# RECLAMATION
*Managing Water in the West*

Desalination and Water Purification Research and Development
Report No. 117

# Microfiltration/Reverse Osmosis Ocean Water Desalination Pilot Plant Project

U.S. Department of the Interior
Bureau of Reclamation
Technical Service Center
Denver, Colorado

West Basin
Municipal Water District
Carson, California

September 2007

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**14. ABSTRACT**

West Basin Municipal Water District conducted an ocean water desalination pilot study at the El Segundo Power Facility in El Segundo, CA. The study determined operating conditions for both ultrafiltration and microfiltration on powerplant influent and effluent (post-condenser) water sources. Numerous reverse-osmosis membranes were tested and were found to achieve the water quality goals.

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# Microfiltration/Reverse Osmosis Ocean Water Desalination Pilot Plant Project

## Final Report

**Prepared for the Bureau of Reclamation Under Agreement No. 05-FC-81-0838**

*by*

**Paul Shoenberger and Phil Lauri**
**West Basin Municipal Water District**

**Steve Alt and Mark Donovan**
**Separation Processes, Inc.**

**U.S. Department of the Interior**
**Bureau of Reclamation**
**Technical Service Center**
**Denver, Colorado**

**West Basin**
**Municipal Water District**
**Carson, California**

**September 2007**

# MISSION STATEMENTS

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The mission of the Bureau of Reclamation is to manage, develop, and protect water and related resources in an environmentally and economically sound manner in the interest of the American public.

West Basin Municipal Water District mission:  To provide a safe and reliable supply of high quality water to the communities we serve.

## Disclaimer



## Acknowledgments

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# ACRONYMS AND ABBREVIATIONS

| | |
|---|---|
| CIP | clean in place |
| CMF-S | continuous microfiltration submerged |
| gpm | gallons per minute |
| DL | detection limit |
| GFD | gallons per square foot of membrane area per day |
| $H_2SO_4$ | sulfuric acid |
| HYD | Hydranautics (membrane manufacturer) |
| MC | maintenance clean |
| MF | microfiltration |
| mg/L | milligrams per liter |
| ng/L | nanograms per liter |
| NaOCl | sodium hypochlorite |
| $NH_3$ | ammonia |
| $NH_4OH$ | ammonium hydroxide |
| NTU | nephelometric turbidity units |
| PDT | pressure decay test |
| psi | pounds per square inch |
| PVDF | polyvinylidene fluoride |
| RO | reverse osmosis |
| SBS | sodium bisulfite |
| SCFM | standard cubic feet per minute |
| SDI | silt density index |
| TDS | total dissolved solids |
| TMP | trans-membrane pressure |
| TOC | total organic carbon |
| UF | ultrafiltration |
| UV 254 | amount of ultraviolet light, at a wavelength of 254 nanometers, absorbed by organic matter in a water sample (reported here as absorbance per centimeter) |
| WBMWD | West Basin Municipal Water District |
| µg/L | micrograms per liter |
| µm | micrometers or microns |
| µS | microsiemens |

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# EXECUTIVE SUMMARY

West Basin Municipal Water District (WBMWD) conducted an ocean water desalination pilot study at the El Segundo Power Facility in El Segundo, CA. The study was very successful, meeting its objectives and providing a body of data not previously available. The study investigated the use of microfiltration (MF) and ultrafiltration (UF) membrane processes as pretreatment to reverse osmosis (RO). The objectives of the study were to evaluate and optimize the performance of MF, UF, and RO operating parameters on powerplant intake water as well as on warmer power plant post-condenser effluent water, and to expose the project to the variability of the ocean itself. The research indicates that these membranes will work effectively at the full scale level with the information and experience gained from this pilot project. The long timeframe of the testing provides confidence in the results.

The study began in 2002 and was separated into two phases of testing: Phase A and Phase B. Phase A testing occurred from June 2002 to June 2004 and Phase B from July 2004 through September 2007.

Phase A was an evaluation of MF and RO performance, establishing operating parameters such as MF backwash frequency and membrane flux rates on powerplant intake water. Phase A testing showed that the Siemens CMF-S MF system provides excellent quality filtrate to be used as a feed to RO, and that the use of chlorine in the MF backwash was beneficial to keeping fouling of the MF membrane under control. Permeate water produced by the RO membranes was consistently of high quality, with TDS generally less than 300 mg/L and boron concentrations between 0.6 and 1 mg/L.

Phase B was separated into three different sub-phases as follows:

- Phase B1 evaluated four “next-generation” or recently developed RO membranes on microfiltered powerplant influent water. These recently developed membranes had the highest boron rejection available.
- Phase B2 evaluated MF and next-generation RO membranes on powerplant effluent and the Zenon UF System on powerplant influent.
- Phase B3 identified two of the four next-generation RO membranes for longer term testing and evaluated all systems on powerplant effluent.

Phase B demonstrated that the optimized Phase A MF operating parameters for influent water were unchanged for the warmer effluent water source. Both the MF and UF produced excellent quality filtrate for use as RO feed water. No differences in RO fouling were observed that could be attributed to differences in filtrate quality between the MF and UF processes.

Operating the RO systems at the elevated temperatures of the effluent stream did result in lower RO feed pressure requirements, but also resulted in higher permeate concentrations of TDS, boron, and other constituents, as expected. The RO systems were also affected by biofouling to a greater extent on the warmer effluent water than on the colder influent water.

The MF, UF, and RO systems operated through several algae bloom events (red tides) during the course of Phase B testing. Periodic testing did not detect the algal toxin domoic acid in any RO permeate samples, despite elevated concentrations in feed water as a result of the red tide events. The ocean water contained domoic acid levels as high as 2 to 3 µg/L during red tide events, yet the RO permeate levels were consistently below the detection limit of 0.002 µg/L. This demonstrated that the RO treatment process is an excellent barrier to this constituent. However, the MF and UF systems did lose some permeability during the more severe algae blooms, which temporarily reduced their filtration capacity.

Data collected on the "next-generation" RO membranes indicated improved performance (lower permeate concentrations of key constituents) over the previous versions tested in Phase A. Each of the newer membranes tested demonstrated the capability of providing permeate water with less than 200 mg/L total dissolved solids (TDS) across the powerplant influent temperature range and water with less than 300 mg/L across the powerplant effluent temperature range. Additionally, differences were noted in salt rejection performance among the various types of new membranes, and these differences provide options to achieve lower chloride or boron concentrations. For example, both the Hydranautics SWC4+ and Dow SWHRLE4040 membranes provided excellent boron rejection, with permeate water levels typically less than 0.7 mg/L. However, SWC4+ produced a permeate water with less than 50 mg/L chloride ion, substantially less than the Dow membrane.

From environmental, financial, operational, and other aspects, the pilot testing provided a wealth of data and information to support and provide confidence in the implementation of full-scale ocean water desalination.

# Background

Ocean water desalination will eventually play a significant role in the water supply equation for Southern California. To date, the use of ocean water desalination in California has been minimal, primarily due to relatively high cost. Recently, with improved performance and costs, microfiltration (MF) and ultrafiltration (UF) have been proposed as alternatives to conventional pretreatment processes for ocean water reverse osmosis (RO). Microfiltration has become a common pretreatment method for RO installations treating municipal wastewater. UF and MF each remove colloidal and suspended particulate matter that would foul RO membranes. A pilot plant program was begun to evaluate the combination of MF and RO, as well as UF and RO, for potential application to ocean water desalination in California for the domestic water supply.

West Basin Municipal Water District's (WBMWD) Ocean Water Desalination Pilot Plant Program tested the capabilities of MF and UF pretreatment in series with a spiral-wound RO system. It developed data to determine the optimum operating conditions and cleaning requirements for MF and UF operating on ocean water, and for the RO process operating on microfiltration filtrate. Phase A of this study consisted of MF followed by RO. In Phase B, a UF system was added in parallel with the MF system, and the results of the RO operation were compared, operating on feed water from the two different pretreatment membrane systems.

The testing occurred at the El Segundo Power Generation Plant (figure 1). Ocean water desalination is energy intensive, and a full-scale ocean water desalination plant collocated with an existing ocean water cooled powerplant has advantages. One potential advantage is that power may be available at relatively low rates "within the fence" of the powerplant. In California, this may result in an energy savings of about 5 cents per kilowatt-hour. In addition, the ocean water desalination plant can also utilize the existing intake and outfall structures that allow ocean water to be brought into the powerplant and returned to the ocean. Furthermore, the salinity of the RO concentrate is reduced by blending with the powerplant discharge water.

Using the existing intake/outfall structure presents two options for the source water to the desalination plant. The plant can either feed from ocean water entering the powerplant or from water that has already been used in the powerplant cooling process and is being returned to the ocean. At the El Segundo Power Generation Plant, there is typically a 14 °F difference between the cool ocean water entering the powerplant and the warmer return water. A membrane desalination plant operating on the warmer return water would have the advantage of decreased energy usage associated with a decrease in water viscosity. On the other hand, the warmer water may promote bacterial growth that may have a higher fouling potential for the membrane treatment processes. Also, the salinity of the treated water would be slightly higher. Phase A of this work included operation on the cooler powerplant influent water. In Phase B, an ultrafiltration membrane process was added and the entire operation was switched to the warmer powerplant effluent water.

**Figure 1.—Pilot test equipment at the El Segundo Powerplant.**

# Conclusions

The following conclusions can be drawn from this 5-year study:

1. The study successfully established the feasibility of utilizing an MF/UF followed by RO process to produce potable quality water. This was demonstrated on Pacific Ocean water taken from either a powerplant intake or the warmer powerplant post-condenser effluent source.
2. Each of the "next-generation" RO membranes tested demonstrated the capability of providing permeate water with less than 200 mg/L TDS across the influent water temperature range and less than 300 mg/L TDS across the effluent temperature range.
3. RO membranes operated effectively at a flux of 8 to 12 gallons per square foot of membrane area per day (GFD) on both the MF and UF filtrate.
4. Analyses for domoic acid with a lower detection limit of 0.002 μg/L did not detect any in the RO permeate, even when elevated concentrations (2–3 μg/L) existed in the raw feed water due to substantial algae bloom events.
5. Both the MF backwash and the RO concentrate waste streams were characterized for disposal options.
6. For the Siemens CMF-S microfiltration system:
    a. A flux of 34 GFD was found to be sustainable on the influent feed source (as established in Phase A) and was shown to be optimum for operation on the effluent source.
    b. Chlorine was added to the backwash and this addition was considered critical to performance achievement.
    c. Optimum MF operating conditions were determined to be:
        i. Flux = 34 GFD
        ii. Backwash frequency = 20 minutes
        iii. Backwash with 20 mg/L NaOCl every backwash
        iv. Clean-in-place (CIP) frequency of every 3 weeks
    d. A periodic heated CIP was required to restore membrane permeability. Non-heated CIP's proved to be inadequate to restore the membrane permeability to within 10 percent of its original level. A successful CIP protocol consisted of:
        i. 2-percent citric acid recirculation/aeration at 36–38 °C followed by
        ii. 400–600 mg/L NaOCl recirculation at 20–22 °C
    e. The filtrate water produced had turbidity and a silt density index suitable for spiral RO membranes when the MF system maintained integrity.

f. Fiber damage from shell fragments was prevented by use of an Arkal pre-filter of 70 microns or less.

g. It was necessary to reduce MF capacity by 25–30 percent during the most severe algae bloom (red tide) events.

7. For the Zenon ZW-1000 Ultrafiltration system:

   a. A flux of 27.5 GFD and was found to be sustainable on the effluent source. While this flux was not demonstrated on the influent source, it is expected to be feasible based on similarities in UF performance between the two sources at other operating conditions.

   b. Chlorine was used in the backwash and maintenance clean was critical to the performance achieved. Heating of the maintenance clean and CIP solutions was beneficial.

   c. Optimum UF operating conditions were determined to be:

      i. Flux = 27.5 GFD
      ii. Backwash frequency = 22 minutes
      iii. 4 mg/L NaOCl to be used in every backwash
      iv. CIP frequency of every 3 weeks

   d. Fiber damage from shell fragments was prevented by use of an Arkal pre-filter of 100 micron or less.

   e. It was necessary to reduce UF capacity by 25–30 percent during the most severe algae bloom (red tide) events.

8. Two sets each of Hydranautics (HYD) SWC-4040 and Dow (Filmtec) SW30-4040 membranes were tested in Phase A, and in each set, Dow membranes initially produced significantly better water quality. Each set of SW30-4040 membranes produced permeate with a conductance of approximately 300 µS — 50 percent lower than for SWC-4040. The first set of membranes suffered from membrane oxidation, which proceeded much more rapidly on the Dow membranes. The second set of Dow membranes also experienced a decrease of salt rejecting properties, albeit less severe, whereas the Hydranautics water quality was more stable.

9. The Dow, Hydranautics, and Toray next-generation RO membranes achieved improved boron rejection compared with the earlier versions tested in Phase A. Boron concentrations were consistently below 1 mg/L and in some cases less than 0.5 mg/L. Hydranautics SWC4+ achieved 20 percent lower chloride concentration than the other membranes.

10. Continuous chlorination and subsequent ammonia dosing was tried, in an attempt to create chloramines, and proved to be unsuitable for full-scale implementation due to the creation of bromamine and the resulting oxidation of the RO membranes. This process was replaced by MF backwash chlorination and continuous sodium bisulfite dosage prior to the RO.

11. No relationship was found between RO operating flux and fouling in the range tested, 8 to 12 GFD. RO operation at any flux within this range was found to be sustainable. The optimum RO flux for this study was found to be 9 GFD. However, this optimum is based upon site-specific parameters such as water quality, energy cost, and capital expenses. A flux of 9 GFD may not be optimal for all ocean water sources.

12. Operation on ocean water from the common powerplant influent introduced additional challenges for the treatment process. The powerplant heat treatment cycles, which clear the influent pipes of shellfish or other marine growth by recirculating ocean water at elevated temperature, result in a period of sluff-off of shells and other particulate matter. A strainer was required in front of the pilot membrane system feed pump to prevent blockage of the pump. Furthermore, an 800-micron strainer in series with a 500-micron strainer proved to be ineffective at preventing sand and crushed shell fragments from reaching the MF and UF systems and puncturing fibers. Required prestraining was determined to be an 800-µm screen followed by a 70- to 100-µm Arkal filter.

# INTRODUCTION

West Basin Municipal Water District (WBMWD) conducted an ocean water desalination study at the El Segundo Power Facility in El Segundo, CA. The study included the operation of microfiltration (MF), ultrafiltration (UF), and reverse osmosis (RO) processes, as described in the pilot test protocol document entitled *Seawater Desalination Pilot Plant Project Microfiltration/Reverse Osmosis Pilot Testing Protocol* (Appendix A).

The objectives of the Ocean Water Pilot Test Program were established in the test protocol and are also presented below. Each of these objectives was tested on both powerplant influent (Phase A) and powerplant effluent water (Phase B):

1. Determine the optimum membrane operating flux, backwash and CIP membrane cleaning frequency for both a MF and a UF system operating on Southern California coastal ocean water. Investigate cleaning formulations and techniques for the removal of contaminants found in ocean water, which foul the MF and UF membranes.
2. Determine the optimum membrane operating flux and CIP membrane cleaning frequency for an ocean water RO system operating on MF filtrate and UF filtrate. Investigate cleaning formulations and techniques for removal of contaminants found in microfiltered and ultrafiltered ocean water, which foul RO membranes.
3. Characterize the MF/UF backwash and RO concentrate streams to develop data suitable for evaluation of waste stream disposal options.
4. Demonstrate the performance, specifically the operating pressure and permeate quality, for the latest generation seawater RO membranes from Dow, Hydranautics, Toray, and Koch operating on MF and/or UF filtrate derived from both the influent water and the warmer effluent water from the powerplant cooling loop.

The data from this pilot study will provide the relationship between operating flux rates and membrane fouling rates for MF, UF, and RO membranes. It will also support the development of updated costs for ocean water desalination in California for the domestic water supply.

# PROCESS DESCRIPTION

The pilot plant is located on the California coast in the city of El Segundo at the El Segundo Power Generation Plant. Ocean water is brought through an existing open intake to the powerplant cooling system (≈200 million gallons per day). Existing treatment by the power station consists of a coarse traveling screen (>1 inch) and intermittent chlorination. Standard powerplant practice consists of two treatment techniques for controlling organic activity in the cooling loop. Chlorination is manually initiated two times per week for a duration of approximately two hours. The addition rate results in a total chlorine concentration at the plant outfall (condenser effluent) of approximately 0.06 mg/L. This dosage translates to a trace chlorine amount (<0.1 mg/L) in the pilot plant feed water. Secondly, approximately every 2 to 3 months the powerplant cooling water is "heat treated" to control biological growth/attachment. Duration of this treatment is 1 hour at 105–120 °F. The pilot equipment was shut down during the heat treatment events.

The feed water to the pilot plant was Pacific Ocean water with an average analysis as indicated in table 1.

Table 1.—Ocean Water Quality
[All values in mg/L except for pH and temperature]

| Constituent | Value | Constituent | Value |
|---|---|---|---|
| Calcium | 407 | Bromide | 64 |
| Magnesium | 1,335 | Boron | 3.8 |
| Sodium | 10,963 | Nitrate (as N) | <25 |
| Potassium | 404 | Fluoride | 0.9 |
| Ammonia (as N) | 0.05 | Silica | <10 |
| Barium | <0.025 | Total Dissolved Solids | 34,500 |
| Strontium | 7.7 | pH | 8.1 |
| Bicarbonate (as $CaCO_3$) | 115 | Total organic carbon | 1.2 |
| Sulfate | 2,537 | Temperature (°C) | 15.5–24 |
| Chloride | 19,080 | Temperature (°F) | 60–75 |

The overall pilot treatment process is indicated in the initial process flow diagram (figure 2). Originally, the first component of the pilot treatment process was a transfer pump, which provided sufficient head for delivery of ocean water through an 800-micron duplex basket strainer to the microfiltration system. The ON/OFF operation of the transfer pump was controlled by the MF system. The strainer design allows cleaning of one basket while the other was in operation, without interruption of the treatment process. Initially, 1 mg/L sodium hypochlorite was injected prior to the microfiltration system by a flow-paced sodium hypochlorite addition system. Data from MF pilot operations at other ocean water pilot sites indicated that the presence of free chlorine would improve the MF performance.

The MF system was a Siemens CMF-S system, using 0.1-micron nominal pore size polyvinylidene fluoride (PVDF) hollow-fiber technology. The PVDF membrane chemistry has a high tolerance of chlorine and other oxidants, providing a wide range of options for the control of biological growth within the system and the prevention of membrane fouling due to organic matter. The CMF-S process consists of four modules submerged in a process tank. The MF filtrate pump applies suction to the lumens (open interiors) of the fibers, drawing water though the walls of the fibers while particulate matter accumulates on the outside surface of the fibers. The CMF-S process includes periodic interruption of filtration for backwashing of the fibers. The filtration period was 15 minutes at the start of the testing. Following the filtration period, the fibers are backwashed by reversing the filtrate flow and introducing an air scour across the membrane's outside surface. Subsequently, the process tank is drained and refilled. The entire backwash operation consumes about 2.5 minutes. A critical MF process parameter is the operating flux (filtrate flow per unit area of membrane). Initially the MF was operated at a filtrate flow setpoint of 20 gallons per minute (gpm) (5 gpm per module; 21.5 GFD instantaneous fluxes).

The UF system was a Zenon ZW1000 utilizing 0.02-micron nominal pore size fibers also made of PVDF. Like the Siemens system, the Zenon ZW1000 technology is submerged and requires a filtrate suction pump to draw water through the fibers. However, the module configuration is different, and Zenon modules or cassettes hold the fibers in a horizontal arrangement. Like the Siemens system, the ZW1000 uses set filtration time periods segregated by brief backwashes or backpulses. The ZW1000 also uses a process called a maintenance clean (MC). The MC is a mini-CIP, during which the unit is shut down for approximately 30 minutes and a chemical solution is recirculated.

MF and UF filtrates were directed to covered break tanks, which serve to equalize flow between the intermittent MF/UF production and the continuous RO process. Provision was made for adding chemicals to the MF filtrate stream before it enters the break tank. The chemical metering pump was suitable for the addition of either ammonium hydroxide or sodium bisulfite, for chloramine formation or dechlorination, respectively. The elimination of free chlorine was necessary to protect the polyamide RO membranes, which are subject to damage from exposure to strong oxidants.

Initially, operation of the pilot included the addition of ammonium hydroxide to the chlorinated feed at this location. The ammonium hydroxide dose was on a mole ratio of 2:1 $NH_3$:HOCl. This ratio provided an excess of ammonia to ensure the combination of all free chlorine. The RO membranes have tolerance for low concentrations of chloramine, but minimal tolerance for free chlorine.

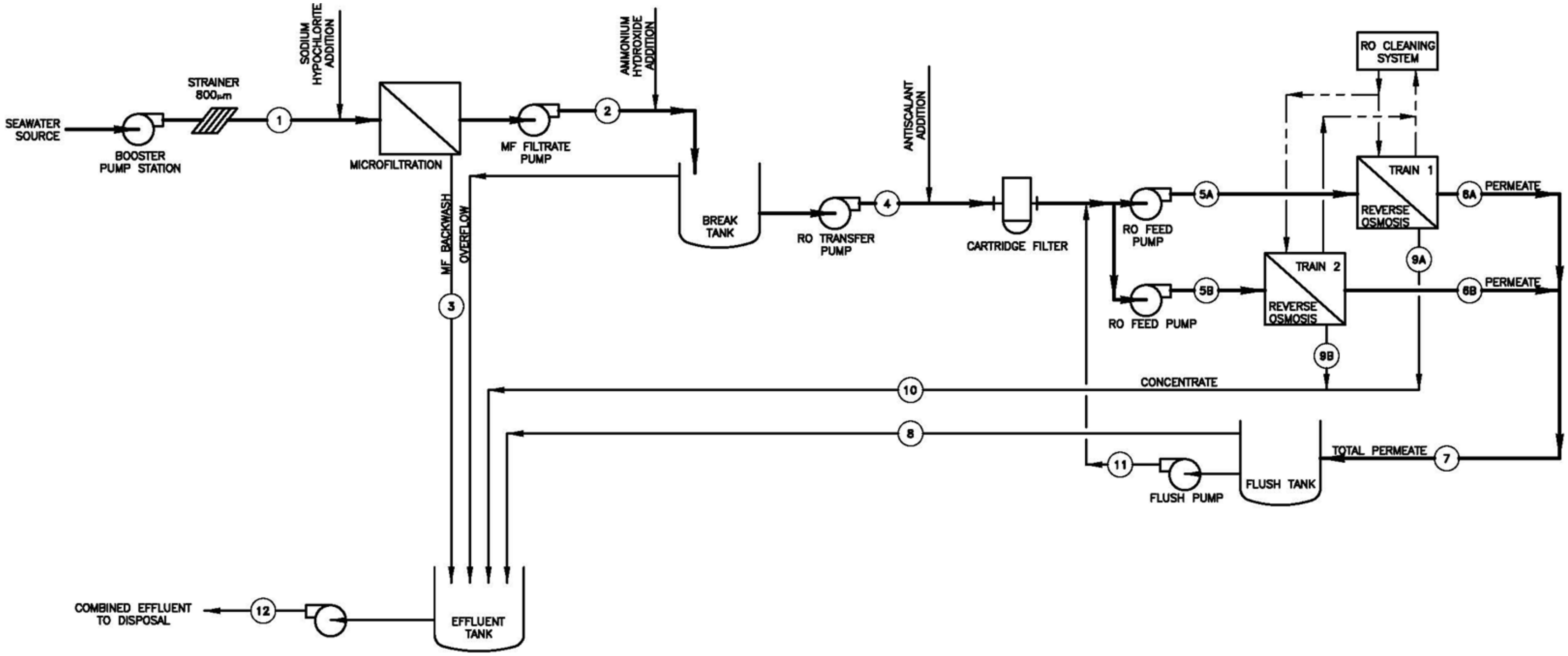
The MF filtrate was then pumped from the break tank by a booster pump to the RO system. The booster pump discharge was approximately 35–50 psi, delivering RO feed water through cartridge prefilters and providing sufficient suction pressure to the RO high-pressure pumps. Excess MF filtrate overflowed the break tank to the combined effluent tank. Permatreat PC-191 antiscalant (3 mg/L) was injected downstream of the RO booster pump. The stream then ran through 20-micron cartridge filters to provide mixing and a barrier to debris introduced at the break tank. No acid was added to the RO feed stream.

Following cartridge filtration, the stream split to feed two identical RO units (Train 1 and Train 2). Each train consisted of a high-pressure pump feeding two 4-inch-diameter pressure vessels in series. Each vessel was capable of holding four elements in series. During this study, a spacer assembly was used in one vessel to allow operation of seven elements in series. Concentrate flow was manually adjusted to the flow setpoint using the concentrate control valve. The RO units were fed using positive-displacement high-pressure pumps. Therefore, permeate flow was manually adjusted to a setpoint using the high-pressure pump recycle control valve. The RO system included ancillary cleaning and flush systems. Upon shutdown, the RO system was automatically flushed with RO permeate.

Hydranautics and Dow were selected to provide RO membranes for Phase A of this study as these two manufacturers have products meeting the treatment requirements and have a substantial share of worldwide reverse osmosis membrane sales.

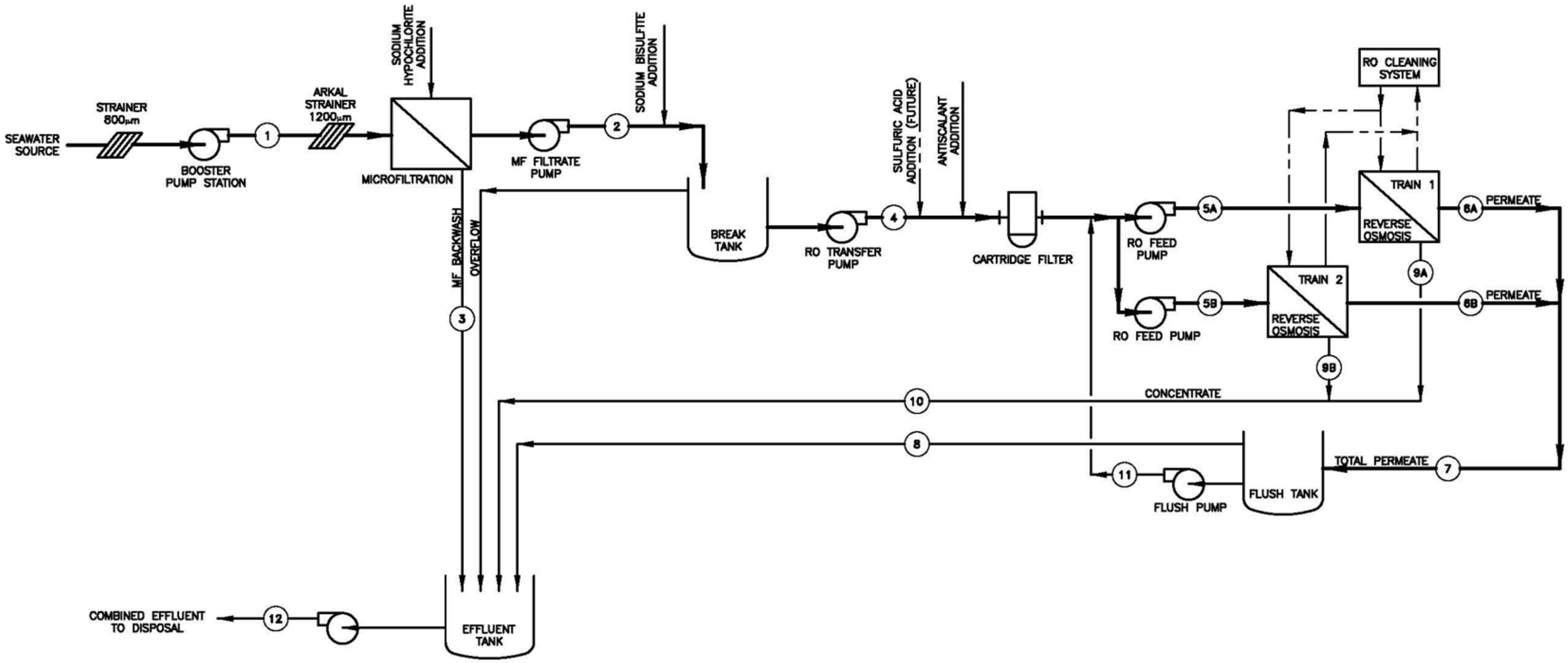
Many process and equipment challenges occurred over the course of this study. Some of these, as described below in the “Process and Equipment Challenges” section, required modifications to the process flow of the pilot equipment. Figure 3 shows the revised Phase A testing process flow diagram for the pilot equipment. The major issues that required process flow modification are discussed below.

Phase B of the testing introduced the Zenon ZW1000 ultrafiltration system, and the ability to run the equipment on the warmer effluent water source. The Phase B testing process flow diagram is presented below as figure 4.



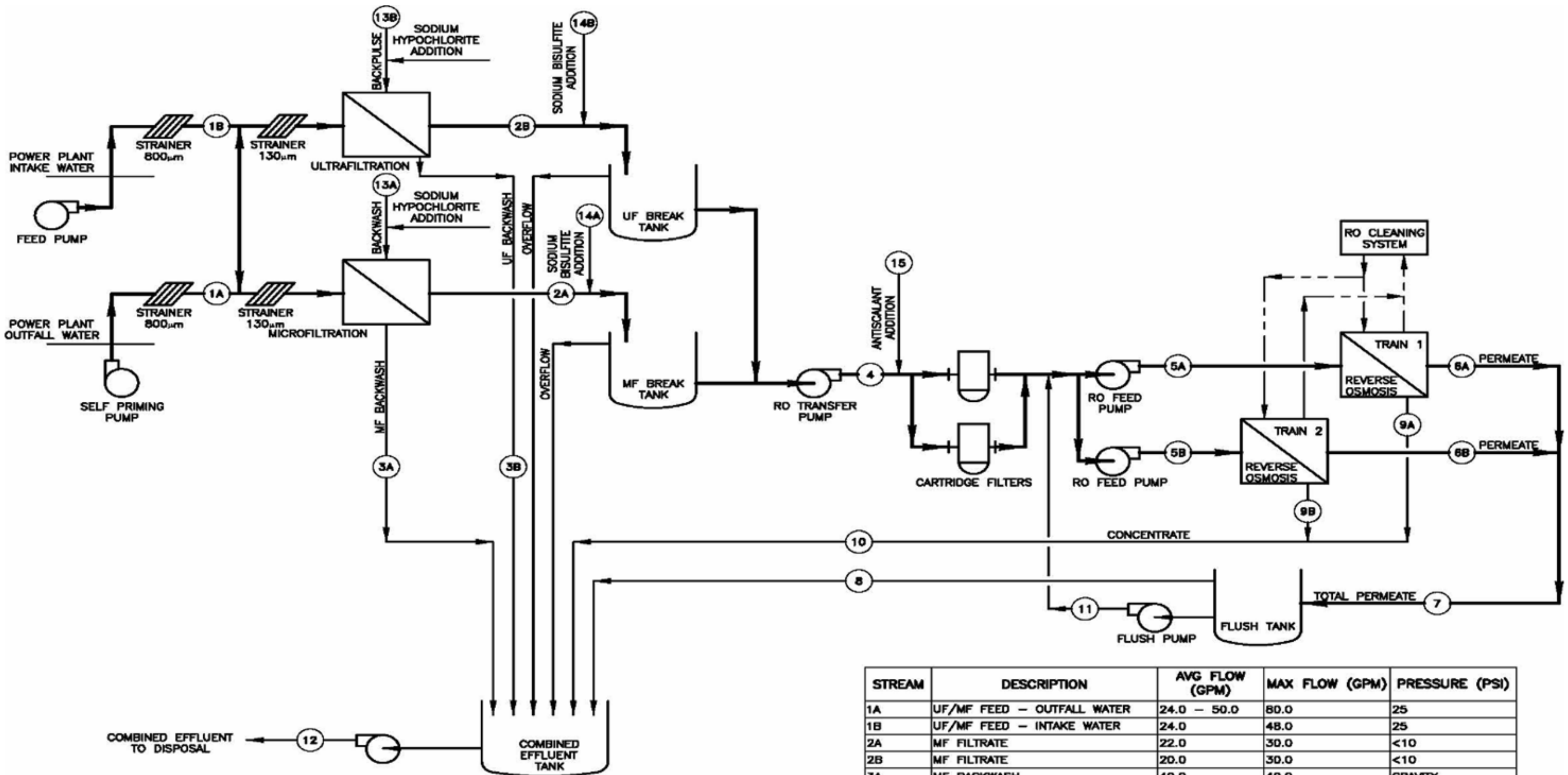

| STREAM | DESCRIPTION | AVG FLOW (GPD) | MAX FLOW (GPM) | PRESSURE (PSI) |
|---|---|---|---|---|
| 1 | MF FEED | 26,250 – 42,600 | 40.0 | 25 |
| 2 | MF FILTRATE | 23,000 – 41,000 | 32.0 | <10 |
| 3 | MF BACKWASH | 1,760 – 3,280 | 40.0 | GRAVITY |
| 4 | RO LOW PRESSURE FEED | 22,400 – 28,600 | 18.0 | 40 |
| 5 | RO HIGH PRESSURE FEED | 11,200 – 13,300 | 9.0 | 900–1,000 |
| 6 | RO TRAIN PERMEATE | 5,800 – 6,720 | 4.5 | 10 |
| 7 | TOTAL RO PERMEATE | 11,200 – 13,300 | 9.0 | 10 |
| 8 | TOTAL RO PERMEATE TO WASTE | 11,200 – 13,300 | 9.0 | GRAVITY |
| 9 | RO TRAIN CONCENTRATE | 5,800 – 6,720 | 4.5 | 10 |
| 10 | TOTAL RO CONCENTRATE | 11,200 – 13,300 | 9.0 | 10 |
| 11 | RO FLUSH | N/A | 10.0 | 15 |
| 12 | COMBINED EFFLUENT | 26,250 – 42,600 | 40.0 | TBD |

**Figure 2.—Initial process flow diagram of the pilot system.**




| STREAM | DESCRIPTION | AVG FLOW (GPD) | MAX FLOW (GPM) | PRESSURE (PSI) |
|---|---|---|---|---|
| 1 | MF FEED | 26,250 – 42,600 | 40.0 | 25 |
| 2 | MF FILTRATE | 23,000 – 41,000 | 32.0 | <10 |
| 3 | MF BACKWASH | 1,760 – 3,280 | 40.0 | GRAVITY |
| 4 | RO LOW PRESSURE FEED | 22,400 – 28,600 | 18.0 | 40 |
| 5 | RO HIGH PRESSURE FEED | 11,200 – 13,300 | 9.0 | 900–1,000 |
| 6 | RO TRAIN PERMEATE | 5,800 – 6,720 | 4.5 | 10 |
| 7 | TOTAL RO PERMEATE | 11,200 – 13,300 | 9.0 | 10 |
| 8 | TOTAL RO PERMEATE TO WASTE | 11,200 – 13,300 | 9.0 | GRAVITY |
| 9 | RO TRAIN CONCENTRATE | 5,800 – 6,720 | 4.5 | 10 |
| 10 | TOTAL RO CONCENTRATE | 11,200 – 13,300 | 9.0 | 10 |
| 11 | RO FLUSH | N/A | 10.0 | 15 |
| 12 | COMBINED EFFLUENT | 26,250 – 42,600 | 40.0 | TBD |

**Figure 3.—Revised process flow diagram, Phase A.**




| STREAM | DESCRIPTION | AVG FLOW (GPM) | MAX FLOW (GPM) | PRESSURE (PSI) |
|---|---|---|---|---|
| 1A | UF/MF FEED – OUTFALL WATER | 24.0 – 50.0 | 80.0 | 25 |
| 1B | UF/MF FEED – INTAKE WATER | 24.0 | 48.0 | 25 |
| 2A | MF FILTRATE | 22.0 | 30.0 | <10 |
| 2B | MF FILTRATE | 20.0 | 30.0 | <10 |
| 3A | MF BACKWASH | 40.0 | 40.0 | GRAVITY |
| 3B | MF BACKWASH | 40.0 | 40.0 | GRAVITY |
| 4 | RO SYSTEM FEED | 17.6 | 20.0 | 40 |
| 5A | RO TRAIN 1 HIGH PRESSURE FEED | 8.2 | 10.0 | 900 |
| 5B | RO TRAIN 2 HIGH PRESSURE FEED | 9.4 | 10.0 | 900 |
| 6A | RO TRAIN 1 PERMEATE | 4.1 | 5.0 | 2 |
| 6B | RO TRAIN 2 PERMEATE | 4.7 | 5.0 | 2 |
| 7 | RO SYSTEM TOTAL PERMEATE | 8.8 | 10.0 | 2 |
| 8 | RO PERMEATE TO WASTE | 8.8 | 10.0 | 2 |
| 9A | RO TRAIN 1 CONCENTRATE | 4.1 | 5.0 | 2 |
| 9B | RO TRAIN 2 CONCENTRATE | 4.7 | 5.0 | 2 |
| 10 | RO SYSTEM TOTAL CONCENTRATE | 8.8 | 10.0 | 2 |
| 11 | RO FLUSH | N/A | 10.0 | 15 |
| 12 | COMBINED EFFLUENT | – | 80.0 | 15 |
| 13A | MF 12% SODIUM HYPOCHLORITE | – | 0.2 GPD | 20 |
| 13B | MF 12% SODIUM HYPOCHLORITE | – | 0.2 GPD | 20 |
| 14A | 38% SODIUM BISULFITE | 0.6–1.0 GPD | 1.0 GPD | 15 |
| 14B | 38% SODIUM BISULFITE | 0.6–1.0 GPD | 1.0 GPD | 15 |
| 15 | RO SYSTEM ANTISCALANT | 0.05–0.1 GPD | 0.1 GPD | 50 |

**Figure 4.—Phase B process flow diagram.**

# MICROFILTRATION OPTIMIZATION AND PERFORMANCE

## Phase A Testing

Operation of the Siemens CMF-S system (figure 5) began in June 2002, with the first month used as an equipment commissioning period. The first stable run started on July 19, 2002. The Phase A MF trials are summarized in tables 2 and 3. The testing is divided between different test "trials" and "runs." A trial is defined here as a significant process change. A run is simply operation between chemical cleaning events, module replacements or operational changes.


