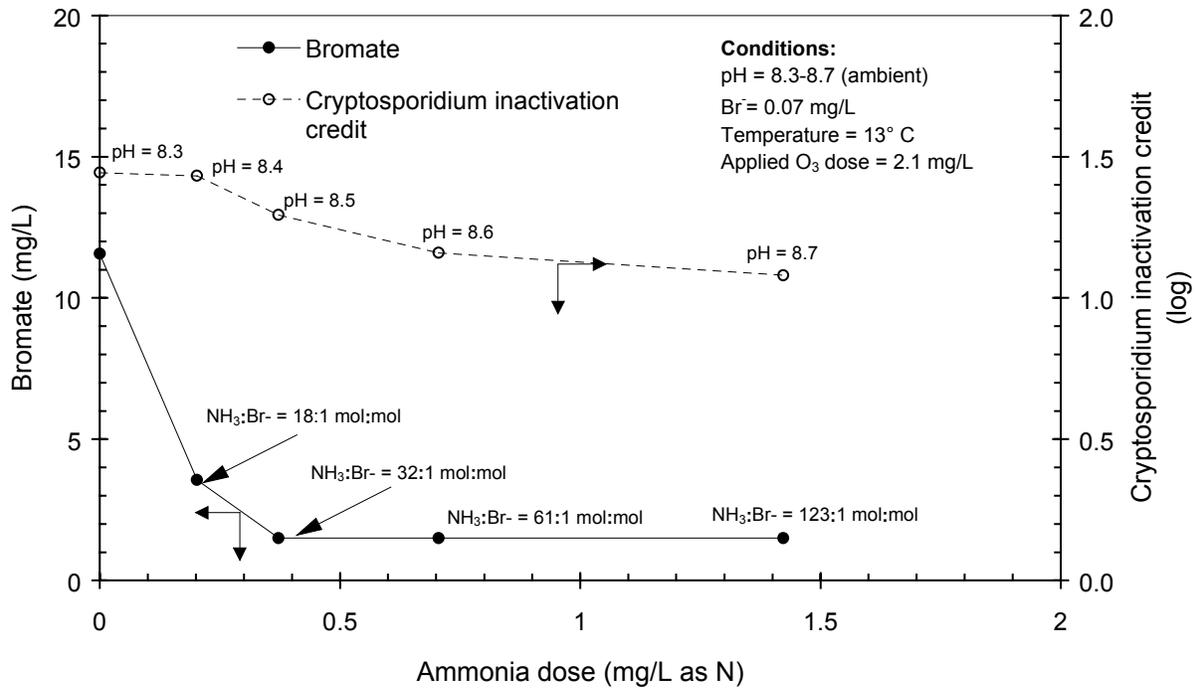


Figure 22. Effect of Ammonia Dose on pH, Bromate Formation, and *Cryptosporidium* Inactivation Credit in CRW (ND plotted at ½ MRL)

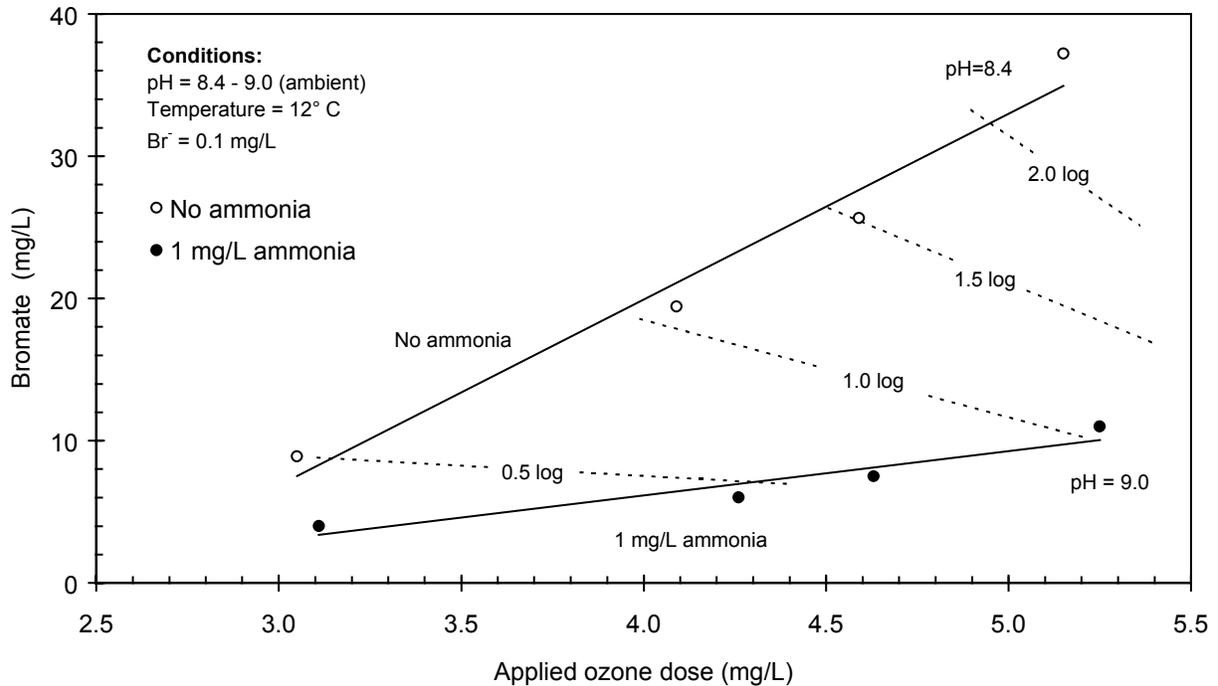


Figure 23. Effect of Ozone Dose and Ammonia Dose on pH and Bromate Formation in SPW (dashed lines show estimates of equal *Cryptosporidium* inactivation)

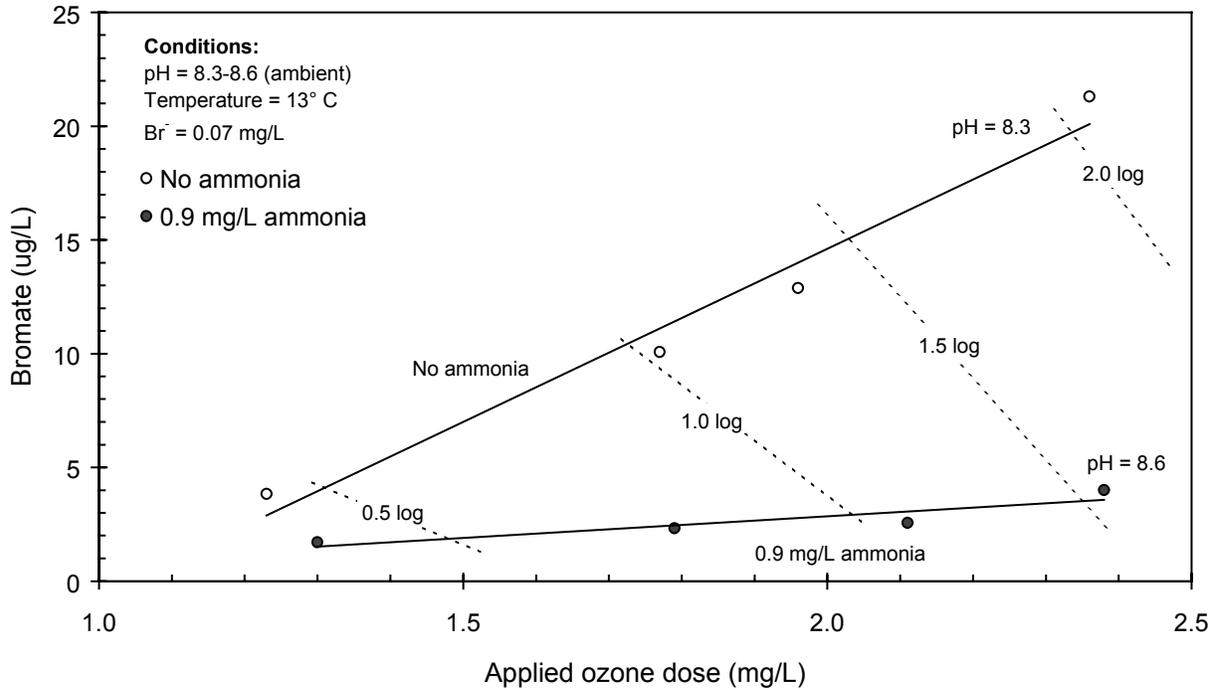


Figure 24. Effect of Ozone Dose and Ammonia Dose on pH and Bromate Formation in CRW (dashed lines show estimates of equal *Cryptosporidium* inactivation; ND plotted at ½ MRL)

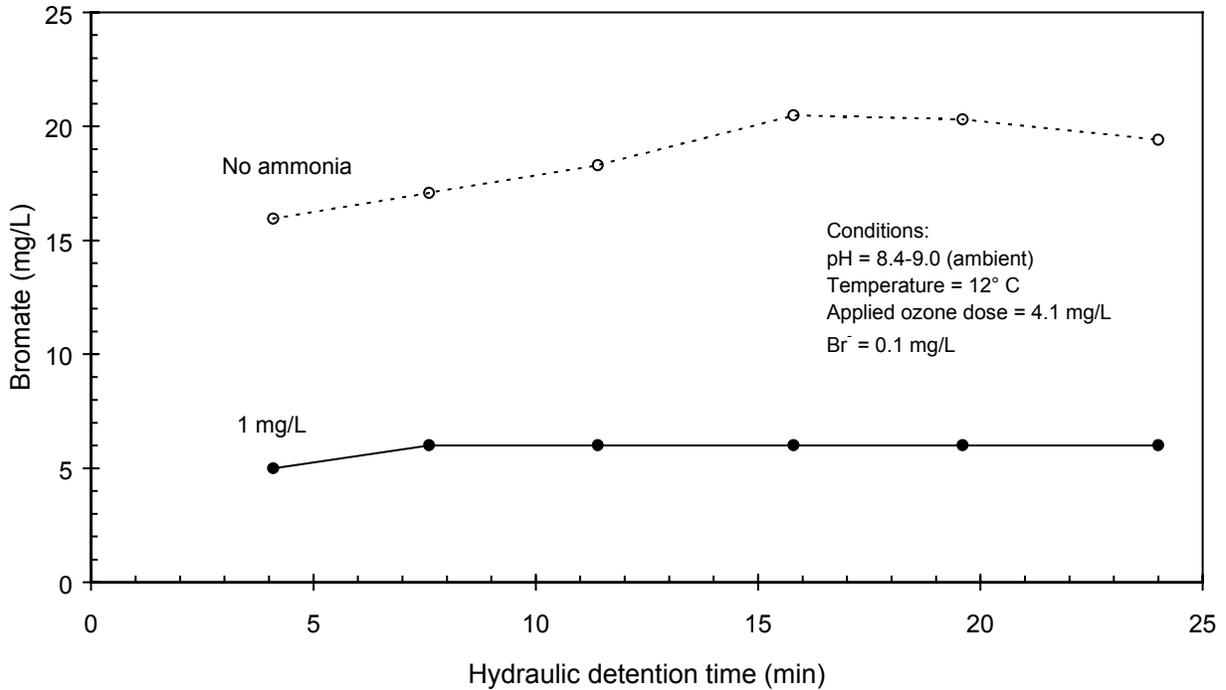


Figure 25. Bromate Formation Kinetics through the Ozone Contactor in SPW

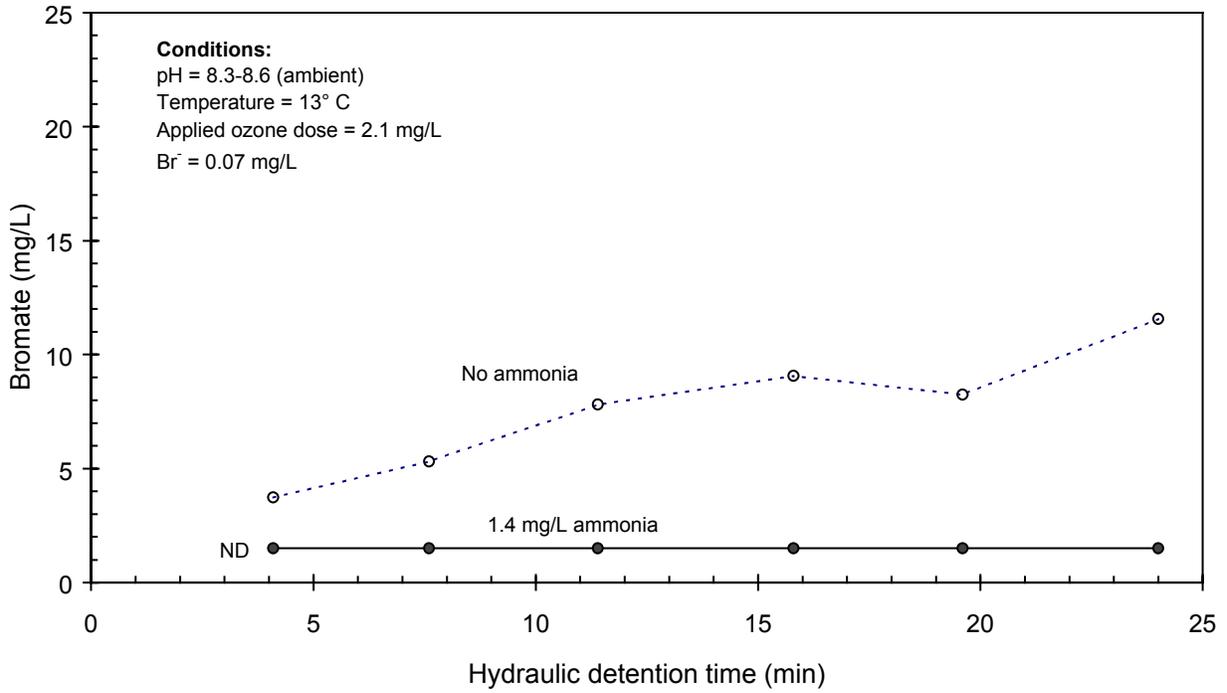


Figure 26. Bromate Formation Kinetics through the Ozone Contactor in CRW

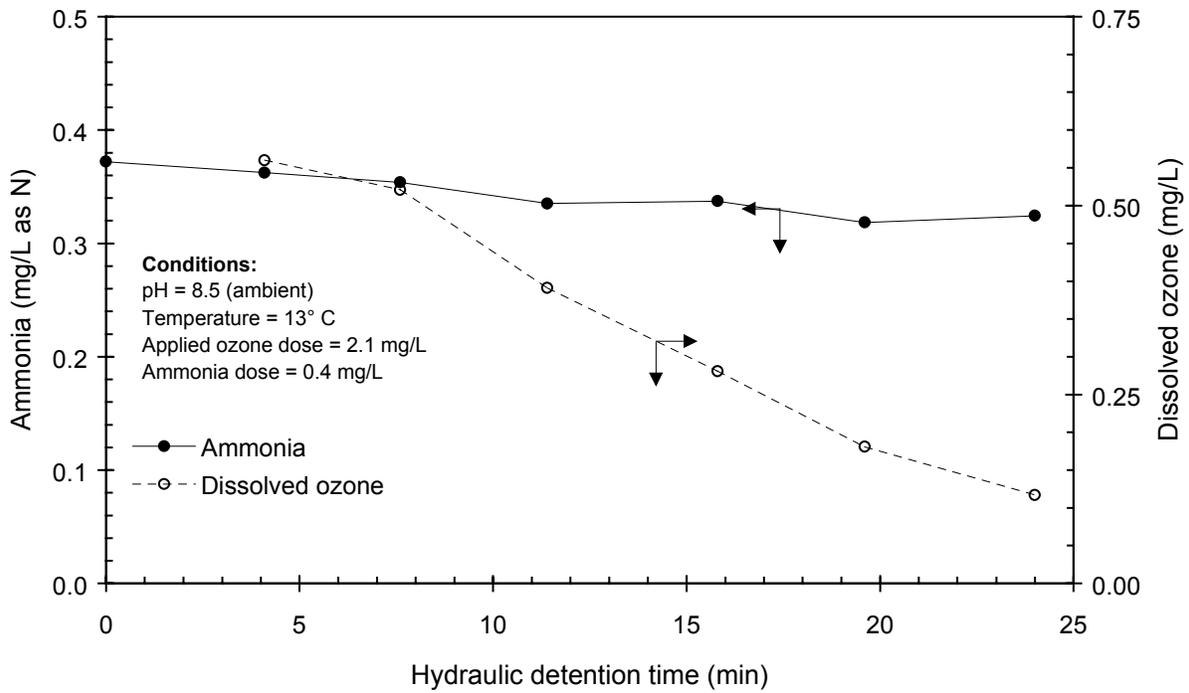


Figure 27. Profile of Dissolved Ozone and Ammonia through the Over/Under Contactor in CRW