**Figure 5.—Siemens CMF-S microfiltration pilot system.**

Table 2.—Phase A MF Testing Trials

| MF Testing Trials | Process Description |
|---|---|
| MF I | Continuous chlorination in MF feed water |
| MF II | Operation without chlorination |
| MF III | Operation with no chlorine in the feed but with chlorination of backwash |
| MF IV | Redesigned MF module, operation with chlorination of backwash |
| MF V | Arkal 130-μm filter in front of MF, operation with chlorination of backwash and redesigned MF module |

Table 3.—Details of Each Phase A Microfiltration Run

| Trial | Run # | Dates | MF Run Hours | Total Filtrate Flow, gpm | Per Module Filtrate Flow, gpm | Flux GFD | Target Feed Chlorination (ppm) | Backwash Frequency, min | Comments |
|---|---|---|---|---|---|---|---|---|---|
| MF I | MF 1 | 7/19/02–8/8/02 | 525–951 | 20 | 5 | 21.5 | 1 | 15 | Unit ran continuously between 525 (7/19) and 951 (8/7) hrs |
| | MF 2 | 8/9/02–9/28/02 | 965–1853 | 22 | 5.5 | 23.6 | 1 | 15 | Stable performance |
| MF II | MF 3 | 10/3/02–10/8/02 | | 22 | 5.5 | 23.6 | 0 | 15 | Ran <1 week before CIP |
| | MF 4 | 10/10/02–10/17/02 | | 22 | 5.5 | 23.6 | 0 | 15 | Ran <1 week before CIP |
| MF III | MF 5 | 10/22/02–11/4/02 | 2263– | 22 | 5.5 | 23.6 | 10 in every backwash | 15 | Ran ~10 days before CIP required |
| | MF 6 | 11/7/02–11/26/02 | 2648–2860 | 22 | 5.5 | 23.6 | 40 in every backwash | 15 | Stable performance |
| | MF 7 | 11/26/02–12/19/02 | 2868–3357 | 22 | 5.5 | 23.6 | 25 in every backwash | 15 | Stable, No CIP before this run |
| | MF 8 | 12/23/02–1/9/03 | 3382–3600 | 24 | 6 | 25.8 | 25 in every backwash | 15 | 1 problematic module replaced, added rinse to protect RO CIP 12/26 request by USF to wet new module |
| | MF 9 | 1/9/03–1/24/03 | 3600–3820? | 24 | 6 | 25.8 | 25 in every backwash | 15 | 1/9, CIP replaced header assembly O-ring. 1/15, Replaced a second original module that had a crack in the potting. Salt density index (SDI) now 2.4. RO membranes replaced |
| | MF 10 | 1/24/03–2/5/03 | 3820?–4028 | 24 | 6 | 25.8 | 25 in every backwash | 15 | Heater broken. CIP not very effective before this run |
| | MF 11 | 2/5/03–2/21/03 | 4028–4242 | 24 | 6 | 25.8 | 25 in every backwash | 15 | Heater broken. CIP not very effective before this run. Electrical problem shutdown 2/11– 2/13 |
| | MF 12 | 2/21/03–3/6/03 | 4242–4513 | 24 | 6 | 25.8 | 25 in every backwash | 15 | Inadvertent daily mini-CIP with chlorine improved performance |
| | MF 13 | 3/6/03–3/11/03 | 4513–4623 | 24 | 6 | 25.8 | 25 in every backwash | 15 | |
| | MF 14 | 3/12/03–4/3/03 | 4650–5100 | | 6 | | 40 in every backwash | 15 | Various flows |
| MF IV | MF 15 | 10/22/03–11/13/03 | 5380–5723 | 18 | 4.5 | 23.6 | 20 in every backwash | 15 | **Restart with redesigned membranes (new module design), increasing permeability** |
| | MF 16 | 1/15/04–03/10/04 | 5840–6296 | 26 | 6.5 | 34 | 20 in every backwash | 15 | Post run CIP performed, over 120 pins added to the four modules. Majority of run w/o Arkal filter due to installation problems |
| MF V | MF 17 | 03/10/04–5/17/04 | 6296–7110 | 26 | 6.5 | 34 | 20 in every backwash | 20 | **Modules Replaced 5/28/04** |
| | MF 18 | 6/8/2004– | 7314– | 26 | 6.5 | 34 | 20 in every backwash | 20 | CIP after very short run. Modules reconditioned |

## Permeability of Original CMF-S Module Design

The Siemens CMF-S system runs at constant flux and thus, as the membrane fouls, the trans-membrane pressure (TMP) required to maintain the throughput rises. However, because transmembrane pressure is also influenced by water temperature and variations in flow, the appropriate method of monitoring membrane fouling is to observe variations in the temperature-corrected permeability or specific flux.

Permeability is the filtrate flux divided by the temperature-corrected TMP and is typically reported in units of GFD/psi. The terminal TMP (the TMP at which membrane cleaning is required) for the CMF-S system is 12 psi. Thus, at a filtrate flux of 22–26 GFD, and a temperature of ~20 °C, the unit should be cleaned when the permeability reaches ~2 GFD/psi. At a flux of 34 GFD, the unit should be cleaned at 2.6 GFD/psi.

### *Trial I—Continuous Prechlorination*

MF runs 1 and 2 were performed with continuous chloramination in the feed water, as indicated in table 3. The MF demonstrated very stable operation during this period as indicated in figure 6. After an initial 3-week run, the MF membrane was cleaned, the flux increased to 24 GFD and the unit was restarted. This 24-GFD run with continuous chloramination lasted over 6 weeks without requiring a chemical cleaning. Variations of TMP, filtrate flux, and permeability in trial I are illustrated in figure 6. The continuous chloramination was discontinued following MF Trial I as the process of chlorination followed by MF followed by ammonia dosing resulted in oxidation of the RO membranes. The bromide ion naturally present in ocean water interfered with the intended formation of chloramine, and bromamine was formed. Bromamine is a stronger oxidant than chloramine, and it damaged the downstream RO membranes. This is discussed further in the "Process and Equipment Challenges" section of this document.

At many other ocean water RO installations on open intakes with conventional filtration pretreatment, operators add a reducing agent, such as sodium bisulfite, and allow significant chlorine contact time to neutralize the oxidant before it contacts the RO membranes. However, as demonstrated by Hamida and Moch (1996), this continuous chlorination/dechlorination process has been shown to enhance the tendency towards biological fouling. Therefore, this process was not considered a viable option for this study.

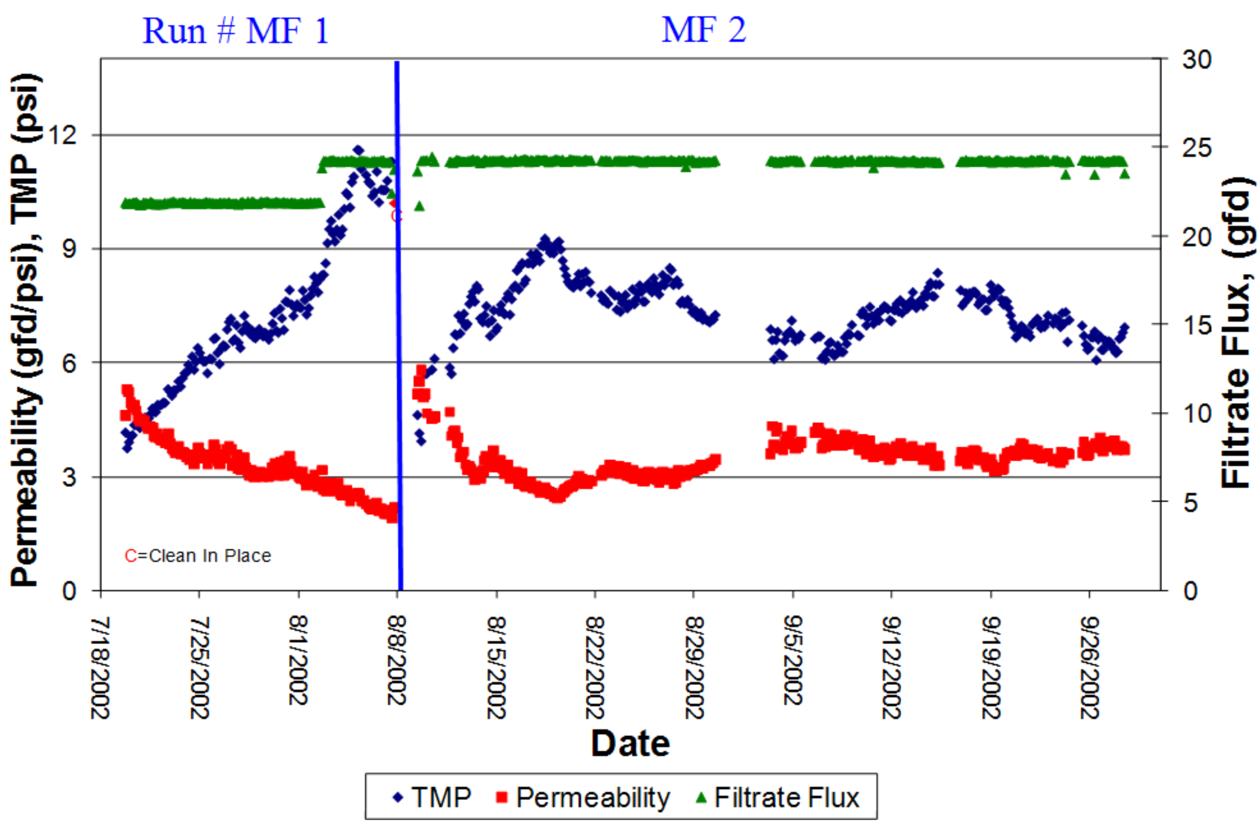



**Figure 6.—Performance of microfiltration system with continuous prechlorination (MF trial I).**

### *Trial II—No Chlorination*

Once prechlorination was abandoned, attempts were made to run the Siemens CMF-S system at the same conditions with no chlorination at all. Rapid fouling was observed in two consecutive runs, as shown in figure 7. Note that neither of these runs lasted more than 10 days before reaching terminal permeability. Operation at 24 GFD was unsuccessful without the chloramination in the feed water, and this result demonstrated how beneficial the oxidant is to the stable performance of the microfiltration membrane process on this feed source.

### *Trial III—Chlorinated Backwashes*

Recognizing the benefit of chlorine to the MF process but accepting that the attempted chloramination of the feed water (Trial I) resulted in an adverse impact to the RO membrane, an alternative approach to the use of chlorine was attempted in MF Trial III, chlorinated backwashes. NaOCl (10 mg/L) was added to every backwash and again rapid fouling was observed, as depicted in figure 8. A stable run condition was finally achieved in run #6 by increasing the dose to 40 mg/L NaOCl in every backwash. This run showed a slow fouling rate over two weeks. When the chlorination was decreased from 40 to 25 mg/L NaOCl for every backwash in run #7, the MF operated for an additional month without requiring a shutdown for a chemical CIP.

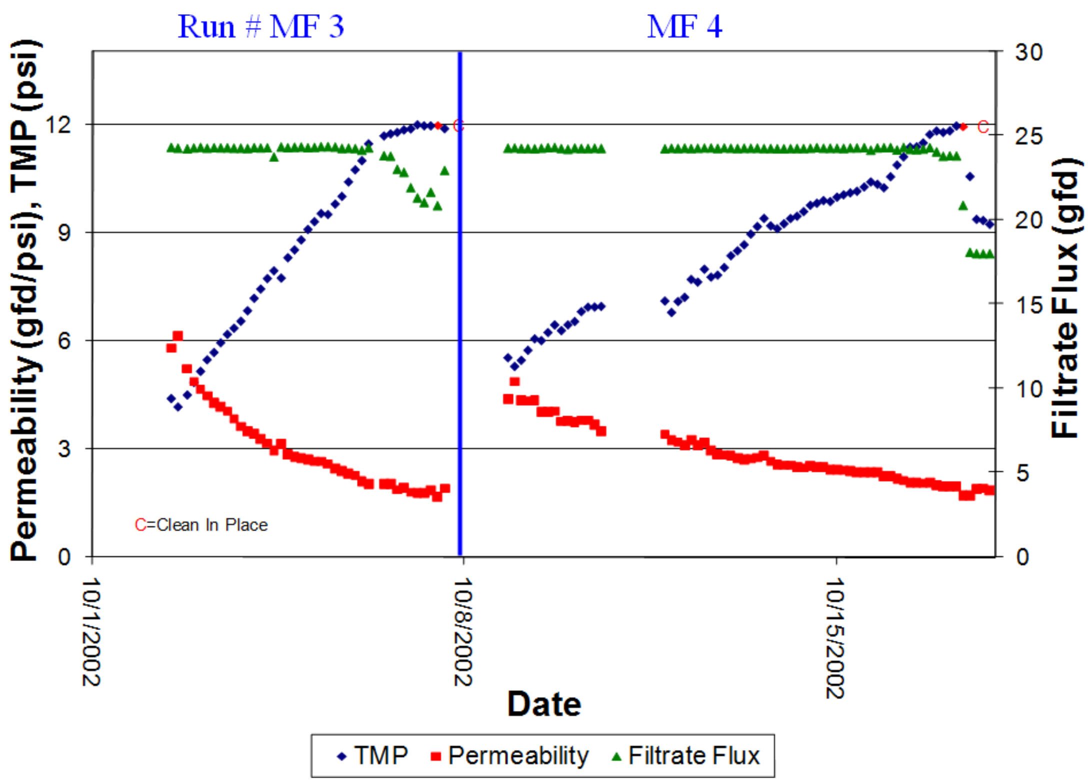



**Figure 7.—Performance of microfiltration system with no chlorination (MF trial II).**

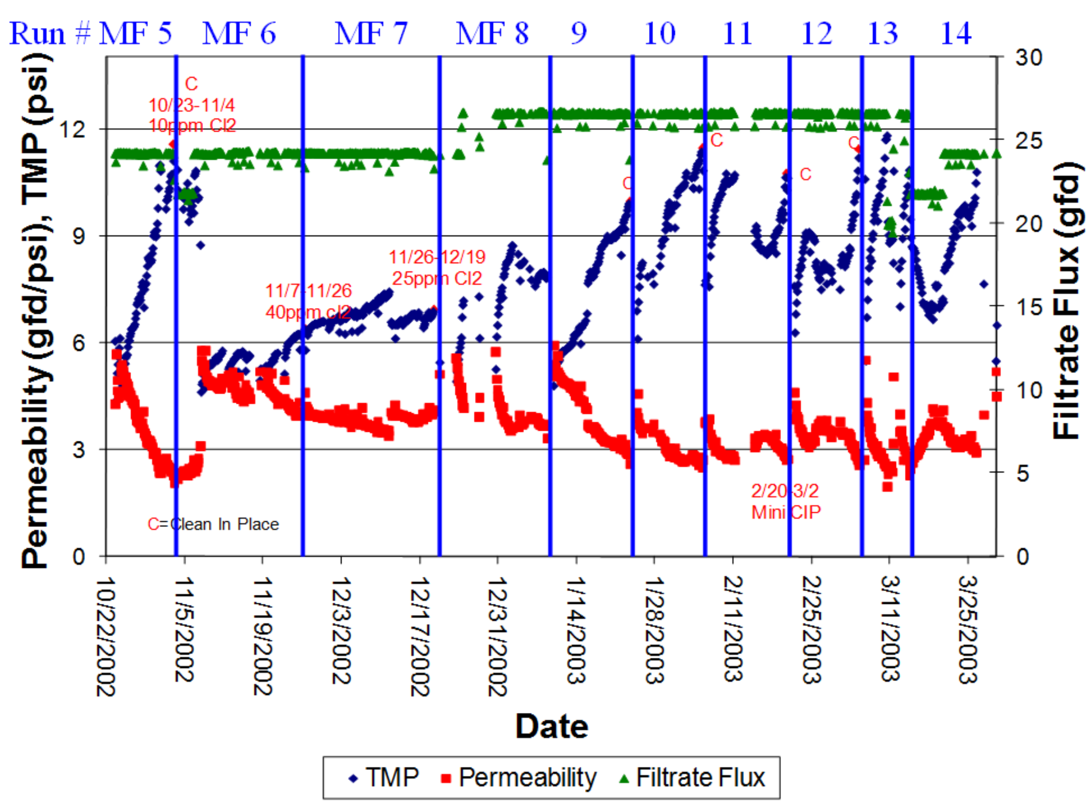



**Figure 8.—Performance of microfiltration system with chlorinated backwashes (MF trial III).**

The filtrate flow was then increased from 22 to 24 gpm for run #8, corresponding to a flux increase from 24 to 25.8 GFD.  Several runs failed to achieve a run time longer than 3 weeks at this flux before a CIP was required.  This result was compounded by the fact that the CMF-S  CIP heater was disabled for a period of time, and the cleanings done to start runs #10 and #11 did not restore the membrane permeability effectively.

Run #13 was started with a fully heated CIP.  However, this run had a very short run time.  Two things were now evident:

1. A filtrate flux of 25.8 GFD was not sustainable with these original CMF-S membranes
2. The membranes had been fouled to the point that the normal heated CIP process did not restore the permeability to a “fully clean” condition or approximately 6 GFD/psi.

During run #14, the filtrate flow and hence the flux rates were varied as shown in figure 9.

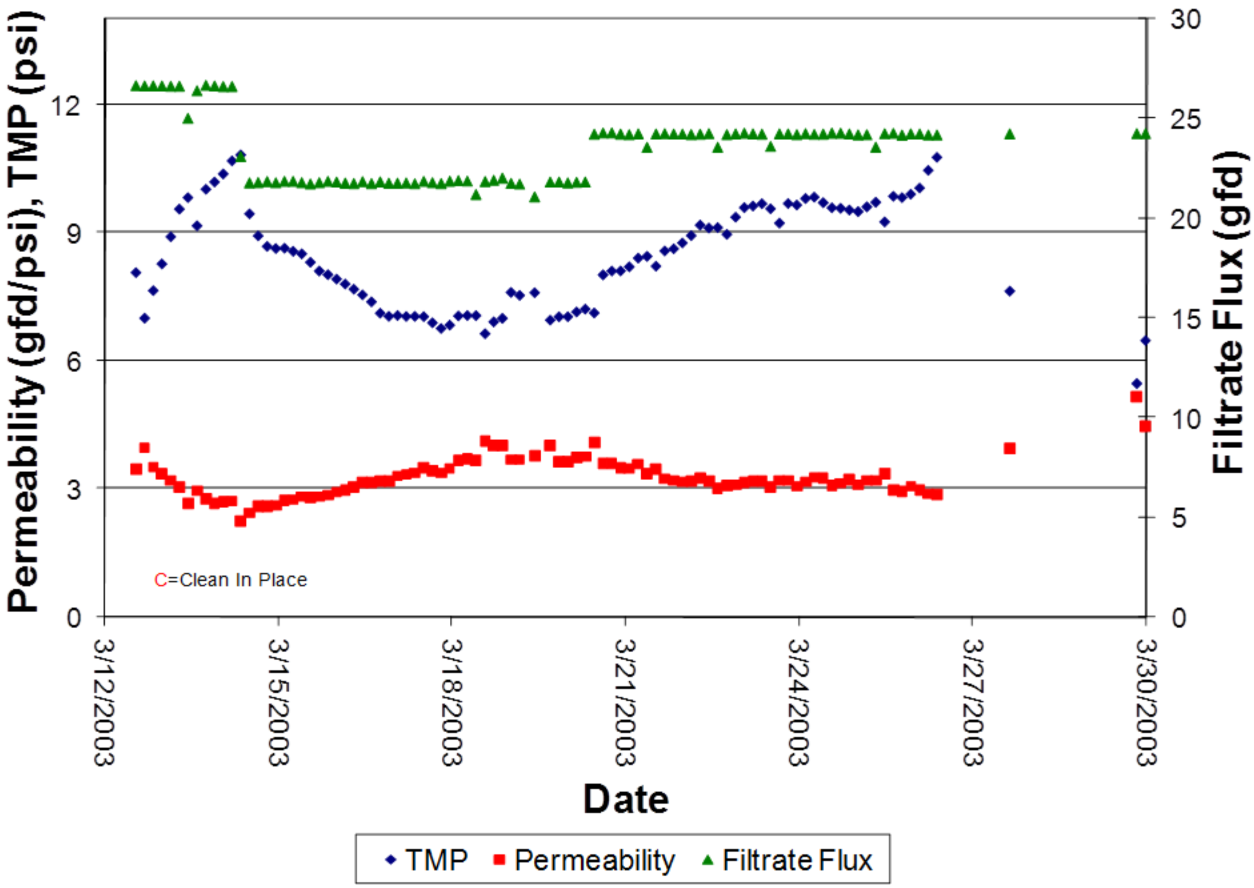



**Figure 9.—Microfiltration run #14 performance.**

The run was started with a filtrate flux of ~25.8 GFD and demonstrated rapid fouling, similar to the previous runs. Dropping the flux down to ~22 GFD improved the permeability results. Subsequently, the flux was increased to ~24 GFD and the fouling rate increased. Close examination of this data reveals that the acceptable filtrate flux on this water is 22–24 GFD with these original CMF-S membranes.

## Cleaning Effectiveness

Examination of figure 8 shows that the "clean" or post-CIP microfiltration permeabilities had declined since January 23, 2003. This is a sign of an ineffective CIP procedure. The problem began when the CMF-S heater failed, and the two subsequent cleanings were performed with cold water on January 23 and February 5, 2003. These cleanings were not effective, as shown in figure 8. The clean permeabilities are only 4 GFD/psi after the cold-water cleanings, whereas with previous heated CIPs, the clean permeabilities were consistently ~6 GFD/psi.

At the completion of run #14, an enhanced CIP process was undertaken in an attempt to restore the clean permeability of the membranes to the ~6 GFD/psi range. The process used hydrochloric acid in addition to the normal citric acid and chlorine. This enhanced process showed improvement, but failed to fully restore the membranes. Examination of the data in figures 6 and 7 demonstrates that the heated CIP was effective at restoring the membrane permeability and it was not until the CMF-S heater failed that the membranes were fouled to the point that not even an enhanced CIP process could restore them. This indicates each CIP solution must be heated to be effective.

Table 4.—Effective Microfiltration Cleaning Procedure.

| Step | Chemical | Temperature (°C) | Procedure |
|---|---|---|---|
| 1 | 2% citric acid | 36–38 | Perform reverse filtration until membrane cell is filled with MF filtrate. Add chemicals, heat solution and aerate every 2 minutes. Perform filtrate recirculation for 30 minutes. Repeat 5-minute aeration/5-minute soak cycles 9 times. |
| 2 | 400–600 mg/L chlorine | 20 | |

## Siemens PVDF Membrane Module Integrity—Original CMF-S Modules (MF Trials I–III)

The Siemens CMF-S unit utilized for this study contains four S10V PVDF modules. Over the course of trials I–III, two of these modules required replacement. The first was replaced on December 10, 2002, due to extensive fiber breakage, and the second on January 7, 2003, after it developed a crack in the epoxy that separated the feed water from the filtrate. Furthermore, fiber breakage was also observed in one of the replacement modules.

Broken fibers were easily detected during the pressure decay test (PDT). During the PDT, the unit was isolated and the lumen (filtrate) side of the modules was drained. Air was then injected to the lumens at 15 psi, and then a valve on the feed side was opened to the air. Intact wetted fibers retain the air pressure, as the pressure decay rate across an intact fiber is diffusion controlled. Broken fibers pass air at a much greater rate than normal diffusion, resulting in a rapid pressure decay rate. The intact Siemens system with no fiber breaks displays a PDT rate of ≤0.5 psi/minute. To quantify the broken fiber problems observed during this study, on March 4, 2003, a PDT was performed on the system, resulting in a decay rate of ~2.3 psi/minute. Thereafter, between 30 and 35 fibers were isolated on one of the four modules in the system. Each original CMF-S module contained ~14,500 fibers. Figure 10 below demonstrates that the unit has had broken fibers over most of trials I–III of the study. Figure 11 shows a visible sign of air passage during a pressure decay test through the crack that developed in the module epoxy.

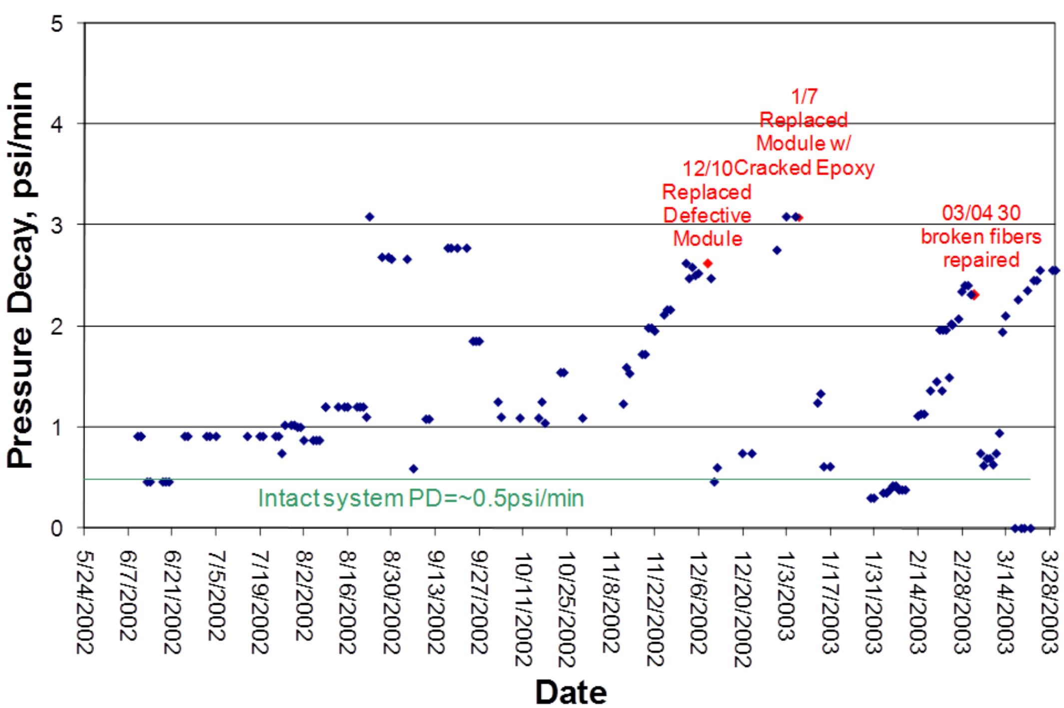



**Figure 10.—Siemens microfiltration unit pressure decay test results Phase A testing.**

Siemens sent their problematic modules to Australia for autopsy to determine the cause of the fiber breakage and epoxy failures. The results from the analysis of the module with the cracked epoxy can be summed up as:

A. The epoxy crack was probably a manufacturing problem resulting from an incorrect epoxy mixing or curing procedure.

B. When the flow distribution screen was removed from the end of the module, particles were found covering 20 mm of the fibers at the bottom. The particles consisted of sand and broken shell fragments that apparently passed through both the 800-µm coarse strainer and the standard 500-µm strainer on the CMF-S unit. It was noted that a number of broken fibers were punctured by what appeared to be sharp objects. It is possible that the broken shell fragments are a cause for some of the fiber breakage problems. The original 500-µm strainer in front of the MF was replaced with a 130-µm Arkal filter to alleviate this problem.

C. Twenty-four fibers were analyzed for fiber break extension or fiber strength. The fiber strength had decreased by 20–40 percent. Scanning electron micrographs (figures 12 and 13) showed that other broken fibers that had sheared appeared to have been stretched before failure.

The fiber stretching and the fact that three of the six modules had no epoxy cracks and minimal fiber breakage, provided evidence of a module manufacturing problem.

Siemens recognized that they had some design and manufacturing issues with their PVDF modules, and they notified West Basin that their module underwent a substantial redesign (table 5), including:

1. Larger fibers (increased diameter and thicker walls)
2. Fewer fibers in each module (different packing density)
3. Reduced fiber area per module

**Figure 11.—Air bubbles emitted from the cracked epoxy during the pressure decay test.**

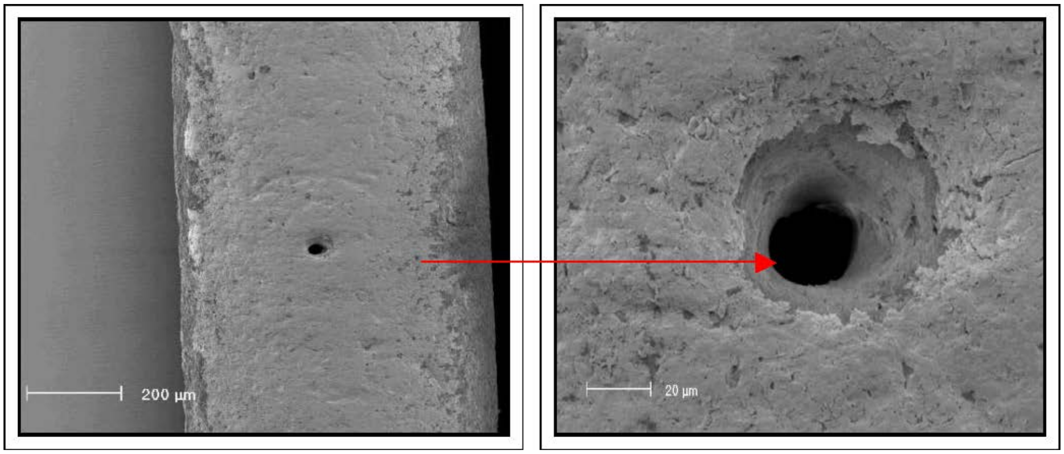



**Figure 12.— Scanning electron micrographs of a hole in a CMF-S module fiber.** The hole was found 490 mm from the top. A closer look shows that it appears to have been caused by a sharp object, or by something wearing into the fiber.

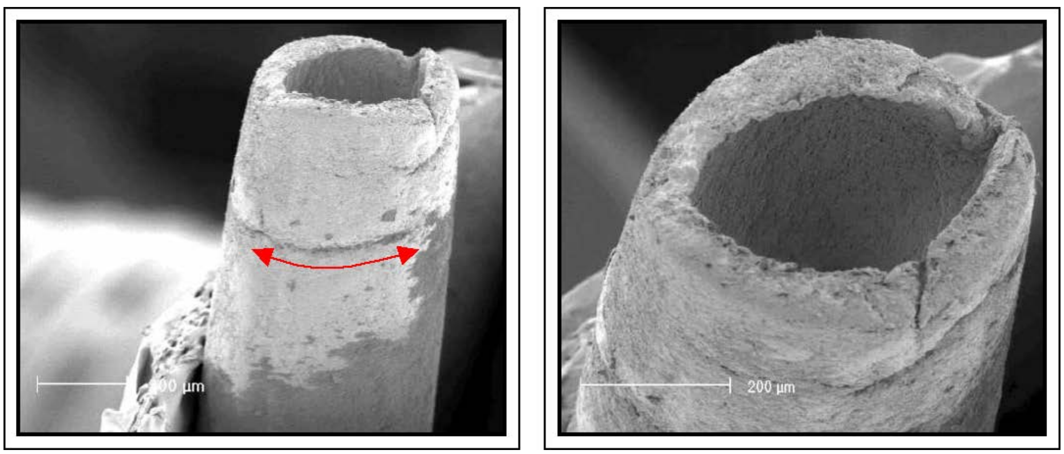



**Figure 13.—Sheared CMF-S fiber shows evidence of stretch failure.** This broken fiber was found 350 mm from the bottom. The fiber has been bent, and the surface appears stretched.

Table 5.—Siemens CMF-S Module Comparison

| Parameter | Original S10V Module Generation A | Redesigned S10V Module Generation B |
|---|---|---|
| Fiber outside diameter, µm | 650 | 800 |
| Fiber inside diameter, µm | 390 | 500 |
| Number of fibers per module | 14,500 | 9,600 |
| Module active membrane area, $m^2$ | 31.1 | 25.3 |

## Arkal Disc Filter System

The Arkal filter operates using a specially designed disc filtration technology. Thin, color-coded polypropylene discs are diagonally grooved on both sides to a specific micron size. These discs are then stacked in columns and compressed on the outside of specially designed spines. When stacked, the grooves on the top of each disc run opposite to the grooves below, creating a filtration element with a statistically significant series of valleys and traps for solids. The stack is enclosed in a corrosion-resistant plastic housing.

The system utilized in this study is a Spin Klin System (figure 14), with two disc filter columns operating in parallel with a third, center housing used for the air assisted backwashing.

During normal filtration mode seawater is fed in parallel through the two disc filter columns and a small volume of filtrate is stored in the third empty housing. After a predetermined time, or on high differential pressure across the discs, a backwash sequence is automatically initiated.

During the backwash process air is fed under pressure into the top of the housing containing the filtered backwash water, and the backwash water is sent to the inside of one of the disc filters to start the backwash process. Inside the disc filter housing, the compression spring holding the discs in place is released and the discs are then able to move freely. Tangential jets of the filtered backwash water are sent through the column of discs in the opposite direction through nozzles at the center of the spine. The discs spin freely, loosening the trapped solids, which are then flushed out through the drain. The freshly cleaned filter column is then operated normally for a brief period of time to collect another volume of filtered backwash water in the third housing, and then the backwash process is repeated on the second filter disc column.

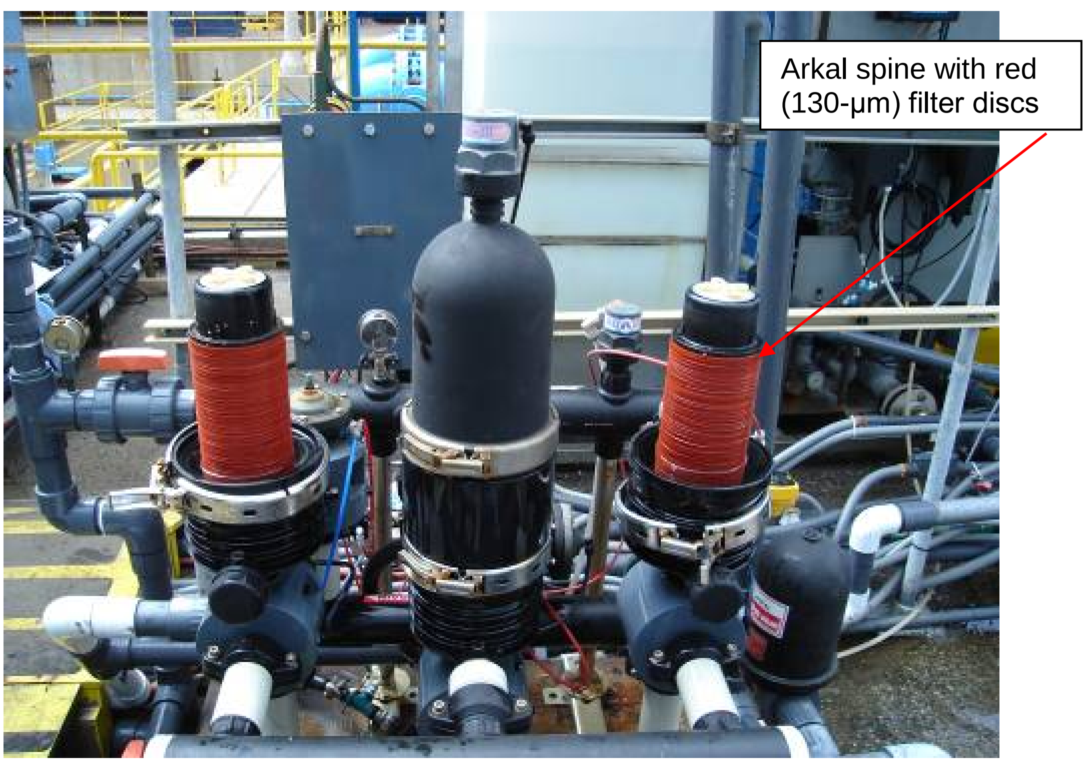



**Figure 14.—Arkal Spin Klin disc filtration system.**

## Performance of Newly Designed CMF-S Modules

### *Trial IV—Redesigned CMF-S Modules Without Arkal Filter*

In October 2003, after a delay in testing due to the reconfiguration of the RO feed pumps (as discussed in the "Process and Equipment Challenges" section below), the trials commenced with the new, improved Siemens CMF-S module. Siemens had postulated that with fewer, larger fibers, the redesigned modules would be more efficient and would be able to run at a higher flux rate and maintain permeability. As figure 15 shows, this proved to be true. The redesigned modules were first run for 8 weeks at the same 24-GFD flux rate as the "original" Siemens modules. No permeability decline (fouling) was observed. The flux was then increased to 34 GFD, and the system stabilized after some initial fouling. Note that the Arkal 130-µm filter was installed for this trial but was bypassed as described in "Process and Equipment Challenges" section.

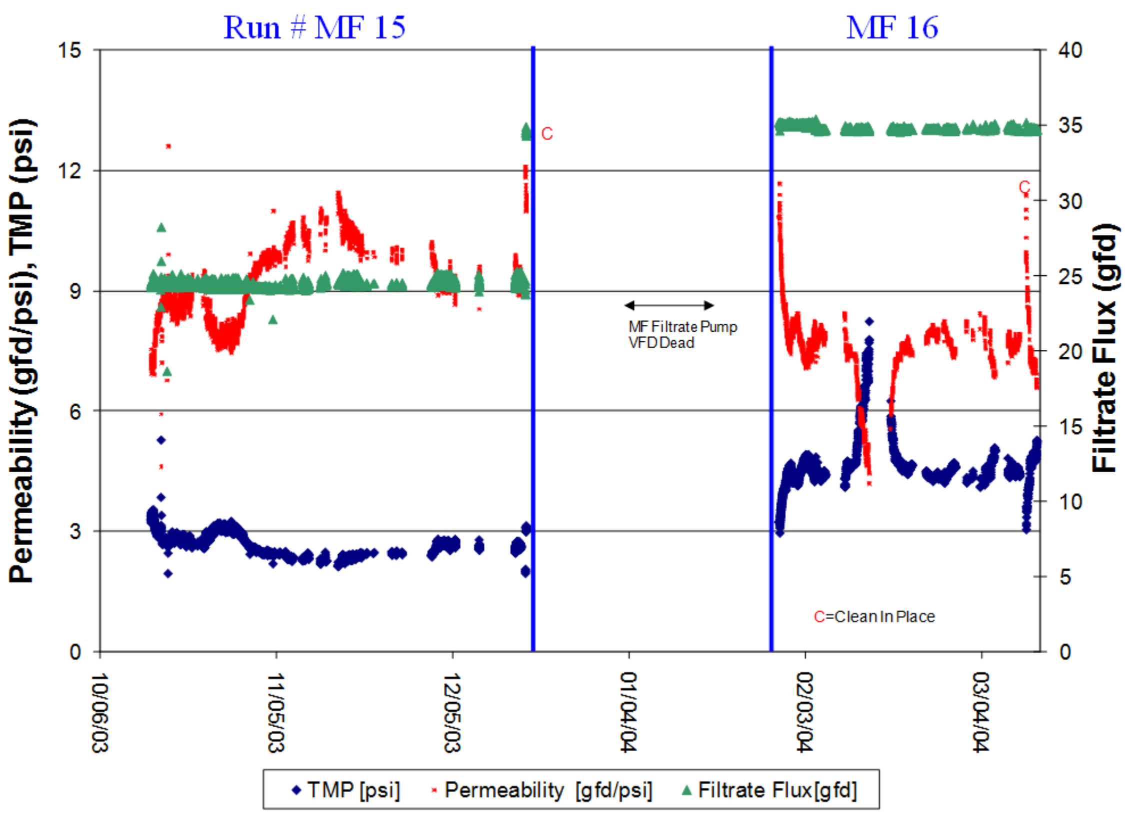



**Figure 15.—Performance of redesigned MF modules (MF trial IV).**

### *Trial V—Performance of New Modules with the Arkal Spin Klin Filter as Pretreatment*

The Arkal Spin Klin 130-µm filter was finally operational on March 10, 2004, and the unit was put on line. Another 34-GFD run was initiated, and the backwash frequency of the Siemens CMF-S unit was decreased from every 15 to every 20 minutes. Figure 16 shows that one run was executed under these conditions and achieved a 3-week run time before a cleaning was required.

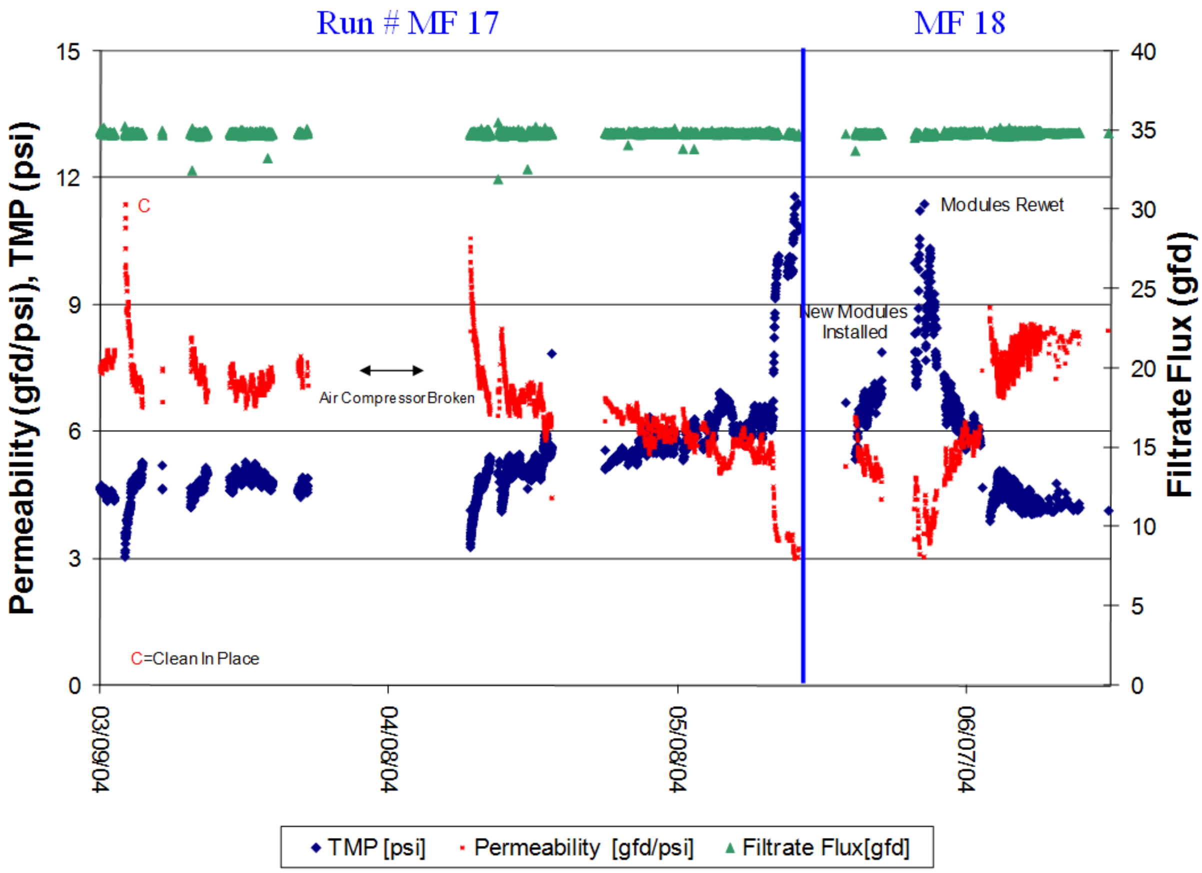



**Figure 16.—Performance of redesigned modules with Arkal filter (trial V).**

Table 6.—Optimized Siemens CMF-S Microfiltration Run Parameters, Phase A

| Parameter | Value |
|---|---|
| Filtrate flow per module (gpm)* | 6.5 |
| Filtrate flux (GFD)* | 34 |
| Filtration time between backwashes (min) | 20 |
| Recovery | 93% |
| Backwash parameters: | |
| Air scour rate (SCFM/module) | 7 |
| Air scour duration (seconds) | 30 |
| Backpulse rate (gpm/module) | 9.9 |
| Air scour + backpulse duration (seconds) | 15 |
| Additional feed to drain volume (gal) | ~25 |
| Rinse duration (seconds) | 15 |
| Refill duration (seconds) | ~35 |
| Backwash chlorination (mg/L)* | 20 |

*Optimized parameters. Non-optimized parameters recommended by Siemens.

## New Redesigned Siemens PVDF Membrane Module Integrity Problems

On May 28, 2004, all four redesigned CMF-S modules were replaced due to numerous fiber breakages. Note that these newly designed modules had been run for at least 300 hours with only the 800-µm strainer as pretreatment, as the 130-µm Arkal Spin Klin filter was bypassed due to installation problems. It was clear nevertheless that the new module design allowed a significantly higher stable operating flux, but it did not maintain integrity with only the 800-µm strainer as pretreatment (figure 17).

The Arkal Spin Klin filter was placed on line in late March, prior to the installation of the second set of redesigned modules (May 28, 2004). Even with the damaged modules still in use, the results following activation of the Arkal 130-µm filter seemed promising, in that (a) the pressure decay did not worsen (figure 17) and (b) the replacement modules held their integrity. Phase A concluded with additional run time using the Arkal filter, in order to determine if this would prevent further MF fiber breakage.

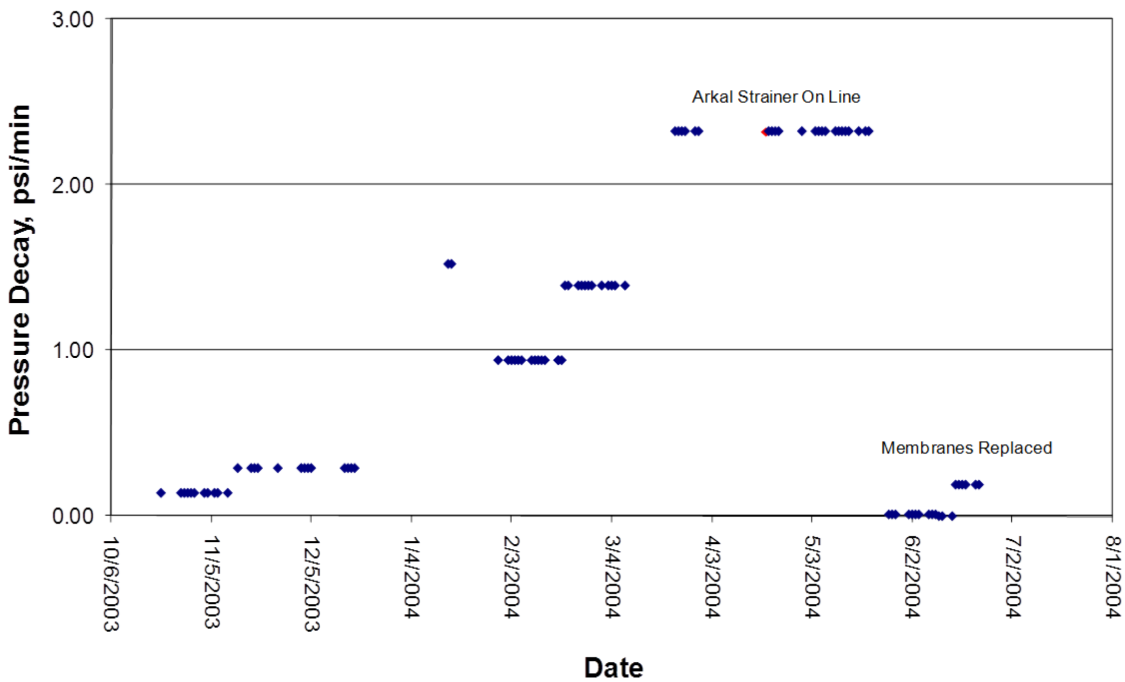



**Figure 17.—PDT results of redesigned CMF-S modules.**

## MF Filtrate Quality

The MF pretreatment is used to condition the raw ocean water to make it suitable for spiral-wound reverse osmosis membranes. This involves particulate matter removal, which is best monitored through measurements of turbidity and silt density. Spiral-wound reverse osmosis membranes operate best when the RO feed water has turbidity less than 1 nephelometric turbidity unit (NTU) and a silt density index (SDI) less than 4.

### *Turbidity*

The presence of suspended material in water causes opacity which is known as turbidity (Kerri, 1994). The raw ocean water and MF filtrate turbidities were measured once per day at the test site. The incoming ocean water turbidity averaged ~1 NTU, with peak values of ~5 NTU. As shown in figures 18 and 19, the MF filtrate turbidity averaged 0.05 NTU and typically was <0.1 NTU, suitable for RO despite the module and fiber problems.

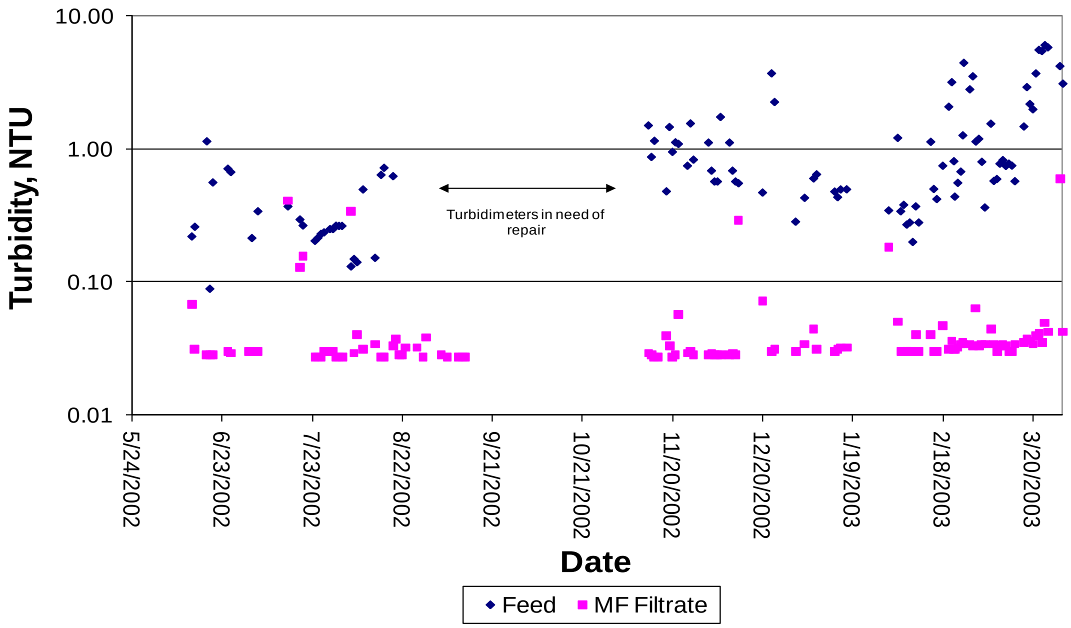



**Figure 18.—Feed water and microfiltration filtrate turbidity—MF trials I–III.**

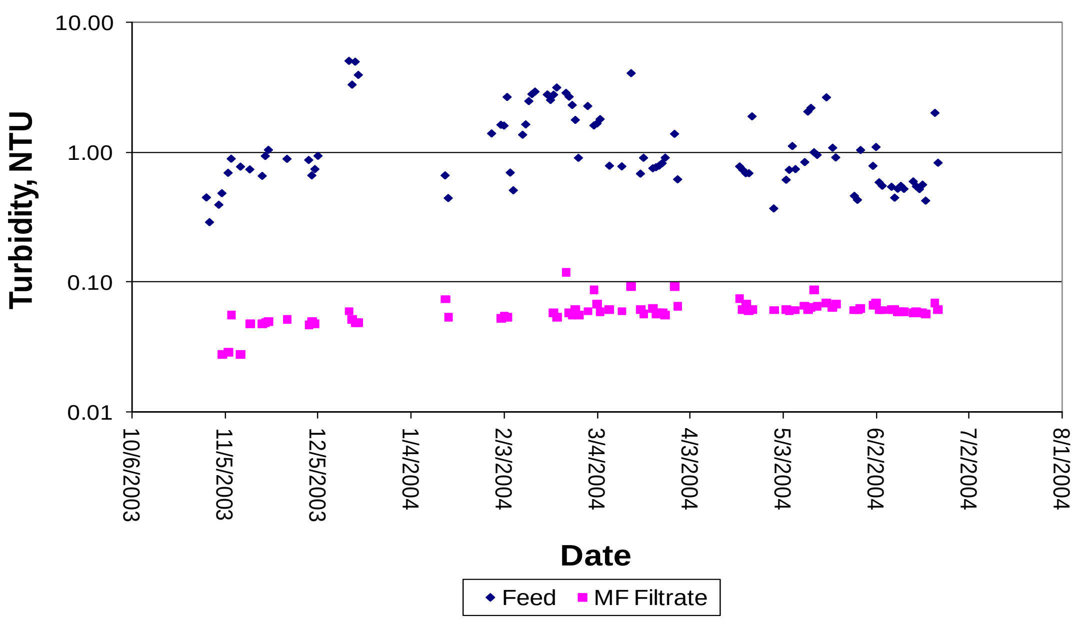



**Figure 19.—Feed water and microfiltration filtrate turbidity—MF trials IV and V.**

## Silt Density Index

The silt density index, or $SDI_{15}$, is a popular method for determining feed-water quality in RO applications. It is based on the difference between (a) the time required to filter a volume of water through a new 0.45-µm filter pad at a feed pressure of 30 psig, and (b) the time required for the same operation after 15 minutes of continuous filtration. Colloidal and suspended matter clogs the filter pad, slowing down the filtration process.

It is important for the feed water to the spiral RO membranes to have an $SDI_{15}$ less than 4 (Hydranautics and Dow). An $SDI_{15}$ greater than 4 represents water that poses an increased risk to RO membrane fouling, declining permeability, and increasing differential pressure.

Multiple attempts to determine the $SDI_{15}$ of the raw ocean water were unsuccessful, as it clogged the SDI pad significantly within 5 minutes and almost completely by the 15-minute mark. The CMF-S system proved to be quite effective at $SDI_{15}$ reduction, typically producing water with an $SDI_{15}$ between 2 and 3. Figures 20 and 21 show the RO Feed $SDI_{15}$ and MF pressure decay. The graph for MF Trials IV and V demonstrates that the $SDI_{15}$ did increase to unacceptable levels when the pressure decay on the MF system exceeded 2 psi/minute. Note that the $SDI_{15}$ reduced to less than 2 after the replacement modules were installed on May 28, 2004. It was therefore important to find a solution to the fiber breakage, not only because of the operating and maintenance efforts required to repair the breaks but also to ensure the water quality leaving the MF system is suitable for spiral-wound reverse osmosis membranes.

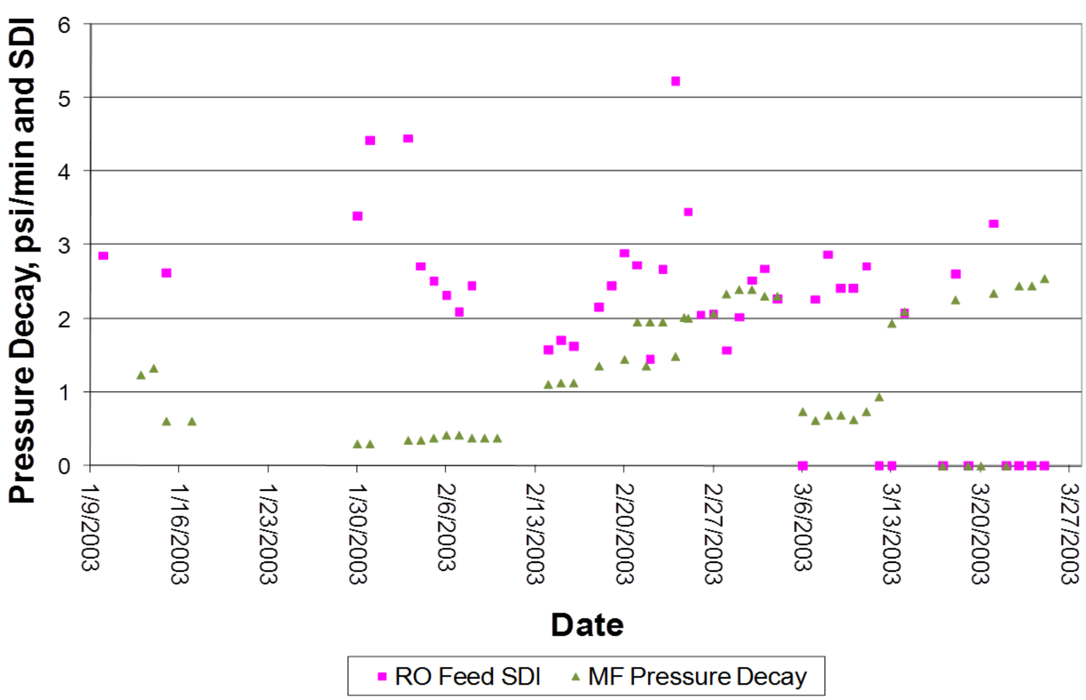



**Figure 20.—CMF-S system pressure decay results and filtrate SDI MF trials I–III.**

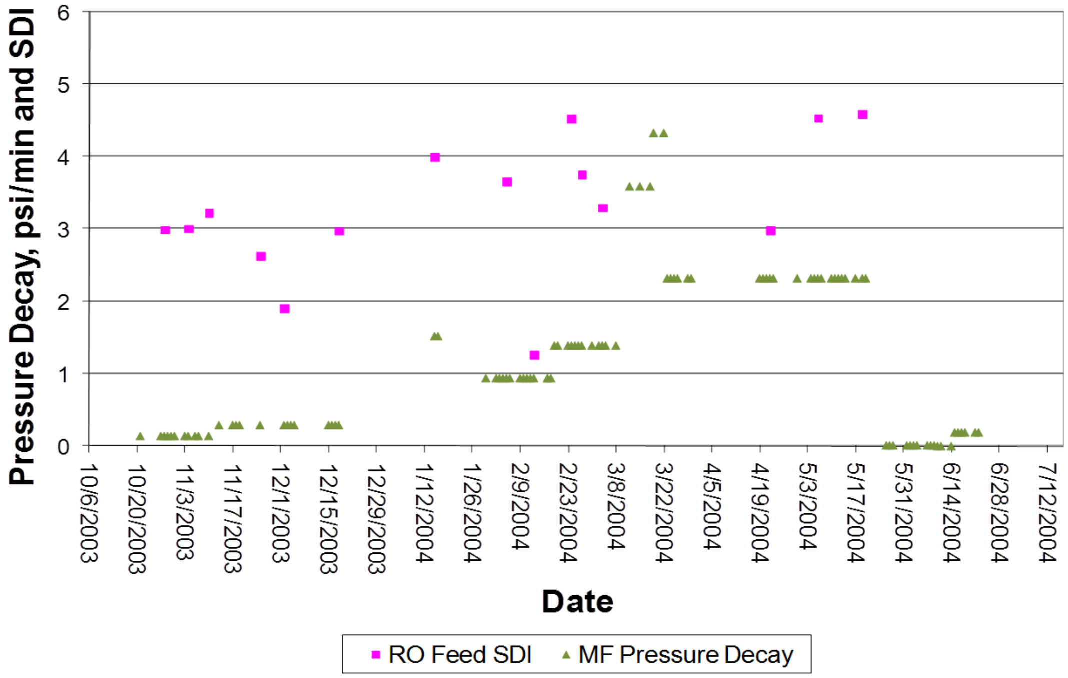



**Figure 21.—CMF-S system pressure decay results and filtrate SDI MF trials IV and V.**

### *MF Filtrate Water Quality Analysis*

Weekly water quality analyses found that the microfiltration system removed a relatively small amount (approximately 10 percent) of total organic carbon (TOC). As expected, inorganic constituents were unaffected.

Table 7.—Microfiltration Water Quality Phase A

| Parameter | Units | DL[1] | MF Testing Phases I – III | | MF Testing Phases IV & V | |
|---|---|---|---|---|---|---|
| | | | Avg. (15 samples) | Std. Dev. | Avg. (15 samples) | Std. Dev. |
| **CMFS FEED** | | | | | | |
| UV 254 [2] | abs/cm | 0.005 | 0.010 | 0.003 | 0.013 | 0.003 |
| Alkalinity (as $CaCO_3$) | mg/L | 2 | 115.3 | 2.1 | 109.3 | 1.3 |
| Calcium | mg/L | 25 | 407.0 | 29.1 | 388.9 | 22.2 |
| Magnesium | mg/L | 25 | 1,335.3 | 103.3 | 1,236.0 | 68.2 |
| Hardness (as $CaCO_3$) | mg/L | 200 | 6,514.9 | 473.7 | 6,060.8 | 313.1 |
| Sodium | mg/L | 25 | 10,963.4 | 733.2 | 10,285.3 | 527.9 |
| Potassium | mg/L | 25 | 403.9 | 31.8 | 394.1 | 26.3 |
| TOC | mg/l | 0.5 | 0.95 | 0.30 | 0.93 | 0.10 |
| DOC | mg/L | 0.5 | 0.67 | 0.12 | 0.60 | 0.11 |
| **CMFS FILTRATE** | | | | | | |
| Alkalinity (as $CaCO_3$) | mg/L | 2 | 115.2 | 6.3 | 108.9 | 4.1 |
| Calcium | mg/L | 25 | 406.2 | 32.8 | 393.3 | 21.6 |
| Magnesium | mg/L | 25 | 1,338.4 | 105.0 | 1,256.7 | 90.2 |
| Hardness (as $CaCO_3$) | mg/L | 200 | 6,525.9 | 490.7 | 6,157.1 | 409.4 |
| Sodium | mg/L | 25 | 10,920.3 | 808.7 | 10,448.7 | 737.0 |
| Potassium | mg/L | 25 | 405.0 | 36.6 | 399.3 | 41.0 |
| TOC | mg/l | 0.5 | 0.87 | 0.18 | 0.84 | 0.11 |

[1] Detection limit.

[2] UV 254 is the amount of ultraviolet light, at a wavelength of 254 nanometers, absorbed by organic matter in the sample. It is reported as absorbance per centimeter. It is not typically determined for the CMFS filtrate.

# MF Backwash (Waste) Characterization

The backwash effluent was sampled weekly for TOC and monthly for turbidity to characterize this waste stream. Results are listed in table 8 below.

Table 8.—Microfiltration Backwash Effluent Stream Characterization

| Parameter | Units | DL | MF Testing Phases I–III | | MF Testing Phases IV & V | |
|---|---|---|---|---|---|---|
| | | | Avg. (15 samples) | Std. Dev. | Avg. (15 samples) | Std. Dev. |
| TOC | mg/L | 0.5 | 1.00 | 0.37 | 1.06 | 0.10 |
| Turbidity | NTU | 0.1 | 7.6 | 3.5 | 11.3 | 0.11 |

# Phase B

Similar to Phase A, in Phase B the CMF-S showed a maximum sustainable flux of 34 GFD, and the filtrate water produced was suitable for use for reverse osmosis.

The following discussion and tables summarize the MF unit run conditions and events for Phases B-1 through B-3 from June 2004 to October 2007.

## Summary of Siemens CMF-S Operating Conditions and Events

### *Phase B-1*

The primary goal of Phase B-1 was to evaluate the operation of the new-generation RO membranes using MF pretreatment and powerplant influent as feed water. This provided the opportunity to gain additional operating experience with the MF process at the design parameters developed in Phase A. As such, operating conditions were maintained as much as possible (response to a severe red tide event is a notable exception), and no further optimization occurred. Phase B-2 commenced with a set of new MF modules (Generation “B”; see subsequent discussion regarding versions of MF modules). However, the operation was impacted by integrity failures during this period, although they were less severe than those experienced in Phase A, prior to the use of the Arkal filter. Table 9 provides a listing of “fiber pinning” events, and more details are provided in the subsequent discussion of filtrate quality and integrity. Problematic hollow fibers can be isolated by “pinning,” or placing a pin in each open end of the fiber, isolating the fiber from the system. These integrity failures were attributed to small shell fragments that got past the 130-micron Arkal filter.

In general, MF operation in Phase B-1 confirmed that the Phase A design parameters could be sustained. The notable exception was the onset of a severe algae bloom, commonly referred to as a red tide, in late spring 2005. Under favorable environmental conditions, phytoplankton can grow rapidly and form very dense populations or “blooms.” *Red tide* is a common name for a phenomenon in which blooms of certain algal species, which contain red-brown pigments, cause the water to appear red. This change in feed-water quality required a reduction in the operating flux setpoint to maintain reasonable process stability and cleaning frequency. Note that during MF run #22 the operating flux was reduced from 34 to 24.5 GFD, then to 20.5 GFD.

Table 9.—Details of Each Phase B1 MF Run

Feed-water source: Influent water. Backwash chlorination was 20 mg/L and backwash frequency was approximately 20 minutes in all cases.

| Run # | Dates | Flux (GFD) | Comments |
|---|---|---|---|
| MF 18 | 6/8/04–9/10/04 | 34 | New Generation "B" modules, Set 2<br>Arkal 130 micron |
| MF 19 | 9/10/04–12/10/04 | 34 | Several fibers were pinned 9/20/04 |
| MF 20 | 12/10/04–3/10/05 | 34 | 1/15/05 Two pins in one module<br>2/8/05 Same module replaced due to damage |
| MF 21a | 3/10/05–4/27/05 | 34 | |
| MF 21b | 4/27/05–6/6/05 | 34 | New MF pilot unit installed 4/27/05.<br>Continued operation with previous membrane set |
| MF 22 | 6/6/05–7/18/05 | 24.5, 20.5 | Severe red tide event in late May–early June |

### *Phase B-2*

The Phase B-2 MF operation was defined by the shift of feed-water source from powerplant influent to the warmer post-condenser effluent (table 10). However, throughout the summer of 2005 severe algae bloom events recurred. This period was marked by operation at reduced flux. After the algae bloom events subsided, operating flux was increased. During this phase of operation, the replaceable discs in the Arkal pre-filter were changed from 130 microns to 100 microns and subsequently to 40 microns, in an effort to eliminate the fiber damage from shell fragments. Operation with the 40-micron discs was problematic due to the dramatic reduction in throughput and the increased plugging rate of the Arkal filter. Therefore, the discs were changed to the 70-micron size toward the end of Phase B-2, which was maintained though the balance of Phase B testing. The MF membrane in Phase B-2 was affected by severe fouling following installation of a set of new Generation "C" membranes (see table 10), that was not recoverable by CIP. Further discussion is provided below in the MF permeability section.

When the powerplant was operating, the post-condenser effluent stream, which fed the MF process in Phase B-2, was warmer than the influent stream. The El Segundo Powerplant is a peaking facility and as such does not operate continuously. Therefore, there were periods when the effluent temperature was similar to the influent temperature. Figure 22 provides a representation of the temperature variation during a sample period of three months.

Table 10.—Details of Each Phase B2 MF Run

Feed-water source: Post-condenser effluent. Backwash chlorination was 20 mg/L and backwash frequency was approximately 20 minutes in all cases.

| Run # | Dates | Flux (GFD) | Comments |
|---|---|---|---|
| MF 22 | 7/18/05–9/5/05 | 20.5 | Reduced flux during algae bloom |
| MF 23 | 9/6/05–9/16/05 | 34 | |
| MF 24 | 9/18/05–9/23/05 | 27, 34 | 9/23, all modules replaced due to fiber integrity issues. Generation "B", Set 3 |
| MF 25 | 9/26/05–10/1/05 | 34, 27 | Prefiltration tightened from 130 to 100 micron Arkal disc filters prior to run |
| MF 26 | 10/19/05–11/23/05 | 27, 32 | 11/30, prefiltration tightened from 100 to 40 micron Arkal disc filters |
| MF 27 | 12/9/05–12/31/05 | 31–32 | Generation "C," Set 1, of modules installed |
| MF 28 | 1/5/06–1/27/06 | 34 | Irreversible fouling of MF modules on 1/26 |
| MF 29 | 1/31/06–3/6/06 | 34, 19 | Fouling problems continued – 2/10, operation reverted to influent source until June, due to effluent feed pump failure |
| MF 30 | N/A | N/A | Fouling problems continued |
| MF 31 | 4/1/06–4/15/06 | 28 | Set 2 of Generation "C" modules installed due to fouling issues.<br>40-micron size proved too tight to allow for sufficient feed flow to the MF unit, so 70-micron disks were installed 4/17/06 |
| MF 32 | 4/29/06–6/8/06 | 28 | |

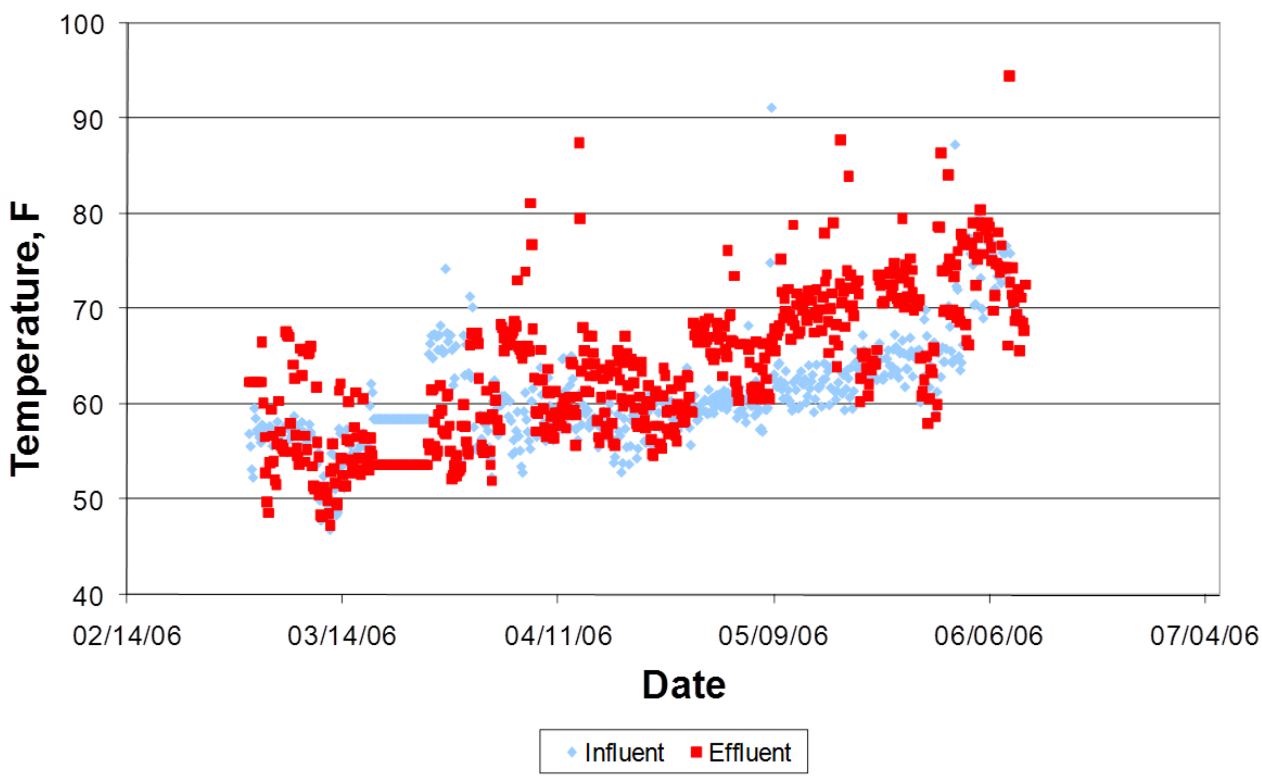



**Figure 22.—Influent and effluent water temperature comparison.**

### *Phase B-3*

With regard to MF operation, Phase B-3 was a continuation of B-2 testing, demonstrating performance of the MF on the warm, post-condenser effluent water. The effluent pump operation was restored prior to the start of MF Run #33. Following correction of backwash chemical dosing, MF Run #33 served as final confirmation of performance at the optimized conditions (30–34 GFD). This performance confirmed that the MF can operate for a 3- to 4-week period before a CIP is required, in the absence of severe algae bloom conditions.

Table 11.—Details of Each Phase B3 MF Run

Feed-water source: Post-condenser effluent. Backwash frequency was approximately 20 minutes in all cases.

| Run # | Dates | Flux (GFD) | Backwash Chlorination (mg/L) | Comments |
|---|---|---|---|---|
| MF 33 | 6/9/06–9/20/06 | 30–34 | 50 for 2 weeks, then 20 in every backwash tank | Very long and stable run, although MF membrane integrity issues developed. |
| MF 34 | 10/1/06–10/9/06 | 32 | 20 in every backwash tank | New modules installed, Generation "C", Set 3. Run stopped short due to equipment relocation |

Over the course of testing, three generations of CMF-S microfiltration membrane modules were tested. Table 12 summarizes the characteristics of each generation. Membrane material remained PVDF and nominal pore size remained 0.1 micron for each generation.

Table 12.—Summary of Siemens CMF-S Modules Tested

| Parameter | Generation A | Generation B | Generation C |
|---|---|---|---|
| Fiber outside diameter, µm | 650 | 800 | 1000 |
| Fiber inside diameter, µm | 390 | 500 | 530 |
| Approximate # of fibers per module | 14,500 | 9,600 | 7,400 |
| Surface area per module, sq. ft. | 335 | 272 | 262 |
| Achievable flux, GFD | 24 | 34 | 34 |
| Permeate flow per module, gpd | 8,040 | 9,248 | 8,908 |

## MF Permeability

Figures 23 through 26 show the performance over the course of June 2004 to October 2006 (Phases B-1 & B-2).

Figure 23 shows that, using influent for feed water, the operating flux of 34 GFD was sustainable for the goal period of 21 days before a CIP was required several times over the course of a year. These results confirmed the Phase A optimized operating parameters for influent operation.

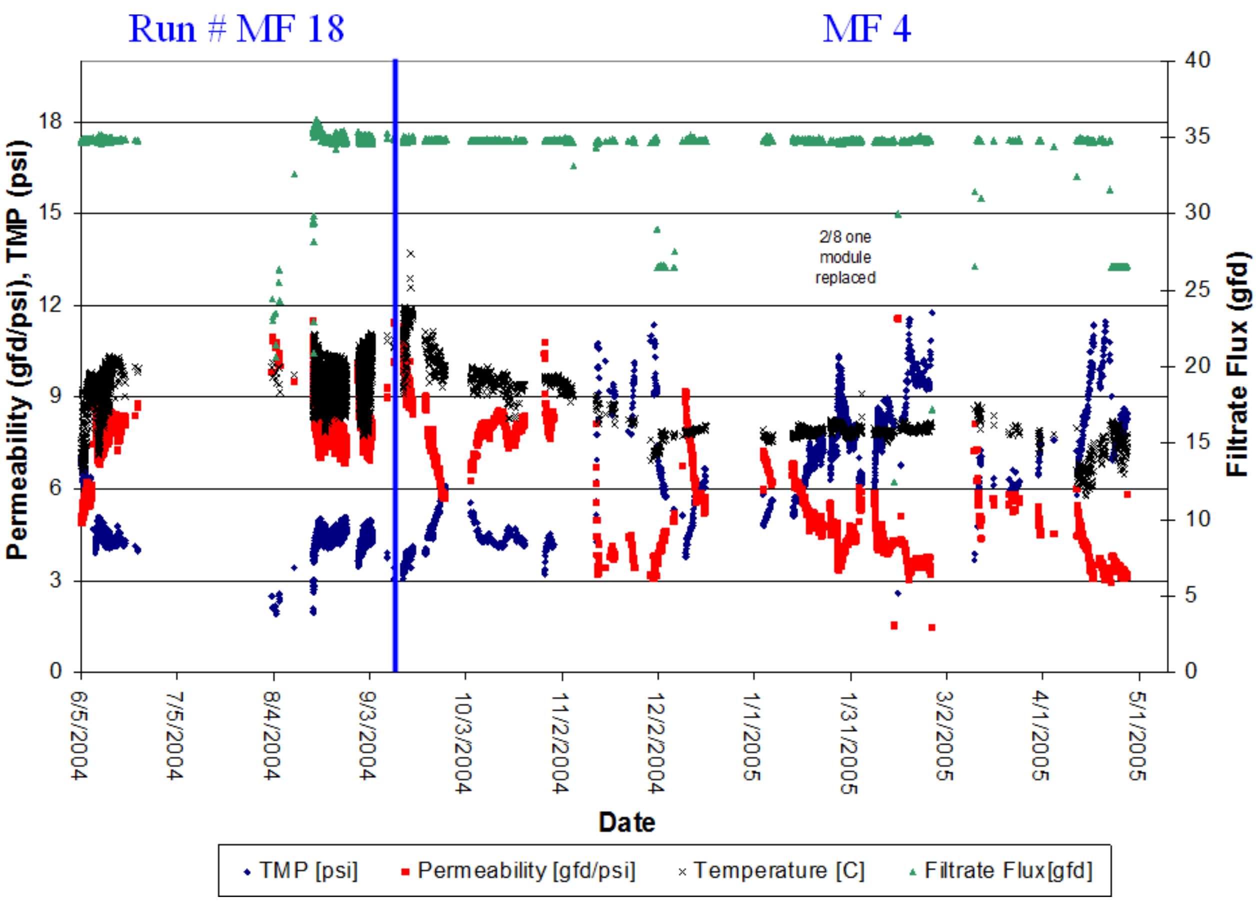



**Figure 23.—CMF-S performance June 2004 – May 2005.**

In late May 2005, a severe algae bloom (red tide) began. As seen in figure 24, the flux rate of the MF unit was reduced in order to maintain operation of the unit. The MF unit was able to operate during this event at a reduced flux rate of approximately 20 GFD, approximately 30 percent less than previous operating flux rates. As the algae bloom conditions subsided in August, the flux rate could be increased back to previous values. During this period of testing, the feed-water source was switched to the warmer powerplant effluent in July.