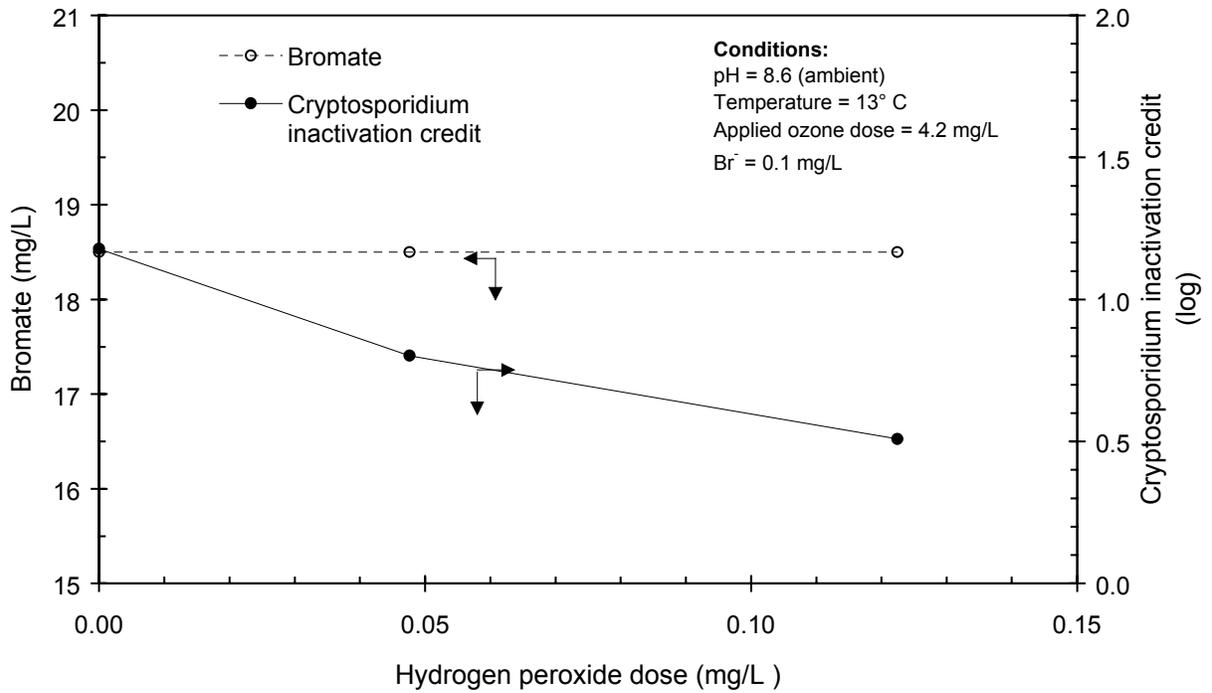


Figure 28. Effect of Hydrogen Peroxide Dose on Bromate Formation and *Cryptosporidium* Inactivation Credit in SPW

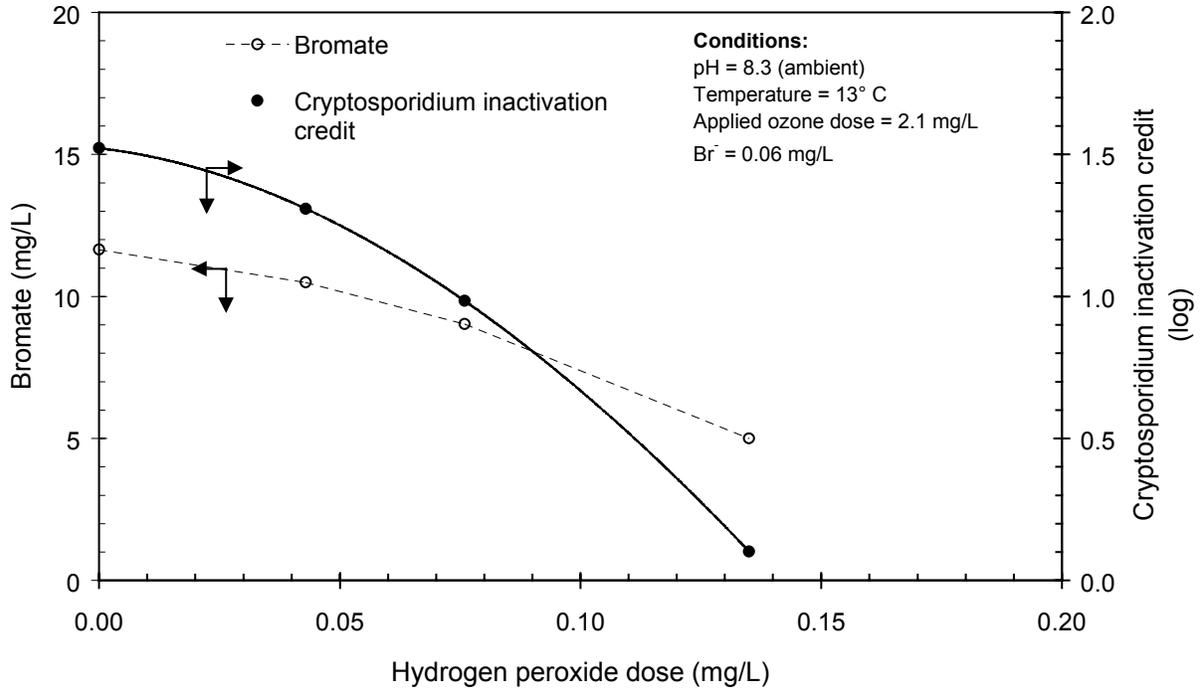


Figure 29. Effect of Hydrogen Peroxide Dose on Bromate Formation and *Cryptosporidium* Inactivation Credit in CRW

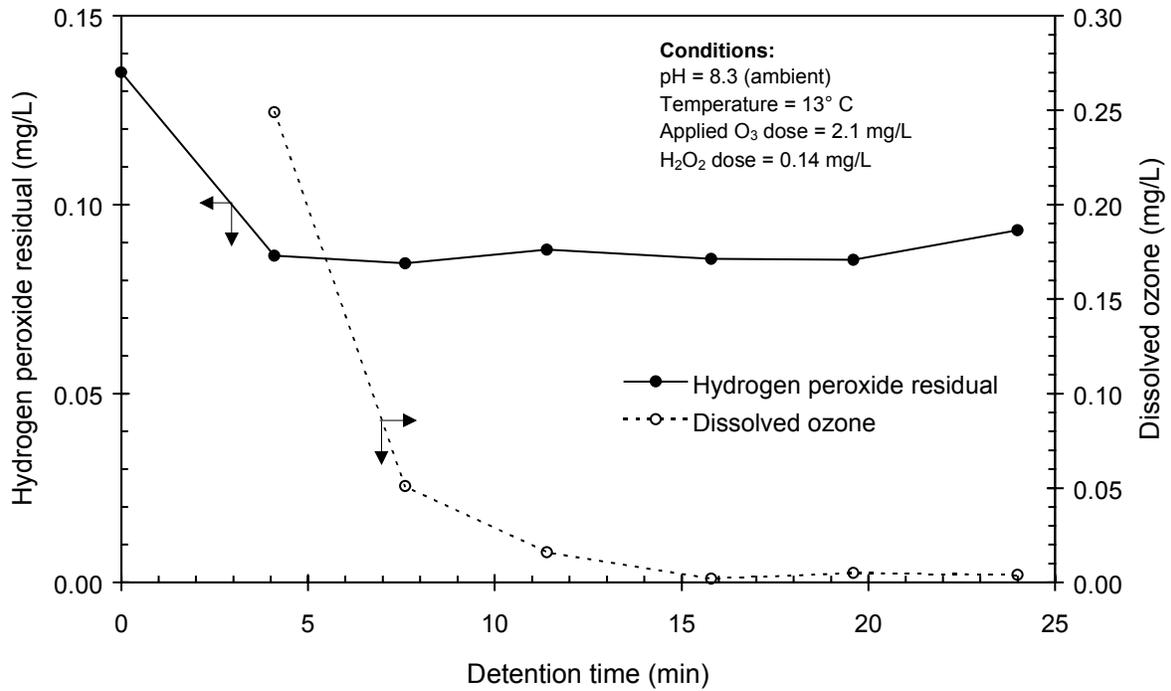


Figure 30. Profile of Dissolved Ozone and Hydrogen Peroxide Through the Over/Under Contactor in CRW

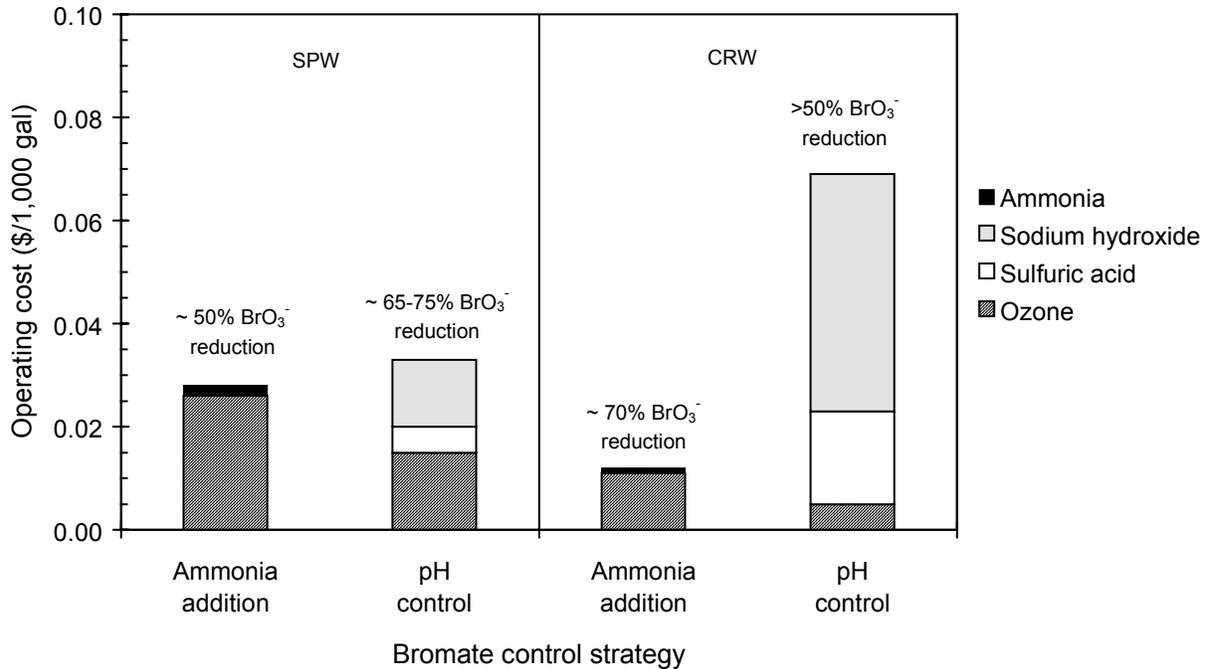


Figure 31. Cost Comparison of Ammonia Addition and pH Control (to comparable bromate control) for Bromate Minimization in SPW and CRW

Table 15. Raw Water Quality

Parameter	Units	Mean value and range*		
		SPW	CRW	Blend (1:1)
Temperature	°C	14 12-17	13	14
pH	--	8.1 8.1 – 8.6	8.3	8.4
TOC	mg/L	3.7 3.0-4.0	2.7 2.6-2.7	3.3 3.3-3.4
SUVA	(UV/TOC) x 100	3.0†	1.5†	--
Alkalinity	mg/L as CaCO ₃	73 63-75	131	103
Bromide	mg/L	0.15 0.09-0.18	0.07 0.06-0.09	0.11

*Range not shown if parameter did not significantly vary

† Characteristic values measured in previous studies.

Table 16. Average Bromate (ug/L) Formed at Estimated 1-log *Cryptosporidium* Inactivation Level*

Water	SPW	SPW	SPW/CRW	CRW
Bromide (mg/L)	~0.40	~0.10	~0.11	~0.07
pH 8	48	23	21	5
pH 7	12	8	7	4
pH 6.5	9	6	2†	nd‡
pH 6	4	2†	nd	nd

*Target inactivation was 1.0 ± 0.2 log *Cryptosporidium* inactivation credit

†Average includes one non-detect value (<3 µg/L)

‡Non-detect

Table 17. Results of Ammonia Addition and Hydrogen Peroxide Addition Experiments

Source Water													
SPW							CRW						
Influent pH	Ozonation pH	Applied Ozone Dose (mg/L)	Crypto. Inactivation Credit (log)	Ammonia Dose (mg/L as N)	Hydrogen Peroxide Dose (mg/L)	Bromate* (□g/L)	Influent pH	Ozonation pH	Applied Ozone Dose (mg/L)	Crypto. Inactivation Credit (log)	Ammonia Dose (mg/L-N)	Hydrogen Peroxide Dose (mg/L)	Bromate* (□g/L)
8.4	8.4	3.1	0.4	-	-	9	8.4	8.4	1.2	0.5	-	-	4
8.4	8.4	4.1	1.1	-	-	19	8.4	8.4	1.8	1.0	-	-	10
8.4	8.4	4.6	1.6	-	-	26	8.4	8.4	2.0	1.4	-	-	13
8.4	8.4	5.2	2.3	-	-	37	8.4	8.4	2.4	2.1	-	-	21
8.4	8.4	4.2	1.0	0	-	18	8.3	8.3	2.1	1.4	0	-	12
8.4	8.5†	4.2	0.7	0.2	-	12	8.3	8.4†	2.2	1.4	0.2	-	4
8.4	8.7†	4.2	0.7	0.4	-	11	8.3	8.5†	2.1	1.3	0.4	-	ND‡
8.4	8.8†	4.2	0.7	0.9	-	7	8.3	8.6†	2.1	1.2	0.7	-	ND
8.4	8.9†	4.2	0.5	2.0	-	5	8.3	8.7†	2.1	1.1	1.4	-	ND
8.6	9.0	3.1	0.2	1.0	-	4	8.3	8.6†	2.4	1.6	0.9	-	4
8.6	9.1	4.3	0.4	1.2	-	6	8.3	8.6†	2.1	1.1	0.9	-	3
8.6	9.0	4.6	0.5	1.1	-	8	8.3	8.6†	1.8	0.6	0.9	-	ND
8.6	9.0	5.3	1.1	0.9	-	11	8.3	8.6†	1.3	0.4	0.8	-	ND
8.6	8.6	4.3	1.2	-	0	19	8.3	8.3	2.1	1.5	-	0	12
8.6	8.6	4.2	0.8	-	0.05	19	8.3	8.3	2.1	1.3	-	0.04	11
8.6	8.6	4.1	0.5	-	0.12	19	8.3	8.3	2.1	1.0	-	0.08	9
8.6	8.6	4.1	0.4	-	0.13	19	8.3	8.3	2.1	0.1	-	0.14	5

*Reported bromate values represent the average of duplicate samples collected at 30-minute intervals.

† pH calculated using the Rothberg, Tamburini and Winsor model (Rothberg and Scuras, 1994).

‡ ND = not detected, MRL for bromate = 3 □g/L.

Table 18. Estimated Operating Costs* for Ammonia and pH Control Bromate Minimization Strategies at 1-Log *Cryptosporidium* Inactivation

	SPW		CRW	
	Ammonia	pH Control	Ammonia	pH Control
Bromate reduction (percent) at 1-log <i>Cryptosporidium</i> inactivation	50	65-75	~70	>50
pH	ambient (~9)‡	7	ambient (~8.5)‡	6.5
Estimated dose (mg/L)				
O ₃	~5†	3.0	2.1	1.0
H ₂ SO ₄	--	15	--	55
NaOH	--	13	--	45
NH ₃	1.0	--	0.5	--
Estimated cost (\$/1,000 gal)				
O ₃	0.026	0.015	0.011	0.005
H ₂ SO ₄	--	0.005	--	0.018
NaOH	--	0.013	--	0.046
NH ₃	0.002	--	0.001	--
Total estimated cost (\$/1,000 gal)	0.028	0.033	0.012	0.069
(\$/ac-ft.)	9.1	11.0	3.9	22.5
Estimated TDS increase (mg/L)	0	22	0	98

* The following costs were used to calculate this table: ozone specific energy, 7.5 kWh/lb; electricity, \$0.08/kWh; H₂SO₄, \$80/ton; NaOH, \$245/ton; NH₄OH, \$407/ton

† Ozone dose in these tests was higher than expected because of high influent pH plus pH increase due to ammonia addition; the actual dose required if pH can be maintained (e.g. with acid addition) to ~8 is estimated to be 4 mg/L or less.