Figure 25 shows the performance of the MF unit from October 2005 to May 2006 (balance of Phase B-2), with feed water continuing to come from powerplant effluent. The MF unit experienced integrity issues during this time period, and several different Arkal prescreen filter disc sizes were tried in an effort to keep shell particles and other debris from damaging the membrane fibers. MF flux rates varied from 19 to 34 GFD during this period, and one episode of irreversible fouling occurred in January to February, necessitating a reduction of the operating flux to 19 GFD. The irreversible aspect of this fouling event was unique in the entire Phase A and B operation. The operating personnel reported the water in the MF basin had an unusual yellow color during this period. A post-mortem analysis of an MF module by the membrane manufacturer indicated the presence of organic and biological matter on the membrane surface, but it was not able to provide a more specific cause of the permeability loss. Lab scale cleaning trials on the fibers indicated the best recovery when cleaning was performed with 0.5 percent sodium percarbonate (40 °C) followed by 0.05-percent $H_2SO_4$ (40 °C). This information was retained for implementation at the pilot plant, should a similar event occur. Operational issues with the effluent feed pump resulted in reverting back to influent water from February 10th until early June.

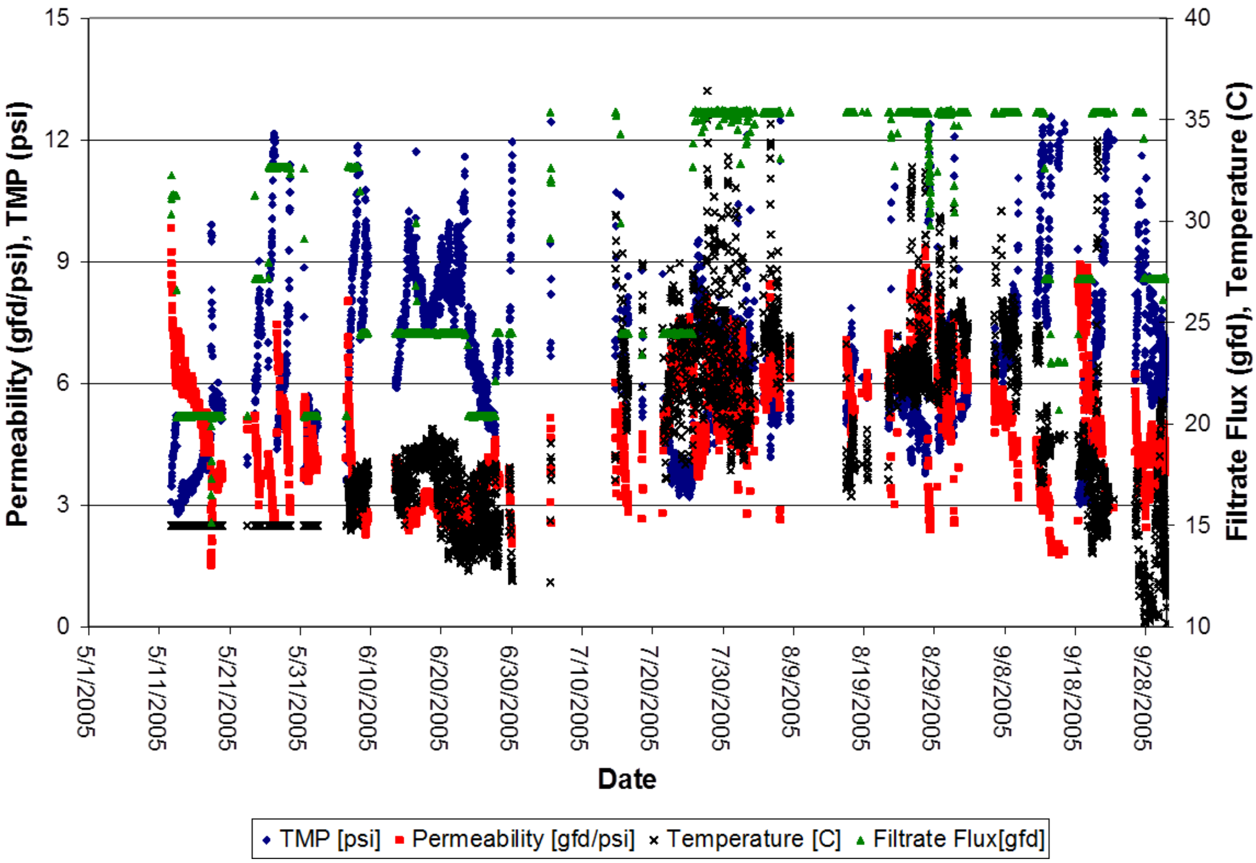



**Figure 24.—CMF-S performance May 2005 – September 2005.**

Figure 26 displays the last interval of the run time for the CMF-S unit. The unit experienced a very long run time during this period, with a flux range of 32–34 GFD. During run #33 the CMF-S maintained 30 GFD for 2 months without requiring a CIP. Subsequently, in late August 2006, the flux rate was increased to 34 GFD and the unit maintained this for another month, again, without requiring a CIP.

Integrity issues developed toward the end of run 33, but this can be attributed to an operational error with the Arkal prescreening system, which allowed raw ocean water containing shell fragments and other debris into the membrane tank. This shows the importance of proper prescreening prior to the MF system.

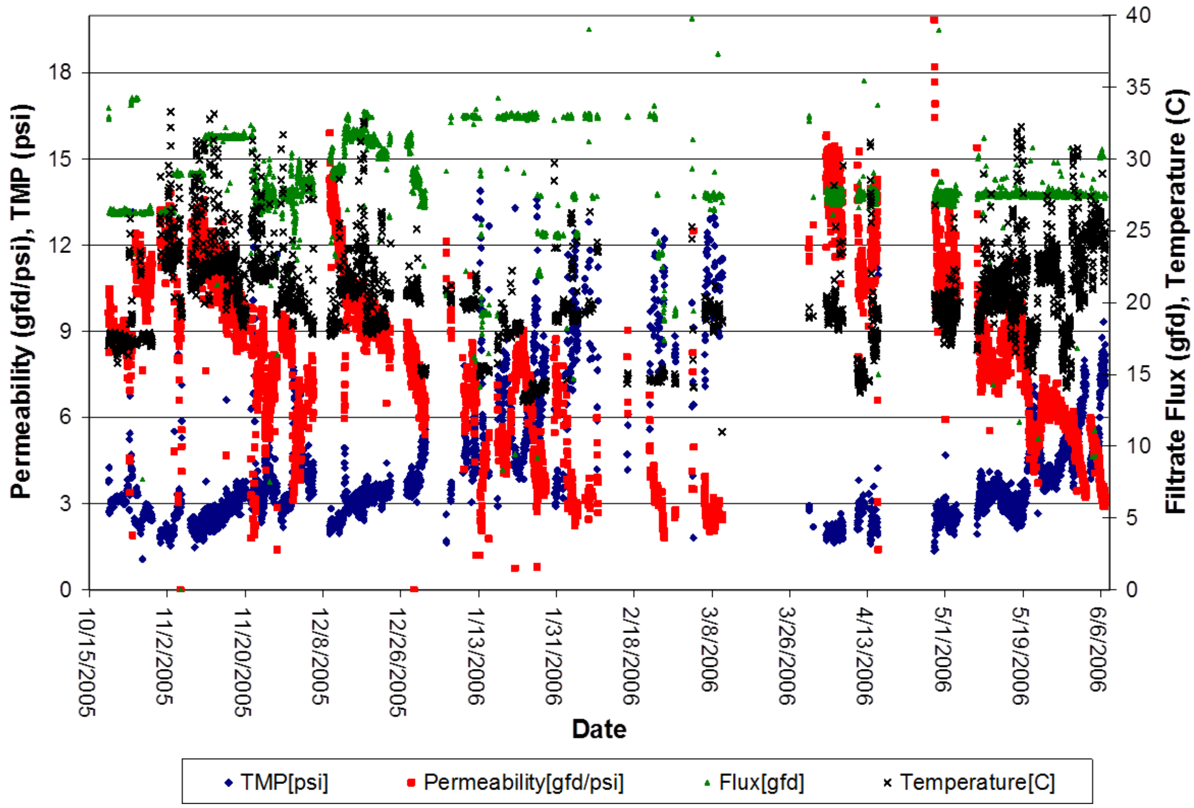



**Figure 25.—CMF-S performance October 2005 – May 2006.**

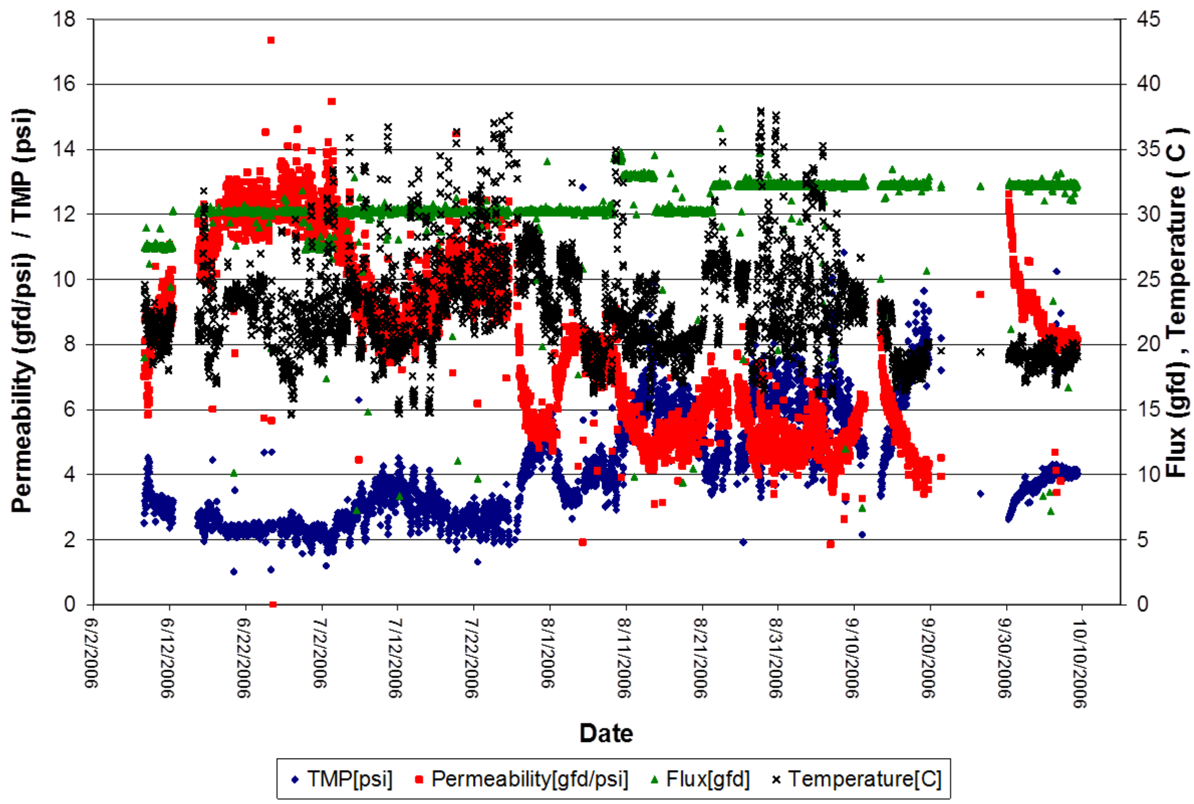



**Figure 26.—CMF-S performance June 2006 – October 2006.**

## MF Filtrate Quality – Phase B

### *Turbidity*

The MF filtrate in general had turbidity less than 0.1 NTU, as seen in figure 27. The values greater than 0.1 can generally be attributed to fiber breakage.

Figure 28 shows the integrity of the MF fibers during Phase B with various grades of prescreening. Note that the major fiber integrity issues in August of 2005 coincide with the highest turbidity values as depicted in figure 27.

Figure 29 shows the same PDT values plotted with the CMF-S filtrate SDI values. In general, the SDI values for the CMF-S system were below 3, with only three measurements in the 4 to 5 range during this period of testing.

Tables 13 and 14 show detailed water quality of both the CMF-S feed and filtrate water, respectively. As in the Phase A testing, the CMF-S system demonstrated approximately 10 percent removal of TOC, and no removal of inorganic constituents.

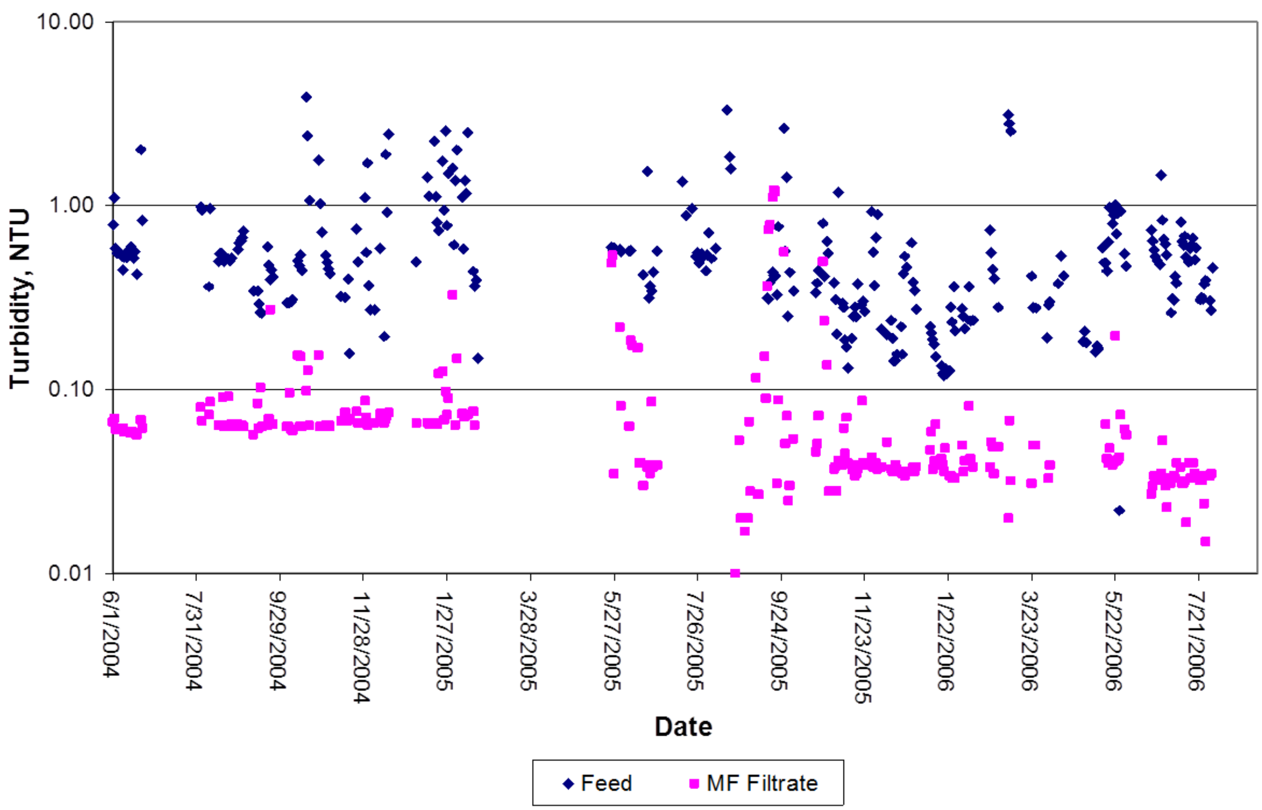



**Figure 27.—Phase B Siemens CMF-S turbidity.**

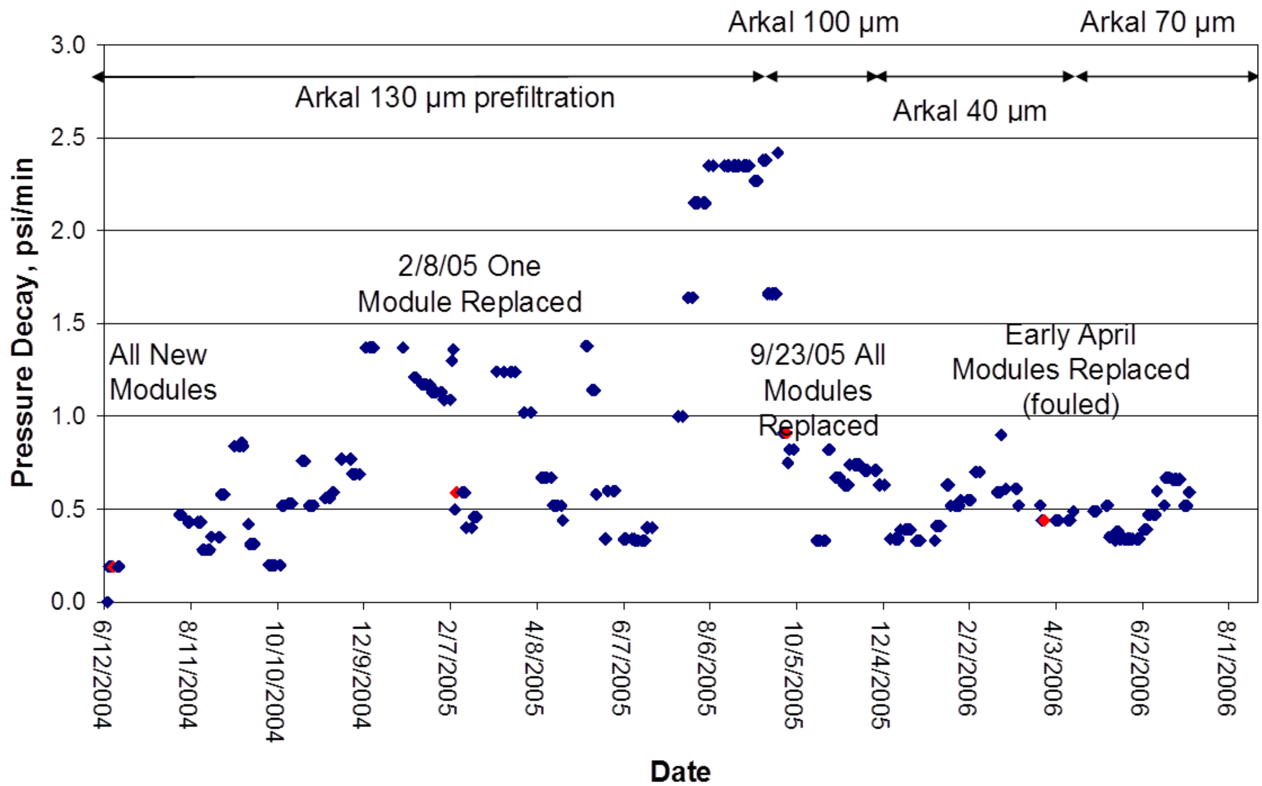



**Figure 28.—Phase B CMF-S pressure decay test results.**

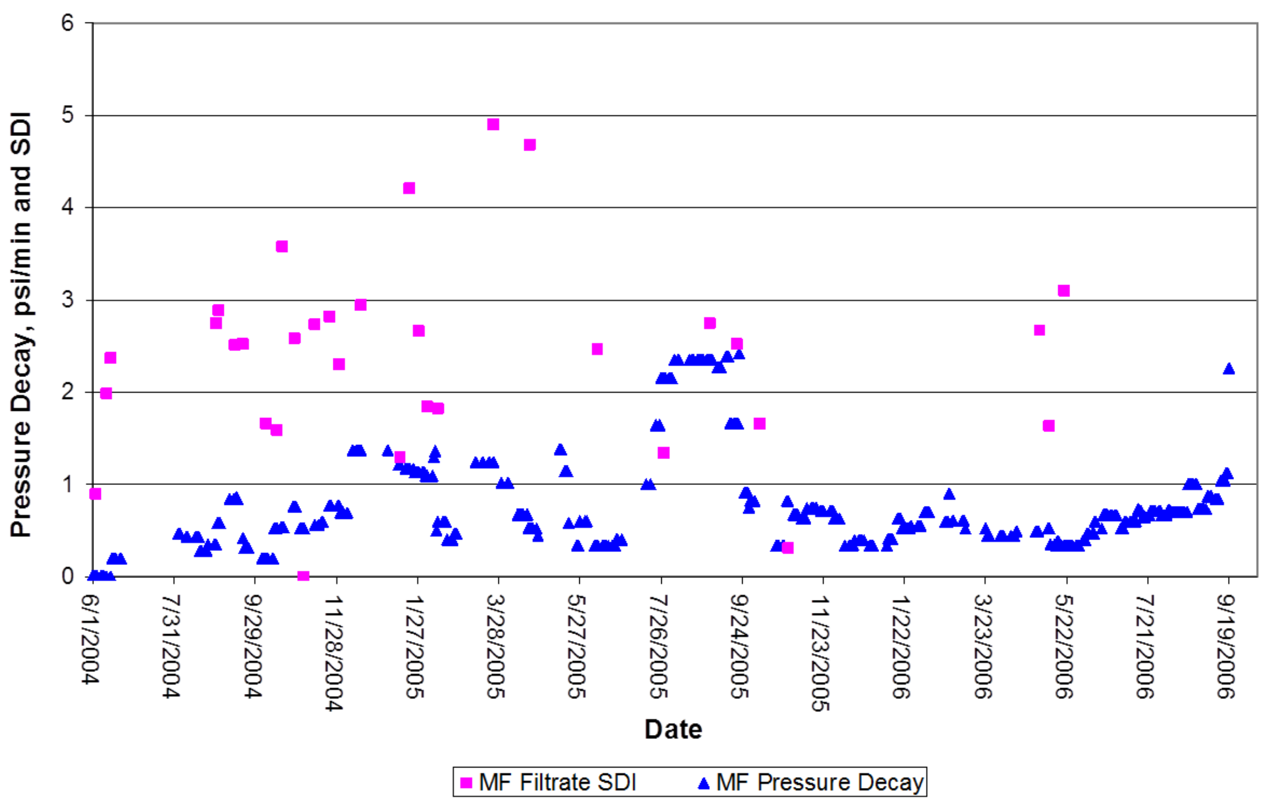



**Figure 29.—Phase B CMF-S PDT and SDI values.**

Table 13.—CMF-S Feed-water Quality, January 2005 – October 2006

| CMF-S Feed | | | Phase B1 | | Phase B2 | | Phase B3 | |
|---|---|---|---|---|---|---|---|---|
| **Parameter** | **Units** | **DL** | **Average** | **Std. Dev.** | **Average** | **Std. Dev.** | **Average** | **Std. Dev** |
| UV 254 | abs/cm | 0.01 | 0.013 | 0.003 | 0.016 | 0.007 | 0.014 | 0.004 |
| Alkalinity (as $CaCO_3$) | mg/L | 2 | 113 | 4.5 | 113 | 4.0 | 113 | 1.2 |
| Calcium | mg/L | 25 | 386 | 18 | 377 | 25 | 387 | 14 |
| Magnesium | mg/L | 25 | 1,245 | 52 | 1,254 | 95 | 1,190 | 65 |
| Hardness (as $CaCO_3$) | mg/L | 200 | 6,089 | 248 | 6,105 | 441 | 5,866 | 294 |
| Sodium | mg/L | 25 | 10,237 | 414 | 10,422 | 716 | 9,830 | 602 |
| Potassium | mg/L | 25 | 372 | 17 | 390 | 32 | 373 | 17 |
| TOC | mg/L | 0.5 | 0.99 | 0.24 | 0.93 | 0.20 | 0.85 | 0.13 |
| DOC | mg/L | 0.5 | 0.65 | 0.12 | 0.63 | 0.12 | 0.70 | 0.07 |

Table 14.—CMF-S Filtrate Water Quality, January 2005 – October 2006

| CMF-S Filtrate | | | Phase B1 | | Phase B2 | | Phase B3 | |
|---|---|---|---|---|---|---|---|---|
| **Parameter** | **Units** | **DL** | **Average** | **Std. Dev.** | **Average** | **Std. Dev.** | **Average** | **Std. Dev.** |
| UV 254 | abs/cm | 0.01 | Typically ND | | Typically ND | | Typically ND | |
| Alkalinity (as $CaCO_3$) | mg/L | 2 | 113 | 4.9 | 113 | 3.9 | 113 | 1.1 |
| Calcium | mg/L | 25 | 386 | 24 | 378 | 25 | 390 | 17 |
| Magnesium | mg/L | 25 | 1,249 | 66 | 1,264 | 94 | 1,203 | 80 |
| Hardness (as $CaCO_3$) | mg/L | 200 | 6,108 | 325 | 6,147 | 432 | 5,930 | 367 |
| Sodium | mg/L | 25 | 10,303 | 508 | 10,509 | 683 | 9,941 | 675 |
| Potassium | mg/L | 25 | 373 | 23 | 390 | 28 | 377 | 20 |
| TOC | mg/L | 0.5 | 0.85 | 0.15 | 0.87 | 0.16 | 0.76 | 0.18 |

## MF Summary

The testing for the Siemens CMF-S system is complete after a total of approximately four years of testing. The sustainable flux rate and filtrate water quality were found to be similar using either powerplant influent or post-condenser effluent as the water source. The optimum flux was determined to be 34 GFD for both water sources, and the filtrate quality was consistently acceptable as feed to the reverse osmosis units. Fiber damage did occur during Phase B testing, and a pre-filter rating of 70 microns or less was found to be effective at preventing damage. The optimized CMF-S operating parameters are included in table 15.

Table 15.—Optimized CMF-S Parameters

| Parameter | Value |
|---|---|
| Filtrate flux | 34 GFD |
| Filtration time between backwashes | 20 minutes |
| Recovery | 93% |
| Backwash parameters | |
| Air scour rate | 7 SCFM/module |
| Air scour duration | 30 seconds |
| Backpulse rate | 9.9 gpm/module |
| Air scour + backpulse duration | 15 seconds |
| Refill duration | ~35 seconds |
| Backwash chlorination | 20 mg/L |

Chlorination of the backwash was found to be vital to maintain the performance achieved.

At the end of Phase B1 and into Phase B2, a severe algae bloom (red tide) event occurred that required the operating flux to be reduced by approximately 30 percent in order to maintain stable operation and a reasonable period between chemical cleanings.

Three generations of MF modules were tested during Phases A and B. The most recent module, Generation C, had the thickest fiber and lowest surface area of all the modules tested, but was least affected by fiber breakage issues. The one fiber breakage incident that did occur with the Generation C modules was believed to be the result of an operational error with the Arkal prescreening unit. The generation C module with the 70-µm Arkal prefilter demonstrated acceptable integrity and would be suitable for full-scale design consideration.

A successful CIP protocol was found to be:

- 2 percent citric acid recirculation/aeration at 36–38 °C followed by
- 400 to 600 mg/L NaOCL recirculation at 20–22 °C

# ZENON ZW1000 ULTRAFILTRATION MEMBRANE SYSTEM PERFORMANCE

Phase B-2 included the addition of a Zenon ultrafiltration (UF) system to the site in May of 2005. The unit was operated on both powerplant influent (Phase B-2) and effluent (Phase B-3) with various operating strategies.

**Figure 30.—Zenon ZW1000 ultrafiltration pilot system.**

Early operation of the UF in 2005 and 2006 achieved a maximum sustainable flux rate of only 16–18 GFD. During this period chlorine was only introduced once or twice a day in the form of a maintenance clean, as outlined in tables 16 and 17. Zenon reported that this operating scheme had been successfully applied with higher operating fluxes at other ocean water locations, but our testing did not confirm this. Unfortunately, the commissioning of the UF pilot coincided with the severe algae bloom of 2005. While there was some concern that the early fouling events associated with the algae bloom may have permanently affected the membrane performance, replacement membranes performed similarly to the first set.

Ultimately a maximum sustainable flux rate of 27.5 GFD was achieved during the final period of testing (summer of 2007) using a chlorinated backwash operating strategy. This operating strategy included more frequent dosing of chlorine to inhibit and remove foulants of the membrane compared to earlier runs.

Tables 16 and 17 summarize the UF unit run conditions during the Phase B testing period.

## UF Permeability

Like the Siemens CMF-S system (and the RO), the UF runs at constant flux and thus, as the membrane fouls, the TMP required to maintain throughput rises. However, because transmembrane pressure is also influenced by water temperature and variations in flow, the appropriate method of monitoring membrane fouling is to observe variations in the temperature-corrected permeability or specific flux.

As shown in the summary tables 16 and 17, early testing in 2005 and 2006 of the Zenon unit on both influent and effluent streams resulted in a sustainable flux rate of 16–18 GFD. Figure 31 shows the details of operation between May and September 2005. The Zenon system was brought on line during the first severe red tide event, making it difficult to achieve long run times during the first two months of operation and resulting in a reduction of operating flux. As figure 31 shows, runs #2 and 3 (May 20 through July 27, 2005) consisted of operation at 20 GFD, and operation at this flux rate did not provide the target 21 days of operation before a CIP was required. The flux was therefore lowered to 18 GFD with run #5 starting on September 14, 2005.

Figure 32 details the continued Phase B-2 operation from November 2005 to March 2006. The unit was switched from powerplant influent to powerplant effluent during UF Run #6 on November 23, 2005 (site operational requirements). Noteworthy in the data from UF Run #6 is that the rate of permeability loss is the same before and after the change to effluent water. The flux rate was 18 to 19 GFD in this period.

Table 16.—Summary of Phase B2 UF Runs

| Feed Source | Run | Dates | Flux (GFD) | Backwash Frequency (minutes) | # of NaOCl MCs per day | NaOCl Concen tration (mg/L) | # of Citric Acid MCs per week | Citric acid Concen-tration (g/L) | Comments |
|---|---|---|---|---|---|---|---|---|---|
| Power-plant influent | UF 1 | 4/15/05–5/20/05 | 23.5 | 25 | 3 | 100 | 1 | 0.5 | Unit commissioned in April and May with 500 sq ft ZW1000 modules with a nominal pore size of 0.02 micron. Material is PVDF. |
| | UF 2 | 5/20/05–7/4/05 | 20.1 | 28 | 3 | 100 | 1 | 0.5 | Late May to early June, red tide event started. |
| | UF 3 | 7/4/05–7/20/05 | 20.1–16 | 28 | 1 | 100 | 1 | 0.5 | Red tide required flux reduction to maintain adequate runtime. |
| | UF 4 | 7/27/05–8/8/05 | 18 | 28 | 2 | 100 | 1 | 0.5 | Powerplant operating issues resulted in short run. |
| | UF 5 | 9/14/05–9/26/05 | 18 | 28 | 2 | 100 | 1 | 0.5 | Equipment shut down midway through run 5 for overall pilot upgrades. |
| | UF 6 | 11/7/05–11/23/05 | 18 | 28 | 2 | 100 | 1 | 0.5 | Zenon unit switched to powerplant effluent during this run. |
| Effluent water | UF 6 | 11/23/05–11/30/05 | 18 | 28 | 2 | 100 | 1 | 0.5 | Fiber breakage occurred in mid/late November, later attributed to manufacturer defect. |
| | UF 7 | 12/2/05–12/24/05 | 18 | 28 | 2 | 100 | 1 | 0.5 | New 500-sq.ft. ZW-1000 modules installed. Upgraded Arkal disk filter from 130 to 40 micron. |
| | UF 8 | 1/11/06–2/1/06 | 19 | 28 | 2 | 100 | 1 | 0.5 | CIP study showed heating CIP solutions to 35–40 °C to be more effective. |
| Power-plant influent | UF 9 | 2/1/06–2/28/06 | 19 | 28 | 2 | 100 | 1 | 0.5 | Runs 8–10 did not quite reach 21 day run target. |
| | UF 10 | 3/1/06–3/29/06 | 19 | 28 | 2 | 100 | 1 | 0.5 | |
| | UF 11 | 3/30/06–5/5/06 | 14 | 34 | 1 | 100 | 0* | N/A | Flux reduced to ensure 21 day run time between cleanings. Arkal filters loosened to 100 micron. |
| | UF 12 | 5/10/06–5/31/06 | 14 | 34 | 1 | 100 | 0* | N/A | |

* Citric acid maintenance cleans were discontinued after run 10.

Table 17.—Summary of Phase B3 UF Runs
Feed source is effluent water

| Run | Dates | Flux (GFD) | Backwash Frequency (min) | # of NaOCl MCs per day | NaOCl concentration in MCs (mg/L) | NaOCl used in backwash | NaOCl backwash concentration (mg/L) | Comments |
|---|---|---|---|---|---|---|---|---|
| UF 13 | 6/2/06–8/9/06 | 14 | 34 | 1 | 100 | No | N/A | Effluent supply pump restored. Run lasted more than 60 days with no CIP. |
| UF 14 | 8/10/06–9/25/06 | 14–18 | 34 | 1 | 100 | No | N/A | Flux increased after extended run time at 14 GFD. |
| UF 15 | 9/26/06–10/15/06 | 16 | 34 | 1 | 100 | Yes | Experimental | Experimental hypochlorite dosing in backwash started in addition to the existing daily hypochlorite maintenance clean. |
| **(None)** | 10/16/06–5/9/07 | N/A | N/A | N/A | N/A | N/A | N/A | **Equipment relocation, down for 6 months** |
| UF 16 | 5/10/07–6/19/07 | 20–25 | 22–24 | 1 @ 110 °F | 100 | Yes | 2 mg/L in every backwash tank | New unit with 600-sq. ft. ZW-1000 modules installed, nominal pore size remains 0.02 micron. Break-in run. |
| UF 17 | 6/20/07–7/18/07 | 25–27.5 | 22 | 1 @ 110 °F | 100 | Yes | 2 mg/L in every backwash tank | Increase of flux during this period. |
| UF 18 | 7/23/07–8/15/07 | 27.5 | 22 | 1 @ 110 °F | 100 | Yes | 2 mg/L in every backwash tank | Demonstration of 27.5 GFD sustainable for 21 days. |
| UF 19 | 8/17/07 through Sept. 2007 | 27.5 | 22 | 1 @ 110 °F | 350 | Yes | 4 mg/L in every backwash tank | Increase in chlorine concentrate in both the backwash and maintenance cleans. Very stable run at 27.5 GFD with little increase in TMP for over 30 days. |

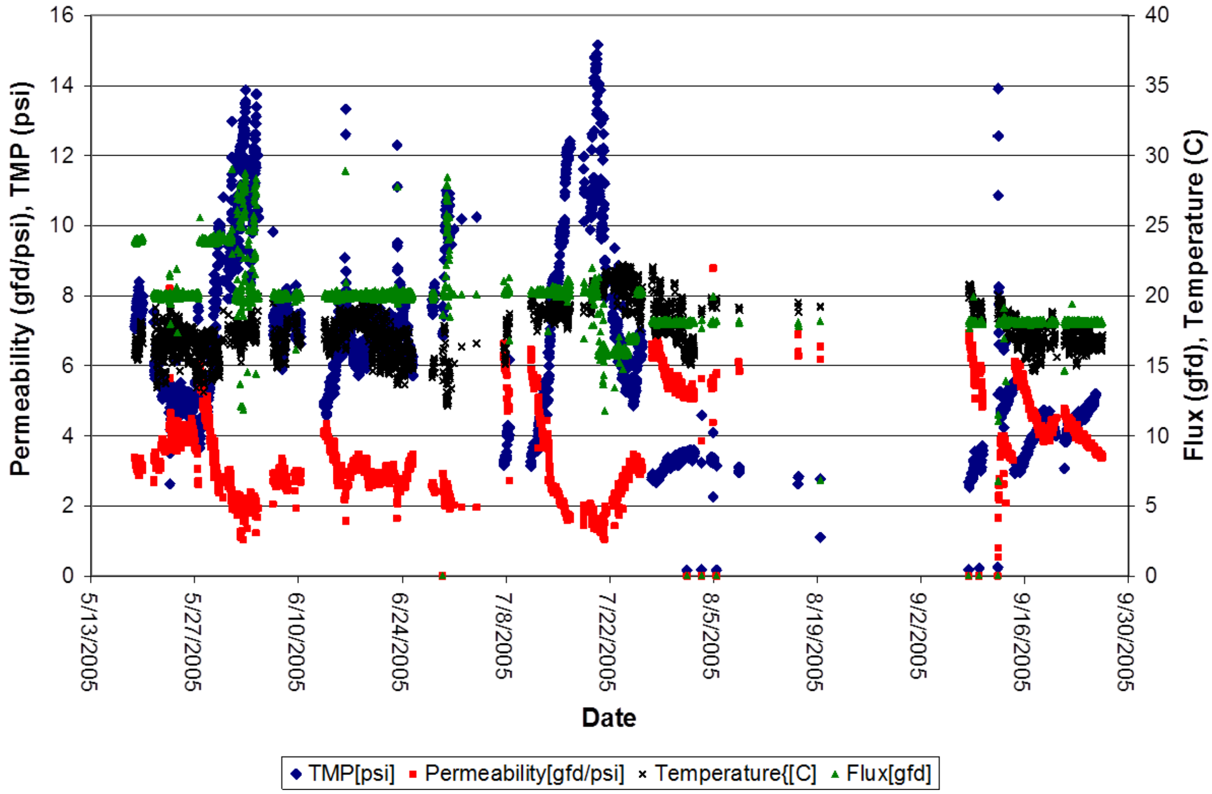



**Figure 31.—Zenon operating performance May 2005 – September 2005.**

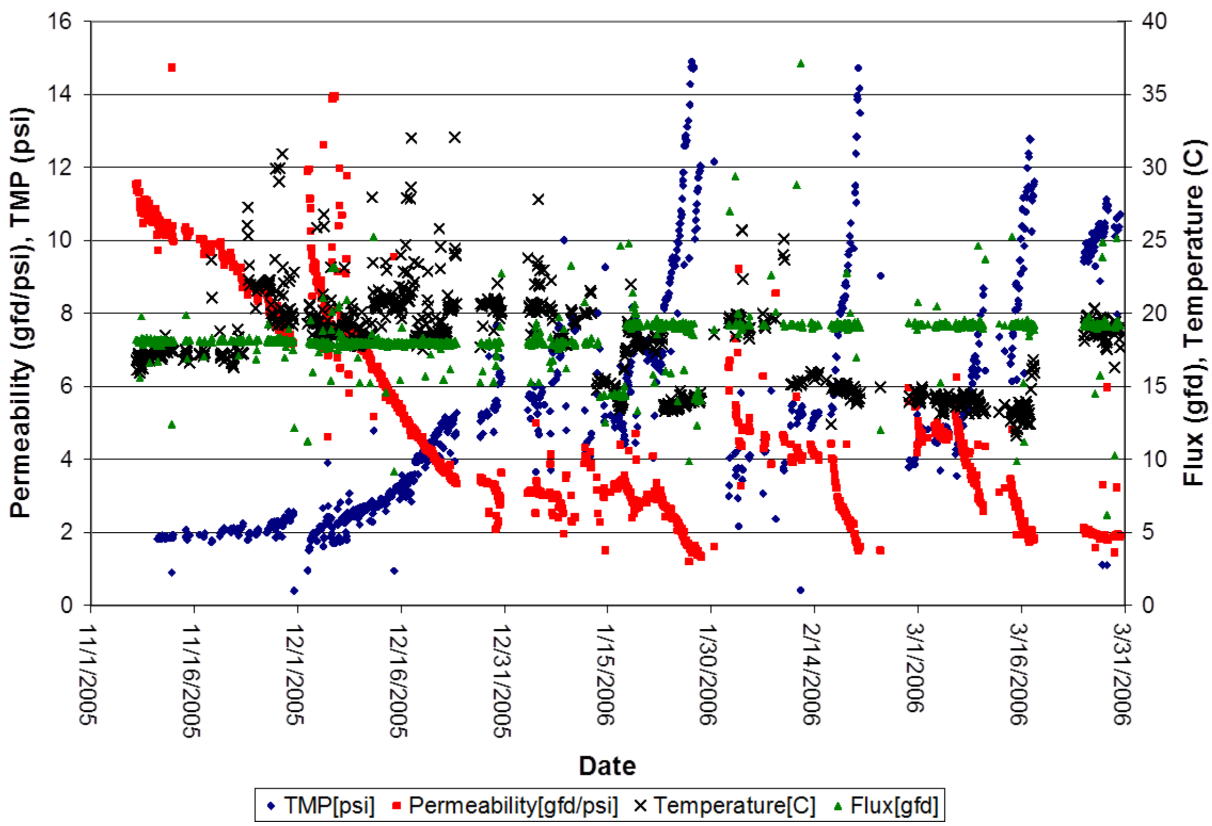



**Figure 32.—Zenon operating performance November 2005 – March 2006.**

Detailed operating performance for April 2006 to October 2006 is shown in figure 33. Flux rate was reduced to 14 GFD for a period of this testing, resulting in extended run times between cleanings. Run #13 exceeded 60 days of run time, indicating a flux of 14 GFD was too low, as the target CIP frequency was 21 days.