‡ Metropolitan distributes water with a pH of 8.0–8.5.

2.4 Disinfection of Reclaimed Water

2.4.1 Chloramination and Ultraviolet Light for Disinfecting Tertiary Effluent

Ultraviolet disinfection is a viable alternative to chloramination for disinfection of reclaimed wastewater for non-potable reuse. However, use of newer, more effective and efficient systems in California has been limited by the acceptance of UV systems by California DHS according to the "California UV Guidelines." The effluent from the UV disinfection process had a consistent coliform concentration of <1 cfu/100 mL, the exception being when the turbidity was greater than 2 nephelometric turbidity units (NTU) and the UV dose less than 100 mWs/cm². The calculated UV dose during this testing was generally less than 140 mWs/cm², which is the minimum design dose for approved standard UV disinfection systems for non-potable reuse in California. The consistent performance of the UV process makes it a reliable and safe alternative to chloramination for tertiary effluent.

2.4.2 Techniques for Altering Transmittance of Reclaimed Wastewater

Both brewed and instant coffees, caffeinated and decaffeinated were found to be effective at increasing UV absorbance consistently across the UV light spectrum. Instant coffee was selected for use in the UV pilot study since it is easier to handle. The effect of coffee addition and transmittance reduction on coliphage inactivation was less extreme than expected when compared with collimated beam tests. Due to this finding, the use of a UV inhibitor compound when pilot testing UV disinfection equipment is a remaining subject of research. The use of coffee for transmittance reduction when testing pilot UV systems with a polychromatic lamp output and comparing to low pressure collimated beam tests needs to be studied further.

2.4.3 Developing UV Dose-Response Curves Using Collimated Beam

A collimated beam test, using coliphage MS-2, should be performed and compared to published curves to assure that the test results are consistent and comparable to other similar studies. Figure 32, for example, shows the results of a collimated beam test with MS-2 conducted by the University of California, Davis. As shown, the collimated beam results for these studies, using filtered secondary effluent from different wastewater treatment facilities, correlate well.

The collimated beam unit achieved seeded coliphage MS-2 inactivation at doses that are higher than those observed with the pilot unit for indigenous coliphage inactivation during Phase I. However, collimated beam tests were not conducted in phase I. Factors that impact Phase II coliphage inactivation results include seeding, batch mixing, and transmittance adjustment of UV influent. As discussed previously, seeding and batch mixing of the UV influent impacts the particle size distribution. Both the total number of particles and those greater than 5 microns were increased after seeding and mixing. The larger particles can harbor or shade the microorganisms and protect them from UV radiation.

Coffee was shown to uniformly absorb the UV light in the range of 200 to 300 nm. For medium-pressure lamps, the intensity of UV emissions for each wavelength in the range of 200 to 300 nm varies substantially. Therefore, the uniform absorption could potentially eliminate or substantially reduce some of these germicidal wavelengths. The impact of coffee addition for transmittance adjustment on medium pressure lamps, therefore, requires further investigation.

2.4.4 Establishing Dose Response Curves for UV Pilot Units

The performance of the Aquionics pilot unit was tested using the total coliform, and both indigenous and seeded coliphage. For the initial 6 weeks of the study, the pilot unit was operated continuously with the Green Acres Project (GAP) filter effluent. The UV transmittance during this phase ranged from 60 to 70 percent. The testing was done with 2 or 3 banks on line at different power sets for the UV banks. The pilot plant influent and effluent samples were collected and analyzed for indigenous total coliform and coliphage.

The second phase of the study included seeded studies with coliphage MS-2. The UV influent transmittance was also reduced to less than 55 percent using coffee. The pilot plant influent and effluent samples were collected and analyzed for total coliform and coliphage MS-2. Collimated beam tests were also conducted on the UV influent samples. The results of this study indicate that:

- The Aquionics pilot unit is capable of achieving both the 7 day median total coliform level of 2.2 MPN per 100 mL, and the 4-log indigenous coliphage inactivation at a dose of 100 mWs/cm², as demonstrated in phase I. This pilot unit performance can be correlated to the collimated beam unit using a calibration coefficient. The dose calculation for Aquionics unit was developed using their proprietary model.

It is typical that the performance of full-scale facilities deviates from collimated beam results due to complex system hydraulics. As a consequence, it is important that all equipment be tested in accordance with the test protocol detailed in Appendix A. Collimated beam tests are crucial in establishing that the results are consistent with other studies. Also collimated beam tests provide the basis for dose estimation and comparison of various UV systems performance. It is important to compare the collimated beam results with the performance of the pilot units using indigenous coliphage and coliform as well as seeded coliphage MS-2.

2.4.5 Evaluating Pulsed UV on Clarified Secondary Effluent

Based on the total coliform results after the Pulsed UV equipment modification, it was concluded that the hydraulic flow problems were improved and previous hydraulic “short-circuiting” was eliminated. The performance of the pulsed UV system is now optimized for further testing at OCWD, including seeded MS-2 and parasite testing.

2.5 Low Pressure Membrane for Reclaimed Water

2.5.1 Membrane Performance on Clarified Secondary Effluent

The following section contains a summary of the performance of microporous and semi permeable membranes during the reclamation of secondary effluent. Data on flux, pressure and water quality are presented for the pressure driven modular systems (manufacturer, Memcor & Pall) and the suction driven immersed system (manufacturer, Zenon). Water quality data only is presented for one pressure driven modular system (manufacturer, KOCH) due to repeated failure of the microporous membrane.

2.5.1.1 Pressure Driven Modular System (Memcor)

Six M10 modules (0.2 μ m polypropylene membranes \times ft²/module) were challenged with clarified secondary effluent containing a 3 ppm chloramine residual over 290-day period from January 6 to October 22, 1997. The 0.2 μ m polypropylene membranes operated for 281 days (97 percent on-line factor) in the dead end mode (0 percent recirculation) at an instantaneous flux of 22 gallons per square foot per day (gfd). The membranes were backwashed every 20 minutes for a total recovery of 87 percent. The membranes were chemically cleaned on 16 occasions to remove fouling layer and restore membrane permeability. The effectiveness of the chemical clean was assessed by the reduction in transmembrane pressure (TMP) and the elapsed time between cleanings. In most cases, a two stage cleaning process (low pH solution followed by a high pH solution) was found to be more effective than a single clean at high pH. A single clean in a warm (104°F) aqueous solution of 1 percent sodium hydroxide containing a non ionic surfactant at pH 12 reduced the TMP to 7.1 ± 2.3 psi and the time interval between chemical cleanings was 13 ± 10 days. In contrast, cleaning the membrane with a warm (104°F) aqueous solution of 2 percent citric acid at pH 2 prior to the caustic clean at pH 12 reduced the TMP to 3.0 ± 1.0 psi with an interval between chemical cleanings of 27 ± 11 days (Table 19a). Low pH cleaning solutions using citric acid, however, were not effective at restoring permeability on all 0.2 μ m polypropylene membranes. A set of 4 M10C modules that had been fouled by metal silicates (see Appendix 1) were cleaned with a 2 percent solution of citric acid followed by a 1 percent solution of sodium hydroxide. Under this cleaning regime, the TMP across the membranes was reduced to 7.3 ± 0.3 with a chemical cleaning interval of 3 days (Table 19a). Similar results were obtained when the double cleaning regimen was supplemented with a 0.5 percent solution of ethylene-diamene-tetra-acetic acid and 0.5 percent hydrogen peroxide. The permeability of the fouled M10C modules was restored by cleaning the membrane with a 1 percent solution of ammonium bifluoride; the metal silicates were soluble in the hydrofluoric acid liberated in an aqueous solution of ammonium bifluoride. Under this cleaning regime, the TMP across the membranes was reduced to 4.0 ± 0.3 with a chemical cleaning interval of 24 ± 8 days (Table 19a).

Table 19. Low Pressure Microporous Membrane System Performance Summary

Table 19a: Pressure driven, shell side active surface, air backwash (20 minute intervals)									
Membrane	Flux (gfd)	Feed Pressure (psi)	Recirculation (% of feed)	Recovery (% of feed)	TMP		Cleaning Interval		
					Initial (psi)	Final (psi)	Days	No. (n)	
M10	22	20	0%	87%	7.1 ±2.3	14.1 ±1.5	13 ±101	9	
M10	22	20	0%	87%	3.0 ±1.0	11.7 ±3.9	27 ±11 ²	7	
M10C4	22	20	0%	87%	7.3 ±0.3	15.1 ±2.3	3 ±03	2	
M10C4	22	20	0%	87%	6.6	15	41	1	
M10C4	22	20	0%	87%	8.5	15.1	4 ²	1	
M10C5	22	20	0%	87%	8.1	15.8	91	1	
M10C5	22	20	0%	87%	4 ±0.28	14.2 ±0.14	23.5 ±7.8 ²	2	
Table 19b: Pressure driven, shell side active surface, liquid backwash (15 minute intervals) + air scour (60 minute intervals)									
Membrane	Flux (gfd)	Feed Pressure (psi)	Recirculation (% of feed)	Recovery (% of feed)	TMP		Runs		
					Initial (psi)	Final (psi)	Length (days)	No. (n)	
13K MWCO PAN	19.5	35	7	93%	15.2 ±3.6	24.6 ±6.5	16.6 ±8.2	7	
0.1 □m PP	32	35	10%	90%	8.2 ±5.7	16.5 ±4.8	15.5 ±7.6	6	
0.1 □m PVDF	32	35	10%	90%	5.1 ±2.4	28 ±4.8	16.6 ±10.9	11	

Membranes cleaned with 0.1 percent NaOH + non-ionic surfactant only

Membranes cleaned with 2 percent citric acid followed by 0.1 percent NaOH + non-ionic surfactant

Membranes cleaned with 2 percent NaOH +0.5 percent EDTA +0.5 percent H₂O₂ in RO Permeate

Runs occurred before one-time cleaning with 1 percent ABF solution to remove silica fouling

Runs occurred after one-time cleaning with 1 percent ABF solution to remove silica fouling

Table 20. Water Quality Data from Memcor MF

Table 20a. M10 Water Quality Summary-Pre fiber repair							
Parameter	Count	Min	Max	Mean	Median	70%	95%
T.coliform	15	0.10	30	10.3	7.0	14.5	30.0
F.coliform	15	0.10	1	0.4	0.1	1.0	1.0
HPC	14	0.20	500	77.7	20.5	46.0	402.5
TSS	15	0.10	1	0.2	0.1	0.1	0.5
SiO2	15	18.00	25	21.4	21.0	22.9	24.6
TOC	15	8.40	10	9.5	9.5	9.9	10.2

Table 20b. M10 Water Quality Summary-Post fiber repair							
Parameter	Count	Min	Max	Mean	Median	70%	95%
T.coliform	21	0.10	2	0.2	0.1	0.1	0.1
F.coliform	21	0.10	0	0.1	0.1	0.1	0.1
HPC	21	0.20	250	16.2	0.2	10.0	20.0
TSS	23	0.10	2	0.3	0.1	0.1	1.4
SiO2	23	9.20	31	22.3	23.6	24.0	26.2
TOC	23	7.68	9	8.4	8.2	8.8	8.9

Table 20c. M10C Water Quality Summary – Post ABF clean							
Parameter	Count	Min	Max	Mean	Median	70%	95%
T.coliform	17	0.10	3	0.6	0.1	1.0	2.2
F.coliform	18	0.10	0	0.1	0.1	0.1	0.1
HPC	18	0.20	3800	240.6	5.0	13.8	867.5
TSS	18	0.10	2	0.2	0.1	0.1	0.4
SiO2	18	16.30	24	20.1	19.9	22.1	24.0
TOC	18	8.23	11	9.5	9.5	10.1	10.7

Filtrate from the system containing the M10 and M10C modules was sampled on 58 occasions during the 281 days of operation (Table 20). Three general water quality trends were apparent; filtrate from the M10 and M10C suitable for processing with RO; the filtrate quality improved after broken fibers were detected and repaired; and, aggressive cleaning with ammonium bifluoride did not impact filtrate quality. The M10 modules were repaired on May 17 after 161 days of operation; four of the six modules contained between three and six broken fibers. In filtrate samples collected during the first 161 days of operation the 95th percentile for total coliform was 30 cfu/100 ml (Table 20a); in samples collected after the M10 modules were repaired the 95th percentile for total coliforms was to <1 cfu/100 ml (Table 20b). Similarly, the 95th percentile for total coliforms in samples collected following cleaning with ABF was 2.2 cfu/100 ml (Table 20c).

Thin film composite reverse osmosis (RO) membranes (TFCHR) operated downstream of the 0.2 μ m polypropylene membranes (M10 and M10C) at a flux of 10.4 gfd and a recovery of 75 percent for approximately 7900 h. The RO membranes were cleaned on three occasions at 1000 h, 5000 h and 5500 h. In the first 1,000 hours the normalized feed pressure (NFP) increased from 120 psi to 175 psi. The first clean reduced the NFP from 175 to 118 psi. Over the next 4,500 hours the NFP increased from 118 psi to 210 psi. After the second clean the NFP was reduced to 165 psi and increased to 205 psi after 500 hours. The final clean reduced the NFP from 205 to 150 psi. Over the next 2,500 hours the normalized feed pressure increased from 150 to 218 psi (Figure 3).

2.5.1.2 Pressure Driven Modular System (Pall)

Four LGV 5010 modules (13,000 MWCO polyacrylonitrile membranes xft^2 /module) were challenged with clarified secondary effluent containing a 3 ppm chloramine residual over 170-day period from March 4 to September 21, 1997. The 13,000 MWCO polyacrylonitrile membranes operated for 116 days (68 percent on-line factor) in the cross flow mode (9 percent recirculation) at an instantaneous flux of 19.5 gallons per square foot per day (gfd). The membranes were chemically cleaned on 7 occasions to remove fouling layer and restore membrane permeability. Chemical cleaning decreased the TMP from 24.6 ± 6.5 to 15.2 ± 3.6 psi. Under these conditions the interval between chemical cleaning was 16.6 ± 8.2 days (Table 19b).

Four XLSV-5200 modules (0.1 μ m polyethylene membranes xft^2 /module) were challenged with clarified secondary effluent containing a 3 ppm chloramine residual over 97-day period from August 1 to November 5, 1997. The 0.1 μ m polyethylene membranes operated for 93 days (96 percent on-line factor) in the cross flow mode (10 percent recirculation) at an instantaneous flux of 32 gallons per square foot per day (gfd). The membranes were chemically cleaned on 6 occasions to remove fouling layer and restore membrane permeability. Chemical cleaning decreased the TMP from 16.5 ± 4.8 to 8.2 ± 5.7 psi. Under these conditions the interval between chemical cleaning was 15.5 ± 7.6 days (Table 19b).

Four XUSV-5203 modules (0.1 μ m polyvinylidene fluoride membranes xft^2 /module) were challenged with clarified secondary effluent containing a 3 ppm chloramine residual over 221-day period from November 7 1997 to June 18, 1998. The 0.1 μ m polyvinylidene fluoride membranes operated for 183 days (82 percent on-line factor) in the cross flow mode (10 percent recirculation) at an instantaneous flux of 32 gallons per square foot per day (gfd). The membranes were chemically cleaned on 11 occasions to remove fouling layer and restore membrane permeability. Chemical cleaning decreased the TMP from 28 ± 4.8 to 5.1 ± 2.4 psi. Under these conditions the interval between chemical cleaning was 16.6 ± 10.9 days (Table 19b).

Filtrate from system containing the LGV 5010 modules, XLSV-5200 modules and the XUSV-5203 modules was sampled on 112 occasions during the 484 days of operation (Table 21). Two general water quality trends were apparent; filtrate from the symmetric 0.1 mm polyethylene and polypropylene membranes was consistently better than the asymmetric 13,000 MWCO polyacrylonitrile (PAN) membranes; moreover, the filtrate quality of the PAN membrane did not improve after broken fibers were detected and repaired. The 95th percentile for total coliforms in filtrate samples collected from the XLSV-5200 and XUSV-5203 modules was 1 cfu/100 ml (Table 21c) and <1 cfu/100 ml (Table 21d) respectively. In contrast the 50th and 70th percentiles for total coliforms in filtrate samples collected from the LGV 3010 modules were 14

cfu/100 ml and 150 cfu/100 ml respectively (Table 21a). The LGV 5010 modules containing the 13000 MWCO PAN membranes were repaired on April 10 after 32 days of operation; the modules contained between approximately seven broken fibers. Fiber repair only marginally improved filtrate water quality by decreasing the 50th percentile for total coliforms to <1 cfu/100 ml (Table 21b).

Thin film composite reverse osmosis (RO) membranes (TFCHR) operated downstream of the 13,000 MWCO polyacrylonitrile membranes at a flux of 10.4 gfd for approximately 700 h. The RO membranes were cleaned once at 400 hours. In the first 400 h the normalized RO feed pressure increased from 100 psi to 185 psi. The chemical clean at 400 h reduced the normalized RO pressure from 185 to 158 psi; this cleaning coincided with the maintenance of the LGV 5010 modules to repair broken fibers. Over the next 300 hours the pressure increased from 158 psi to 185 psi (Figure 4). After 700 hours a new set of TFCHR RO membranes were installed in the RO system and the LGV 5010 modules were replaced with XLSV-5200 module. The second set of TFCHR RO membranes operated downstream of the 0.1 μ m polyethylene membrane for 1,250 hours and the normalized feed pressure increased from 100 to 120 psi. The use of the integral 0.1 μ m polyethylene membranes removed the coliforms, suspended solids and other colloidal particles that contribute to fouling of the RO surface. Consequently, the transition from the compromised 13,000 MWCO polyacrylonitrile membrane to the uncompromised 0.1 μ m polyethylene membrane lowered the silt density index and reduced the total coliform count in the RO feed water. After 1,250 hours the XLSV-5200 modules were replaced with XUSV-5203 modules containing 0.1 μ m polyvinylidene fluoride (PVDF) membranes. The RO membranes were cleaned and operated downstream of the PVDF membranes at 10.4 gfd for 4,000 hours. Under these conditions the normalized feed pressure increased from 118 psi to 225 psi. Table 22 presents data for the microporous membrane filtrate (RO feed) and RO permeate.

2.5.1.3 Pressure Driven Modular System (KOCH)

Four sets of KOCH PM-100 modules were operated on clarified secondary effluent using a protocol which differed from the normal protocol outlined in section 3.1. The fiber lumens (active surface) were backwashed using RO permeate containing either free chlorine residual (60 to 200 mg/L) or hydrogen peroxide (150 to 1600 mg/L). The more aggressive backwash approach was used to mitigate membrane fouling. Table 23 presents data on the filtrate water quality for the three 43 inch and one 72 inch modules. In general, the 50th percentile for total and fecal coliforms for three of the five filtrate sampling sets (Table 23a to e) was greater than 1 cfu/100 ml. This implies that one out of two samples collected tested positive for both total and fecal coliform. Moreover; isolation and repair of the fibers did not significantly improve filtrate quality; the 75th percentile for fecal decreased from 295 cfu/100 ml (n = 7 samples; Table 21a) to 135 cfu/100ml (n = 4; Table 23c). The presence of fecal coliforms can only be attributed to fiber breakage and not re-growth. It was possible, however, that use of strong oxidants (aggressive backwash) coupled with the membrane symmetry and module configuration used in the KOCH system increased the incidence of fiber failure leading to the passage of coliforms (both total and fecal) across the membrane. The water quality data for the KOCH system relative to other systems tested on the secondary effluent is discussed in 3.5, Discussion and Analysis.

Table 21. Water Quality Data from Pall MF

Table 21a. PAN Water Quality Summary - Pre Fiber Repair							
Parameter	Count	Min	Max	Mean	Median	70%	95%
T.coliform	9	0.10	600	109.6	14.0	150.0	440.0
F.coliform	9	0.10	18	4.2	0.1	9.0	14.8
HPC	8	0.20	9000	1536.4	43.0	1300.0	6620.0
SIO2	9	17.00	24	19.9	20.0	20.9	23.3
TOC	9	7.98	10	9.1	8.8	9.8	10.1

Table 21b. PAN Water Quality Summary - Post Fiber Repair							
Parameter	Count	Min	Max	Mean	Median	70%	95%
T.coliform	11	0.10	10000	1174.6	0.1	210.0	6150.0
F.coliform	11	0.10	260	45.6	0.1	20.5	230.0
HPC	11	0.20	15000	2215.9	700.0	1600.0	9000.0
SIO2	11	10.20	26	22.4	23.8	24.2	25.2
TOC	11	8.02	9	8.7	8.8	9.1	9.2

Table 21c. PE Water Quality Summary							
Parameter	Count	Min	Max	Mean	Median	70%	95%
T.coliform	10	0.10	1	0.2	0.1	0.1	0.6
F.coliform	10	0.10	0	0.1	0.1	0.1	0.1
HPC	7	0.20	140	28.7	0.2	30.0	113.0
SiO2	7	22.40	28	23.8	23.0	23.9	27.0
TOC	7	8.29	9	8.6	8.6	8.7	8.8

Table 21d. PVDF Water Quality Summary							
Parameter	Count	Min	Max	Mean	Median	70%	95%
T.coliform	82	0.10	0	0.1	0.1	0.1	0.1
F.coliform	82	0.10	0	0.1	0.1	0.1	0.1
HPC	28	0.20	460	44.3	12.5	50.0	126.5
SiO2	31	13.60	25	19.3	18.6	20.0	23.9
TOC	31	8.34	13	9.7	9.5	10.0	11.0

Table 22. Water Quality Data for Pall RO Effluent

Table 22a. PAN Water Quality Summary-Pre fiber repair							
Parameter	Count	Min	Max	Mean	Median	70%	95%
B	9	0.2	0.3	0.3	0.3	0.3	0.3
Br	9	0.2	0.4	0.2	0.2	0.2	0.4
Cl	9	2.4	12.5	4.5	2.9	3.7	10.7
FCOLIF	9	1.0	5.0	1.4	1.0	1.0	3.4
HPCSP	8	2.0	230.0	46.8	4.0	18.5	188.0
NH3-N	9	0.7	2.3	1.2	1.1	1.1	1.9
NO3-N	9	0.1	0.3	0.1	0.1	0.1	0.2
ORG-N	9	0.1	0.2	0.1	0.1	0.1	0.2
SIO2	9	1.0	1.9	1.1	1.0	1.0	1.5
TCOLIF	9	1.0	27.0	5.3	1.0	2.8	20.6
TDS	9	18.0	44.0	31.6	34.0	40.0	44.0
TKN	9	0.7	2.5	1.2	1.1	1.2	2.0
TOC	9	0.1	0.5	0.3	0.3	0.4	0.5
TOTALK	9	8.8	26.8	14.1	11.8	13.4	23.7

Table 22b. PAN Water Quality Summary-post							
Parameter	Count	Min	Max	Mean	Median	70%	95%
B	9	0.2	0.3	0.2	0.2	0.2	0.3
Br	9	0.2	0.2	0.2	0.2	0.2	0.2
Cl	9	2.9	7.6	5.1	4.8	5.5	7.1
FCOLIF	9	1.0	3.0	1.2	1.0	1.0	2.2
HPCSP	4	2.0	90.0	46.8	47.5	90.0	90.0
NH3-N	9	0.4	2.0	1.3	1.4	1.6	2.0
NO3-N	9	0.1	0.4	0.2	0.2	0.3	0.4
ORG-N	9	0.1	0.4	0.2	0.2	0.2	0.4
SIO2	9	1.0	1.0	1.0	1.0	1.0	1.0
TCOLIF	9	1.0	80.0	13.9	2.0	11.4	55.2
TDS	9	12.0	62.0	26.0	22.0	25.2	50.8
TKN	9	0.8	2.2	1.4	1.6	1.6	2.1
TOC	9	0.2	0.9	0.5	0.6	0.7	0.9
TOTALK	9	11.0	26.2	15.8	15.0	15.8	22.6

Table 22c. PE Water Quality Summary							
Parameter	Count	Min	Max	Mean	Median	70%	95%
B	2	0.3	0.3	0.3	0.3	0.3	0.3
Br	2	0.2	0.2	0.2	0.2	0.2	0.2
Cl	2	4.0	4.1	4.1	4.1	4.1	4.1
FCOLIF	2	1.0	1.0	1.0	1.0	1.0	1.0
NH3-N	2	1.7	2.2	2.0	2.0	2.1	2.2
NO3-N	2	0.2	0.2	0.2	0.2	0.2	0.2
ORG-N	2	0.1	0.1	0.1	0.1	0.1	0.1
SIO2	2	1.0	1.0	1.0	1.0	1.0	1.0
TCOLIF	2	1.0	1.0	1.0	1.0	1.0	1.0
TDS	2	13.0	15.0	14.0	14.0	14.4	14.9
TKN	2	1.7	2.3	2.0	2.0	2.1	2.3
TOC	2	0.7	0.9	0.8	0.8	0.9	0.9
TOTALK	2	15.4	15.7	15.6	15.6	15.6	15.7

Table 22d. PVDF Water Quality Summary							
Parameter	Count	Min	Max	Mean	Median	70%	95%
B	26	0.1	0.3	0.2	0.3	0.3	0.3
Br	26	0.2	0.4	0.2	0.2	0.2	0.2
Cl	26	3.3	83.1	7.3	4.1	4.9	6.0
FCOLIF	26	1.0	1.0	1.0	1.0	1.0	1.0
NH3-N	26	0.7	2.1	1.1	1.2	1.2	1.6
NO3-N	26	0.1	0.4	0.2	0.2	0.2	0.4
ORG-N	26	0.1	0.5	0.2	0.1	0.2	0.5
SIO2	26	1.0	1.0	1.0	1.0	1.0	1.0
TCOLIF	26	1.0	1.0	1.0	1.0	1.0	1.0
TDS	26	4.0	72.0	19.7	15.0	21.5	60.5
TKN	26	0.8	2.1	1.3	1.3	1.4	1.8
TOC	26	0.1	1.0	0.4	0.3	0.4	0.9
TOTALK	26	9.7	274.0	29.0	15.5	20.5	53.0

Table 23. Water Quality Data from Koch MF

Table 23a. Water Quality Summary-43 inch Pre Fiber Repair							
Parameter	Count	Min	Max	Mean	Median	75%	95%
T.coliform	7	0.10	20000	3475.7	120.0	2100.0	15200.0
F.coliform	7	0.10	400	141.9	2.0	295.0	376.0
HPC	6	0.20	380000	63738.4	355.0	1407.5	285425.0
TSS	7	13.00	75	26.2	20.0	20.2	58.6
SIO2	7	13.00	75	26.2	20.0	20.2	58.6
TOC	7	1.99	9	8.0	8.9	9.2	9.5

Table 23b. Water Quality Summary 72-inch							
Parameter	Count	Min	Max	Mean	Median	70%	95%
T.coliform	7	0.10	12000	1909.9	31.0	650.0	8730.0
F.coliform	7	0.10	200	53.8	0.1	88.0	191.0
HPC	7	40.00	48000	12141.4	2500.0	16500.0	42600.0
TSS	7	11.10	31	21.5	21.3	23.5	28.7
SIO2	7	11.10	31	21.5	21.3	23.5	28.7
TOC	7	5.42	10	8.2	8.8	9.1	9.5

Table 23c. Water Quality Summary-43-inch Post Fiber Repair							
Parameter	Count	Min	Max	Mean	Median	70%	95%
T.coliform	4	1.00	22000	4857.3	2000.0	3850.0	16990.0
F.coliform	4	0.10	200	70.3	18.0	135.0	194.0
HPC	4	20.00	21000	8002.9	1500.0	15500.0	20400.0
TSS	3	5.70	25	19.7	22.9	23.9	25.0
SIO2	3	5.70	25	19.7	22.9	23.9	25.0
TOC	3	4.32	9	7.6	8.2	8.5	8.6

Table 23d. Water Quality Summary-43-inch Second Set							
Parameter	Count	Min	Max	Mean	Median	70%	95%
T.coliform	6	0.10	75	24.2	2.1	50.5	72.8
F.coliform	6	0.10	11	1.9	0.1	0.1	8.3
HPC	6	0.20	1500	275.1	45.5	56.3	1139.8
TSS	6	17.10	23	21.4	22.1	22.2	22.5
SIO2	6	17.10	23	21.4	22.1	22.2	22.5
TOC	6	8.22	11	9.6	9.8	9.9	10.8

Table 23e. Water Quality Summary-43-inch Third Set							
Parameter	Count	Min	Max	Mean	Median	70%	95%
T.coliform	36	0.10	58000	4817.7	1250.0	4700.0	20000.0
F.coliform	36	0.10	8600	251.1	2.0	11.0	83.3
HPC	10	80.00	9000	1768.0	820.0	1275.0	6435.0
TSS	15	17.90	55	24.7	23.3	24.7	37.2
SIO2	15	17.90	55	24.7	23.3	24.7	37.2
TOC	15	6.82	15	9.4	8.5	9.9	14.0

2.5.1.4 Suction Driven Immersed System

Four ZW150 M10 cassettes configured with ADC membranes (0.08 μ m approximately 25 percent porosity \times ft²/module) were loaded into a 1280 ft³ tank challenged with clarified secondary effluent containing a 3 ppm chloramine residual over a 96-day period from January 6 to June 10, 1997 (63 percent utility). The ADC membranes operated under suction in the dead end mode at an instantaneous flux of 14.6 gfd (

Table 24). The ratio between filtrate collected under suction from the lumen side of the membrane to overflow from the tank was initially 9:1 (90 percent recovery). The membranes were cleaned under these conditions on 10 occasions; the first six cleaning attempts used a one percent solution of sodium hydroxide. The 1 percent sodium hydroxide was applied in what is referred to as an "*in situ* clean", since the concentrated secondary effluent was not drained from the tank prior to cleaning. The *in situ* clean produced an initial vacuum on the lumen side of the fiber to 9.4 ± 3.6 inches of Hg and a 8.2 ± 7.8 day cleaning interval (

Table 24). On four occasions the secondary effluent was drained from the tank prior to cleaning (referred to as an "empty tank clean"). Under empty tank conditions the membranes were cleaned twice with 1 percent sodium hydroxide and twice with a 2 percent solution of citric acid followed by a 1 percent solution of sodium hydroxide. The empty tank clean regime reduced the vacuum to 5.5 ± 3.6 inches of mercury (in Hg), however, the use of the 2 percent citric acid solution prior to the 1 percent sodium hydroxide solution extended the interval between cleaning from 3.5 ± 0.5 to 20 ± 1 days.

One ZW500 cassette configured with CFA membranes (0.08 μ m approximately 50 percent porosity \times ft²/module) was challenged with clarified secondary effluent containing a 3 ppm chloramine residual over 103-day period from June 10 to October 22, 1997 (76 percent utility). The CFA initially operated at a flux of 38.8 gfd; under these conditions the vacuum increased from 7.2 ± 2.8 to 13.1 ± 3 inches of mercury (in Hg) in approximately 8 days (

Table 24). Decreasing the flux to a value between 19.4 and 22.5 gfd, decreased the initial vacuum to 5 to 7 in Hg; operation at a lower flux also extended the cleaning interval from 8 to 24 days.

On one occasion both the ADC and CFA membranes operated at a recovery of 97 percent by increasing the ratio between filtrate collected under suction from the lumen side of the membrane to overflow from the tank to 32:1. Both membranes were cleaned using a 1 percent sodium hydroxide solution applied under empty tank conditions. Increasing the process recovery from 90 percent to 97 percent did not significantly alter the initial vacuum or interval between cleanings (

Table 24). Filtrate from the suction driven system containing the ZW 500 modules configured with the ADC and CFA membranes was sampled on 20 occasions during the 199 days of operation (

Table 25 presents water quality data for the RO permeate.

Table 25). In general, the filtrate water quality was suitable for use as feed water for reverse osmosis.