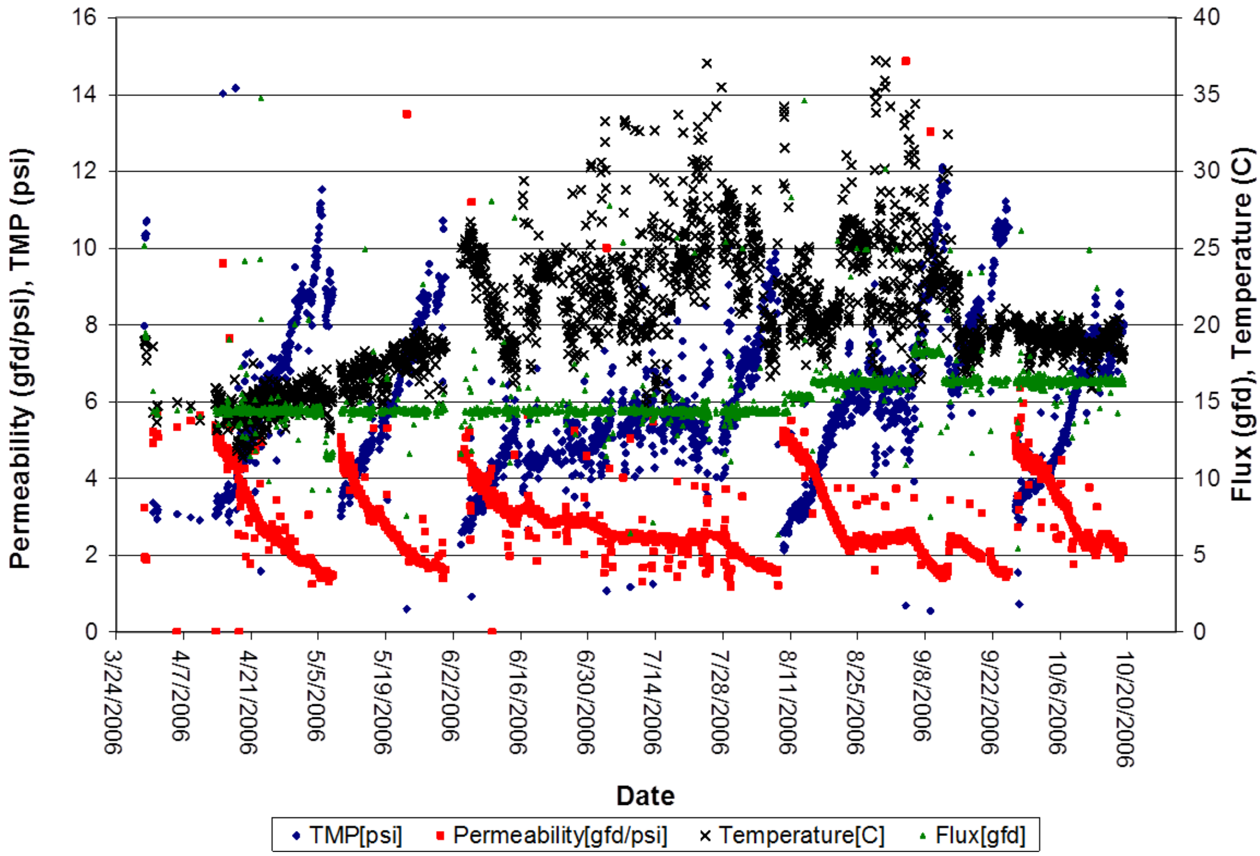



**Figure 33.—Zenon operating performance April 2006 to October 2006.**

In May of 2007, as part of the pilot equipment relocation effort, an upgraded Zenon pilot system was installed at the site. The new unit uses a total of three 600-sq.-ft. ZW-1000 membrane cassettes. The membrane material remains PVDF with a nominal pore size of 0.02 micron. Several changes in operating strategy were implemented with this new round of testing in an effort to bring the flux rate up to a value that was more competitive with the previous Siemens MF system. The most significant changes included the use of chlorine in every backwash in addition to the use of heated, chlorinated maintenance cleans once a day. The Zenon unit was operated on effluent water during this Phase B3. Figure 34 shows the details of this time period and confirms that the changes provided a drastic improvement in performance, as a stable flux rate of 27.5 was achieved.

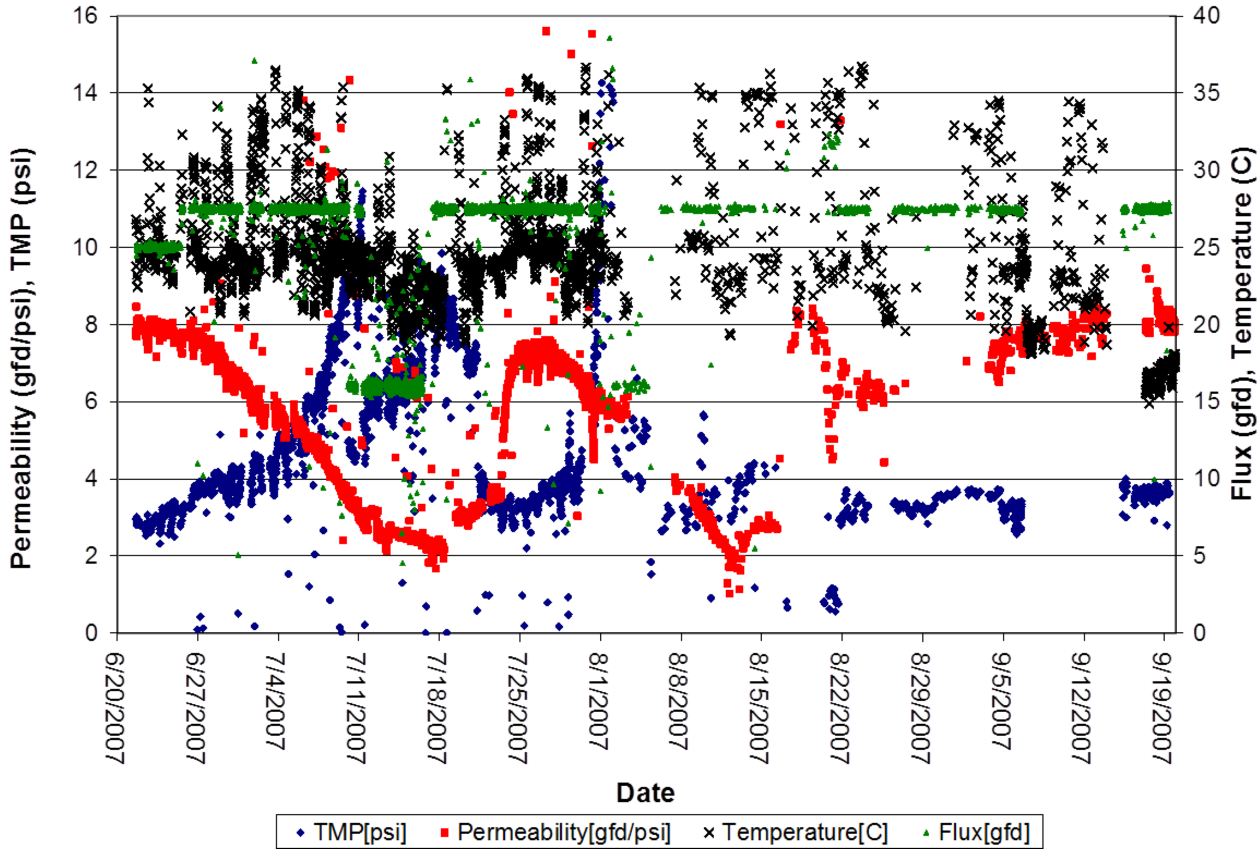



**Figure 34.—Zenon performance June 2007 – September 2007.**

The Zenon unit was restarted in late May 2007, with Run 16 considered a "break-in" period. Runs 17 and 18 were operated under the following conditions, with adjustments to flux rates made periodically:

- Instantaneous flux rate: 25–27.5 GFD
- Recovery: ~93 percent
- Backwash frequency: ~22 minutes
- Backwash type: Chlorinated backwash (2 mg/L in membrane tank) with air scouring
- Daily maintenance clean: 100 mg/L chlorine solution in membrane tank heated to 40 °C, 30-minute soak

During Run #19 starting on August 17, 2007, the hypochlorite concentration in the backwashes was increased from 2 to 4 mg/L, and that in the maintenance cleaning was increased from 100 to 350 mg/L. These increases resulted in much more stable operation, and there was little change in permeability and TMP over approximately 30 days of testing.

## UF Water Quality

Tables 18 and 19 show detailed water quality results for the CMF-S feed and the filtrate, respectively. On average, the Zenon system has also demonstrated approximately 10 percent removal of TOC.

Table 18.—Zenon Feed-water Quality, May 2005 – July 2007

| Zenon ZW 1000 Feed | | | Phase B1 | | Phase B2 | | Phase B3 | |
|---|---|---|---|---|---|---|---|---|
| **Parameter** | **Units** | **DL** | **Average** | **Std. Dev.** | **Average** | **Std. Dev.** | **Average** | **Std. Dev.** |
| UV 254 | abs/cm | 0.01 | 0.015 | 0.005 | 0.014 | 0.005 | 0.018 | 0.009 |
| Alkalinity (as $CaCO_3$) | mg/L | 2 | 115 | 6.1 | 113 | 1.9 | 114 | 1.2 |
| Calcium | mg/L | 25 | 390 | 32 | 377 | 27 | 391 | 24 |
| Magnesium | mg/L | 25 | 1,230 | 64 | 1,263 | 111 | 1,206 | 70 |
| Hardness (as $CaCO_3$) | mg/L | 200 | 6,039 | 329 | 6,142 | 509 | 5,942 | 339 |
| Sodium | mg/L | 25 | 10,124 | 447 | 10,407 | 826 | 9,955 | 652 |
| Potassium | mg/L | 25 | 373 | 24 | 389 | 31 | 377 | 22 |
| TOC | mg/L | 0.5 | 1.04 | 0.22 | 0.94 | 0.24 | 1.43 | 0.85 |
| DOC | mg/L | 0.5 | 0.71 | 0.08 | 0.59 | 0.06 | 0.97 | 0.36 |

Table 19.—Zenon Filtrate Water Quality, May 2005 – July 2007

| Zenon ZW 1000 Filtrate | | | Phase B1 | | Phase B2 | | Phase B3 | |
|---|---|---|---|---|---|---|---|---|
| **Parameter** | **Units** | **DL** | **Average** | **Std. Dev.** | **Average** | **Std. Dev** | **Average** | **Std. Dev.** |
| UV 254 | abs/cm | 0.01 | Typically ND | NA | Typically ND | NA | Typically ND | NA |
| Alkalinity (as $CaCO_3$) | mg/L | 2 | 115 | 6.1 | 113 | 2.1 | 113 | 5.2 |
| Calcium | mg/L | 25 | 391 | 35 | 381 | 22 | 394 | 23.9 |
| Magnesium | mg/L | 25 | 1,234 | 49 | 1,272 | 97 | 1,213 | 72.7 |
| Hardness (as $CaCO_3$) | mg/L | 200 | 6,059 | 279 | 6,191 | 434 | 5,979 | 350.6 |
| Sodium | mg/L | 25 | 10,187 | 385 | 10,514 | 700 | 10,042 | 721.1 |
| Potassium | mg/L | 25 | 379 | 28 | 399 | 32 | 379 | 22.9 |
| TOC | mg/L | 0.5 | 0.95 | 0.11 | 0.86 | 0.17 | 1.3 | 0.83 |

Figure 35 displays the feed and filtrate turbidity of the Zenon UF unit in 2005 and 2006. The feed turbidity during the most recent testing was typically on the order of 1 NTU, with filtrate turbidity typically less than 0.1 NTU. Erratic turbidity values (>0.1 NTU) in May–September 2006 were attributed to inconsistent flow to the turbidity meter.

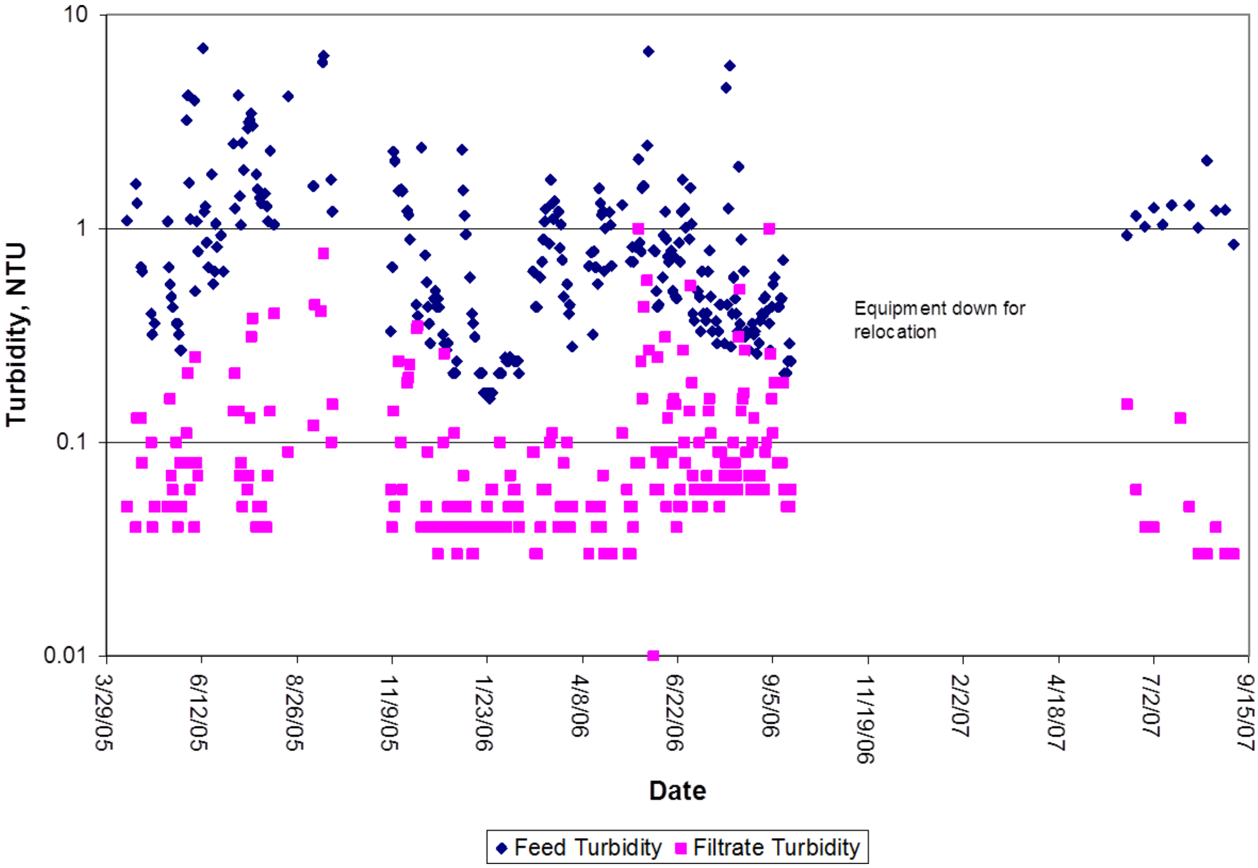



**Figure 35.—Zenon UF turbidity.**

Figure 36 documents the Zenon system membrane integrity from April 2005 to September 2006. The Zenon system had only a single integrity problem: a couple of fibers in one module sheared in half in November 2005. After membrane autopsy, this event was deemed a membrane manufacturing defect, not an operational issue associated with feed-water quality.

The SDI of the UF filtrate (figure 37) shows that it was consistently acceptable as feed to the RO system. Data from June 2007 through September 2007 shows no fiber integrity issues.

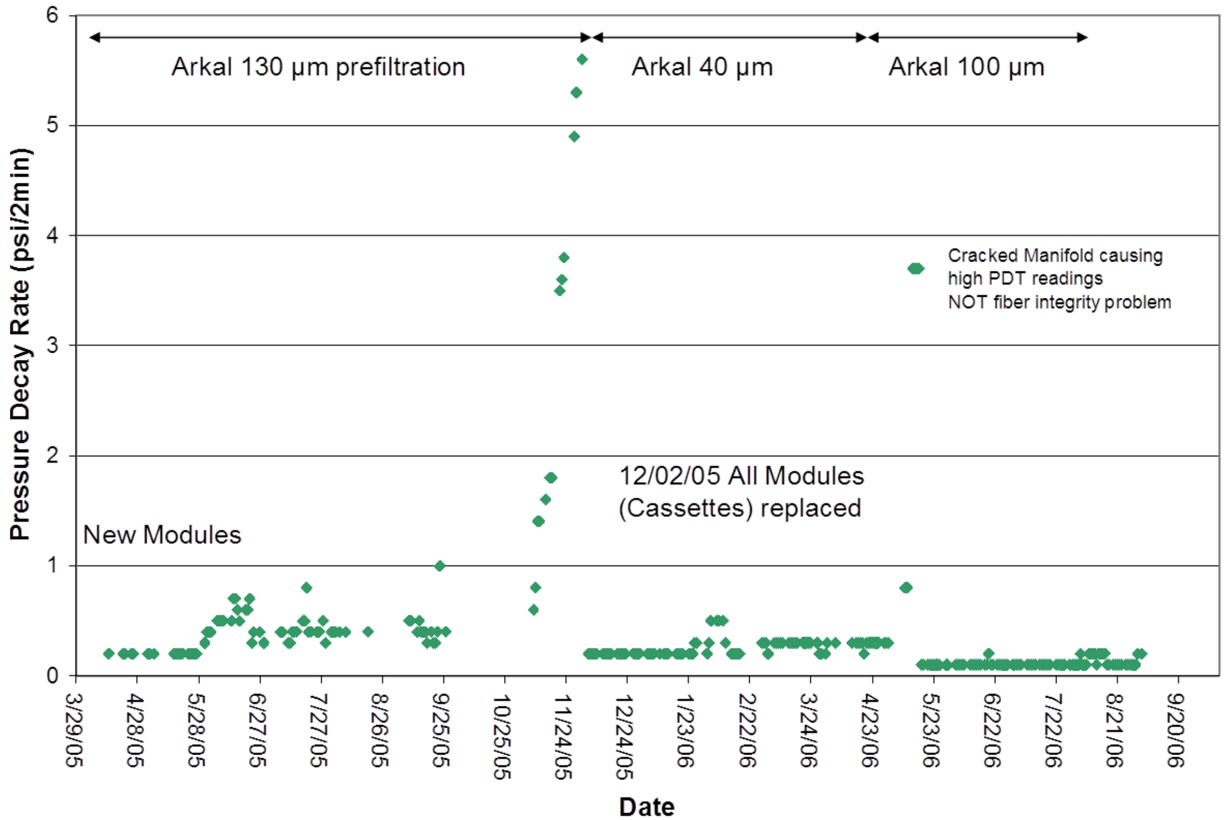



**Figure 36.—Zenon ZW1000 pressure decay test results.**

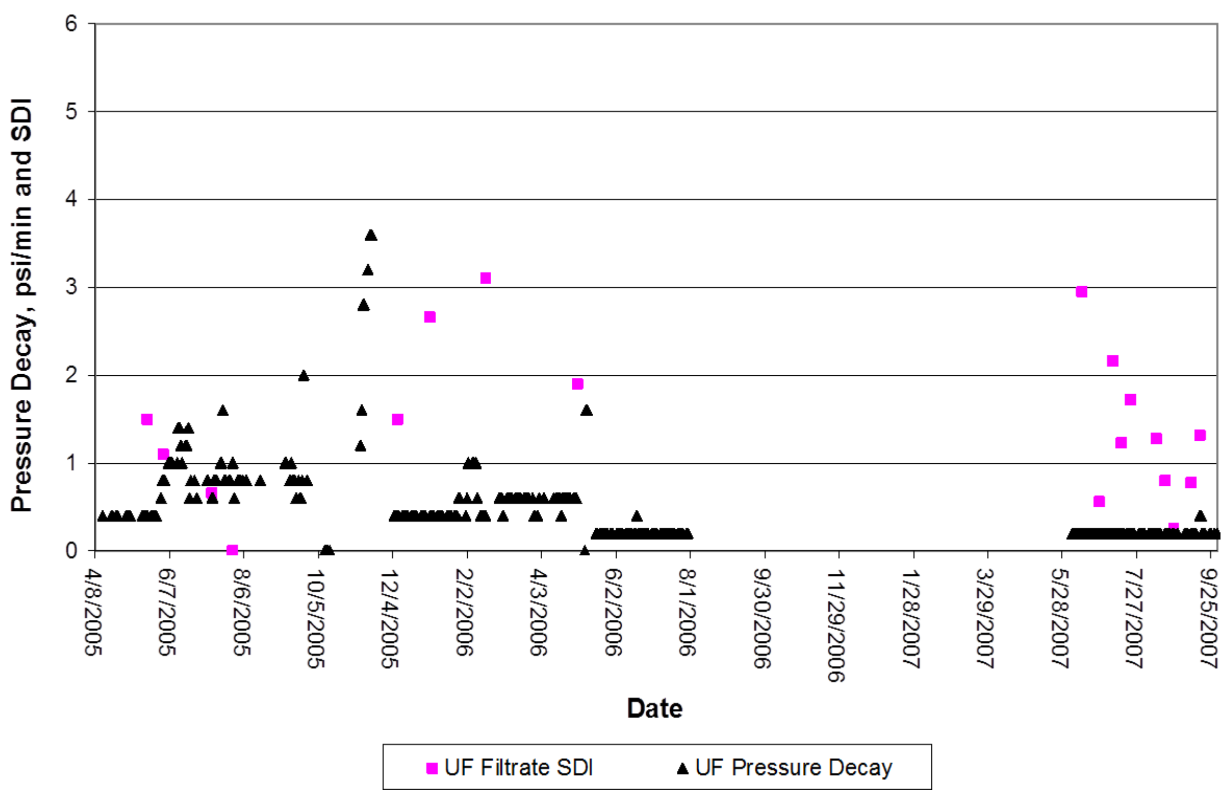



**Figure 37.—Zenon ZW1000 PDT and SDI values.**

## UF Summary

The Zenon ZW1000 system was tested on both powerplant influent and effluent for a period of approximately 2 years. The sustainable flux rate and filtrate water quality were observed to be similar using both the powerplant influent and the post-condenser effluent water sources. The most recent period of testing with the 600-$ft^2$ membrane, from June 2007 to September 2007, produced the most favorable results with respect to sustainable flux rate. The use of chlorination in every backwash combined with a daily heated chlorinated maintenance clean has resulted in a sustainable flux rate of 27.5 GFD. Other successful operational parameters are listed in table 20 below.

Membrane integrity was very good on the Zenon system. The use of a pre-filter rating of 100 microns or less was effective at protecting the UF membrane from damage due to particulates, including shell fragments. UF filtrate quality was excellent throughout the testing period, as indicated by turbidity, filtrate SDI, and ultimately the performance of the downstream RO process.

A successful CIP protocol for the ZW1000 on this water was found to be:

- 2 percent citric acid recirculation/aeration at 40 °C followed by
- 500 mg/L NaOCL recirculation at 40 °C.

Table 20.—Optimized Zenon ZW1000 Operating Parameters

| Parameter | Value |
|---|---|
| Filtrate flux (GFD) | 27.5 |
| Filtration time between backwashes (min) | 22 |
| Recovery | 93% |
| Backwash parameters | |
| Air scour rate (SCFM/module) | 3 |
| Air scour duration (seconds) | 30 |
| Backpulse rate (gpm/module) | 8.7 |
| Backpulse duration (seconds) | 30 |
| Refill duration (seconds) | ~50 |
| Backwash chlorination (mg/L) | 2 |
| Maintenance clean frequency | 1/day |
| Maintenance clean chlorination (mg/L) | 100 |
| Maintenance clean duration (min) | 30 |

# REVERSE OSMOSIS OPTIMIZATION AND PERFORMANCE

## A Note About the RO Membranes

The RO membranes utilized in this study are 4 inches in diameter. These membranes are smaller than the 8-inch-diameter membranes that would be used in a full-scale desalination facility. The reduced scale of the pilot membranes was necessary in order to reduce the flow requirement of the RO system. They are representative smaller versions of the 8-inch membranes and provide equivalent engineering data, except that their salt-rejection capabilities are not equivalent to those of the standard 8-inch products. Therefore, the RO manufacturers were asked to “cherry-pick” their 4-inch-diameter inventory and supply membranes that were representative to their 8-inch counterparts in both flux and rejection properties.

This was true for the Phase A membranes and the Phase B RO membranes discussed below.

**Figure 38.—Reverse osmosis test equipment.**

# Phase A Testing

All Phase A testing of the RO used a microfiltered feed-water source. Phase A of the RO testing can be grouped into the trials listed below in tables 21 and 22.

# RO Trial I Testing

The reverse osmosis unit consists of two independent trains. Each train has two pressure vessels operating in series, with three elements in the lead vessel and the four in the trailing vessel, for a total of seven 4-inch-diameter seawater elements. This configuration simulates a single stage in a full-scale RO system. To prevent precipitation of sparingly soluble salts in the RO system, 3 mg/L of antiscalant is added continuously to the feed water downstream of the RO feed tank.

The original pretreatment process, an attempt to create chloramines in ocean water, damaged the RO membranes in RO trial I. In many MF/RO membrane facilities operating on wastewater, chlorine is added to the feed water to enhance the membrane performance. Ammonia, naturally occurring or added to the wastewater, combines with the chlorine to form chloramines. The intent is to have a combined oxidant that would improve the fouling rate of both the MF and RO processes. This chloramination followed by MF and a subsequent RO process has been used successfully at many wastewater reclamation facilities, including WBMWD's 20-million-gallon-per-day water recycling plant, located 2 miles east of this study's test site. The ammonia reacts with free chlorine or HOCl to form chloramines.

However, two items complicate the formation of chloramine on ocean water. First, ammonia is not present in ocean water and thus must be added. Second, the presence of bromide ($Br^-$) in ocean water interferes with the reactions. The Pacific Ocean water source used in this study has around 64 mg/L of $Br^-$. $Br^-$ substitutes for $Cl^-$ such that the chlorine addition to ocean water actually produces hypobromous acid (HOBr) instead of HOCl. This is discussed further in the "Process and Equipment Challenges" section of this document.

Table 21.—Phase A RO Testing Trials

| RO Testing Trial | Details |
|---|---|
| RO I | Operation with ammonium hydroxide addition pretreatment in an attempt to form chloramines, subsequent sodium bisulfite (SBS) pretreatment – RO membranes oxidized |
| RO II | SBS pretreatment, operation at 8 GFD |
| RO III | SBS pretreatment, operation at 9 GFD |
| RO IV | SBS pretreatment, operation at 11 GFD |

Table 22.—Details of Each Phase A Reverse Osmosis Run

| Trial | Run # | Dates | MF Filtrate Chemical | RO Feed Antiscalant ppm | Hydra-nautics Flux, GFD | Hydra-nautics Recovery | Filmtec Flux, GFD | Filmtec Recovery | Notes |
|---|---|---|---|---|---|---|---|---|---|
| RO I | RO 1 | 7/15/02–9/6/02 | 1 ppm $NH_4OH$ | 3 | 8 | 50 | 8 | 50 | RO membranes show signs of oxidation |
| | RO 2 | 9/1/02–9/28/02 | 1.5 ppm $NH_4OH$ | 3 | 8 | 50 | 8 | 50 | Adjusted $NH_4OH$ dose. RO membranes continue to degrade |
| | RO 3 | 9/29/02–10/23/02 | None | 3 | 8 | 50 | 8 | 50 | Rapid MF fouling |
| | RO 4 | 10/23/02–11/24/02 | 1 ppm SBS | 3 | 8 | 50 | 8 | 50 | Memcor chlorinated backwash oxidizing RO |
| | RO 5 | 11/25/02–12/16/02 | 2–3 ppm SBS | 3 | 8 | 50 | 8 | 50 | Increase SBS |
| | RO 6 | 12/17/02–1/15/03 | 2–3 ppm SBS | 3 | 8 | 50 | 8 | 50 | Both RO pumps repaired, recycle modification |
| RO II | RO 7 | 1/15/03–3/9/03 | 2–3 ppm SBS | 3 | 8 | 50 | 8 | 50 | 1/15 — Replaced both HYD and Dow RO membranes |
| RO III | RO 8 | 3/9/03–4/3/03 | 3 ppm SBS | 3 | 9 | 50 | 9 | 50 | Increased RO Flux |
| | RO 9A | 10/21/03–11/19/03 | 3 ppm SBS | 3 | 9 | 50 | 9 | 50 | Installed RO feed pump VFD |
| | | 11/19/03–1/15/04 | 3 ppm SBS | 3 | 9 | 50 | 9 | 50 | Infrequent operation to MF/feed flow problems. CIP 12/5 |
| | RO 9B | 1/30/04–2/18/04 | 3 ppm SBS | 3 | 9 | 50 | 9 | 50 | |
| RO IV | RO 10 | 2/18/04–6/10 | 3 ppm SBS | 3 | 11 | 50 | 11 | 50 | Increased RO flux |

As depicted in figures 39 and 40, this chlorination, MF, ammonia addition, RO process failed to protect the RO membranes from oxidation. The specific flux and permeate conductivity of the Dow membranes started rising almost immediately. The Hydranautics membranes proved to be more resistant, but after ~100 days of operation it was clear that the salt passage or permeate conductivity of this membrane was rising as well. On September 1, 2002, the $NH_4OH$ addition rate was increased 50 percent to 1.5 mg/L in an effort to ensure that excess ammonia was present and prevent the presence of free chlorine. This did not alleviate the problem, and the permeate conductivity continued to rise. In response to the RO deterioration, on October 3, the continuous chlorination in front of the MF was discontinued. Subsequently, attempts were made to run without any chlorine in the process and rapid MF fouling was observed (MF Trial II). Chlorine in the 20–40 mg/L range was then used in the MF backwash, an intermittent operation. An additional "rinse" step was added to the MF backwash to ensure no chlorine carryover to the RO. This operational scheme, combined with the addition of sodium bisulfite in front of the RO, was used in the remainder of the trials.

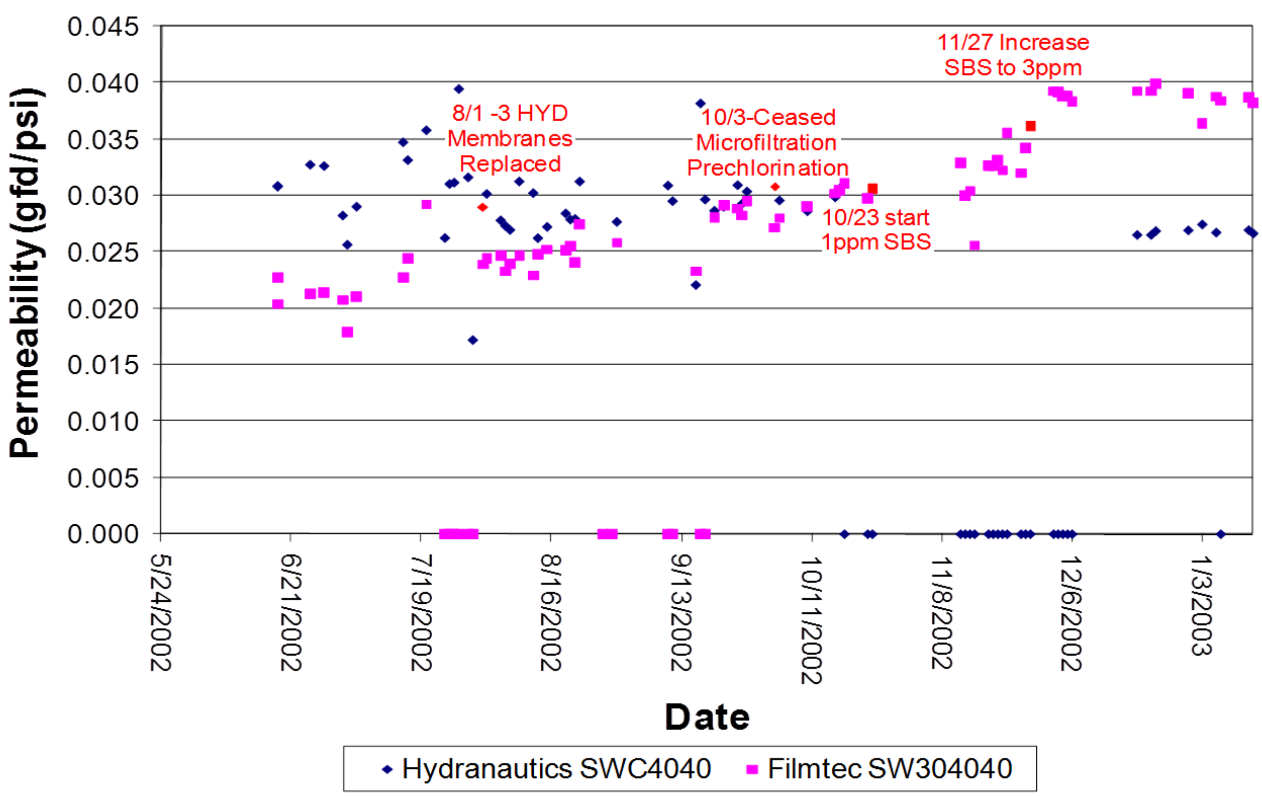



**Figure 39.—Increasing permeability of RO membranes due to oxidation (RO trial I).**

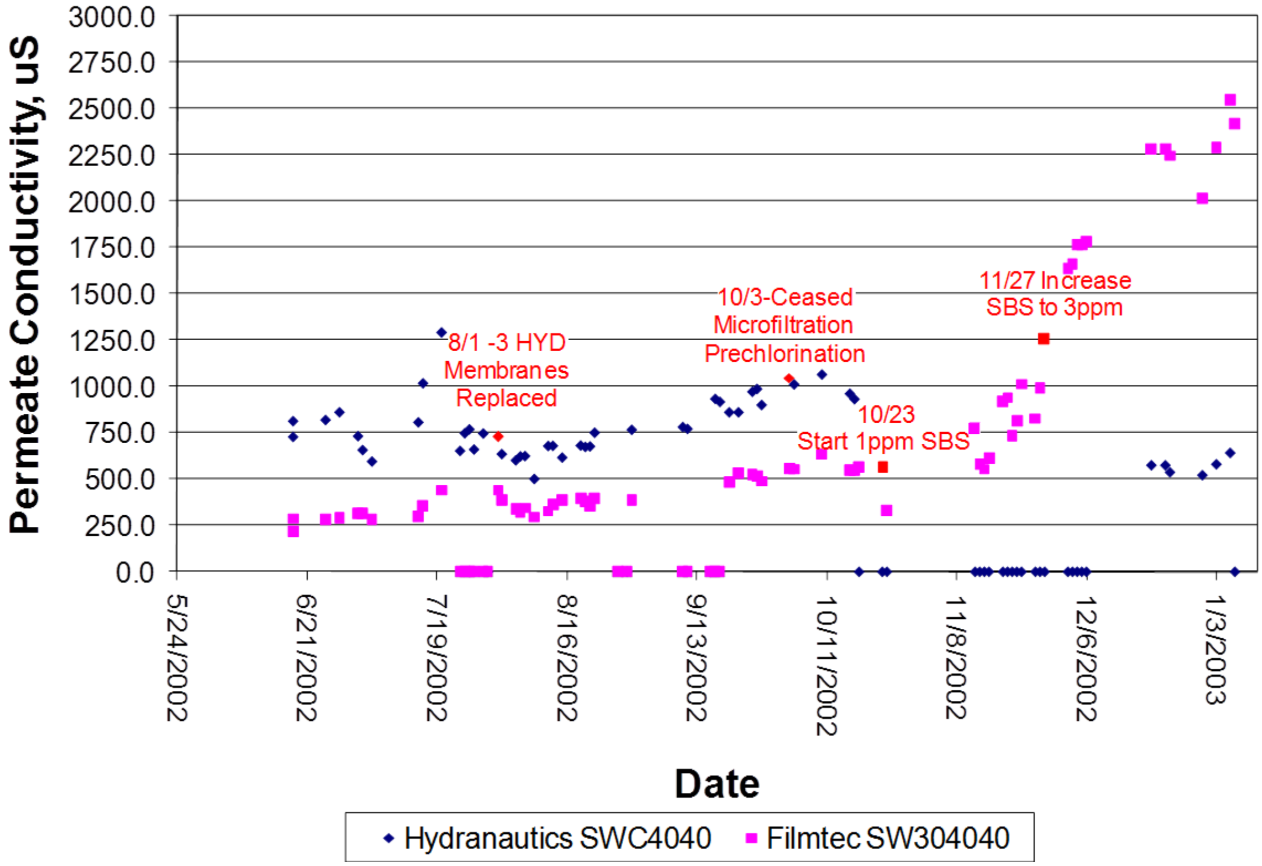



**Figure 40.—Increasing permeate conductivity of RO membranes due to oxidation (RO trial I).**

From October through December 2002, the RO was run with the damaged membranes in an attempt to find a pretreatment strategy that would allow the MF to maintain reasonable flux rates and run times without further RO oxidation. The RO membranes were replaced on January 15, 2003, and trial II of the RO testing commenced on MF filtrate water with 3 mg/L sodium bisulfite protecting the RO. This was continued for the remainder of the trials. Note that the use of sodium bisulfite for reduction of trace free chlorine is distinctly different from the continuous chlorination/dechlorination approach that has been found to result in RO biofouling.