Table 24. Summary Suction Low Pressure Microporous Membrane System Performance

System Three: Suction driven, shell side active surface, continuous aeration of shell liquid backwash (15 minute intervals)								
Membrane	Flux (gfd)	Feed Pressure (psi)	Recirculation (% of feed)	Recovery (% of feed)	Vacuum		Cleaning Interval	
					Initial (in Hg)/Final (in Hg)	Days/ Runs (n)		
ADC	14.6	10	0%	90%	9.4 ±3.6	14.2 ±2.8	8.2 ±7.81	6
ADC	14.6	10	0%	90%	5.5 ±3.6	14.3 ±1.2	3.5 ±0.52	2
ADC	14.6	10	0%	90%	5.5 ±3.6	14.3 ±1.2	20 ±13	2
ADC	14.6	10	0%	97%	7	15.7	11	1
CFA	38.8	10	0%	90%	8.2 ±0.3	14.4 ±1.2	15 ±2.81	2
CFA	38.8	10	0%	90%	7.2 ±2.8	13.1 ±3	8.3 ±3.52	3
CFA	38.8	10	0%	90%	10	12	3	1
CFA	19.4	10	0%	90%	5	5	10	1
CFA	22.5	10	0%	97%	7	15.8	11	1

Notes: Operation changes and cleaning intervals

1. Membranes cleaned with 0.1 percent NaOH + non-ionic surfactant in situ
2. Membranes cleaned with 0.1 percent NaOH + non-ionic surfactant under empty tank conditions
3. Membranes cleaned with 2 percent citric acid followed by 0.1 percent NaOH + non-ionic surfactant under empty tank conditions

Table 25 presents water quality data for the RO permeate.

Table 25. Zenon RO Permeate Water Quality Fed from Secondary Effluent

Table 25a - ADC Membrane							
Parameter	Count	Min	Max	Mean	Median	70%	95%
B	11	0.2	0.4	0.3	0.3	0.3	0.4
Br	11	0.2	0.2	0.2	0.2	0.2	0.2
Cl	11	1.5	5.1	3.0	2.5	3.9	4.9
FCOLIF	11	1.0	1.0	1.0	1.0	1.0	1.0
HPCSP	10	2.0	340.0	67.9	10.0	68.5	254.5
NH3-N	11	0.7	2.5	1.5	1.3	1.8	2.4
NO3-N	11	0.1	0.3	0.2	0.1	0.2	0.3
ORG-N	11	0.1	1.0	0.3	0.1	0.2	1.0
SIO2	11	1.0	1.6	1.1	1.0	1.0	1.3
TCOLIF	11	1.0	87.0	18.4	1.0	1.0	76.5
TDS	11	8.0	40.0	22.7	18.0	34.0	39.0
TKN	11	0.9	2.5	1.6	1.3	1.9	2.4
TOC	11	0.1	7.9	1.0	0.4	0.4	4.2
TOTALK	11	9.6	17.2	12.6	12.3	13.2	16.8
Table 25b - CFA Membrane							
Parameter	Count	Min	Max	Mean	Median	70%	95%
B	9	0.2	0.3	0.3	0.3	0.3	0.3
Br	9	0.2	0.2	0.2	0.2	0.2	0.2
Cl	9	3.6	4.9	4.1	4.0	4.2	4.7
FCOLIF	9	1.0	1.0	1.0	1.0	1.0	1.0
HPCSP	1	85.0	85.0	85.0	85.0	85.0	85.0
NH3-N	9	0.5	1.5	1.0	1.0	1.1	1.4
NO3-N	9	0.2	0.4	0.3	0.3	0.3	0.4
ORG-N	9	0.1	0.8	0.2	0.1	0.2	0.6
SIO2	9	1.0	1.0	1.0	1.0	1.0	1.0
TCOLIF	9	1.0	1.0	1.0	1.0	1.0	1.0
TDS	9	11.0	106.0	36.9	22.0	36.4	89.2
TKN	9	0.5	1.8	1.2	1.1	1.5	1.8
TOC	9	0.3	0.9	0.5	0.5	0.6	0.8
TOTALK	9	10.9	51.3	21.6	15.4	17.9	46.4

Thin film composite reverse osmosis (RO) membranes (TFCHR) operated downstream of the 0.08 μm ADC and CFA membranes at a flux of 10.4 gfd and a recovery of 75 percent for approximately 3900 h. The normalized feed pressure increased from 100 to 150 psi after 400

hours of operation. The membranes were rinsed in RO permeate and the NFP returned to 120 psi and increased to 185 psi after an additional 500 hours operation. At 900 hours the membranes were rinsed in RO permeate and the feed pH was decreased from 8.0 to 7.0; under these conditions the normalized feed pressure increased from 120 to 175 after 3,000 hours of operation. Options for Treating Microporous Membrane Residuals, Backwash

Backwash recovery experiments were conducted on a system configured with 0.2 μ m polypropylene membranes operating with an air backwash followed by a feed flush on the shell side of the membrane. For such as system, process recovery was dependent on flux, backwash interval and the volume of feed used to flush the shell side of the membrane during the backwash. At a backwash interval of 15 minutes the process recovery increased from 80 percent to 85 percent as the flux increased from 22.3 to 30.3 gfd. at a backwash interval of 30 minutes the recovery increased from 88 percent to 92 percent as the flux increased from 22.3 to 30.3 gfd. Backwash for the following experiments was generated at a recovery of 85 percent using a flux of 22.3 gfd and a backwash interval of 18 minutes. Filtrate collected from the first stage MF (Table 26a) and the second stage MF (Table 26b) contained <1 mg/L suspended solids, a turbidity of 0.1 NTU, and <1 CFU/100mL of total and fecal coliform. In both cases the removal of turbidity and coliform bacteria was independent of feed water quality (Table 26a and b).

In contrast the removal efficiency of the settling process was dependent on feed concentration (Table 27a and Table 27b). Moreover, the backwash from both the first and second stage process was not inherently settleable. Backwash from the first stage system had a turbidity of 5.23 to 6.0 ntu (Table 27) whereas the turbidity of the settled backwash ranged from 1.61 to 3.50. Maximum reduction in turbidity of the first stage backwash ranged from 54 percent (6.0 NTU to 2.76 NTU) using 10 mg/L ferric chloride with 0.5 mg/L anionic polymer (Table 27a), to 69 percent (5.23 NTU to 1.61 NTU) using 10 mg/L polyaluminum chloride with 0.5 mg/L anionic polymer (Table 27b). In addition to the poor settling properties, the first stage backwash was not inherently filterable. The feed fouling index (a measure of the filterability of a source water) increased from 0.003 to 3.1×10^6 as the suspended solids concentration increased from 0.01 to 25 mg/L. The cleaning interval for the first and second stage MF system decreased from approximately 25 days to 2 days as the flux increased from 13.4 to 26.8 gfd.

Table 26. Evaluation of Microfilter Backwash Residuals

Table 26a - Generation of Microfilter Backwash			
First Stage Filtration Process Stream			
Parameter	Secondary Effluent	MF Filtrate	MF Backwash
TSS (mgL-1)	5	<1	25
TOC (mgL-1)	10.6	9.4	22.4
SO ₂ (mgL-1)	22.5	21.4	22.5
Turbidity (NTU)	2 – 5	0.1	15
Coliform (cfu/100 ml)	170000	<1	4.9 x 10 ⁶
Fecal coli (cfu/100 ml)	11000	<1	-
Indig. Phage (PFU/mL)	38	<1	1.2 x 10 ²

Table 26b. Treatment of Microfilter Backwash Residuals			
Second Stage Filtration Process Stream			
Parameter	MF Backwash 2nd stage feed	2nd stage MF Filtrate	2nd Stage MF Backwash
TSS (mgL-1)	25	<1	933
TOC (mgL-1)	22.4	9.84	139
Turbidity (NTU)	15	0.1	150
coliform (cfu/100 ml)	4.9 x 10 ⁶	<1	4.2 x 10 ⁷
Indig. Phage (PFU/mL)	1.2 x 10 ²	<1	1 x 10 ³

Table 27. Backwash Settling Studies

Table 27a - Settling Studies using Ferric Chloride and Polymer						
Ferric Chloride (mg/L)	Polymer (mg/L)	Time Floc Observed	Floc Size*	Initial Turbidity (NTU)	Final Turbidity (NTU)	Source Water Used
5	0.25	25 sec	B	6.0	2.80	1st stage BW
10	0.5	17 sec	B – C	6.0	2.76	1st stage BW
15	1	15 sec	C – D	6.0	3.50	1st stage BW
5	0.25	60 sec	A	21.8	10.83	2nd stage BW
10	0.5	40 sec	A – B	21.8	9.60	2nd stage BW
15	1	20 sec	B – C	21.8	9.18	2nd stage BW

Table 27b - Settling Studies using Alum and Polymer						
Alum (mg/L)	Polymer (mg/L)	Time Floc Observed	Floc Size*	Initial Turbidity (NTU)	Final Turbidity (NTU)	Source Water Used
5	0.25	10 sec	A	5.23	3.11	1st stage BW
10	0.5	8 sec	A – B	5.23	1.61	1st stage BW
15	1	4 sec	B – C	5.23	3.96	1st stage BW
5	0.25	12 sec	B – C	51.3	14.44	2nd stage BW
10	0.5	10 sec	B	51.3	17.90	2nd stage BW
15	1	5 sec	A	51.3	17.56	2nd stage BW

Notes:

Rapid Mix: 100 rpm for 1 minute

Slow Mix: 70 rpm for 7.5 minutes, 40 rpm for 7.5 minutes, and 20 rpm for 5 minutes

Settling Time: 30 minutes

* Floc Size: A = 0.3 – 0.5 mm, B = 0.5 – 0.75 mm, C = 0.75 – 1.0 mm

2.5.2 Microporous Membrane Use for Management of Peak Flow Conditions

Operation of low pressure microporous membranes at elevated flow rates, flux, was assessed for clean membranes to establish a fouling profile at design and peak flows. The membranes were also operated with a step transition from design to peak flux to simulate membrane performance during a storm event.

A pressure driven modular system configured with a 0.2 μm microporous membrane processes operated continuously, at 100 percent utility, on clarified secondary effluent, at a design flux between 19.7 and peak operation flux of 38 gfd. Under these conditions the interval between chemical cleaning ranged from 1 to 4 weeks. It was possible, however, to increase the design flux by 50 percent and 100 percent for periods of 24 to 48 hours without irreversibly fouling the membranes or impacting filtrate water quality.

In general, increasing flux increased the initial transmembrane pressure and decreased the interval between fouling. The increase in transmembrane pressure however, did not require an increase in the pressure on the upstream side of the membrane, only an increase in flow rate. It was possible to operate at elevated fluxes without exceeding the normal operating conditions of the feed pump. Flux versus time profiles for the transition from design flux to peak operation flux are presented for pressure driven modular systems configured with 0.2 μm and 0.1 μm membranes and suction driven immersed membrane configured with 0.08 μm membrane.

The filtrate water quality at peak operating flux was indistinguishable from the filtrate quality at normal design flux for two of the three membranes tested in the peak shaving experiment, (Table 28a, b and c). The exception was the filtrate from the suction driven 0.08 μm membrane, which had a total coliform count of 44 cfu/100 ml (Table 28a). The feed and filtrate turbidity profiles for the design and peak operation were very similar to the filtrate profiles during the transition from design to peak operation.

Transient flow conditions produced a temporary increase in flux from 19.7 to 29.0 gfd (50 percent).

Table 28. Water Quality at Design and Peak Flux

Table 28a –Suction driven 0.08 μm membrane with continuous aeration				
Constituent	15 gfd (Design Flux)		30 gfd (Max Peak Flux)	
	Influent	Filtrate	Influent	Filtrate
Total coliform (cfu/100 ml)	2x10 ⁶	4	2x10 ⁵	44
Fecal coliform (cfu/100 ml)	2.8x10 ⁵	<1	2x10 ⁴	<1
Suspended Solids (mg/L)	14	<1	5.8	<1
Total Organic Carbon (mg/L)	10.1	7.93	9.83	7.41

Table 28b Discharge pressure driven 0.2 μm membrane				
Constituent	19 gfd (Design Flux)		29 gfd (Elevated Flux)	
	Influent	Filtrate	Influent	Filtrate
Total coliform (cfu/100 ml)		<1	1.9 x 10 ⁶	<1
Fecal coliform (cfu/100 ml)		<1	7.3 x 10 ⁵	<1
Suspended Solids (mg/L)		<1	5.6	<1
Total Organic Carbon (mg/L)		8.67	12.4	10.2

Table 28C Discharge pressure driven 0.1 μm membrane								
Constituent	33.7 gfd		44.3 gfd		55 gfd		65.6 gfd	
	Influent	Filtrate	Influent	Filtrate	Influent	Filtrate	Influent	Filtrate
Tot.coli (cfu/100 ml)			4 x 10 ⁵	<1	4 x 10 ⁵	<1	8 x 10 ⁵	<1
Fec.coli (cfu/100 ml)			1 x 10 ⁵	<1	4 x 10 ⁴	<1	1 x 10 ⁵	<1
Sus. Solids (mg/L)			-	-	-	-	-	-
Tot.Org.Carbon (mg/L)			-	-	12.2	9.9	10.8	8.8

1. Water quality is based on single 1 liter grab sample, unless otherwise noted.

2.5.3 Reclamation of Unclarified Secondary and Primary Effluent

An evaluation of an immersed membrane system was used in order to investigate the replacement of secondary clarification in wastewater. A reverse osmosis test unit was set up downstream of the suction driven microfilter operating on effluent from an air activated sludge process.

2.5.3.1 Microporous Membrane Performance

The immersed membranes were introduced in a 1,000 gallon tank with the outer surface of the fiber in contact with the feed water. The contents of the tank were continually aerated at 75 cfm/module and operated at 90 percent recovery. The lumen side of the fibers was connected at the top and bottom to a permeate collection manifold that was plumbed to the suction side of a centrifugal pump. The suction side vacuum was manually adjusted in order to achieve desired fluxes. Cassettes were added and removed to control flux rates. When a higher flux was desired, cassettes would be removed and permeate flow would be adjusted to achieve the target flux. Permeate was collected on the lumen side of the fibers. The membranes were backwashed with a reverse flow from the lumen to the feed side of the membrane with microfiltrate. The microporous unit operated on a 10 minute backwash interval for a duration of 30 seconds. When the immersed membrane system was fed by unclarified secondary effluent, cleaning occurred once per week with 300 ppm sodium hypochlorite backpulsed through the membranes followed by a backpulse of clean water. When primary effluent was feeding the reactor, the cleaning occurred more often, approximately two to three times a week. Cleanings lasted approximately 2 hours.

During the operation of the immersed membrane system on unclarified secondary, the unit ran under three different fluxes (Table 29). The MF was operated on unclarified secondary from April 20, 1998 to September 24, 1998. On June 11, 1998, after operating 375 hours, the ADC membranes were replaced with CFA membranes which are characterized with twice the porosity. Initially, the system operated on three cassettes at 39 gpm filtrate flow with a flux of 12.5 gfd. The unit operated at a vacuum pressure between 4 and 7 inches of mercury. The first set of experiments operated under an airflow of 75 cfm/module. The next series of runs required the removal of one cassette increasing the flux. The airflow was lowered by one third to compensate for fewer cassettes being on line. Simultaneously, the permeate flow remained at its current rate of 39 gpm. With two cassettes on line and a permeate flow of 39 gpm the flux was raised to 18.7 gfd. This flux is similar to the rate the unit was run at OCWD on clarified secondary effluent. The unit was run at this rate for approximately 2 weeks. Based on the vacuum pressure increase it was determined if the weekly chlorine cleaning could be suspended during these tests. The unit operated at a vacuum pressure between 6.4 and 13.4 inches of mercury. To achieve the highest flux, the MF unit operated on one cassette at 27 gpm filtrate flow for a flux of 25 gfd. The operating vacuum pressure was between 14.1 and 19.1 inches of mercury. After elevating the flux of the immersed microporous membranes, the next test involved running the unit without the weekly chlorine cleaning and only cleaning when a desired vacuum pressure was achieved. Backwash loading rate was reduced as well. The unit was run with three cassettes on line and an effluent flow of 39 gpm. The MF provided adequate pretreatment for the RO successfully removing solids from the feed water.

Filtrate contained less than 1 mg/L of suspended solids and on average less than 5 cfu/100 mL of total coliform. The coliform count slightly increased when the backwash loading was increased. Vacuum pressure increased steadily as the tests were performed (Table 29).

The immersed membrane system operated on primary effluent from October 28, 1998 till the present. The membrane was replaced on January 21, 1999 with an OCP membrane. The flux was adjusted but another variable was introduced: the amount of mixed liquor suspended solids was adjusted by increasing and decreasing the waste rate.

The vacuum pressures for the MF were reduced when the CFA membranes were replaced with the OCP membranes. The OCP membranes treated the primary at a flux of 12 gfd with an operating vacuum pressure between 3 and 3.5 Inches of mercury. The new membranes operated at a lower flux due to their design and increased surface area. The MF provided adequate removal of solids (0.5 mg/L suspended solids).

2.5.3.2 Reverse Osmosis Membrane Performance

The RO unit operated initially with a feed pressure of 114 psi. At the conclusion of the run, the RO was operating at pressures slightly above 300 psi. Throughout this run, no cleaning was performed. The RO operated continuously while the MF was adjusted for flux studies. The RO showed evidence of fouling as time passed and as flux increased within the MF. Thin film composite reverse osmosis membranes (TFCHR) operated downstream of the ADC and CFA membranes at a flux of 10.4 gfd for approximately 2,500 h. Following this run time of 2,500 hours, the RO was chemically cleaned. The use of the integral 0.08 μ m membranes removed the coliforms, suspended solids and other colloidal particles that contribute to fouling of the RO surface.

The RO was cleaned once initially when the MF was fed primary effluent. It maintained a relatively constant pressure of approximately 150 psi throughout the run. In the last 1,000 hours of operation, a steep increase in RO feed pressure reaching close to 300 psi.

The performance of microporous and semi-permeable membranes on unclarified secondary effluent and primary effluent was inferior to the performance on clarified secondary effluent. In general, microporous membrane permeability decreased as the concentration of suspended solids and biochemical oxygen demand in the feed water increased. Feed pressure for the reverse osmosis feed pressures also increased even though most of the precursors for RO fouling, such as suspended solids and microorganisms, were removed by the microporous membrane. It is possible to operate microporous and semipermeable membranes in wastewater treatment plants, however, the merits of this approach should be based on overall reduction in cost (capital and O&M) for wastewater treatment process rather than the cost of wastewater reclamation.

Table 29. Zenon WQ Effluent Fed by UCS

Flux	Initial Vacuum Pressure	Final Vacuum Pressure	Average Water Temp	No of Runs	Start Date	End Date
	inches of Hg	inches of Hg	F			
12.51	2.7	5.2	78.92	8	04/20/98	07/20/98
18.71	6.4	10.2	83.6	4	07/21/98	08/26/98
251	13.1	13.3	84	1	08/26/98	09/04/98
12.52	7.9	8.7		2	09/08/98	09/24/98

Notes:

1 week clean; 75 cfm/module; 10 minute run with a 30 second backwash

Reduced backwash loading

2.5.4 Optimum Features of Low Pressure, Microporous Membranes

Several trends emerged from this study that identified design features of low pressure microporous membranes when used as pretreatment for reverse osmosis. These include:

2.5.4.1 Effect of Compromised Pretreatment Membrane and Restoration of Membrane Integrity

An uncompromised microporous membrane will remove some of the species, such as bacteria and suspended solids, from secondary effluent that have the potential to foul the RO membrane. As a result, feed pressures for the polyamide thin film composite RO membrane will generally remain below 140 psi for the first 750 hours of operation. However, if the membrane is compromised, the corrected feed pressure flux for a polyamide thin film composite RO membrane increase from 110 to 190 psi. The loss of fiber integrity resulted in an increase in the SDI and the concentration of total coliforms in the RO feed water. The silt density index in the feed measured at approximately 100 hour intervals increased from 1.2 to 5; of the six samples collected, 4 were greater than 3. Presence of a 3 ppm combined chlorine residual in the pretreatment membrane filtrate masked the presence of total coliforms in most of the samples, however, by the end of the experiment the filtrate contained as much as 600 total coliform cfu/100 ml. Consequently, the optimum pretreatment membrane will be configured to minimize the incidence of fiber failure.

Most commercial MF and UF systems have a method for detecting broken fibers or o-rings that can compromise system integrity. One advantage of the hollow fiber configuration is that faulty fibers can be readily identified and isolated, and the repaired element can be reused. For example, the M10 modules evaluated in this study were approximately 5 years old. In May of 1997 all six elements were examined for broken fibers and faulty o-rings. The elements were repaired and put back into service. The fecal and total coliform data from this period (Table 33) can be expressed in terms of the probability, expressed as a percent, that the filtrate sample would contain less than a given number of colony forming units per (cfu) 100 ml (Table 30). Prior to repair of the elements, the probability that the level of total coliforms in the filtrate sample was less than 1 cfu/100 ml was 14 percent. The inference is that 96 percent of the samples tested positive for total coliforms. It is possible that some of these counts could be due to regrowth on sample taps, however, the presence of fecal coliforms in the filtrate cannot be

attributed to regrowth. This would confirm that the integrity of the system was compromised. After the elements were repaired the probability that the filtrate contained less than one fecal and total coliform cfu/100 ml was 100 percent and 95 percent respectively (Table 30, Membrane A). It is important to note that the presence of the 3 ppm combined chlorine residual would definitely mask some of the coliforms in the filtrate; it follows that it was possible that filtrate quality could be attributed the efficacy of the combined chlorine rather than physical removal by the membrane. Nevertheless, it was possible to improve the microbial quality of the filtrate by identifying and isolating compromised fibers in a membrane element.

2.5.4.2 Effects of Membrane Symmetry and Module Design on Filtrate Quality

The probability that a given MF or UF filtrate sample contained total and fecal coliforms could be influenced by a variety of factors. These include, but are not limited to, the membrane material, the symmetry of the hollow fiber membrane; the type and concentration of chemicals used in the backwash and cleaning solutions; the backwash and cleaning frequency; and the method of potting the hollow fibers. Some data on the influence of these factors on microbial breakthrough has been collected during this study (Table 30). It is important to note, however, that it was not possible to run a controlled study. To this end it is not possible to compare the probability of microbial breakthrough as a function various membrane and element properties. Nevertheless, some combinations of membrane symmetry and potting techniques could decrease the probability of microbial breakthrough during the filtration of secondary effluent. It might be possible that under some conditions a symmetric membrane is more robust than an asymmetric membrane in a statically potted element with an elastomer overlay (Table 30. Membrane B versus membrane C). Similarly an asymmetric membrane might be more robust when an elastomer overlay is used in a statically potted membrane (Table 30. Membrane B versus membrane D). Although this data should be interpreted cautiously, an argument could be made that the optimum membrane/module combination would be symmetric membranes potted under dynamic conditions or potted under static conditions with an elastomer overlay. Filtrate samples collected from systems configured with this membrane module/combination statistically contained less microbial indicators than filtrate collected from other membrane/module combinations. However, because of the lack of control elements. A more rigorous analysis of the effect of membrane symmetry and element fabrication techniques on system integrity would require customized modules containing the membrane prepared using different potting techniques and operated under identical conditions.

2.5.4.3 Effect of Pretreatment Membrane Pores Size on Reverse Osmosis Pressures

An uncompromised UF or MF membrane will reduce the concentration of the fouling precursors to various degrees depending on the membrane pore size. The best performance from an RO system was achieved downstream of a 0.08 μ m membrane. The corrected feed pressure for RO membrane after 500 hours of operation decreased from 160 to 130 psi, as the pore size of the pretreatment membrane decreased from 0.2 to 0.08 μ m. This pressure differential persisted until the experiments with the 0.08 μ m were terminated after 3,800 hours of operation. The energy efficiency associated with a decrease of 30 psi is approximately 0.34 kWh per 1,000 gallons (110 kWh/acre foot).

It is difficult, however, to generalize about the mechanism of RO fouling and the effect of microporous membrane pore on RO operating pressures. This is evident from the data for two sets of identical polyamide thin film composite RO membranes operating at an instantaneous flux of 10.4 gfd downstream of a 0.1 μm and 0.2 μm MF membrane. After 2,000 hours of operation the RO feed pressures downstream of the 0.1 μm and 0.2 μm membrane was 145 and 175 psi respectively. However, between 2,000 and 3,700 hours of operation the RO feed pressure pressures converged at 200 psi; eliminating the 30 psi pressure differential. In retrospect it would have been beneficial to clean the RO membrane downstream of the 0.1 μm membrane after 2,500 hours to maintain a higher normalized flux. In the short term <2,000 hours, the temporal increase in RO feed pressure was greater downstream of 0.2 μm membranes compared with 0.1 μm and 0.08 μm). This would argue for the use of tighter pretreatment membranes to reduced fouling on the RO membrane. However, other forms of fouling that cannot be controlled by the removal of microorganisms, particles and small colloids must operate for RO systems. This fouling could be attributed to the adsorption of macromolecules or growth of microorganisms on the membrane surface and should be considered when developing strategies for the operation, particularly the cleaning set point, of the RO during the reclamation of secondary effluent.

2.5.4.4 Effect of Membrane Porosity on Flux and Permeability

The instantaneous flux and average initial permeability of the various microporous membranes (pore size $>0.08 \mu\text{m}$) increased with increasing membrane surface porosity (Table 31). The average permeability of microporous membranes operated in the suction mode (immersed system) increased from 5 gfd/psi to 8 gfd/psi as the membrane porosity increased from approximately 25 percent to 50 percent. In contrast, the average permeability of membranes contained in pressure vessels increased from 5.5 gfd/psi to 6.2 gfd/psi as the porosity increased from approximately 30 to 40 percent to 50 to 60 percent. In theory, a high flux and high permeability microporous membrane has the capacity to treat more effluent at lower pressures for the same membrane area. In practice, the benefits are moot, as membrane costs vary between manufactures; membrane area does not necessarily scale linearly with equipment footprint requirements and microporous membrane energy requirements are typically 1/10th to 1/8th of the RO energy requirements.

2.5.4.5 Effect of Operation in Suction Mode with Continuous Aeration

Although UF and MF remove the particulate fouling precursors, it is possible for other forms of fouling, such as inorganic scaling, to occur on the RO membrane. The precipitation of inorganic salts at recoveries above 75 percent can be effectively prevented through the addition of 3.0 mg/L of threshold inhibitor to the UF or MF treated secondary provided that the pH is held below 7.4. However, an inorganic scale developed on membrane elements in the final pass of an RO system operating downstream of an immersed MF membrane. The shell side of the MF fibers were continuously scoured with air at 3.3×10^{-3} scfm/ft². The normalized flux for the polyamide thin film composite membrane decreased from 36 gfd to 22 gfd after 390 hours of operation at which point the RO membranes were rinsed with product water. The flux initially recovered to 28 gfd then declined to 17 gfd after another 500 hours of operation. This fouling was attributed to the precipitation of carbonate salts. The continuous aeration on the shell side of the MF fiber raised the pH of the clarified secondary effluent from 7.3 to 8.0; presumably by

stripping dissolved gasses, such as carbon dioxide, which shifted the carbonate/bicarbonate equilibrium. At the higher pH, the precipitation of carbonate salts could not be prevented by the threshold inhibitor. It was possible to reverse the fouling by rinsing the membrane and lowering the pH of the MF filtrate to pH 7.0 through the addition of 93 percent sulfuric acid.

Table 30. Probability (as percent) of Presence of Coliforms in Microporous Membrane Filtrate

Probability (as percent) of Presence of Coliforms in Microporous Membrane Filtrate									
		<1		<2.2		<10		<100	
	n	Fecal	Total	Fecal	Total	Fecal	Total	Fecal	Total
Membrane A: Symmetric structure + dynamically potted element									
Pre-maintenance	15	64%	14%	100%	31%	100%	66%	100%	100%
Post-maintenance	22	100%	95%	100%	100%	100%	100%	100%	100%
Membrane B: Asymmetric structure + statically potted element with elastomer overlay									
	16	64%	37%	65%	40%	87%	46%	100%	70%
Membrane C: Symmetric structure + statically potted element with elastomer overlay									
	43	100%	100%	100%	100%	100%	100%	100%	100%
Membrane D: Asymmetric structure + statically potted element									
	20	26%	5%	47%	11%	54%	18%	63%	35%

(n = Sample No.)

Table 31. Microporous Membrane Flux (gfd) and Permeability (gfd/psi) as a Function of Porosity

Microporous Membrane Flux (gfd) and Permeability (gfd/psi) as a Function of Porosity						
	Discharge Side ²			Suction side ³		
	Flux	Permeability ⁴		Flux	Permeability ⁴	
Porosity (%) ¹		Single ⁵	Double ⁶		In situ ⁵	Empty ⁶
25%	- NDC	- NDC	- NDC	14.6	2.9	5.0
30 – 40%	22	3.1	5.5	- NDC	- NDC	- NDC
50%	- NDC	- NDC	- NDC	38.8	9.0	10.0
50%	- NDC	- NDC	- NDC	19.4	- NDC	7.9
50%	- NDC	- NDC	- NDC	22.5	- NDC	6.12
50 – 60%	32	- NDC	6.2	- NDC	- NDC	- NDC

Notes:

Membrane porosity based on manufacturer's information or estimate

Membranes located on discharge side of pump

Membranes located on suction side of pump

Permeability after chemical cleaning and expressed as gfd/psi

Single clean using 1 percent NaOH only

Double clean using 2 percent citric acid followed by 1 percent NaOH

Membranes cleaned in situ (tank full of secondary effluent)

Membranes cleaned in empty tank (tank drained prior to clean)

NDC: No data collected for this membrane under these conditions

2.5.5 Treatment and Disposal of Residuals

The residual waste stream generated by the microporous membranes on clarified secondary effluents that are not inherently settleable. This is to be expected given that the suspended solids in the clarified secondary effluent represent the fraction of the secondary mixed liquor that would not settle after 2 hours detention time in the secondary clarifier. The settling properties of the backwash were improved by the use of the trivalent metal coagulants and anionic polymer; however, the turbidity of the settled backwash was dependent on the influent solids concentration. Moreover, the settled backwash would require additional processing; with microfiltration, the waste stream could be reused.

In addition to the poor settling properties, the Feed Fouling Index (FFI) for residual waste stream indicates that the backwash solids were not inherently filterable. As a rule of thumb, membrane processes are not suitable for feed streams with an FFI of 20 or greater. The FFI for typical backwash was greater than 10⁶. Nevertheless, it was possible to treat the backwash with microfilter without fouling the membrane. At a given flux over the range 12.5 to 28 gfd the interval between chemical cleaning for both the first stage (clarified secondary effluent), and second stage (first stage backwash), was found to be independent of the concentration of suspended solids in the feed water. The interval between chemical cleaning for both stages decreased as the system flux increased. Moreover, filtrate water quality, expressed as turbidity, total and fecal coliform, and total suspended solids, was indistinguishable for the first and second stage process and independent of the influent concentration.

The best option for the treatment of residuals generated by low pressure microporous membranes is the use of a second stage membrane to concentrate the backwash. The use of a second stage microporous membrane reduced the volume of backwash by 85 percent and produced an effluent that was suitable for further processing with reverse osmosis for indirect re-use, or disinfection for direct no-potable reuse. It may also be possible to reduce the amount of backwash further by using a third stage MF to treat backwash from the second stage MF units. Once the volume of backwash is reduced to a manageable level it can be disposed of into a sewer, routed to the head of the plant, sent to a DAF thickener, or sent to a solids digester.

2.5.6 Replacement of Secondary Clarifier

MF membranes are being fitted to the end of treatment schemes to refine the effluent prior to disposal and are being fitted directly into primary and secondary treatment basin to allow for increased loading rates. Microporous membranes can be incorporated into membrane bioreactors that operate under aerobic conditions that function similarly to activated sludge/clarification/filtration systems in that air is used in the digestion of organics. These membranes can prove to be efficient in removal of solids. Membrane bioreactors, however, can achieve large loading rates, and produce cleaner effluent compared with conventional processes.

The ability to properly pretreat the feed for reverse osmosis treatment is yet to be completely determined. There exists a high fouling rate when the RO is fed by pretreated primary effluent and pretreated unclarified secondary. Further investigation will allow a better understanding of these processes role in the treatment of wastewaters.