## RO Permeability

Like the MF and UF, the RO system is run at constant flux and thus, if the membrane fouls, the pressure required to maintain throughput rises. The membrane permeability is monitored by the calculation of specific flux, which is the operating flux divided by the temperature-corrected net driving pressure. This way, changes in the membrane properties due to fouling can be observed regardless of changes in the operating conditions (e.g., temperature, flux, etc.).

Figure 41 shows that the permeability of the Hydranautics membrane was fairly stable following the replacement of the RO membranes (RO trial II). Dow membranes, on the other hand, showed a slight increase in specific flux and—as will be discussed in the next section—in permeate conductivity as well. These trends are consistent with membrane oxidation. However, the Hydranautics membranes did not show these signs of oxidation, and these two types of membranes were running side by side on the same feed water. It is possible that small amounts of chlorine (or bromine), not reduced by the sodium bisulfite, reached the RO system, and the Hydranautics membranes may be more resistant to oxidation. Likewise, an examination of figures 39 and 40, above reveals that the Dow membranes also deteriorated much faster than the Hydranautics membranes during the Trial 1 testing, in which the failed chloramination process (chlorine and ammonia added to the feed water) presumably oxidized the membranes. RO Trial III commenced in March 2003 operating at 9 GFD.

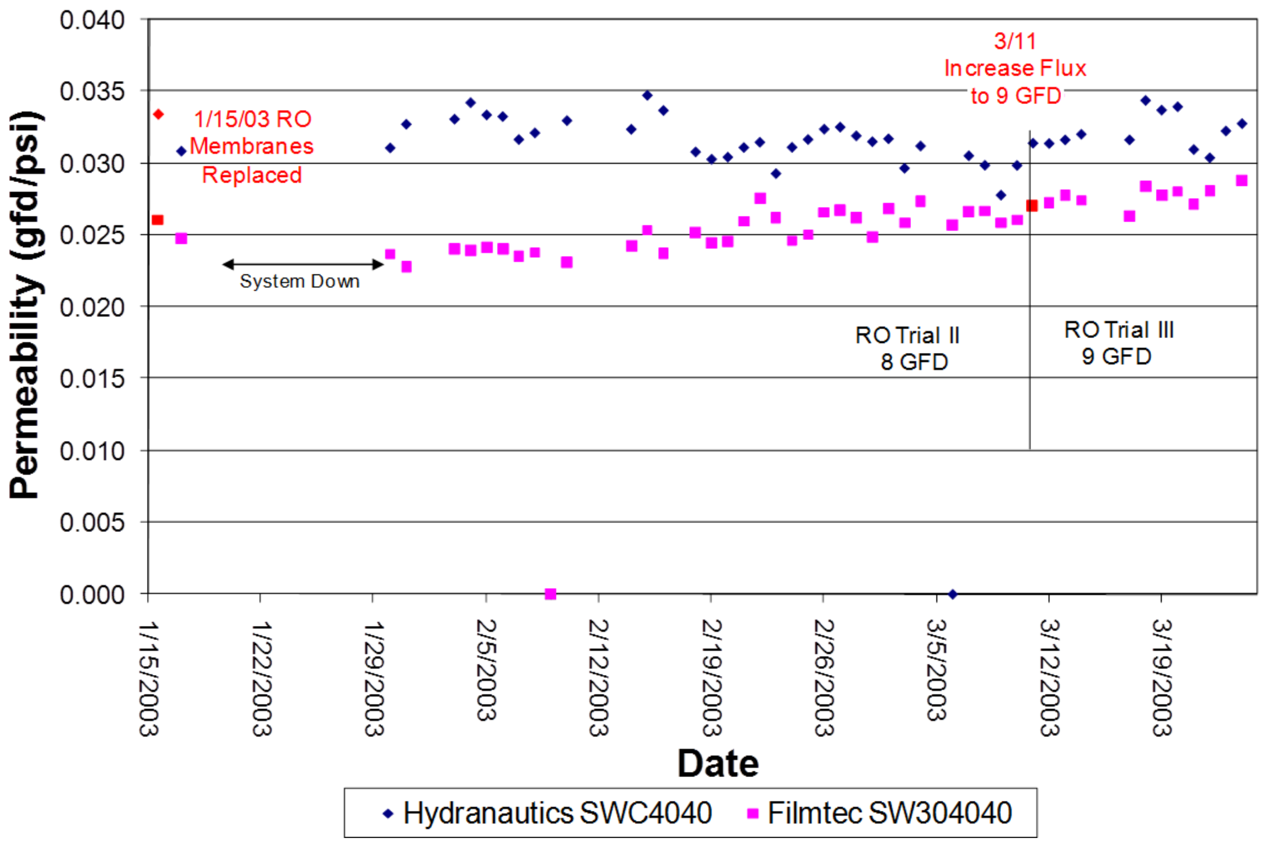



**Figure 41.—Reverse osmosis membrane permeability trial II and beginning of trial III.**

Between April and October 2003, the trials were halted to make some mechanical changes to the RO system, namely moving the high pressure pumps to a separate skid and the addition of variable frequency drives. These changes are discussed further in the "Process and Equipment Challenges" section. Testing was resumed in October 2003. A drop in permeability was immediately observed, and so the membranes were cleaned on December 5, 2003. The permeability decline was probably due to bacteriological growth in the RO membranes during the period of shutdown. For most of the shutdown, the membranes were periodically run and then flushed with RO permeate water. However, the RO retrofit occurred over a period of 2 months in the summertime, the power to the unit was out, and thus the membranes could not be flushed. After cleaning, the permeability was restored to pre-shutdown values and the system operated at 9 GFD flux. The flux was increased to 11 GFD on February 18, 2004. Comparing the permeability between January 15, 2003, and June 2, 2004 (the beginning of the period in figure 41 and the end of the period in figure 42), demonstrated that both the Hydranautics and Dow membranes did not decrease in permeability over the course of the testing. Thus, no significant fouling was observed on these RO membranes over approximately 3,100 hours of testing.

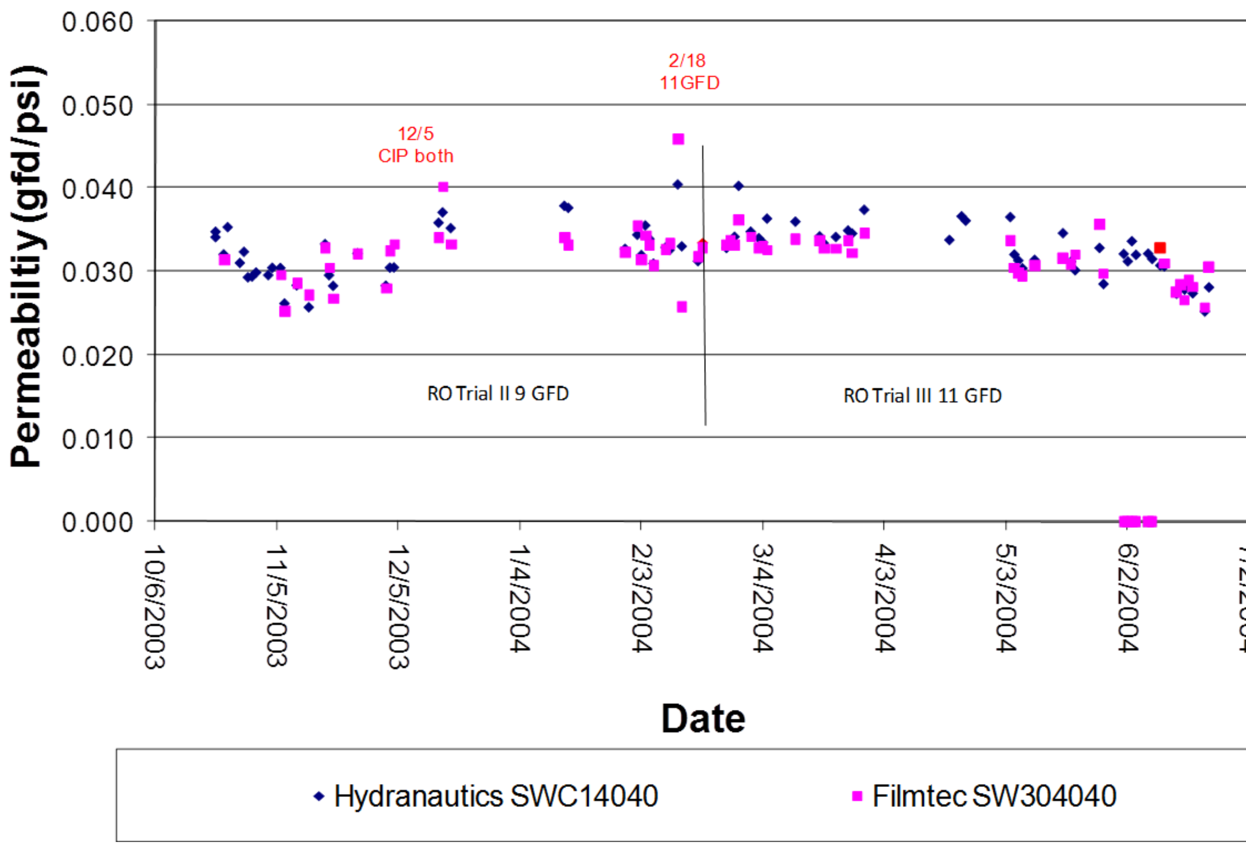



**Figure 42.—Reverse osmosis membrane permeability, end of trial III and trial IV.**

On June 10, 2004, the RO flux rate was increased to 12 GFD. Further testing was required at this flux rate, and at the end of Phase A, the optimized RO run parameters were as follows:

Table 23 - Optimized RO Parameters, Phase A Testing

| Parameter | Value |
|---|---|
| RO operating flux (GFD)* | 8–11 |
| Recovery | 50% |
| Sodium bisulfite dose (mg/L)* | 3 |
| Antiscalant dose (mg/L) | 3 |

*Optimized parameters.

## RO Permeate Quality

Over the course of the Phase A testing, two sets of RO membranes from each RO manufacturer were tested, and for each set, the Dow SW30-4040 initially produced water of significantly better quality (lower concentration of most constituents) than the Hydranautics SWC-4040. RO permeate quality was continuously measured via conductivity, and biweekly samples were taken for individual analysis.

### *Conductivity*

Figure 43 demonstrates that the conductivity of the permeate produced by the Dow membrane was initially significantly lower than that of Hydranautics. However, during trial II, the conductivity of Dow permeate rose and the Hydranautics permeate conductivity gradually declined. By the beginning of Trial III of the RO testing, the two membranes were producing water with similar conductivity. At the end of Trial IV of the testing, each membrane was producing permeate water of about 550 µS at a flux of 11 GFD and 18 °C feed-water temperature (figure 44).

### *Individual Ion Analyses*

Tables 24, 25, and 26 summarize the average results of the laboratory analysis performed on the RO streams for each trial of the Phase A testing. The following were evident:

1. For each Trial (flux), each RO membrane produced permeate with TDS < 300 mg/L. Note that this treatment process did not include stabilization of the RO permeate, which would be necessary for distribution of potable water.

2. The Dow membrane initially produced water with substantially lower concentrations of both boron and TDS than the Hydranautics membrane. The Dow membrane continued to produce lower concentrations, but the gap between the two membranes lessened as the testing progressed. Boron levels were constantly below 1.5 and 1.0 mg/L for the Hydranautics and the Dow, respectively.

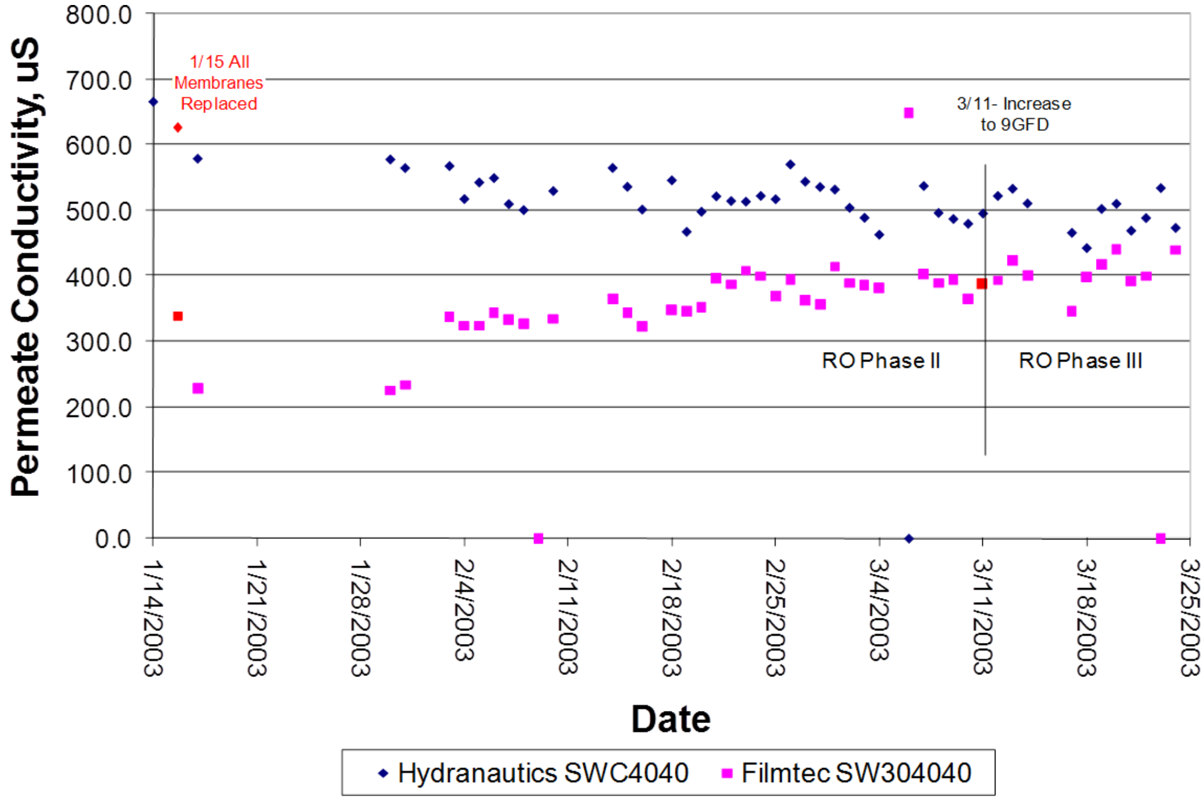



**Figure 43.—Reverse osmosis membrane conductivity trial II and beginning of trial III.**

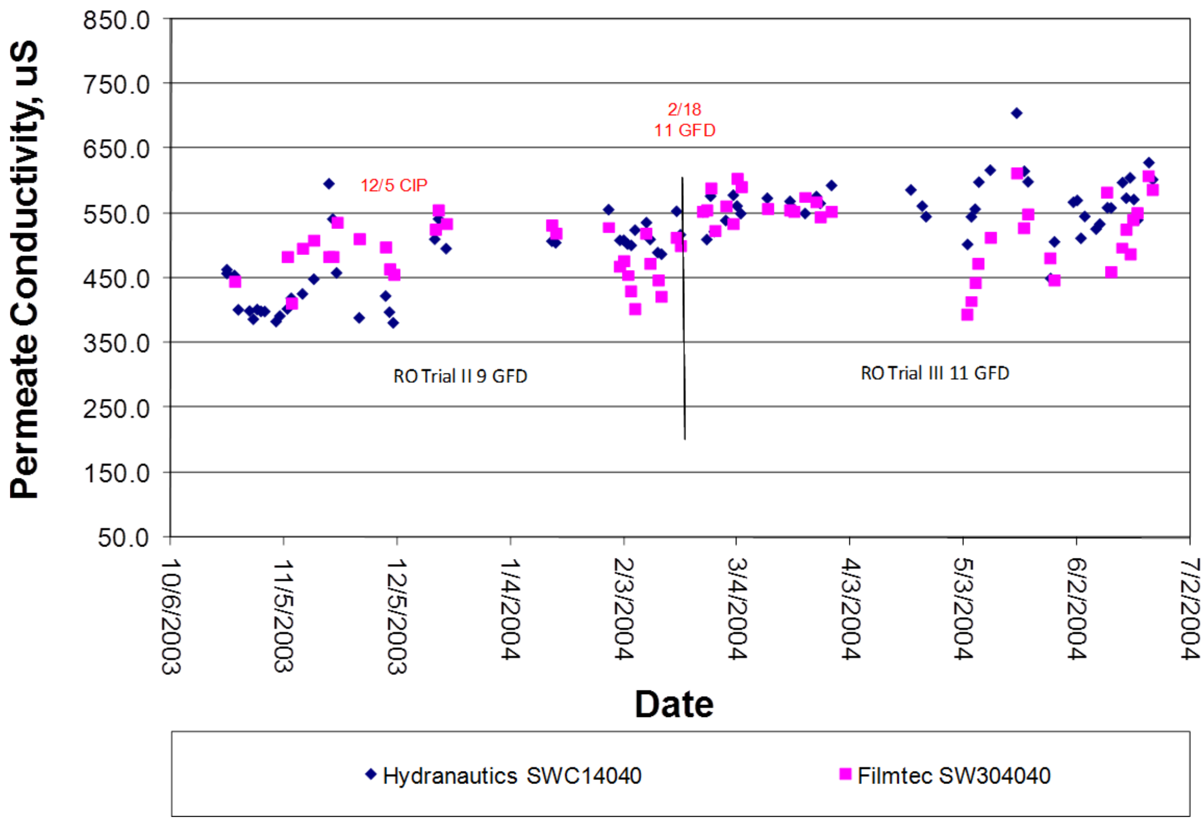



**Figure 44.—Reverse osmosis membrane conductivity end of trial III and trial IV.**

Table 24.—Average RO Membrane Water Quality for Trial II

Flux rate = 8 GFD

| Parameter | RO Feed | Permeate | | Concentrate | | Units |
|---|---|---|---|---|---|---|
| | | Train 1, HYD | Train 2, DOW | Train 1, HYD | Train 2, DOW | |
| TDS | 34,750 | 230 | 150 | 69,000 | 67,000 | mg/L |
| Lab pH* | 8.1 | 6.9 | 6.5 | 7.9 | 7.9 | pH |
| Alkalinity (as $CaCO_3$) | 115 | <2 | <2 | 212 | 214 | mg/L |
| Bicarbonate (as $CaCO_3$) | 114 | <2 | <2 | 210 | 212 | mg/L |
| Carbonate (as $CaCO_3$) | 1.3 | <0.1 | <0.1 | 1.5 | 1.6 | mg/L |
| Hydroxide (as $CaCO_3$) | 0.06 | <0.01 | <0.01 | 0.04 | 0.04 | mg/L |
| Sulfate | 2,533 | <10 | <10 | 5,538 | 5,463 | mg/L |
| Chloride | 18,875 | 111 | 70 | 35,325 | 34,975 | mg/L |
| Nitrate (as N) | <25 | <0.5 | <0.5 | <25 | <25 | mg/L |
| Nitrite (as N) | <25 | <0.5 | <0.5 | <25 | <25 | mg/L |
| Bromide | 63 | <0.25 | <0.25 | <100 | <100 | mg/L |
| Calcium | 395 | 0.6 | 1.1 | 739 | 724 | mg/L |
| Magnesium | 1,360 | 2.0 | 2.6 | 2,504 | 2,460 | mg/L |
| Hardness (as $CaCO_3$) | 6,586 | 9.4 | 13.1 | 12,156 | 11,937 | mg/L |
| Ca Hardness (as $CaCO_3$) | 986 | 1.5 | 2.8 | 1,846 | 1,807 | mg/L |
| Sodium | 11,175 | 77 | 46 | 20,600 | 20,400 | mg/L |
| Potassium | 398 | 2.7 | 1.9 | 779 | 756 | mg/L |
| Fluoride | 0.9 | <0.1 | <0.1 | 1.2 | 1.2 | mg/L |
| Strontium | 7.6 | 0.011 | 0.018 | 14.6 | 14.5 | mg/L |
| Barium | <0.025 | <0.025 | <0.025 | <0.025 | <0.025 | mg/L |
| Boron | 3.7 | 1.2 | 0.6 | 6.6 | 6.9 | mg/L |
| Silica | <10 | <10 | <10 | <10 | <10 | mg/L |
| Ammonia (as N) | <0.1 | <0.1 | <0.1 | <0.1 | <0.1 | mg/L |
| TOC | 0.9 | <0.5 | <0.5 | 1.7 | 1.7 | mg/L |

Notes: Avg. temperature 22 °C, four samples
Maximum TDS: 290 HYD, 160 Dow
Maximum boron: 1.3 HYD, 0.7 Dow

Table 25.—Average RO Membrane Water Quality for Trial III
Flux rate = 9 GFD

| Parameter | RO Feed | Permeate | | Concentrate | | Units |
|---|---|---|---|---|---|---|
| | | Train 1, HYD | Train 2, DOW | Train 1, HYD | Train 2, DOW | |
| TDS | 34,167 | 185 | 178 | 64,667 | 64,667 | mg/L |
| Lab pH* | 8.0 | 6.6 | 6.6 | 7.8 | 7.8 | pH |
| Alkalinity (as $CaCO_3$) | 112 | <2 | <2 | 205 | 205 | mg/L |
| Bicarbonate (as $CaCO_3$) | 111 | <2 | <2 | 204 | 204 | mg/L |
| Carbonate (as $CaCO_3$) | 1.1 | <0.1 | <0.1 | 1.2 | 1.3 | mg/L |
| Hydroxide (as $CaCO_3$) | 0.05 | <0.01 | <0.01 | 0.03 | 0.03 | mg/L |
| Sulfate | 2,538 | <10 | <10 | 5,265 | 5,160 | mg/L |
| Chloride | 18,967 | 100 | 95 | 35,050 | 33,950 | mg/L |
| Nitrate (as N) | <25 | <0.5 | <0.5 | <200 | <200 | mg/L |
| Nitrite (as N) | <25 | <0.5 | <0.5 | <200 | <200 | mg/L |
| Bromide | 66 | <0.25 | <0.25 | <100 | <100 | mg/L |
| Calcium | 378 | 0.6 | 0.9 | 718 | 724 | mg/L |
| Magnesium | 1,260 | 1.5 | 2.4 | 2,410 | 2,457 | mg/L |
| Hardness (as $CaCO_3$) | 6,133 | 7.1 | 11.2 | 11,716 | 11,925 | mg/L |
| Ca Hardness (as $CaCO_3$) | 944 | 1.4 | 2.2 | 1,792 | 1,808 | mg/L |
| Sodium | 10,383 | 68 | 63 | 19,867 | 20,133 | mg/L |
| Potassium | 384 | 2.3 | 2.3 | 719 | 743 | mg/L |
| Fluoride | 1.0 | <0.1 | <0.1 | 1.3 | 1.3 | mg/L |
| Strontium | 7.6 | 0.01 | 0.02 | 14 | 14 | mg/L |
| Barium | <0.025 | <0.010 | <0.010 | <0.025 | <0.025 | mg/L |
| Boron | 3.5 | 1.1 | 0.8 | 6.6 | 6.6 | mg/L |
| Silica | <10 | <1 | <1 | <10 | <10 | mg/L |
| Ammonia (as N) | <0.1 | <0.1 | <0.1 | <0.1 | <0.1 | mg/L |
| TOC | 0.9 | <0.5 | <0.5 | 2.2 | 2.1 | mg/L |

Notes: Avg. temperature 22 °C, five samples
Maximum TDS: 240 HYD, 230 Dow
Maximum boron: 1.2 HYD, 1.0 Dow

Table 26.—Average RO Membrane Water Quality for Trial IV
Flux rate = 11 GFD

| Parameter | RO Feed | Permeate | | Concentrate | | Units |
|---|---|---|---|---|---|---|
| | | Train 1, HYD | Train 2, DOW | Train 1, HYD | Train 2, DOW | |
| TDS | 34,800 | 200 | 160 | 71,400 | 68,600 | mg/L |
| Lab pH* | 8.0 | 7.1 | 6.8 | 7.7 | 7.8 | pH |
| Alkalinity (as $CaCO_3$) | 108 | <2 | <2 | 205 | 205 | mg/L |
| Bicarbonate (as $CaCO_3$) | 107 | <2 | <2 | 204 | 204 | mg/L |
| Carbonate (as $CaCO_3$) | 1.0 | <0.1 | <0.1 | 1 | 1 | mg/L |
| Hydroxide (as $CaCO_3$) | 0.0 | <0.01 | <0.01 | 0 | 0 | mg/L |
| Sulfate | 2,492 | <10 | <10 | 5,370 | 5,276 | mg/L |
| Chloride | 18,580 | 112.8 | 93.1 | 35,000 | 34,460 | mg/L |
| Nitrate (as N) | <25 | <0.5 | <0.5 | <200 | <200 | mg/L |
| Nitrite (as N) | <25 | <0.5 | <0.5 | <200 | <200 | mg/L |
| Bromide | 58 | <0.25 | <0.25 | <100 | <100 | mg/L |
| Calcium | 409 | <0.5 | 0.6 | 790 | 779 | mg/L |
| Magnesium | 1,304 | 1.0 | 1.3 | 2,514 | 2,498 | mg/L |
| Hardness (as $CaCO_3$) | 6,392 | 4.3 | 6.4 | 12,326 | 12,231 | mg/L |
| Ca Hardness (as $CaCO_3$) | 1,021 | <1.2 | 1.5 | 1,974 | 1,945 | mg/L |
| Sodium | 10,480 | 75.2 | 57.3 | 20,240 | 20,040 | mg/L |
| Potassium | 418 | 2.7 | 2.1 | 792 | 784 | mg/L |
| Fluoride | 0.9 | <0.1 | <0.1 | 1.3 | 1.3 | mg/L |
| Strontium | 7.6 | 0.0 | 0.0 | 14.8 | 14.6 | mg/L |
| Barium | <0.025 | <0.010 | <0.010 | <0.025 | <0.025 | mg/L |
| Boron | 3.2 | 1.1 | 0.8 | 5.8 | 6.0 | mg/L |
| Silica | <10 | <1 | <1 | <10 | <10 | mg/L |
| Ammonia (as N) | <0.1 | <0.1 | <0.1 | <0.1 | <0.1 | mg/L |
| TOC | 1.2 | <0.5 | <0.5 | 2.5 | 2.2 | mg/L |

Notes: Avg. temperature 21 °C, five samples
Maximum TDS: 220 HYD, 190 Dow
Maximum boron: 1.2 HYD, 0.9 Dow

## Phase A Reverse Osmosis Membrane Performance vs. Manufacturers' Projected Performance

Both Dow and Hydranautics have RO projection software programs that provide engineering information required for RO system design, including required feed pump pressure and anticipated permeate water quality, etc. Table 27 provides a comparison of the performance of each membrane versus that predicted by the projection software programs.

Table 27.—RO Performance vs. Predicted

| RO Trial | Membrane | Flux (GFD) | Projected | | | Actual | | |
|---|---|---|---|---|---|---|---|---|
| | | | Feed psi (psig) | Permeate TDS (mg/L) | Permeate Boron (mg/L) | Feed psi (psig) | Permeate TDS (mg/L) | Permeate Boron (mg/L) |
| II | Hydranautics SWC-4040 | 8 | 850 | 275 | NA* | 810 | 230 | 1.2 |
| III | Hydranautics SWC-4040 | 9 | 871 | 246 | NA* | 840 | 185 | 1.1 |
| IV | Hydranautics SWC-4040 | 11 | 930 | 190 | NA* | 870 | 200 | 1.1 |
| II | Dow SW30-4040 | 8 | 850 | 230 | 0.80 | 850 | 160 | 0.6 |
| III | Dow SW30-4040 | 9 | 879 | 205 | 0.74 | 870 | 230 | 0.8 |
| IV | Dow SW30-4040 | 11 | 950 | 161 | 0.6 | 905 | 190 | 0.8 |

* Hydranautics software did not predict boron rejection at time of this analysis.

Both membranes provided lower concentrations in the permeate than predicted by the manufacturer's software in initial operation, but higher concentrations in later phases. This is believed to be the result of changes to membrane performance and not inaccuracies in the software at the listed higher flux conditions.

The overall permeate concentrations for both membranes operating in Trials II–IV showed increases that are considered abnormal. These include both the steady increase over a period of operation, as observed with the Dow in Trials II and III, and the step increase observed at the start of Trial IV. A verification of the membrane performance at Trial II conditions (8 GFD) was planned for Phase B of the testing.

## RO Concentrate (Waste) Characterization

The RO concentrate stream was sampled biweekly for the parameters listed above in tables 24, 25, and 26 in order to characterize the RO waste stream. The recovery of the RO was 50 percent for the duration of the testing period.

## Phase B RO Testing

Phase B provided abundant information on new generation RO membranes regarding permeability and water quality. Data was also gathered to help develop strategies for operating on both powerplant influent and effluent, as well as during seasonal water quality events such as red tides and biofouling episodes.

Although Phase A provided valuable RO performance data on two leading seawater RO membranes, substantial development occurred in several manufacturers' product lines in the period from the start of Phase A to the start of Phase B. For that reason, the test plan of Phase B called for evaluation of four "next generation" or newly developed membranes. Phases B1 and B2 consisted of testing four next-generation membranes on powerplant influent and effluent water, respectively. The two membrane models considered to have demonstrated the best performance in Phases B1 and B2 were selected for long-term operation in Phase B3. Interestingly, the criteria for "best" performance saw an evolution, which affected the selection process. The two next-generation RO membranes initially selected for Phase B3 were Toray TM810 and Dow SW30 HR LE-4040. Selection criteria were initially based upon permeability and boron rejection characteristics. Subsequent review of product water quality goals for various proposed full-scale facilities identified chloride concentrations as a controlling constituent in defining the level of desalination required for several of the projects. In response to this issue, the Toray product was replaced with the Hydranautics SWC4+ membrane, which had demonstrated the highest chloride rejection of all membranes previously tested. This selection provided Phase B-3 with the membrane that most efficiently removed boron (Dow) and the one that achieved the lowest chloride concentration (Hydranautics).

Tables 28, 29, and 30 list the operating parameters of the RO membranes during the Phase B period of testing.

Table 28.—Summary of Phase B1 RO Runs

Feed source is powerplant influent

| Run # | Dates | Pretreatment Chemical | Anti-scalant (mg/L) | Membrane A | | | Membrane B | | | Comments |
|---|---|---|---|---|---|---|---|---|---|---|
| | | | | Membrane Type | Flux (GFD) | % Re-covery | Membrane Type | Flux (GFD) | % Re-covery | |
| RO11 | 6/10/04–11/16/04 | 3 mg/L SBS | 3 | Hydranautics SWC1-4040 Set B | 12 | 50% | Dow SW30-4040 Set B | 12 | 50% | Flux increased from 11 to 12 GFD to investigate performance at higher flux. |
| RO12 | 11/17/04–12/10/04 | 3 mg/L SBS | 3 | Hydranautics SWC1-4040 Set B | 8 | 50% | Dow SW30-4040 Set B | 8 | 50% | Flux reduced back to 8 GFD to compare performance vs. previous runs. |
| RO13 | 12/17/04–2/24/05 | 3 mg/L SBS | 3 | None | NA | NA | Toray TM810 | 10, 12 | 50% | Begin testing of next-generation RO membranes |
| RO14 | 2/25/05–4/27/05 | 3 mg/L SBS | 3 | None | NA | NA | Koch 1820SS | 10, 12 | 50% | |
| RO15 | 5/15/05–7/17/05 | 3 mg/L SBS | 3 | Dow SW30HRLE-4040 | 10, 12 | 50% | Hydranautics SWC4+ 4040 | 10, 12 | 50% | Red tide event started in late May/early June. RO membranes experienced fouling |

Table 29.—Summary of Phase B2 RO Runs

Feed source is powerplant effluent

| Run # | Dates | Pretreatment Chemical | Anti-scalant (mg/L) | Membrane A | | | Membrane B | | | Comments |
|---|---|---|---|---|---|---|---|---|---|---|
| | | | | Membrane Type | Flux (GFD) | % Re-covery | Membrane Type | Flux (GFD) | % Re-covery | |
| RO16 | 7/18/05–12/5/05 | 3 mg/L SBS | 3 | Dow SW30HRLE-4040 | 10, 12 | 50% | Hydranautics SWC4+ 4040 | 10, 12 | 50% | |
| RO17 | 12/06/05–5/20/06 | 3 mg/L SBS | 3 | Toray TM810 | 12 | 50% | Koch 1820SS | 10, 12 | 50% | Operation reverted to influent water Feb 10th due to feed pump issues. RO fouling occurred in mid-March, coinciding with another algae bloom. |

Table 30.—Summary of Phase B3 RO Runs

Feed source is powerplant effluent

| Run # | Dates | Pretreatment Chemical | Anti-scalant (mg/L) | Membrane A | | | Membrane B | | | Comments |
|---|---|---|---|---|---|---|---|---|---|---|
| | | | | Membrane Type | Flux (GFD) | % Re-covery | Membrane Type | Flux (GFD) | % Re-covery | |
| RO18 | 5/23/06–8/1/06 | 3 mg/L SBS | 3 | Dow SW30HRLE -4040 | 12 | 50% | Toray TM810 Set B | 12 | 50% | Dow SW30HRLE-4040 and Toray TM810 selected for further testing |
| RO19 | 8/1/06–10/15/06 | 3 mg/L SBS | 3 | Dow SW30HRLE -4040 Set B | 12 | 50% | Toray TM810 Set B | 12 | 50% | RO HP Pump failure required new set of Dow membranes to be installed. Biofouling of Toray membranes occurred, CIP restored performance |
| RO20 | 6/11/07 through Sept. 2007 | 3 mg/L SBS | 3 | Dow SW30HRLE -4040 Set B | 12 | 50% | Hydranautics SWC4+ 4040 | 12 | 50% | Hydranautics installed for further evaluation based on possible need for higher chloride and boron removal. Biofouling occurred for both trains. |

## RO Permeability

Figure 45 displays the permeability of all membranes tested in Phases B1 and B2. First, from June through November 2004, the Dow SW30-4040 and Hydranautics SWC1-4040 membrane were further evaluated to compare their performance at 12 GFD to their previous performance at flux rates of 8, 9, and 11. Unfortunately, an operational error with the sodium bisulfite pump allowed free chlorine to come in contact with both sets of membranes, resulting in membrane oxidation in early August. This is shown by the increase in permeability for these two membranes.

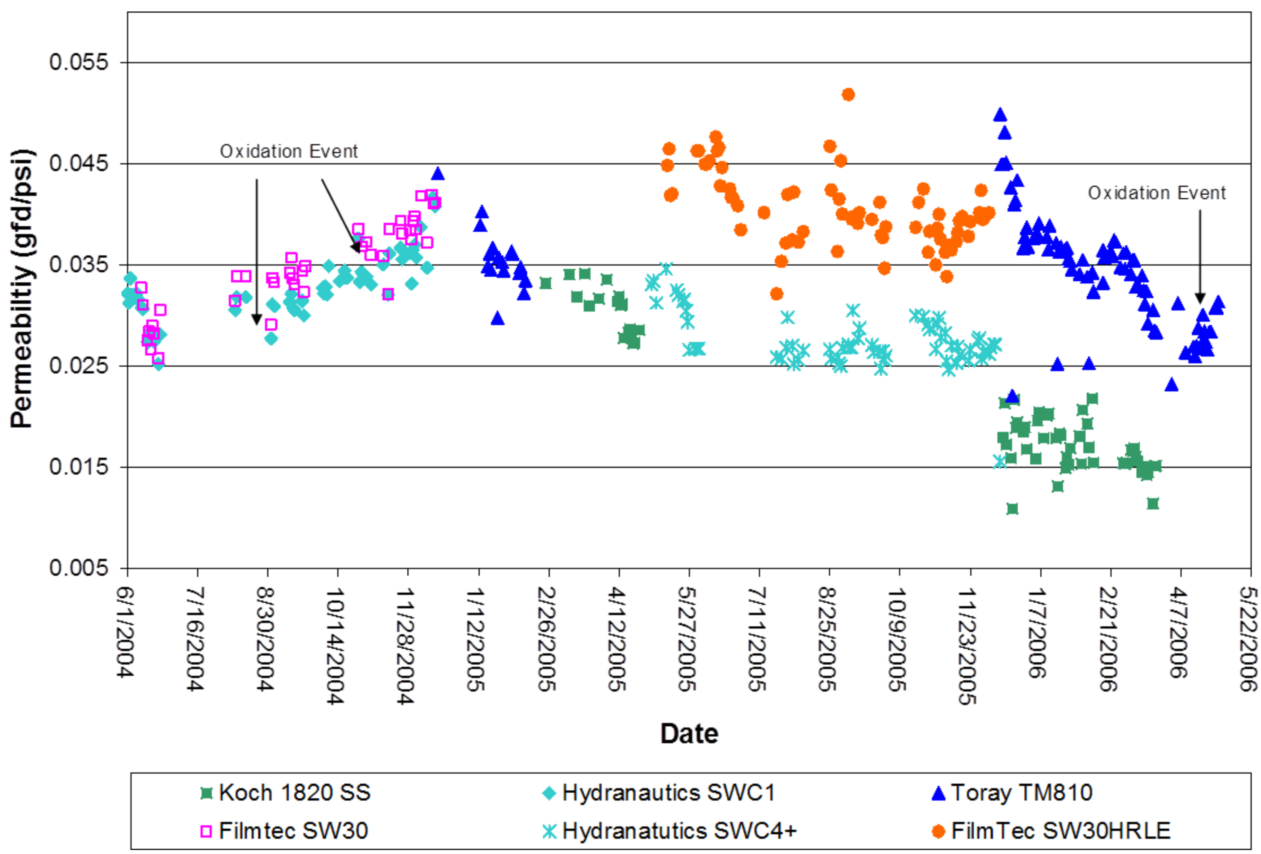



**Figure 45.—Phase B1 and B2 RO permeability**

The Toray TM810 next-generation RO membrane was tested at both 10 and 12 GFD from December 2004 to February 2005 to collect data on powerplant influent water. The Toray membrane showed strong performance with respect to both permeability and permeate quality.

In March and April 2005, data was collected from the Koch 1820SS membrane operating on influent water. The average permeability was slightly lower than for the Toray and the Dow, and average permeate concentrations were higher than those of all other next-generation membranes. This membrane had a comparatively poor performance.

From May through July 17, 2005, the next generation Dow (Filmtec) SW30HRLE and Hydranautics SWC4+ membranes were operated in parallel on influent water pretreated by microfiltration. On July 18th, the feed-water source was switched to effluent water to start Phase B2, and these membranes remained operating on effluent water until December 2005. During this period of testing, a severe red tide event occurred that started at the end of May and subsided in mid-August. Both sets of membranes lost some permeability during this time frame, and it is possible that dissolved organic matter produced by the algae bloom passed through the MF membrane and fouled the RO membranes.