2.5.7 Cost Analysis for Reclaimed Water with Membrane Technology

2.5.7.1 Cost Variables for Reclaimed Water in Membrane Plants

For a large reclamation project such as the Groundwater Replenishment System (GWRS), operation and maintenance costs of the membrane and disinfection processes constitute only 26 percent of the unit cost of reclaimed water (Table 32a). The most significant component unit cost was associated with debt service on the project capital, which amounted to 59 percent of the total annual cost (Table 32b). Using single variable perturbation approach, it was possible to measure the sensitivity of the unit cost to a variety of process variables. Not surprisingly, the unit cost of reclaimed water was most sensitive to changes in interest rates (3.43 percent), the cost of the RO equipment (1.67 percent), power costs (1.29 percent), and the cost of the RO pretreatment equipment (1.27 percent). Moreover, it was apparent that the unit cost of water was relatively insensitive to a 10 percent change in most of the membrane process variables. For example, with power costs fixed at \$0.06/kWh, a 10 percent increase in RO operating pressures only increased the unit cost of water by 0.56 percent (Table 32c). Consequently, the reduction in RO pressures through the use of thin film composite membranes and “tighter” pretreatment membranes should be viewed as only a small component of the unit cost of reclaimed water. In fact, the unit cost of water for an indirect reuse project, such as the GWRS project, would be more sensitive to time on-line (0.87 percent) than the effective membrane life, membrane replacement cost and process maintenance cost.

Table 32. Summary of GWR Cost Model and Sensitivity Analysis

32a - Major capital items as a percentage of total project capital cost (approx. \$275 million)				
Advanced Water Treatment Equipment (installed)				35%
Plant Buildings and Structures				7%
Conveyance System				30.4%
Engineering, Administration and Legal				11.1%
Contingencies				16.5%
32b - Annual cost components as a percentage of unit cost				
Amortized Capital				59%
Conveyance (power and maintenance)				5%
AWT Process (chemicals, power, maintenance)				26%
Staff (Operators)				4.4%
Regulatory Monitoring and Reporting				4.6%
32c - Sensitivity Analysis (Sensitivity expressed as percentage change in unit cost)				
Variables	Units	Initial	Perturbed	Sensitivity
Annual Interest rate	%	6	6.6	+3.43%
Equipment Reverse Osmosis	\$/gal/day	0.9	0.99	+1.67%
RO Pre-treatment (MF/UF)	\$/gal/day	0.4	0.44	+1.27%
Power	\$/kWh	0.06	0.066	+1.29%
Down time	days/year	35.6	39.2	+0.87%
Process maintenance	\$/1000 gal	0.15	0.165	+0.58%
RO operating pressure	psi	180	198	+0.56%
RO membrane life	years	5	5.5	-0.51%
UF/MF membrane life	years	5	5.5	-0.48%
RO membrane cost	\$/element	700	770	+0.46%
Monitoring and reporting	\$/year	1.4 x 106	1.54 x 106	+0.46%
MF/UF membrane cost	\$/element	650	715	+0.43%
Chemicals	\$/1000 gal	0.06	0.066	+0.43%
Labor	\$/h	35	38.5	+0.41%
Building Costs	\$/ ft ²	100	110	+0.34%
MF/UF operating pressure	psi	30	33	+0.15%
UV Energy	kW/lamp	4	4.4	+0.12%
Disinfection	\$/gal/day	0.08	0.088	+0.04%

2.5.7.2 Impact on Water and Sewage Rates for a “Typical” Residence

It is difficult to determine the exact economic value of a wastewater reclamation project in which membrane processes were used to reclaim 100,000 acre feet of wastewater per year, because many of its benefits are societal and environmental which go beyond just the economics of the project.

In 1998 the Orange County Water District prepared a “white paper” on the economic benefits of the Groundwater Replenishment System. The white paper was an attempt to identify and quantify some of the more obvious benefits associated with reclamation projects. The objective of the final component of task 7.0 was to adapt the information in the white paper to estimate the impact of a large reclamation project on monthly water and sewer rates in the areas served by the Orange County Water and Sanitation Districts.

2.5.7.3 Cost/Benefit Ratio for OCWD

The estimated capital cost of the GWRS project is \$275 million. Several grants will be applied towards the capital costs and will reduce the total capital cost by \$25 million. OCWD and OCSD have agreed to split the remaining cost, each paying \$125 million. OCWD will be responsible for the operation and maintenance costs.

Costs and benefits for OCWD were taken from the White Paper published by OCWD and dated April 23, 1998, Appendix Three. From this information, an annualized benefit of \$15.1 million was calculated for OCWD. In 1996/97, 331, 406 acre-feet (AF) of groundwater was extracted from the basin. Assuming an annual benefit of \$15.1 million for the entire basin, this would amount to a saving of \$45.56/AF, or \$0.105/CCF, of groundwater. Part of this total savings may be directly passed on to the consumer as a decrease in the Replenishment Assessment, while part of this savings, such as salinity management, would be achieved through reduction in other consumer costs (Table 33).

2.5.7.4 Estimate of Impact on Water Rates

The 1997 Orange County Water Rates Survey was used to determine the impact on rates for the average household in OCWD’s service area. Information on the number of connections to Single Family Residences (SFRs) and the total water sales to SFRs by agency was used to calculate the Average Water Use per SFR. The rate structure was applied to calculate the Average Monthly SFR Water Bill for each agency. Applying the cost savings only to the pumped groundwater used per household (excluding the water imported from Metropolitan Water District (MWD), a yearly cost per household was calculated for each agency. From this, it was possible to determine an average cost per household for the OCWD service area. In general, the amount of the savings is contingent upon the location of SRF in the service area. The percentage savings per SFR ranged from 0.4 to 8.9 percent depending on the percentage of groundwater served by the each city (Table 34).

2.5.7.5 Cost/Benefit Ratio for Sanitation District

OCSD is a joint sponsor of the GWRS and is continuing to evaluate their role in the GWRS. While GWRS offers several benefits to OCSD, not all costs and benefits have been defined. OCSD is in the process of completing an update of their strategic plan. The strategic plan will establish a policy for the level of treatment which is provided for the wastewater before it is discharged to the ocean. The level of treatment impacts the costs and benefits of GWRS to OCSD users, as well as the costs of wastewater treatment in the future. Therefore, the costs and benefits for OCSD users are based on the best information available at this time.

OCSD's share of the capital of the GWRS project is \$125 million. Another potential cost to OCSD would be to treat waste from the GWRS treatment plant. Two significant waste streams will be produced: backwash from the microfiltration process and brine concentrate from the reverse osmosis (RO) treatment units. The backwash from the microfilters will consist primarily of suspended solids. Because these solids will have already been through the treatment process at Plant No. 1 prior to going to the GWRS project, and because the concentration is high (1,667 mg/L TDS), these solids can be added back into the treatment process as the Dissolved Air Flotation (DAF) thickeners. The solids handling facilities at Plant No. 1 have enough extra capacity to absorb the additional loading for Phase 1 of the GWRS project. OCSD would need to pay the additional O&M cost of processing these solids, approximately \$0.5 million per year. The brine from the RO units will be similar to the brine produced by Water Factory 21. This brine is currently discharged directly to the outfall, and this process should be continued, since the brine would have a negative effect on the treatment processes at Plant No. 1.

The primary benefit of the GWRS would be the delay of construction of a new ocean outfall. The discharge capacity of the 120-inch ocean outfall is rated at 480 mgd. OCSD also has a 78-inch outfall that can be used for emergency discharge only. As a final backup, OCSD can discharge to the Santa Ana River. OCSD has exceeded the nominal discharge capacity of the 120-inch outfall several times in past years during peak storm events. Because the actual discharge capacity of the outfall varies based on tide levels and the amount of buildup in the outfall, the actual capacity was not exceeded, and OCSD has avoided discharging to the 78-inch outfall. However, because flows are projected to increase, a new outfall will be needed in the future. In determining when the new outfall would be required, peak storm flows were not considered. Peak storm events occur so rarely that, even if the capacity of the outfall is exceeded, it can be considered an emergency event. Instead, the daily peak flow for the maximum month was used. This was obtained by looking at the projected average daily flow and multiplying by a maximum month peak factor of 1.1 and a daily diurnal peak factor of 1.32. Using this flow, it was determined that a new outfall would be needed by the year 2009. If the GWRS project is built, a portion of flow would not be discharged through the outfall. The GWRS would take an average of 70 mgd, depending on seasonal variations, and a peak flow of 100 mgd. This would allow the outfall to be delayed until about 2016, a delay of 7 years. This savings would be approximately \$63 million.

The GWRS provides benefits to OCSD that are difficult to quantify or are intangible. One of these is a possible delay in the expansion of secondary treatment facilities at Plant 1. OCSD discharges to a 120-inch ocean outfall under an NPDES permit which limits the total annual mass emission of solids to 20,000 metric tons (mtons). Currently, OCSD discharges an effluent

that is 50 percent primary effluent and 50 percent secondary effluent. This blend allows the permit to be met. However, with the projected flow increases over the next several years, this mass emission limit may be reached in 2005, unless the percentage of secondary effluent in the final discharge is increased. By increasing the amount of secondary effluent in the blend, OCSD can avoid reaching the mass limit until 2012, when additional secondary treatment capacity will be needed. However, if the GWRS project is built, the EPA will allow OCSD to reopen their NPDES permit with the possibility of increasing the mass emission limit to 25,000 mtons/yr. If the limit is increased, secondary treatment expansion could be delayed at least 10 years. The cost savings would be approximately \$60 million in capital costs and \$70 million in operation and maintenance costs, a total of \$130 million. However, because the EPA has not guaranteed that the mass emission limit would be raised with construction of the GWRS project, no credit for the secondary treatment expansion delay has been claimed in this analysis (Table 35). The costs and benefits. Using this analysis, the project has a net annualized cost of \$5.1 million (Table 35).

2.5.7.6 Estimate of Impact on Sanitation Rates

OCSD recently completed a rate study and adopted a new rate structure. The estimated number of Equivalent Dwelling Units (EDUs) in the OCSD service area was determined to be 922,946. A SFR is equal to one EDU. To determine the effect of the GWRS project on each SFR, the annual yearly benefit/cost ratio was divided by the total number of EDUs in OCSD's service area. It is estimated that the annual savings for an EDU in the OCSD service area as a result of the implementation of the GWR System would be 7.5 percent (Table 36).

Table 33. GWRS Cost-Benefit Analysis for Orange County Water District

OCWD Costs		
Item	Cost	Total Annualized Cost (million)
Capital Costs	\$125 M	\$9.8
Operation and Maintenance	\$15.0 M/year	\$15.0
Total Cost to OCWD		\$24.8
OCWD Benefits		
Item	Benefit	Total Annualized Cost (million)
Reduced Salinity Management	\$16.3 M/year	\$16.3
Avoid Expansion of WF21	\$49.8 M	\$3.9
Avoid Annual O&M of WF21	\$7.7 M/year	\$7.7
Avoid Pipeline from Diemer Bypass to bring MWD water to Spreading Basins	\$3.8 M	\$0.3
Avoid Purchase of Imported Water for Spreading Basins	\$11.7 M/year	\$11.7
Total Benefit to OCWD		\$39.9

1. Costs are April 1998, ENR LA = 6,679

Table 34. Effect of GWRS on Orange County Water District Rates

Agency	Avg SFR Monthly Water Bill ⁽²⁾	% OCWD Pumped Groundwater	Avg Monthly Savings Per SFR ⁽¹⁾	Avg Yearly Savings Per SFR	% Savings Per SFR
City of Anaheim	\$35.87	75%	\$2.23	\$26.77	6.2%
East Orange CWD Retail	\$42.74	75%	\$2.68	\$32.16	6.3%
City of Fountain Valley	\$28.79	79%	\$1.93	\$23.19	6.7%
City of Fullerton	\$30.26	75%	\$1.63	\$19.54	5.4%
City of Huntington Beach	\$21.21	72%	\$1.06	\$12.74	5.0%
Irvine Ranch WD	\$7.64	45%	\$0.42	\$5.06	5.5%
City of La Palma	\$40.06	75%	\$1.90	\$22.85	4.8%
Los Alisos WD	\$27.41	7%	\$0.12	\$1.45	0.4%
Mesa Consolidated WD	\$27.48	77%	\$1.30	\$15.61	4.7%
City of Orange	\$32.43	79%	\$2.01	\$24.01	6.2%
Orange Park Acres Mut. WC	\$103.47	75%	\$4.46	\$53.50	4.3%
City of Santa Ana	\$31.23	75%	\$1.90	\$22.79	6.1%
City of Santa Ana	\$59.07	50%	\$1.19	\$14.32	2.0%
Santiago CWD	\$34.18	92%	\$2.53	\$30.39	7.4%
City of Seal Beach	\$59.07	100%	\$5.26	\$63.13	8.9%
Serrano WD	\$35.97	66%	\$1.81	\$21.69	5.0%
Southern California WC	\$31.57	76%	\$1.82	\$21.86	5.8%
City of Tustin	\$28.67	86%	\$1.42	\$17.09	5.0%
City of Westminster	\$30.64	54%	\$1.48	\$17.77	4.8%
Yorba Linda WD	\$27.38		\$1.50	\$18.04	5.5%
AVERAGE for OCWD					

Avg. Monthly Saving per SFR = Avg. Monthly Water used per SFR X percent OCWD Pumped Groundwater X GWRS Cost Saving per CCF of Groundwater

Based on data from the 1997 Orange County Water Rates Survey

Table 35. GWRS Cost-Benefit Analysis for OCWD – Strategy 1 ⁽¹⁾

OCSD Costs		
Item	Cost	Total Annualized Cost
Capital Costs		\$9.8 million
Operation and Maintenance	\$125.0 million	\$0.0 million
Additional Solids O&M	\$0.0 million/year	\$0.5 million
Total Cost to OCSD	\$0.5 million	\$10.3 million
OCSD Benefits		
Item	Benefit	Total Annualized Savings
Delay in Outfall Construction	\$63.0 million	\$4.9 million
Avoided Effluent Pumping ⁽²⁾	\$0.3million	\$0.3 million
Total Benefit to OCSD		\$5.2 million

1. Costs are April 1998, ENR Ia = 6,679
2. Based on 70 mgd of flow @ TDH = 45 ft., power = \$0.06/kWh

Table 36. Effect of GWRS Project on OCSD Rates

Analysis	
No. of EDU's	922,946
Annual Benefit (Cost)	(\$5.1 million)
Average Yearly Sewer Bill	\$73.32 (1)
Average Yearly Savings (Cost) per SFR	(\$5.53)
percent Savings (Cost) per SFR	(7.5%)

1. Based on 1995/96 dollars

2.6 Electron Beam for Destruction of Trace Organics

The OCWD Water Factory 21 pilot test showed the capability of high energy electron injection as a water treatment alternative. Costs to install and operate the system are competitive, and potentially much lower than conventional and emerging treatment alternatives. A distinct advantage of the process is the application of oxidizing and reducing chemistry, making it possible to treat compounds resistant to other destructive processes. In addition, since treatment is direct, there is no creation of waste streams, nor the need to employ mass transfer steps (like stripping) prior to contaminant destruction.

The testing further showed the ability to meet, or approach MCLs on a variety of priority contaminants applied simultaneously for treatment. Since the concentrations of these priority contaminants was much higher than typically expected in wellhead treatment applications, the results suggest wellhead treatment is a viable application of the technology. Further work to demonstrate this capability is strongly suggested in the immediate future.

Treatment results on NDMA showed the ability to treat this specific contaminant effectively at low cost. The results additionally suggest further improvements in treatment efficiency can be derived by simple adjustments in water chemistry. These conditions are candidates for future applied demonstration projects.

Native water chemistry has been shown to impact treatment performance, and adjusting native water chemistry is expected to further improve results both in treatment efficiency and operating costs. A further series of demonstrations, focused on adjusting native water chemistry would be highly recommended.

MTBE only tests showed the ability to meet MCLs at lower energy requirements than competing oxidation technologies, and the ability to simultaneously treat reaction intermediates. A preliminary comparison of results indicates the ability to meet treatment guidelines for MTBE and reaction intermediates TBA/TBF at energy rates slightly greater than required for MTBE only. Since treatment of reaction intermediates is controlled by reducing chemistry, high energy electron injection has a distinct competitive advantage with its unique oxidizing/reducing chemistry. Further applied demonstrations focused on series treatment (maximizing oxidizing species availability for direct MTBE reduction, followed by reducing chemistry for reaction intermediates may further reduce total energy required for the process.

The OCWD Water Factory 21 tests suggest the better applications for the electron beam process will be in high flow rate, single or multiple constituent treatment scenarios where destruction of the contaminants is a desired endpoint. Future work is recommended to further demonstrate these capabilities, refine economics and improve awareness of this technology alternative.

3.0 Conclusions and Recommendations

Based on results obtained from this project, the following major conclusions were reached for each of the tasks. Detailed discussions on individual task research activities can be found in Section 2.0 and in Appendixes.