In December of 2005, the Toray TM810 and Koch 1820SS membranes were reinserted into the system for continued testing on Phase B-2 powerplant effluent. The Toray membranes started up with higher permeability and higher conductivity than when operated in Phase B-1, and, after substantial trouble-shooting, two elements were replaced in the tail end of the system. Overall permeability and permeate conductivity returned to previous (Phase B-1) values when the new membranes were installed. The Koch membranes started up with lower permeability than when operated in Phase B-1. This could possibly be due to biogrowth that occurred in the membranes as they were in storage for 6 months. On February 10th, the system reverted back to influent water operation due to a malfunction of the effluent water supply pump. In mid-March, both sets of membranes lost some permeability. This event coincided with an algae bloom, confirmed by elevated levels of domoic acid present in the feed water as well as by satellite imagery of the Santa Monica Bay source water.

Light energy used in photosynthesis by algae cells and higher plants is absorbed by a number of photosynthetic pigments with absorption spectra covering a large range of the available light energy. The most prominent pigments that absorb this energy are chlorophyll *a* and chlorophyll *b*. Therefore, elevated levels of chlorophyll *a* in the ocean water coincide with increased algal activity. The following website monitors the chlorophyll *a* levels in the southern California ocean water:

http://www.sccoos.org/data/ocm/ocm_regions.php?r=3

Figure 46 below depicts normal chlorophyll *a* activity. This satellite image was taken in September 2006. Figure 47 depicts the chlorophyll *a* levels during the algal bloom in April 2006.

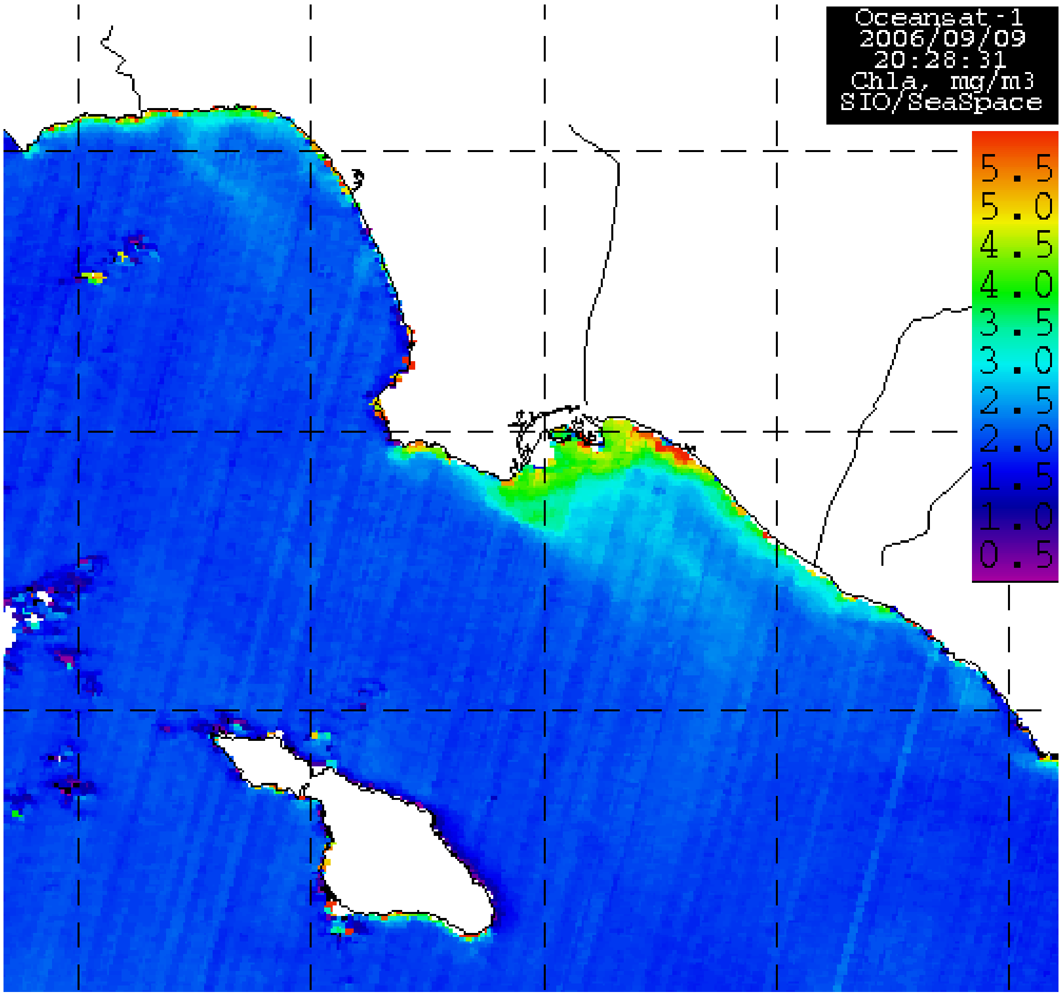



**Figure 46.—Chlorophyll *a* levels off the coast of southern California in September 2006.**

An offsite cleaning trial was performed on the Koch membranes, which is discussed further below. Separately, in an effort to eliminate the presence of biogrowth, the MF/RO break tank was cleaned with a sodium hypochlorite solution. Upon restarting the Toray membranes, some residual chlorine was present in the feed water, which oxidized the Toray membranes.

Phase B3 began in June 2006 with the Dow SW30HRLE membrane and the Toray TM810 membrane. The high permeability and high boron rejection characteristics of these two membranes warranted their selection for further long-term study.

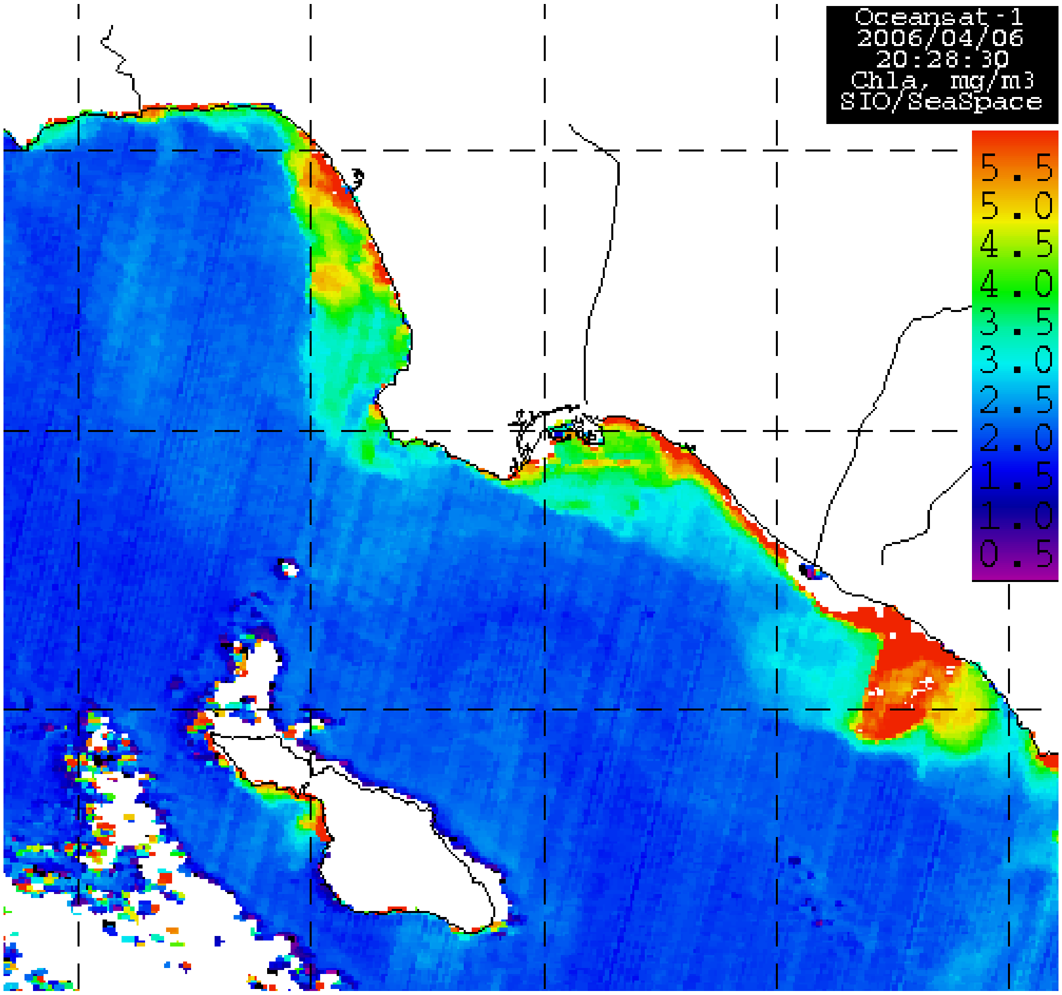



**Figure 47.—Elevated chlorophyll *a* levels off the coast of southern California in April 2006.**

The Toray TM810 and Dow SW30HRLE membranes were operated from June to October 2006 on powerplant effluent. The failure of a high-pressure feed pump seal leaked oil into the feed water and resulted in damage to the first set of Dow membranes, so a second set was installed and started up in August of 2006. Figure 48 shows the performance of the Toray membrane from August to early October 2006 before the entire pilot operation was shut down and relocated. The Toray membranes started to show signs of fouling in August 2006, and the trend continued in September. It was discovered that the MF/RO break tank had experienced biogrowth, which was the most likely contributor to the biofouling in the RO Trains. A membrane cleaning consisting of a 2-percent citric acid cleaning solution (pH ~2) heated to 35–38 °C followed by a caustic cleaning solution with 2-percent Avista P111 membrane cleaner (pH ~10.5) heated to 35–38 °C was successful in restoring performance.

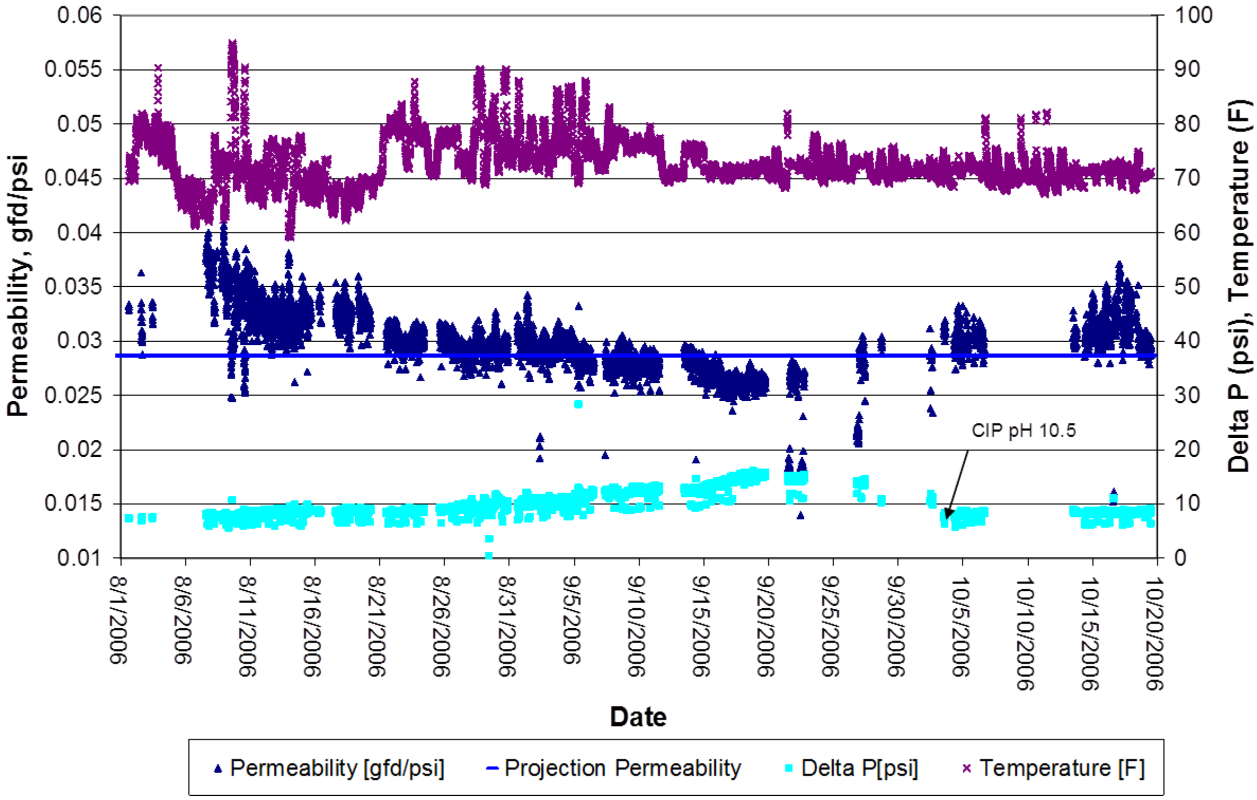



**Figure 48.—Toray TM810 permeability, August 2006 – October 2006.**

Figure 49 illustrates the Dow SW30HRLE membrane operation from August to October 2006. Mechanical issues as discussed in the "Process and Equipment Challenges" section of this document limited the run time during this period, but a loss in permeability was observed for the Dow membranes.

Phase B3 restarted in June 2007 with a new set of Hydranautics SWC4+ membranes to further evaluate the low TDS permeate quality seen in previous testing, along with the previous set of Dow SW30HRLE membranes. When the Dow RO membranes were brought back on line in June 2007, the permeability declined. In early September 2007, a CIP was performed applying the same citric acid and caustic cleaning procedure that was used successfully on the Toray membranes in September 2006 (see above). However, it failed to restore permeability for the Dow SW30HRLE.

Permeability started to decline more thereafter, and a visual inspection of the membranes in early September confirmed the presence of biogrowth in both sets of RO membranes SW30HRLE and Hydranautics SWC4+. Based on the poor results of the previous cleaning formulation at a pH of 10.5, a different cleaning formulation was tried at the end of September. Avista P112 is a commercial membrane cleaning product used to clean biofouling from RO membranes. In late September 2007 a 2 percent solution of P112 was used with the addition of NaOH to bring the pH of the cleaning solution up to 12, and the solution was heated to 30–35 °C. (Temperature guidelines for each membrane manufacturer at high pH were followed.) This formulation had encouraging results, as the pressure drop across both RO trains decreased, and the permeability of each RO train increased. The Hydranautics membrane showed a larger increase in permeability than the Dow, but initial data for the Dow membranes suggests that an additional cleaning step may be able to be remove more foulant.

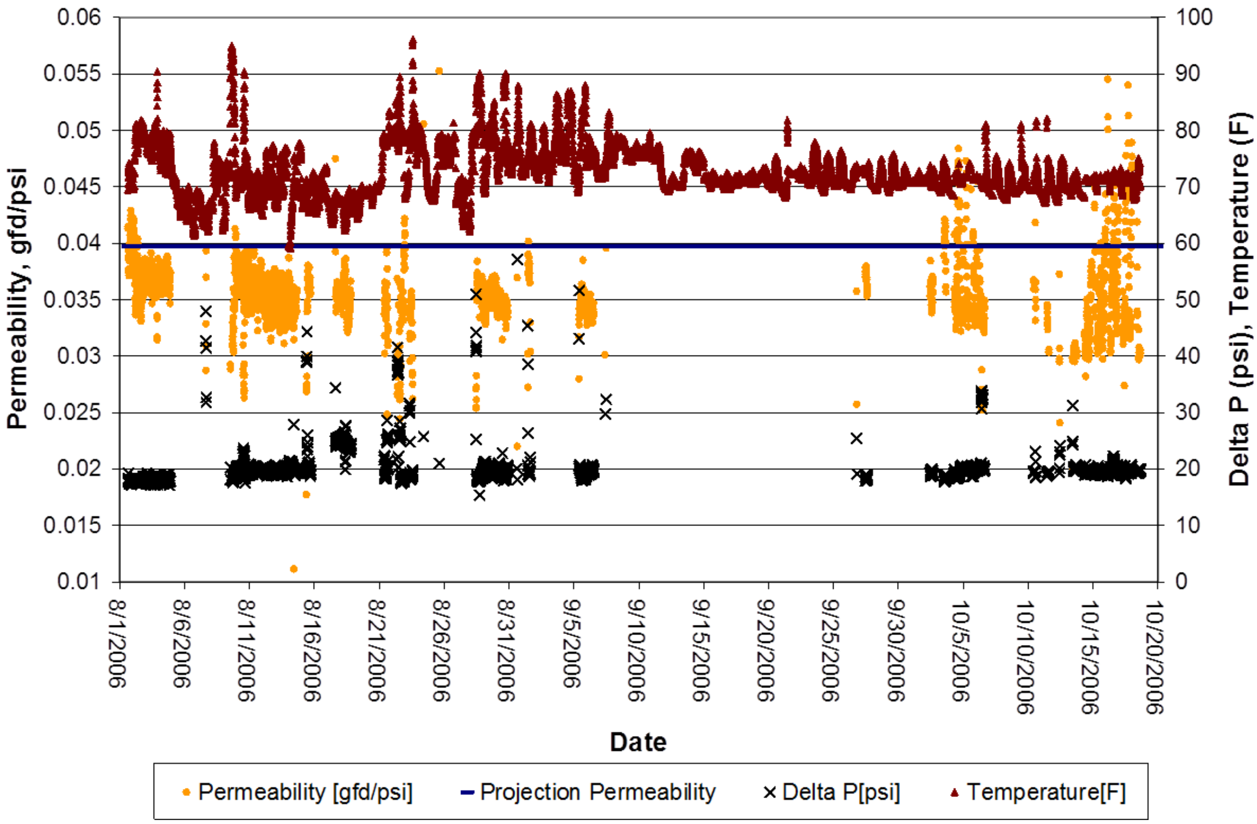



**Figure 49.—Dow SW30HRLE permeability, August 2006 – October 2006.**

Figures 50 and 51 show the performance from June through September 2007 of both the Dow and Hydranautics membranes.

The required feed pressures associated with the startup permeability values for the RO membranes tested in Phase B are shown in table 31. These feed pressures vary widely within the group. It is noteworthy that these pressures increased, in some cases substantially, as a result of the previously discussed fouling events. The membrane requiring the highest pressure (SWC4+) also had the lowest permeate chloride concentration, which may be an acceptable trade-off in some applications.

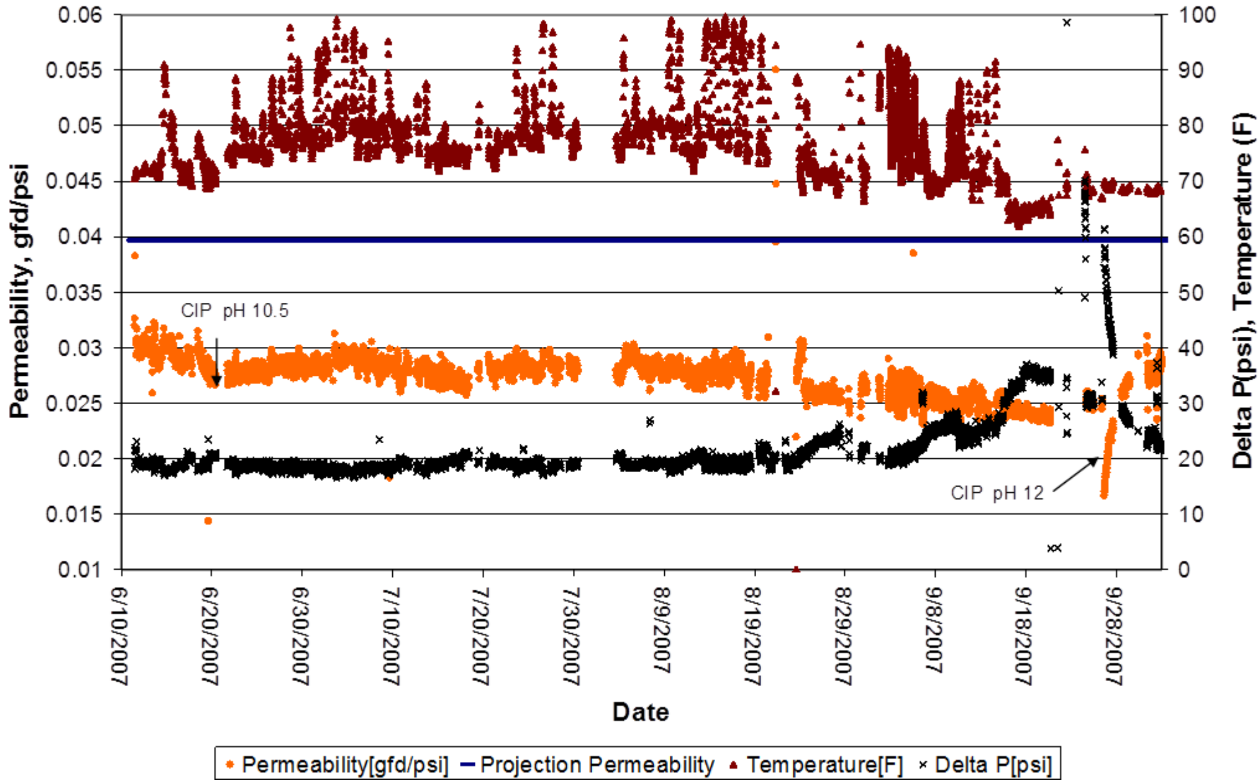



**Figure 50.—Dow SW30HRLE permeability, June 2007 – September 2007.**

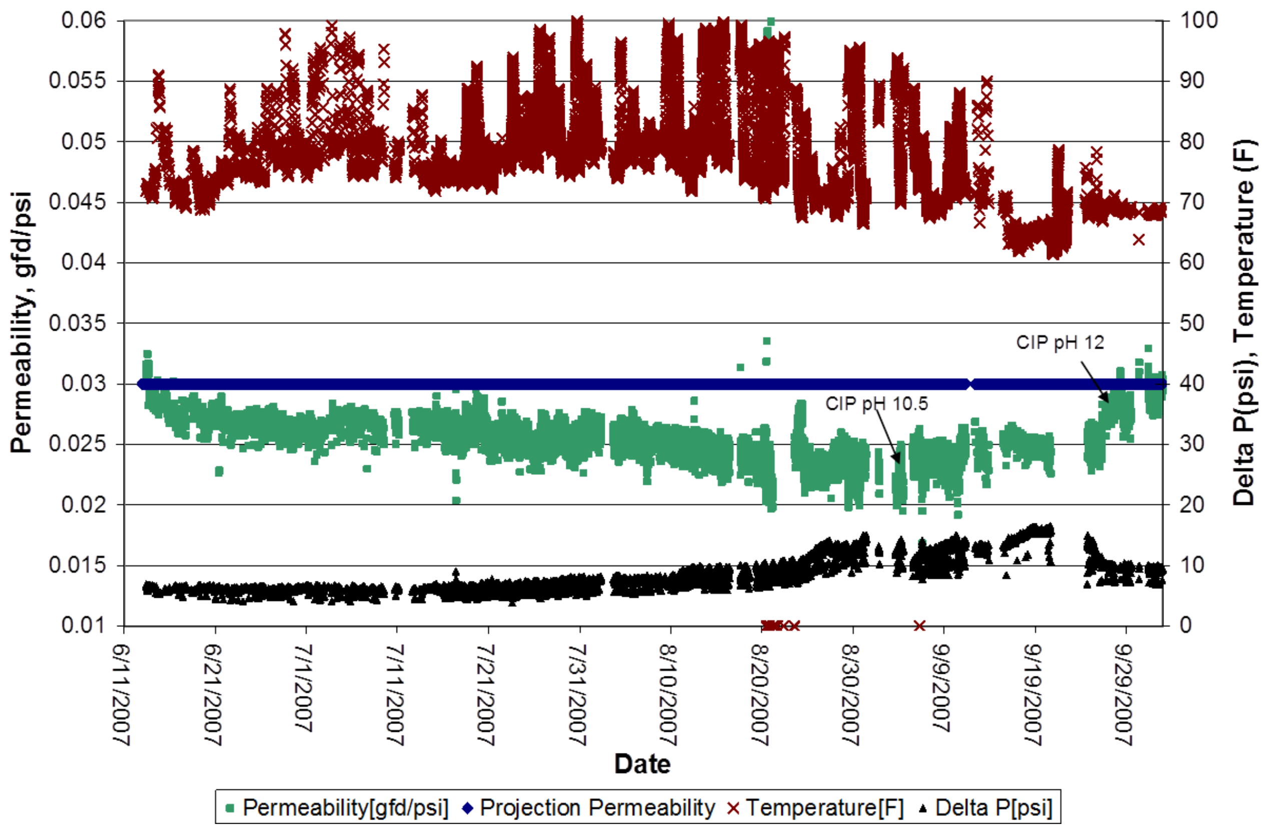



**Figure 51.—Hydranautics SWC4+ permeability, June 2007 – September 2007.**

Table 31.—Startup Feed Pressure Requirements

| Membrane | Feed Pressure Normalized to 25 °C (psi) | |
|---|---|---|
| | 10 GFD | 12 GFD |
| HYD SWC4+ | 910 | 985 |
| DOW SW30HRLE | 755 | 800 |
| Toray TM810 | 810 | 865 |
| Koch 1820SS | 840 | 900 |

## Summary of RO Fouling

The following is a summary of the reverse osmosis fouling events experienced in Phase B, with the details of each occurrence below:

- Four distinct RO fouling events occurred during the 3+ years of Phase B testing.
- Two of the events occurred during algae blooms, with one event on powerplant influent water at a temperature of approximately 65 °F and the other on influent water with an average temperature range of 60–65 °F. The CIP procedure using a commercial membrane cleaner with pH 12 proved more effective at restoring permeability than using either a generic formulation of pH 11 or a commercial cleaner of pH 11.
- The third event was on powerplant effluent water at 72–78 °F, with no algae bloom in effect but with biogrowth present in the break tank. The CIP utilizing the commercial cleaner at pH 10.5–11 proved to be effective at restoring permeability.
- The final event also occurred on powerplant effluent water, with an elevated temperature range of 75–90 °F. There was a continuous abundance of algae in the ocean during this time frame, and visual inspection of RO membranes indicated a biofouling layer was present in the RO membranes and throughout the RO system piping. The commercial membrane cleaner at an elevated pH of 12 was effective at restoring permeability to the Hydranautics membrane.

The first fouling event occurred in late May and early June of 2005 on the Dow SW30HRLE and Hydranautics SWC4+ membranes. This fouling coincided with a severe algae bloom in the ocean water where the pilot plant is located. The feed-water source was influent water, with an average temperature of approximately 65 °F. A two-step cleaning procedure was used for this first fouling event. In step 1, a 2-percent citric acid solution (pH ~2) was applied, heated to 35–38 °C. Step 2 used a high-pH solution with a generic formulation of:

- 1 percent sodium tripolyphosphate,
- 1 percent tetrasodium EDTA
- 1 percent trisodium phosphate
- The pH was adjusted to 11 and the solution was heated to 35–38 °C.

This generic formulation is commonly used for cleaning RO membrane; however, it failed to restore permeability. No other formulations were evaluated at this time.

The second fouling event occurred in March of 2006 on the Toray TM810 and Koch 1820SS membranes. This fouling also coincided with an algae bloom that was verified by presence of domoic acid in the feed water and by satellite imagery. The feed-water source was influent water with an average temperature range of 60–65 °F. In anticipation of difficulty in cleaning these membranes, two Koch elements (Serial # 4010 and 4042) were sent to Avista Technologies for a cleaning study. The study consisted of using commercial membrane cleaners P111 (2-percent solution, pH 11) and P112 (1-percent solution, pH 12), both heated to 35 °C. The P111 cleaner improved #4042 permeability by 23 percent, and the P112 cleaner improved #4010 permeability by 27 percent, bringing the flow within 16 percent of its original performance. This cleaning trial was very encouraging.

The third fouling event occurred in August and September of 2006 on new sets of Dow SW30HRLE and Toray TM810 membranes. This was a biofouling event, as green biogrowth was found in the break tank between the MF and RO units. The feed-water source was effluent water, and water temperature was elevated to an approximate range of 72–78 °F. Since there was no evidence of an algae bloom during this time, and biogrowth was found in the break tanks, the Toray membrane was cleaned using 2-percent citric acid (pH ~2) and Avista P111 (pH~ 10.5), both heated to 35–38 °C. This cleaning proved to be successful in restoring permeability. This procedure could not be performed on the Dow membranes due to the timing of the pilot plant relocation.

The fourth and final fouling event occurred in August and September of 2007 on the same set of Dow membranes mentioned in the previous paragraph, and a new set of Hydranautics SWC4+ membranes. The feed-water source was effluent water, with an average temperature range of approximately 75–90 °F. The powerplant was running consistently during the summer of 2007, and temperature spikes occasionally reached 100 °F. There was also a persistent abundance of algae in the ocean water during RO operation from June through September, and visual inspection of the RO membranes prior to cleaning revealed a layer of biofouling in the RO membranes and the RO system piping. First the Hydranautics membrane was cleaned using 2-percent citric acid (pH ~2) and Avista P111 (pH ~10.5), both heated to 35–38 °C, and the procedure failed to restore performance. Biogrowth continued in the system until another CIP was implemented on the Hydranautics membrane approximately 2 weeks later. This procedure used a 2-percent citric acid solution (pH ~2) heated to 35 °C and a 1-percent Avista P112 solution (pH 12) heated to only 30 °C per Hydranautics specifications on operating limits at elevated pH. This cleaning proved to be successful at restoring permeability of the Hydranautics membrane back to startup values.

The same formulation was then applied to the Dow membranes, with the only difference being heating the P112 solution to 35 °C, per Dow specifications. This procedure did have some effect on restoring permeability, but the operating data after the cleaning suggests that more foulant could be removed.

## RO Permeate Water Quality

The permeate conductivity for each of the next-generation RO membranes tested is displayed below in figure 52. The graph shows that the Hydranautics SWC4+ showed the highest overall rejection rate (lowest permeate conductivity) of all membranes tested, followed by the Dow (Filmtec) SW30HRLE and the Toray TM810, respectively. The Koch 1820SS membrane showed the lowest rejection among the next-generation RO membranes.

It should be noted that there were two operational errors, previously mentioned, one of which resulted in oxidation of the Dow SW30-4040 and Hydranautics SWC1-4040 in the summer of 2004, and another in the spring of 2006 that oxidized the Koch 1820SS and Toray TM810 membranes. The high conductivity in the permeate from each of these membranes (greater than 450 µS/cm) can be seen in figure 52.

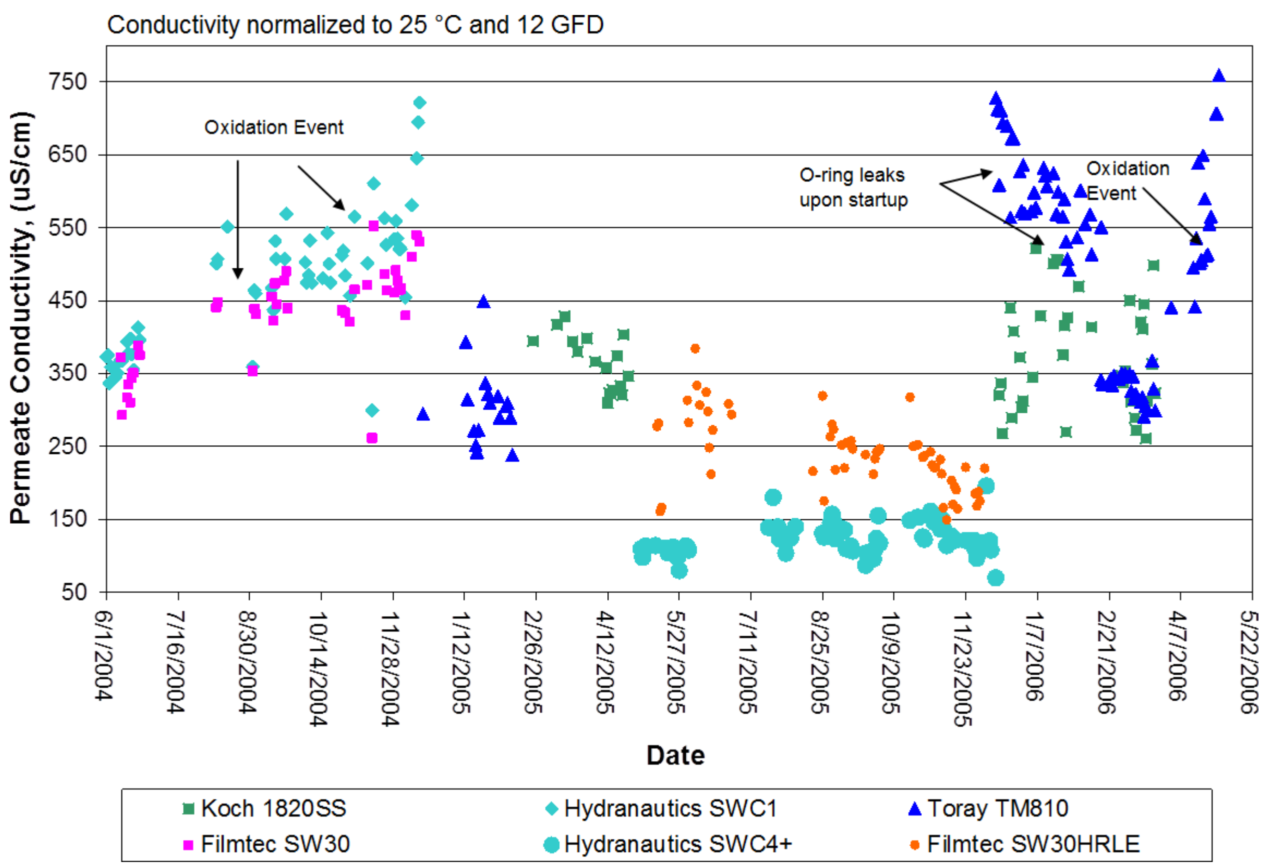



**Figure 52.—Summary of RO conductivity.**

Another noteworthy point relates to the re-installation of the Toray and Koch membranes in December 2005. After substantial troubleshooting involving O-ring leaks in the Toray membrane, two new elements were installed on February 16, 2006, and permeate conductivity returned to the values seen in previous testing.

One important aspect of RO membranes is their response to changes in feed-water temperature. When the temperature of the feed water is elevated, salt passage through the membrane increases, resulting in an increased overall TDS concentration in the RO permeate. This higher salt passage at elevated temperatures will result in elevated levels of individual ions such as chloride and boron. The permeability of the membrane also increases with elevations in feed-water temperature (although at a different rate than salt passage), resulting in less operating pressure required to achieve the same flux. Figure 53 shows a window of operation for the Dow SW30HRLE membrane as the temperature increased. Note the decrease in feed pressure required to maintain a constant flux and the increase in permeate TDS concentration. This window only shows the response through a temperature band of 65–80 °F. The actual operating window (as noted on the figure) extends to a greater temperature range. This results in a greater range of feed pressure, permeate TDS, and individual ion concentrations. Measured permeate boron and chloride concentrations as a function of temperature are displayed in figures 54 and 55, respectively.