3.1 Colorado River Water Salinity Reduction

3.1.1 Conclusions

- While RO using conventional treatment with ozone/biofiltration required cleaning three times sooner than using conventional treatment alone, the RO membranes operated at lower pressures (i.e. lower energy consumption). The higher operating pressure in the conventional treatment study may have been attributed to an incomplete membrane cleaning prior to testing. In order to completely remove all deposits from fouled RO membranes, both acidic and caustic cleaning should be employed.
- Both conventional treatment with and without ozone/biofiltration produced an effluent water quality suitable for use with RO.
- Biologically active filters sloughed off bacteria and bacteriological material that impair the performance of downstream salinity reduction technologies. The use of chloramines to disinfect the water does not adequately protect against biological fouling of RO membranes when using biologically active filters.
- Carbon aerogel CDI technology exhibited moderate salinity removal approximately~50 percent, low capacity 6-8 mg TDS per gram of carbon aerogel, and slow regeneration, greater than >1 hour. These problems may be a result of the CDI design and construction and not inherent to the carbon aerogel material.iii It was suspected that the epoxy mounting of the carbon aerogel material to the titanium support eliminates much of the available surface area for sorption. In addition, a small percentage of salts may become sorbed to the epoxy layer, rather than the carbon aerogel

Farmer, J.C.; J. H. Richardson, & D.V. Fix (1996). Desalination with Carbon Aerogel Electrodes -DRAFT. Report UCRL-ID-125298, Rev. 1. Lawrence Livermore National Laboratory (LLNL), Livermore, Calif. (Dec 4, 1996).

Filteau, G.H. Nitrate Removal from Contaminated Ground through Reverse Osmosis, AWWA Membrane Technology Conference, Baltimore, Maryland, 1993.

B. R. Wilson; P.F. Roessler; E.V. Dellon; M. Abbaszadegan; C.P. Gerba, "Coliphage MS-2 as a UV Water Disinfection Efficacy Test Surrogate for Bacterial and Viral Pathogens," In Proc. AWWA Water Qual Tech. Conf., Toronto, Ontario, Canada (1992).

K. G. Linden and A. A. Mofidi, "Disinfection Efficiency and Dose Measurement for Medium Pressure and Pulsed-UV Disinfection Systems: an Unsolicited Proposal to the American Water Works Association Research Foundation" (University of North Carolina, Charlotte and Metropolitan Water District of Southern California, 1998).

layer. Therefore, results were hindered by an extra resistance to ion diffusion of the aerogel material into the bulk phase.

3.1.2 Recommendations

- Significant energy savings can be realized with the use of ultra-low-pressure RO membranes over the previous generation of low-pressure RO membranes. In addition, through the use of pre-existing conventional treatment plants, municipalities throughout California may be able to forgo building expensive, new pretreatment facilities prior to RO. This will allow for the greater use of membrane technologies to treat previously untapped water supplies and increase water availability and reliability.
- More research is required before these technologies can be used in California. Additional research is needed to optimize the use of chemical coagulants and membrane antiscalants to decrease the cleaning frequency of membrane processes. In addition, larger diameter RO elements are required if membrane technology is to be implemented on a larger scale, greater than 30 million gallons per day (mgd).
- Due to CDI's early state of technical development and poor performance, all future CDI evaluations should be conducted at bench-scale using deionized water supplemented with artificial ionic matrices. This change in scope will enable a more basic understanding of the electrosorption capabilities of carbon aerogel materials. Results from these bench-scale studies will enable water utilities to more adequately gauge CDI's potential as a salinity reduction technology.

3.2 Pulsed UV for Disinfection and Membrane Fouling Control

3.2.1 Conclusions

- Pulsed UV disinfects *Cryptosporidium parvum* in waters with low UV absorbance (≥ 4 -log₁₀ inactivation of oocysts at UV doses <20 mJ/cm²). This corresponds to 99.99 percent *Crypto.* inactivation.
- Pulsed UV disinfects HPC bacteria in waters with low UV absorbance (≥ 3 -log₁₀ inactivation at UV doses >12 mJ/cm²) corresponding to 99.9 percent bacteria inactivation.
- Pulsed UV disinfects MS-2 coliphagem, a virus, in water with low UV absorbance (≥ 6 -log₁₀ inactivation at UV doses >50 mJ/cm²) corresponding to 99.9999 percent virus inactivation.
- Pulsed UV in combination with chloramines adequately disinfected the effluent from biological filters and controlled bacterial regrowth or repair over a duration of 3 days.
- The combination of pulsed UV/chloramines produced low levels of THMs (<3 μg/L) and HAAs (<11 μg/L).

3.2.2 Recommendations

- The above conclusions were derived primarily from bench-scale results. Further research is needed to validate preliminary results and to produce full scale pulsed UV reactors that produce drinking water in a variety of water matrices, water quality, and operating environments.

- If positive results are achieved in full scale pulsed UV reactors, this technology could be used as an additional barrier to pathogenic organisms in drinking water treatment. This technology would be significantly less costly than ozone or membrane *Cryptosporidium* reduction technologies.

3.3 Bromate Formation and Control

3.3.1 Conclusions

- Lowering pH prior to ozonation was effective in reducing bromate formation, even at *Cryptosporidium*-level disinfection conditions, however the technology is expensive because of the large chemical dosages required for pH control. This method also increases the TDS of the water which may create problems for homeowners and industry through corrosion and scaling. Elevated TDS may also limit the use of reclaimed water for irrigation and groundwater recharge.
- Ammonia is a promising bromate control strategy for Colorado River water at ozone doses required for enhanced *Cryptosporidium* disinfection. Its use was less effective in reducing bromate formation in California State Project water.
- The effectiveness of ammonia for controlling bromate is source-water specific and substantially influenced by pH, alkalinity, and bromide concentration. Seasonal variations in these parameters must be considered.
- The operating costs of adding ammonia, at ambient pH, versus lowering the pH to achieve a comparable bromate minimization are approximately 17 percent less for CRW and 20 percent less for SPW.
- Hydrogen peroxide is not an effective bromate control strategy at the ozone doses required for 99 percent inactivation of *Cryptosporidium*.

3.3.2 Recommendations

- Without subsequent investigation, municipalities may consider implementing this strategy for bromate control. However, issues of elevated TDS and increases in operational costs must be taken into consideration.
- Careful consideration must be given to several issues prior to full-scale implementation: (1) the increase in pH; (2) the effect on secondary disinfection and/or breakpoint chlorination; (3) the effect on nitrification (e.g., in downstream biologically active filters); and (4) the effect on disinfection of bacteria leaving biological filters. Additional studies are needed in this area before ammonia may be implemented as a bromate control strategy.
- Results of this project will serve as a paradigm for utilities currently using or implementing ozone and faced with the need to minimize the formation of bromate. The drinking water industry benefits by the improved ability to use ozone to control *Cryptosporidium* or other micropollutants, e.g. taste-and-odor compounds, pesticides, or methyl-tertiary-butyl-ether (MTBE) instead of implementing more expensive technologies to meet multiple water quality objectives.

3.4 Disinfection of Reclaimed Water

3.4.1 Conclusions

- Ultraviolet disinfection is a viable alternative to chloramination for disinfection of reclaimed wastewater for non-potable reuse. It offered superior coliform reduction compared to chloramination for disinfection of tertiary effluent.
- Sodium thiosulfate would not be a good choice for UV transmittance adjustment when testing UV systems with a polychromatic output, since it would not block all the germicidal wavelengths produced. Chemicals such as sodium thiosulfate are typically used for this purpose with low-pressure mercury lamps, which have a monochromatic output. The medium pressure UV lamps, however, are polychromatic with emissions extending across the UV spectrum. Therefore, a substance capable of absorbing UV light evenly across this spectrum was required for reducing the transmittance in the pilot plant influent.
- A 99.99 percent inactivation of indigenous, native to the environment, coliphage was achieved at an UV dose of about 90 mWs/cm². As most of the UV effluent coliphage levels were close to or below the detection limit, it is not possible to establish any correlation between coliphage inactivation and the turbidity of the water entering the UV system.
- Pilot plant achieved 99.99 percent and 99.9 percent virus inactivation with a dose of about 180 and 100 mWs/cm², respectively. The corresponding collimated beam doses were about 135 and 95 for 99.99 percent and 99.9 percent virus inactivation. These results indicate that for the dose ranges expected for reuse applications, the Aquionics calibration coefficient would be in the range of 0.75 to 0.95, compared to the collimated beam unit.
- The Aquionics pilot unit is capable of achieving both a 7 day median total coliform level of 2.2 most probable number (MPN) per 100 mL, and 99.99 percent indigenous, native to the environment, coliphage inactivation at a dose of 100 mWs/cm², as demonstrated in phase I. The pilot unit's performance can be correlated to the collimated beam unit with a calibration coefficient. The dose calculation for the Aquionics unit was developed using their proprietary model.

3.4.2 Recommendations

- The establishment of a standard method for evaluation of UV disinfection systems will allow regulators, design engineers, utilities, and researchers to make comparisons between UV disinfection studies, and ultimately between vastly different UV systems.
- UV disinfection equipment testing performed in accordance with a standard method can be compared equally.
- Re-evaluation of low pressure, high intensity UV system using established UV testing protocol.
- Expansion of UV disinfection work to include parasite testing/inactivation.
- Continuation of pulsed UV evaluation to include seeded studies with phage and parasites.

3.5 Low Pressure Membrane for Reclaimed Water

3.5.1 Conclusions

- Performance of microporous and semi-permeable membranes on unclarified secondary effluent and primary effluent was inferior to that of the secondary effluent. In general, microporous membrane permeability decreased as the concentration of suspended solids and biochemical oxygen demand (BOD) in the feed water increased. Feed pressure to the reverse osmosis (RO) system also increased even though most of the precursors for RO fouling, e.g. suspended solids and microorganisms, were removed by the microporous membrane.
- The best option for treatment of residuals generated by low-pressure microporous membranes is the use of a second stage membrane to concentrate the backwash. This would reduce the volume of backwash by 85 percent and produce an effluent that was suitable for further processing with RO for indirect reuse, or disinfection for direct non-potable reuse.
- Microporous membranes can be incorporated into membrane bioreactors that operate under aerobic conditions that function similarly to activated sludge/clarification/filtration systems in which air is used in the digestion of organics.

3.5.2 Recommendations

Further research is needed to:

- Test oxidant tolerant reverse osmosis membranes and develop more aggressive cleaning regimens to remove organic foulants from RO membranes.
- Investigate the effect membrane geometry and module potting techniques on the incidence of fiber failure in microporous membrane systems.
- Evaluate the performance of a low pressure microporous pretreatment system that incorporates the optimum design features identified in this project.
- Evaluate the feasibility of using a microporous (MF) membrane to separate suspended and soluble BOD and a semipermeable RO membrane to concentrate the soluble biochemical oxygen demand (BOD) prior to stabilization in an anaerobic reactor.
- Quantify the true operating costs of microporous membrane process used as pretreatment for RO based on the scale up of a promising system identified in this project.

3.6 E-Beam Destruction of Trace Organics

3.6.1 Conclusions

- The OCWD Water Factory 21 pilot test showed the capability of high energy electron injection as a water treatment alternative. Costs to install and operate the system are competitive, and potentially much lower than conventional and emerging treatment alternatives. A distinct advantage of the process is the application of oxidizing and reducing chemistry, making it possible to treat compounds resistant to other destructive

processes. In addition, since treatment is direct, there is no creation of waste streams, nor the need to employ mass transfer steps, like stripping, prior to contaminant destruction.

- The testing further showed the ability to meet, or approach Maximum Contaminant Levels (MCLs) on a variety of priority contaminants applied simultaneously for treatment. Since the concentrations of these priority contaminants was much higher than typically expected in wellhead treatment applications, the results suggest wellhead treatment is a viable application of the technology. Further work to demonstrate this capability is strongly suggested in the immediate future.
- Treatment results on NDMA showed the ability to treat this specific contaminant effectively at low cost. The results additionally suggest further improvements in treatment efficiency can be derived by simple adjustments in water chemistry. These conditions are candidates for future applied demonstration projects.
- The OCWD Water Factory 21 tests suggest the better applications for the electron beam process will be in high flow rate, single or multiple constituent treatment scenarios where destruction of the contaminants is a desired endpoint. Future work is recommended to further demonstrate these capabilities, refine economics and improve awareness of this alternative technology.

3.6.2 Recommendation

- The OCWD Water Factory 21 tests suggest the better applications for the electron beam process will be in high flow rate, single or multiple constituent treatment scenarios where destruction of the contaminants is a desired endpoint. Future work is recommended to further demonstrate these capabilities, refine economics and improve awareness of this technology alternative.

4.0 Glossary

ac-ft	acre-foot
ADC	
AFM	atomic force microscopy
Amps	amperes
AOC	assimilable organic carbon
ASTM	American Society for Testing and Materials
AWWA	American Water Works Association
AWWARF	American Water Works Association Research Foundation
BDOC	biodegradable dissolved organic carbon
BOD	biochemical oxygen demand
BOM	biodegradable organic matter
Br	Bromide
°Celsius	degree Celsius
CDI	Capacitative deionization
CEC	California Energy Commission
CF	concentration factor
CFU	Concentration factor unit?
cfu	colony forming units
cfu/mL	colony forming units per milliliter
cm	centimeter
<i>Crypto.</i>	<i>Cryptosporidium</i>
CRW	Colorado River water
CT	disinfectant concentration X contact time
DAF [t]	dissolved air flotation thickener
DBP	disinfection by-product
DOC	dissolved organic carbon
EDS	energy dispersive spectroscopy
EDU	equivalent dwelling units
EPRI	Electric Power Research Institute
ETS	Edison Technology Solutions
°F	degree Fahrenheit
FFI	feed fouling index

ft²	square foot
g	gram
gal	gallon
GAP	Green Acres Project
gfd	gallons per square foot per day
gpm	gallons per minute
GWRS	Ground Water Replenishment System
HAA	haloacetic acids
Hg	mercury
HPC	heterotrophic plate count
hr	hour
hrs	hours
Hz	Hertz (1/sec)
KI-IO₃	potassium iodide-iodate
KW	kilowatt
kWh	kilowatt hour
LLNL	Lawrence Livermore National Laboratory
LTESWTR	Long-Term, Enhanced, Surface Water Treatment Rule
MCL	maximum contaminant level
MCLG	maximum contaminant level goal
mg/L	milligram per liter
mJ/cm²	millijoule per square centimeter
mL	milliliter
mm	millimeter
µg/L	microgram per liter
Yg/L	microgram per liter?
µs	microseconds
µs/cm	microsiemen per centimeter
Ys/cm	microsiemen per centimeter
MRL	minimum reporting level
MS-2	male-specific coliphage (virus)
MTBE	methyl-tertiary-butyl
ether	

MWDSC	Metropolitan Water District of Southern California
mWs/cm ²	units of watt seconds per squared centimeter
MWCO	Define in Section 2.5.1.2.
NA	not available
ND	not detectable
NDMA	pg. 9 (specific contaminant)
nm	nanometer
NOM	naturally occurring matter
ntu / NTU	nephelometric turbidity unit
OCWD	Orange County Water District
O&M	Operation and Management
ORP	oxidative reduction potential
<i>parvum</i>	<i>Cryptosporidium parvum</i>
pH	hydrogen ion concentration
ppm	part per million
psi	pound per square inch
PCR	polymerase chain reaction
PFBHA	O (2,3,4,5,6-pentafluorobenzyl) hydroxylamine
PIER	Public Interest Energy Research
PQL	practical quantitation level
R2A	heterotrophic bacteria growth media
RO	reverse osmosis
SCE	Southern California Edison
SDS	simulated distribution system
SEM	scanning electron microscopy
SDI	silt density index
SIO ₂ define	
Table 20: writes it as SiO ₂ ; or is it SiO ₂ ? ; or as in Table 25, which lists it as SIO ₂ ?	
SPC	salt permeation coefficient
SPW	State Water Project

SUVA Define in Table. 15	
SWTR	Surface Water Treatment Rule
TDS or tds?	total dissolved solids
THM	trihalomethanes
TOC	total organic carbon
T&O	taste and odor
USEPA	U.S. Environmental Protection Agency
UV	ultraviolet
UV-VIS	UV visible
UV₂₅₄	ultraviolet light absorbance at 254 nanometers
V	volt
Ws/cm²	units of watt seconds per squared centimeter
WPC	water permeation coefficient