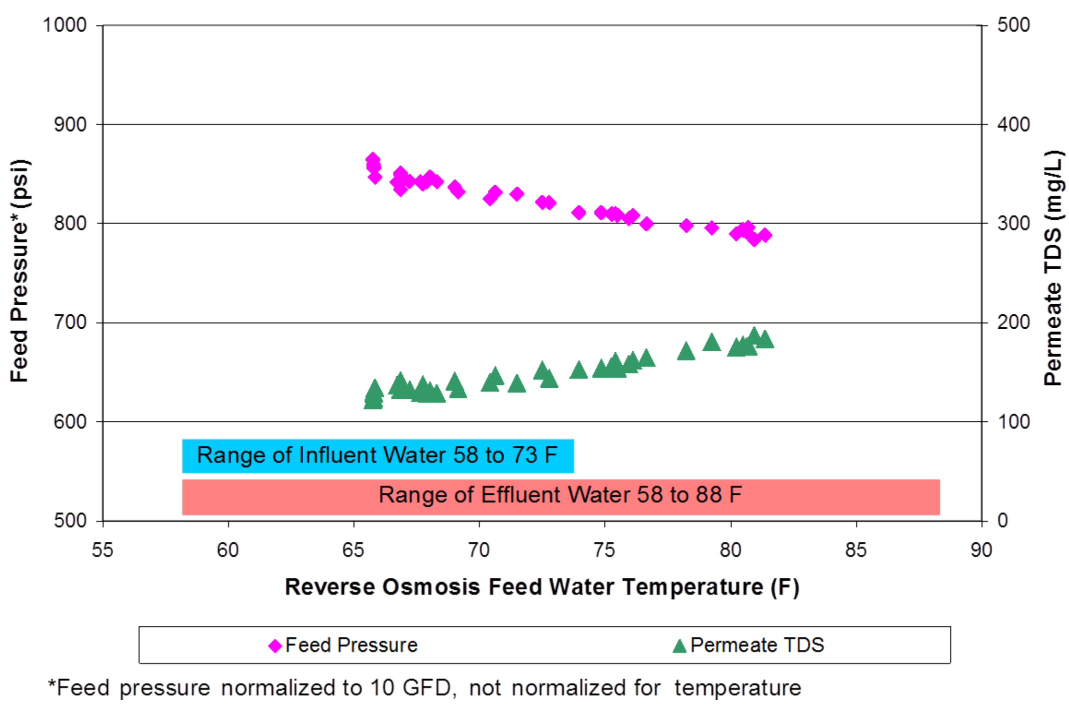



**Figure 53.—Temperature effects on the Dow SW30HRLE RO membrane.**

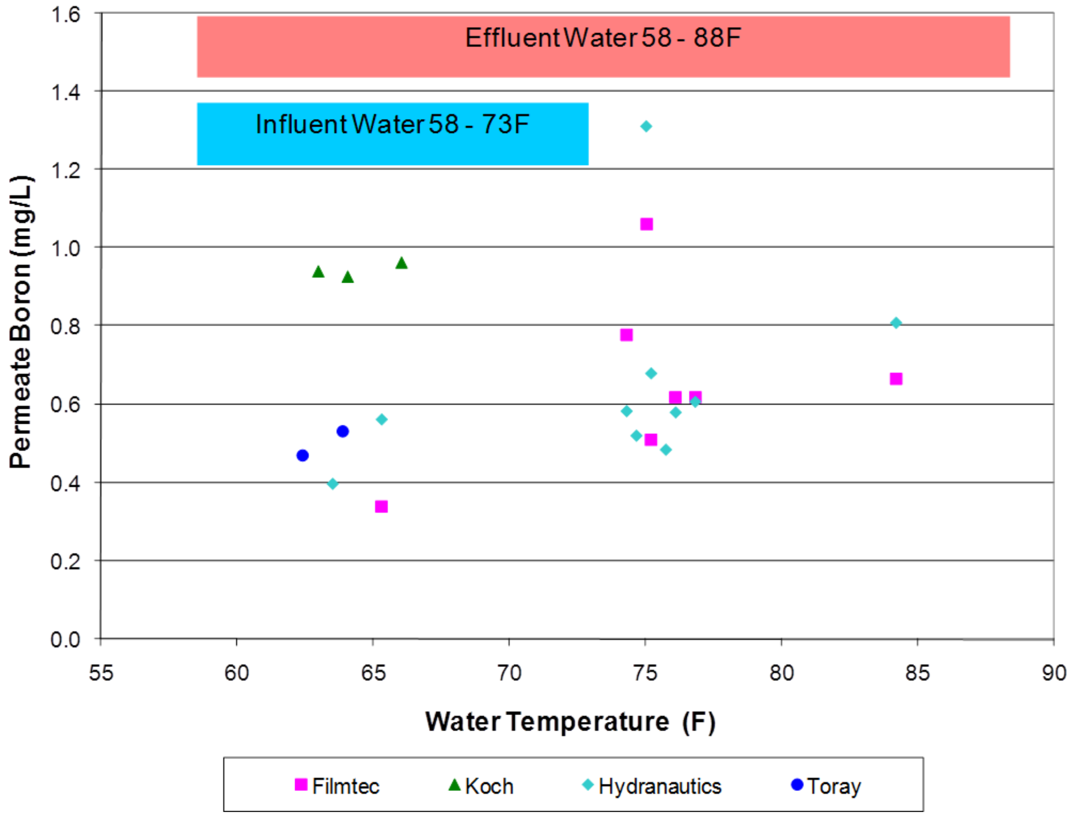



**Figure 54.—Permeate boron concentration vs. temperature at 12 GFD.**

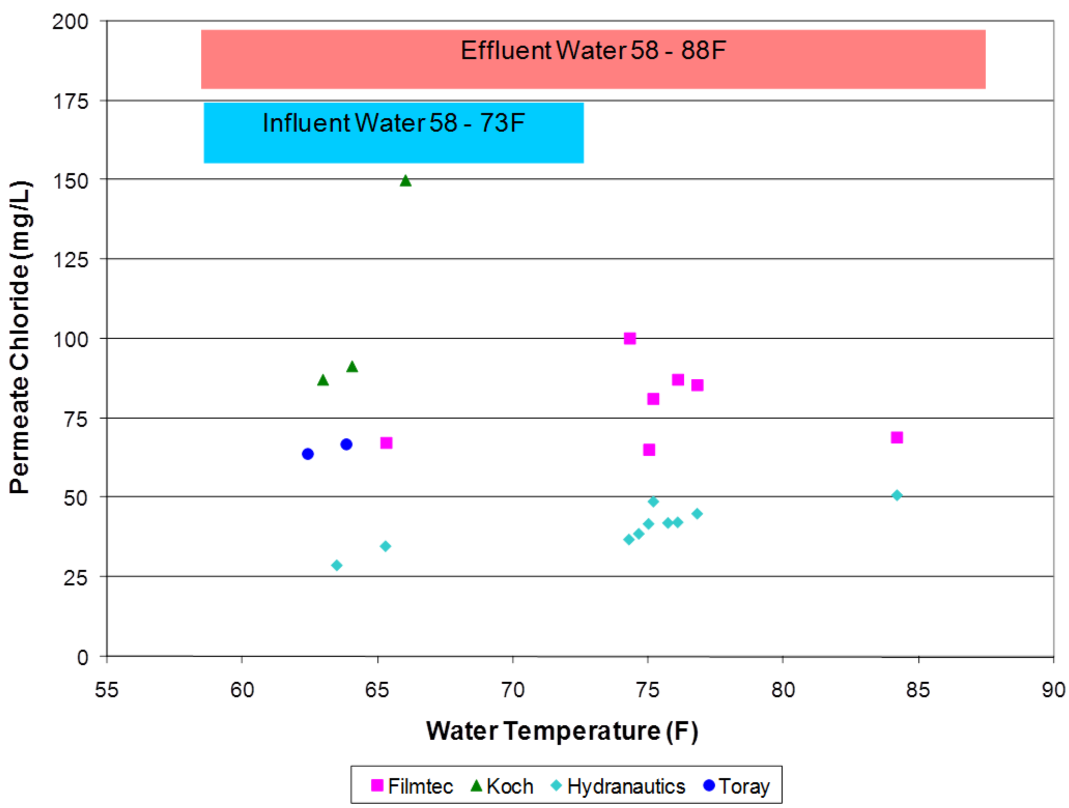



**Figure 55.—Permeate chloride concentration vs. temperature at 12 GFD.**

Tables 32 and 33 display the average feed, permeate, and concentrate water quality from the operation of the membranes. Of particular interest in the RO permeate are TDS, boron, and chloride. The current notification level for boron from the California Department of Public Health is 1 mg/L. Chloride levels less than 100 mg/L may also be deemed important for full-scale RO plants due to horticultural concerns. Note that the values shown in the following tables are averaged over a temperature range, and those values will change accordingly with changes in temperature.

Table 32.—Average Water Quality, June 2004 – July 2006, Hydranautics and Dow

| Parameter | Hydranautics SWC4+ (12 GFD, Avg. Temp. 23.4 °C) | | | Dow SW30HRLE (12 GFD, Avg. Temp. 24.1 °C) | | | Units |
|---|---|---|---|---|---|---|---|
| | RO Feed | RO Permeate | RO Con-centrate | RO Feed | RO Permeate | RO Con-centrate | |
| TDS | 33,889 | 69.8 | 68,400 | 33,857 | 128.9 | 57,714 | mg/L |
| Lab pH | 8 | 6.4 | 7.8 | 8 | 6.6 | 7.8 | pH |
| Alkalinity (as $CaCO_3$) | 111 | <2 | 217 | 111 | 2.5 | 187 | mg/L |
| Bicarbonate (as $CaCO_3$) | 110 | <2 | 216 | 110 | 2.4 | 186 | mg/L |
| Carbonate (as $CaCO_3$) | 1 | <0.1 | 1.26 | 1.06 | <0.1 | 1.11 | mg/L |
| Hydroxide (as $CaCO_3$) | 0.05 | <0.01 | 0.03 | 0.05 | <0.01 | 0 | mg/L |
| Sulfate | 2,629 | <2 | 5,752 | 2,636 | 4.5 | 5,120 | mg/L |
| Chloride | 19,944 | 41.3 | 39,200 | 20,057 | 79 | 33,557 | mg/L |
| Nitrate (as N) | <25 | <0.5 | <200 | <25 | <0.5 | <200 | mg/L |
| Nitrite (as N) | <25 | <0.5 | <200 | <25 | <0.5 | <200 | mg/L |
| Bromide | 57 | <0.25 | <100 | 57 | 0.3 | <100 | mg/L |
| Calcium | 382 | <0.1 | 763 | 379 | 0.5 | 639 | mg/L |
| Magnesium | 1,252 | 0.1 | 2,493 | 1,240 | 1.7 | 2,087 | mg/L |
| Hardness (as $CaCO_3$) | 6,111 | 0.5 | 12,172 | 6,053 | 8.2 | 10,190 | mg/L |
| Ca hardness (as $CaCO_3$) | 954 | <0.25 | 1,905 | 947 | 1.3 | 1,595 | mg/L |
| Sodium | 10,667 | 25.3 | 21,340 | 10,671 | 47.8 | 18,171 | mg/L |
| Potassium | 389 | 1 | 775 | 389 | 1.9 | 658 | mg/L |
| Fluoride | 0.9 | <0.1 | 1.5 | 0.9 | <0.1 | 1.5 | mg/L |
| Strontium | 7.3 | <0.002 | 14.5 | 7.4 | 0.011 | 12.4 | mg/L |
| Barium | <0.025 | <0.01 | <0.025 | <0.025 | <0.01 | <0.025 | mg/L |
| Boron | 3.5 | 0.65 | 6.8 | 3.4 | 0.63 | 5.9 | mg/L |
| Silica | <10 | <1 | <10 | <10 | <1 | <10 | mg/L |
| Ammonia (as N) | <0.1 | <0.1 | <0.1 | <0.1 | <0.1 | <0.1 | mg/L |
| TOC | 1 | <0.5 | 2.3 | 1 | <0.5 | 1.8 | mg/L |

Table 33.—Average Water Quality, June 2004 – July 2006, Toray and Koch

| Parameter | Toray TM810 (12 GFD, Avg. Temp. 20.2 °C) | | | Koch 1820SS (12 GFD, Avg. Temp. 17.8 °C) | | | Units |
|---|---|---|---|---|---|---|---|
| | RO Feed | RO Permeate | RO Con-centrate | RO Feed | RO Permeate | RO Con-centrate | |
| TDS | 34,000 | 127.5 | 64,000 | 32,500 | 140 | 62,500 | mg/L |
| Lab pH | 8 | 6.9 | 7.7 | 8 | 6.7 | 7.8 | pH |
| Alkalinity (as $CaCO_3$) | 111 | 2.5 | 206 | 104 | <2 | 205 | mg/L |
| Bicarbonate (as $CaCO_3$) | 109 | 2.5 | 205 | 103 | <2 | 203 | mg/L |
| Carbonate (as $CaCO_3$) | 0.98 | <0.1 | 0.98 | 0.97 | <0.1 | 1.21 | mg/L |
| Hydroxide (as $CaCO_3$) | 0.05 | <0.01 | 0.03 | 0.05 | <0.01 | 0.03 | mg/L |
| Sulfate | 2,570 | 5.1 | 5,433 | 2,535 | 11.8 | 5,595 | mg/L |
| Chloride | 19,025 | 75 | 35,500 | 18,450 | 89 | 33,900 | mg/L |
| Nitrate (as N) | <25 | <0.5 | <200 | <25 | <0.5 | <200 | mg/L |
| Nitrite (as N) | <25 | <0.5 | <200 | <25 | <0.5 | <200 | mg/L |
| Bromide | 52 | 0.3 | <100 | 63 | <0.25 | <100 | mg/L |
| Calcium | 385 | 0.6 | 722 | 385 | 0.2 | 735 | mg/L |
| Magnesium | 1,225 | 1.9 | 2,250 | 1,280 | 0.5 | 2,420 | mg/L |
| Hardness (as $CaCO_3$) | 6,007 | 8.8 | 11,068 | 6,232 | 2.6 | 11,801 | mg/L |
| Ca hardness (as $CaCO_3$) | 962 | 1.5 | 1,803 | 961 | 0.4 | 1,835 | mg/L |
| Sodium | 10,020 | 43.7 | 18,333 | 10,350 | 60.9 | 19,100 | mg/L |
| Potassium | 384 | 1.7 | 724 | 370 | 2.1 | 750 | mg/L |
| Fluoride | 0.9 | <0.1 | 1.4 | 0.9 | <0.1 | 1.5 | mg/L |
| Strontium | 6.9 | 0.012 | 13.2 | 6.9 | 0.004 | 12.9 | mg/L |
| Barium | <0.025 | <0.010 | <0.025 | <0.025 | <0.010 | <0.025 | mg/L |
| Boron | 3.4 | 0.5 | 6.2 | 3.5 | 0.92 | 6.5 | mg/L |
| Silica | <10 | <1 | <10 | <10 | <1 | <10 | mg/L |
| Ammonia (as N) | <0.1 | <0.1 | <0.1 | <0.1 | <0.1 | <0.1 | mg/L |
| TOC | 1 | <0.5 | 2.1 | 0.8 | <0.5 | 1.8 | mg/L |

Figures 56 and 57 illustrate the Dow and Hydranautics performance from June – September 2007 with respect to both raw and normalized conductivity (normalized for flow and temperature variations). The temperature of the post-condenser effluent water varied greatly when the powerplant was operating, with temperatures reaching 100 °F at times. When the temperature of the feed water is elevated, salt passage through the membrane increased, resulting in increased overall raw conductivity values as seen in figures 56 and 57. These raw values were then normalized to account for fluctuations in temperature in order to properly trend the conductivity of the RO permeate. The conductivity values for both the Dow SW30HRLE and Hydranautics SWC4+ are lower than the manufacturer's projections across the broad temperature range.

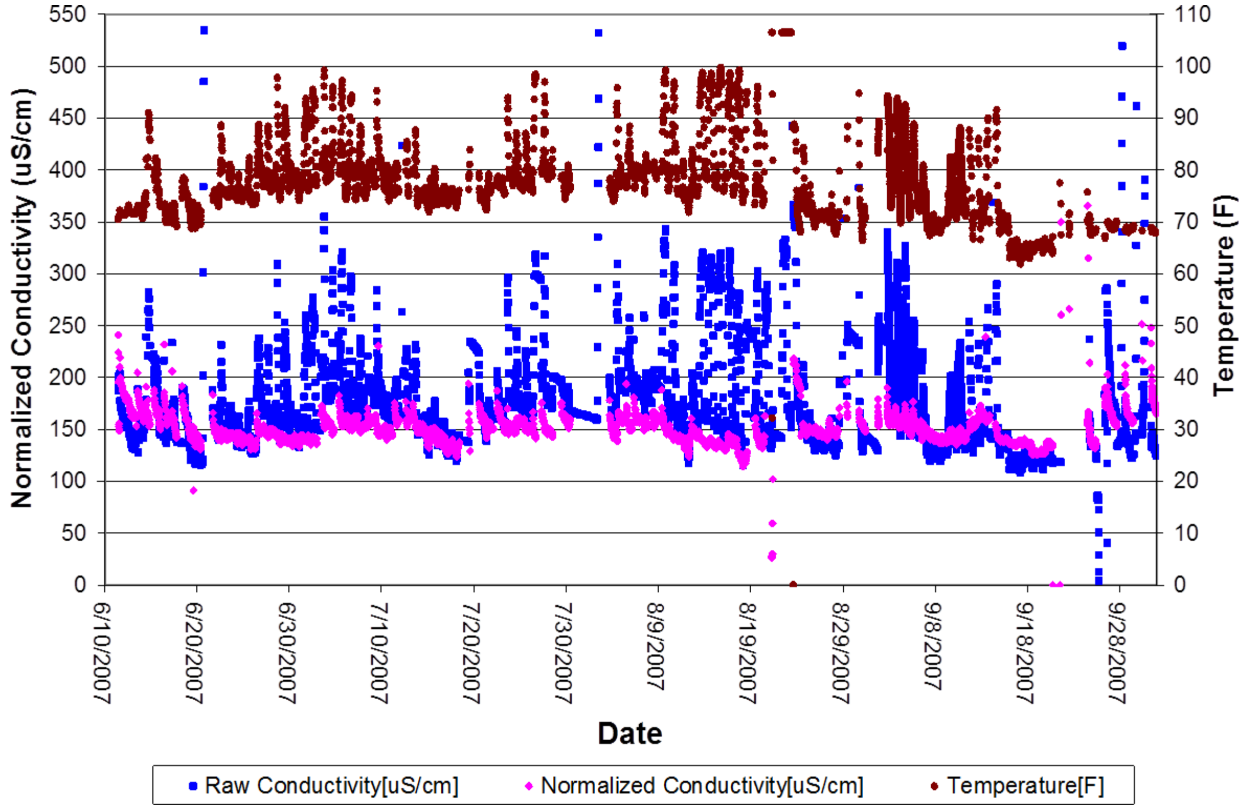



**Figure 56.—Dow SW30HRLE permeate conductivity, June 2007 – September 2007.**

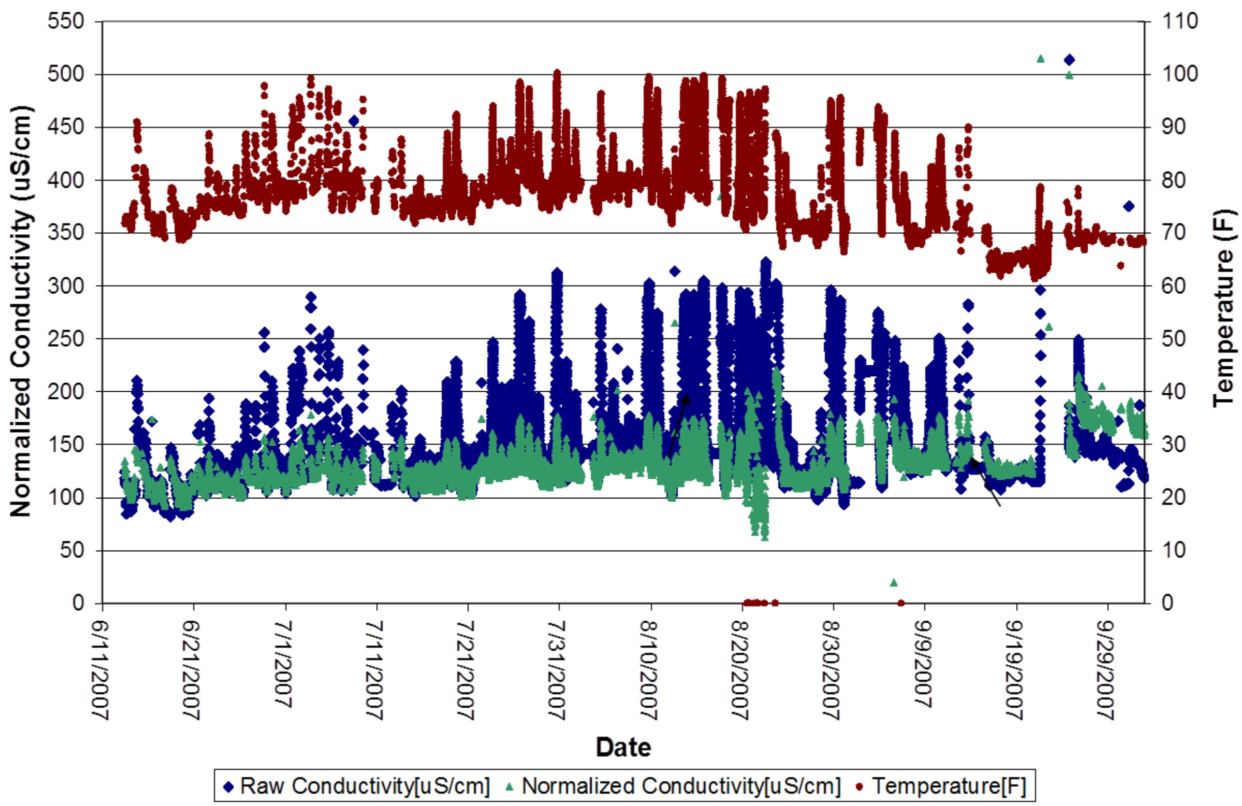



**Figure 57.—Hydranautics SWC4+ permeate conductivity, June 2007 – September 2007.**

Water samples were collected throughout the period of testing for detailed analyses. The flux rate of the RO membranes was varied to 8, 10, and 12 GFD to obtain data on permeate water quality at these different flux rates. At each flux rate, two sets of samples were collected, and the average data is shown in tables 34 and 35 below. The TDS, chloride, and boron concentrations are also compared to the manufacturers' projected performance at those conditions. Both the Dow and Hydranautics membranes lost some of their permeability after startup, which had some impact on overall rejection characteristics as well. This could be part of the reason for seeing actual RO permeate values significantly lower than projected in certain instances.

### *Algal Toxins*

Another important water quality aspect of ocean water desalination has to do with the presence of algal toxins in the water. One such toxin produced by the marine diatom *Pseudonitschia* is domoic acid, which can cause Amnesic Shellfish Poisoning (ASP) in humans and has been responsible for the death of marine mammals such as sea lions and seals along the southern California coast. This toxin accumulates in shellfish and small fish such as sardines and anchovies, which when consumed by humans and sea mammals can result in ASP.

As part of the pilot study, samples of raw water and RO permeate were collected regularly and analyzed for the presence of domoic acid by the University of Southern California. Figure 58 shows levels of particulate and dissolved domoic acid present the raw ocean water for Phase B of testing. Not once during Phase A or Phase B of testing did domoic acid appear in RO permeate. This is to be expected since the domoic acid molecule ( molecular weight 311) is large enough to be rejected by the RO membrane.

## RO Summary

The RO membranes tested operated effectively at 8 to 12 GFD flux rate on either MF or UF filtrate.

Phase A testing was based on the established seawater RO membranes available at the time, Hydranautics SWC1-4040 and Dow SW30-4040. Each of these membranes demonstrated the capability of providing permeate water with less than 300 mg/L TDS from the influent water throughout its temperature range.

Phase B testing provided valuable information on four next-generation RO membranes. Of the membranes tested—Dow SW30HRLE, Hydranautics SWC4+, Toray TM810, and Koch 1820SS—all but the Koch product warranted consideration for further testing. The lower boron rejection and lower permeability of the Koch were the major factors for this membrane not being considered for Phase B3 testing. Each of the "next-generation" membranes tested demonstrated the capability to produce water with less than 200 mg/L TDS across the influent temperature range, and less than 300 mg/L across the effluent temperature range.

Table 34.—Dow Average Feed Water and Permeate Water Quality, June to August 2007

| Parameter | Filmtec, 8 GFD (Avg. Temp. 25.2 °C) | | | Filmtec, 10 GFD (Avg. Temp. 28.3 °C) | | | Filmtec, 12 GFD (Avg. Temp. 22.2 °C) | | | Units |
|---|---|---|---|---|---|---|---|---|---|---|
| | RO Feed | RO Permeate | Projected Permeate | RO Feed | RO Permeate | Projected Permeate | RO Feed | RO Permeate | Projected Permeate | |
| TDS | 37,000 | 107 | 262 | 38,500 | 105 | 260 | 36,000 | 64 | 139 | mg/L |
| Lab pH | 8.1 | 7.1 | | 8.2 | 7.1 | | 8.2 | 7.3 | | pH |
| Alkalinity (as $CaCO_3$) | 113 | <2 | | 115 | <2 | | 116 | <2 | | mg/L |
| Bicarbonate (as $CaCO_3$) | 112 | <2 | | 113 | <2 | | 114 | <2 | | mg/L |
| Carbonate (as $CaCO_3$) | 1.3 | <0.1 | | 1.5 | <0.1 | | 1.7 | <0.1 | | mg/L |
| Hydroxide (as $CaCO_3$) | 0.06 | <0.01 | | 0.071 | <0.01 | | 0.08 | <0.01 | | mg/L |
| Sulfate | 2,580 | 2.5 | | 2,590 | 2.5 | | 2,630 | 2.5 | | mg/L |
| Chloride | 19,450 | 60.2 | 153 | 19,100 | 61 | 152 | 19,350 | 38.3 | 81 | mg/L |
| Nitrate (as N) | <25 | <0.1 | | <25 | <0.1 | | <25 | <0.1 | | mg/L |
| Nitrite (as N) | <25 | <0.1 | | <25 | <0.1 | | <25 | <0.1 | | mg/L |
| Bromide | 67 | <0.2 | | 58 | <0.2 | | 61 | <0.2 | | mg/L |
| Calcium | 422 | 0.29 | | 419 | 0.24 | | 416 | 0.2 | | mg/L |
| Magnesium | 1,335 | 0.94 | | 1,355 | 0.83 | | 1,240 | 0.6 | | mg/L |
| Hardness (as $CaCO_3$) | 6,551 | 4.6 | | 6,626 | 4 | | 6,144 | 3.2 | | mg/L |
| Ca Hardness (as $CaCO_3$) | 1,054 | 0.7 | | 1,046 | 0.6 | | 1,038 | 0.5 | | mg/L |
| Sodium | 11,000 | 38.7 | | 11,100 | 38 | | 10,300 | 22.7 | | mg/L |
| Potassium | 409 | 1.51 | | 416 | 1.5 | | 392 | 0.9 | | mg/L |
| Fluoride | 0.85 | <0.1 | | 1 | <0.1 | | 0.9 | <0.1 | | mg/L |
| Strontium | | | | 8 | 0.0048 | | | | | mg/L |
| Barium | <0.025 | <0.010 | | <0.025 | <0.010 | | <0.025 | <0.010 | | mg/L |
| Boron | 4 | 0.6 | 0.92 | 3.9 | 0.63 | 0.87 | 4.1 | 0.35 | 0.59 | mg/L |
| Silica | <10 | <1 | | <10 | <1 | | <10 | <1 | | mg/L |
| Ammonia (as N) | <0.1 | <0.1 | | <0.1 | <0.1 | | <0.1 | <0.1 | | mg/L |
| TOC | 3.4 | <0.5 | | 3 | <0.5 | | 3 | <0.5 | | mg/L |

Table 35.—Hydranautics Average Feed Water and Permeate Water Quality, June to August 2007

| Parameter | Hydranautics, 8 GFD (Avg. Temp. 25.2 °C) | | | Hydranautics, 10 GFD (Avg. Temp. 28.3 °C) | | | Hydranautics, 12 GFD (Avg. Temp. 22.2 °C) | | | Units |
|---|---|---|---|---|---|---|---|---|---|---|
| | RO Feed | RO Permeate | Projected Permeate | RO Feed | RO Permeate | Projected Permeate | RO Feed | RO Permeate | Projected Permeate | |
| TDS | 37,000 | 91 | 194 | 38,500 | 91 | 169 | 36,000 | 58 | 111 | mg/L |
| Lab pH | 8.1 | 6.3 | | 8.2 | 6.3 | | 8.2 | 6.4 | | pH |
| Alkalinity (as $CaCO_3$) | 113 | <2 | | 115 | <2 | | 116 | <2 | | mg/L |
| Bicarbonate (as $CaCO_3$) | 112 | <2 | | 113 | <2 | | 114 | <2 | | mg/L |
| Carbonate (as $CaCO_3$) | 1.3 | <0.1 | | 1.5 | <0.1 | | 1.7 | <0.1 | | mg/L |
| Hydroxide (as $CaCO_3$) | 0.06 | <0.01 | | 0.071 | <0.01 | | 0.08 | <0.01 | | mg/L |
| Sulfate | 2,580 | <2 | | 2,590 | <2 | | 2,630 | <2 | | mg/L |
| Chloride | 19,450 | 51 | 113 | 19,100 | 49 | 99 | 19,350 | 31 | 65 | mg/L |
| Nitrate (as N) | <25 | <0.1 | | <25 | <0.1 | | <25 | <0.1 | | mg/L |
| Nitrite (as N) | <25 | <0.1 | | <25 | <0.1 | | <25 | <0.1 | | mg/L |
| Bromide | 67 | <0.2 | | 58 | <0.2 | | 61 | <0.2 | | mg/L |
| Calcium | 422 | 0.12 | | 419 | 0.13 | | 416 | 0.1 | | mg/L |
| Magnesium | 1,335 | 0.39 | | 1,355 | 0.32 | | 1,240 | 0.3 | | mg/L |
| Hardness (as $CaCO_3$) | 6,551 | 1.9 | | 6,626 | 1.5 | | 6,144 | 1.6 | | mg/L |
| Ca Hardness (as $CaCO_3$) | 1,054 | 0.3 | | 1,046 | 0.3 | | 1,038 | 0.7 | | mg/L |
| Sodium | 11,000 | 32 | | 11,100 | 31 | | 10,300 | 18.4 | | mg/L |
| Potassium | 409 | 1.49 | | 416 | 1.4 | | 392 | 0.8 | | mg/L |
| Fluoride | 0.85 | <0.1 | | 1 | <0.1 | | 0.9 | <0.1 | | mg/L |
| Strontium | | | | 8 | 0.0023 | | | | | mg/L |
| Barium | <0.025 | <0.010 | | <0.025 | <0.010 | | <0.025 | <0.010 | | mg/L |
| Boron | 4 | 0.63 | 0.57 | 3.9 | 0.67 | 0.49 | 4.1 | 0.29 | 0.35 | mg/L |
| Silica | <10 | <1 | | <10 | <1 | | <10 | <1 | | mg/L |
| Ammonia (as N) | <0.1 | <0.1 | | <0.1 | <0.1 | | <0.1 | <0.1 | | mg/L |
| TOC | 3.4 | <0.5 | | 3 | <0.5 | | 3 | <0.5 | | mg/L |

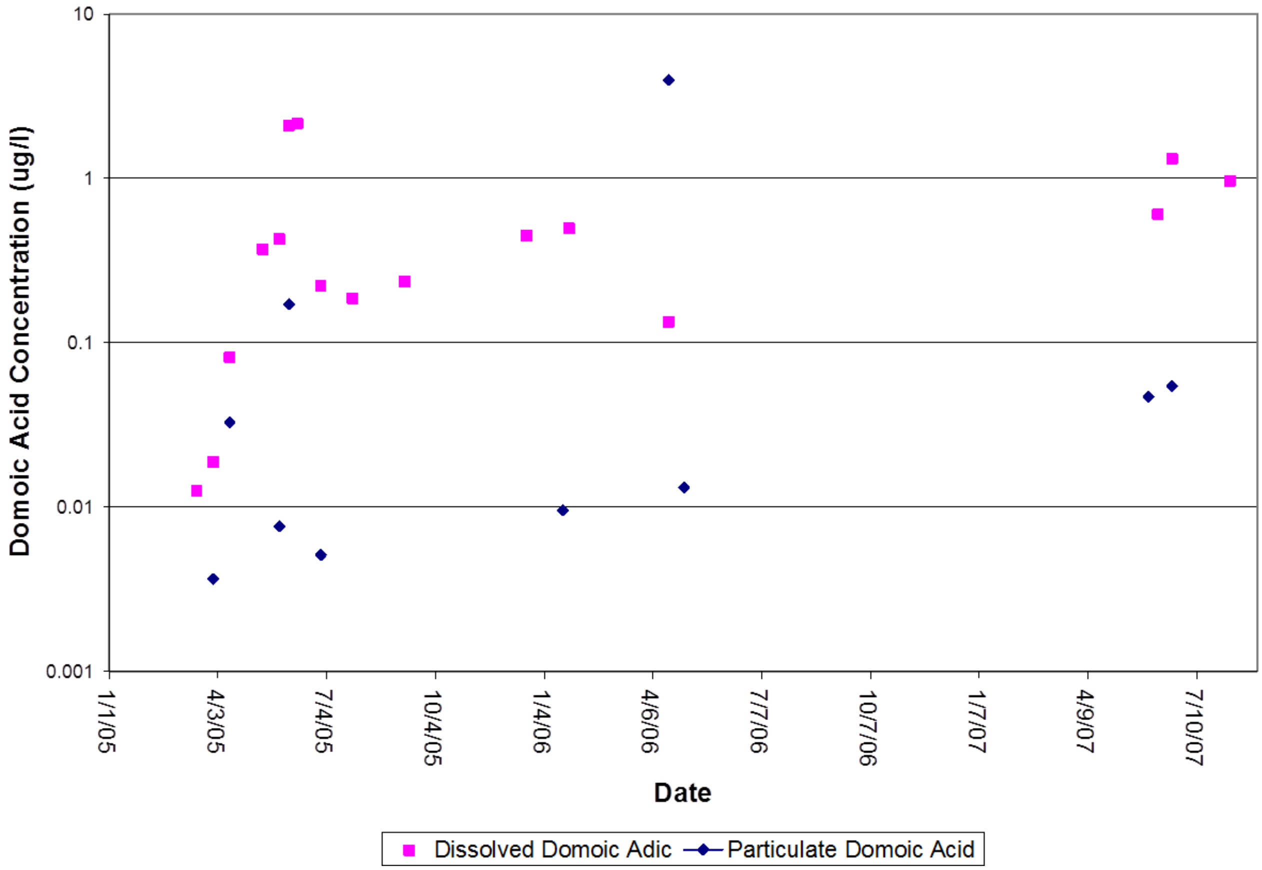



**Figure 58.—Domoic acid levels in ocean water, 2005–2007.**

The difference in chloride rejection versus boron rejection among the membranes tested was unexpected and noteworthy for those developing full-scale implementation of ocean water RO. The Hydranautics SWC4+ achieved permeate boron concentrations similar to those of the Dow and Toray membranes, but produced substantially lower chloride concentrations, albeit at higher operating pressure. This membrane would be of interest in those projects where chloride is the critical constituent for meeting treatment objectives.

Phase B also provided operational data on powerplant influent and the warmer powerplant effluent stream. Operation at the higher temperatures resulted in higher permeate concentrations and lower feed pressure requirements, as to be expected. The magnitude of these changes was shown above in figure 53.

Operation of the powerplant during the summer months coincided with an increase in algal biomass in the ocean. The increased abundance of marine microorganisms combined with the elevated temperatures of the post-condenser effluent seemed to exacerbate RO biofouling. Cleaning trials over the course of Phase B testing indicated that high-pH cleaning formulations with a pH of 12 are necessary to remove some forms of biogrowth. Note that the Phase A testing required no RO cleanings. The main difference between Phase A and B was the water temperature.  Operating on warmer effluent water increased the biofouling of the RO membranes.

The presence and removal of the algal toxin domoic acid by RO membranes was investigated during Phase B.  The RO membranes showed excellent removal of both particulate and dissolved domoic acid from the raw ocean water.  None was detected in any of the permeate samples tested, even when levels of domoic acid in the feed water were considered high relative to average concentrations.  The lower detection limit in the test for presence of domoic acid is 0.002 μg/L.

# PROCESS AND EQUIPMENT CHALLENGES

## Bromamines vs. Chloramines and the Oxidation of the RO Membranes

Membrane processes are susceptible to a phenomenon called membrane "fouling." Fouling, quite simply, is the loss of water permeability or throughput due to the accumulation of one or more foreign substance on the surface of the membrane (AWWA, 1999). As a result of the loss of permeability, fouled membranes require more pressure than clean membranes to produce an equivalent amount of product water. Fouling rates are typically the driving factor in the selection of the operating flux of a membrane system. One of the primary goals of this pilot study is to assess the membrane fouling rates at different operating fluxes.

Previous ocean water microfiltration testing demonstrated that the addition of chlorine to the feed water enhanced the microfiltration membrane performance. The chlorine or oxidant inactivates the microorganisms that can foul the MF membranes. However, thin-film reverse osmosis membranes contain polymers that are destroyed by strong oxidants such as free chlorine. In many previously established ocean water RO installations having open intakes with conventional filtration pretreatment, a reducing agent, such as sodium bisulfite is added after significant chlorine contact time to neutralize the oxidant before it contacts the RO membranes. However, this continuous chlorination/dechlorination process has been shown to actually enhance the tendency towards biological fouling of the RO membrane (Hamida and Moch, 1996).

Many MF/RO membrane facilities operating on wastewater use a different approach to control membrane fouling. In these facilities, chlorine is added to the feed water to enhance the membrane performance. Ammonia, naturally occurring or added to the wastewater, combines with the chlorine to form chloramines. The intent is to have a combined oxidant that would improve the fouling rate of both the MF and RO processes. This chloramination to MF to RO process has been used successfully on many wastewater reclamation facilities, including WBMWD's 20-million-gallon-per-day water recycling plant, located 2 miles east of this study's test site. The ammonia reacts with free chlorine or HOCl to form chloramines. The following reactions apply:

Reaction 1, Addition of sodium hypochlorite:

$$NaOCl + H_2O \rightarrow HOCl + NaOH$$

Reaction 2, Formation of monochloramine:

$$NH_4OH + HOCl \rightarrow NH_2Cl + 2H_2O$$

Reaction 3, Formation of dichloramine:

$$NH_2Cl + HOCl \rightarrow NHCl_2 + 2H_2O$$

Reaction 4, Formation of trichloramine:

$$NHCl_2 + HOCl \rightarrow NCl_3 + 2H_2O$$

Chloramines are weaker oxidants than HOCl or $OCl^-$ (free chlorine), and RO membranes are tolerant of a few mg/L chloramines. Furthermore, it has been demonstrated that the presence of chloramines in the water enhances the membrane performance by inhibiting membrane fouling.

This chloramination process was attempted on ocean water during this study. However, two items complicated the formation of chloramine on this water source. First, ammonia is not present in ocean water and thus must be added. Second, the presence of bromide ($Br^-$) in ocean water interferes with the reactions above. The Pacific Ocean water source used in this study has ~64 mg/L of $Br^-$. $Br^-$ substitutes for $Cl^-$ in reactions 1–4 listed above such that the chlorine addition to ocean water actually produces hypobromous acid (HOBr) instead of HOCl. Furthermore, subsequent ammonia addition creates bromamines instead of chloramines due to chemical kinetics. The following reactions apply:

Reaction 5, Addition of NaOCl to ocean water:

$$NaOCl + Br^- \rightarrow HOBr + Cl^-$$

Reaction 6, side reaction with chloramines:

$$NH_2Cl + Br^- + 2H_2O \rightarrow HOBr + NH_4OH + Cl^-$$

Reaction 7, subsequent ammonia addition:

$$NH_4OH + HOBr \rightarrow NH_2Br + 2H_2O$$

Reaction 8, dibromamine formation:

$$NH_2Br + HOBr \rightarrow NHBr_2 + H_2O$$ (White, 1999)

To protect the RO membranes from oxidation by HOBr, the molar ratio of $NH_3$:HOCl addition should be about 2:1 or greater. A 1 mg/L NaOCl addition and subsequent 1mg/L $NH_4OH$ addition utilized in this pilot study represents an $NH_3$:HOCl molar ratio of 2.1:1. However, HOBr and bromamines are stronger oxidants than their chlorine equivalents, HOCl and chloramines. There was little information or data on the exposure of thin film composite reverse osmosis membranes to bromamines. The chloramination process was selected for this study to determine the success of enhancing the MF/RO desalination operation on open intake ocean water.

As depicted in figures 38 and 39, this chlorination, followed my MF, followed by ammonia addition, followed by RO process failed to protect the RO membranes from oxidation. The specific flux and permeate conductivity of RO Train #1

(Dow membranes) started rising almost immediately. Train 2 (Hydranautics) proved to be more resistant, but after ~100 days of operation it was clear that the salt passage or permeate conductivity of this membrane was rising as well. On September 1, 2002 the $NH_4OH$ addition rate was increased by 50 percent to 1.5 mg/L. This did not alleviate the problem, and the permeate conductivity continued to rise. In response to the RO deterioration, the continuous chlorination in front of the MF was discontinued on October 3. Subsequently, attempts were made to run without any chlorine in the process, and rapid MF fouling was observed. Chlorine in the 20–40 mg/L range was then used in the MF backwash, an intermittent operation. An additional “rinse” step was added to the MF backwash to ensure no chlorine was carried over to the RO. This operational scheme, combined with the addition of sodium bisulfite in front of the RO was used in the remainder of the trials.

## Powerplant Heat Treatment Cycles

The pilot trails were started in June 2002. Soon thereafter, the powerplant performed a heat treatment or “heat treat” cycle. Approximately every one to three months the powerplant influent that feeds the pilot equipment was “heat treated” to control biological growth/attachment. The heat treat consists of recirculation of ocean water at 105–120 °F. During the heat treatment, barnacles/shells and organic matter die and are removed from the walls of the process piping. The pilot plant is turned off during this time to prevent this material and the high-temperature water from reaching the membrane systems. However, there is a significant “release period” after the end of the heat treatment, during which shells and other particulate matter are discharged from the piping walls. This caused repeated clogging of the booster pump impeller as well as the pilot feed line and resulted in shutdowns of the pilot process. To alleviate this problem, the 800-μm strainer was relocated to a position in front of the booster pump. However, some of the particulate matter was small enough to pass through the 800-μm strainer, the booster pump, and the 500-μm strainer on the Siemens CMF-S unit. This particulate matter was discovered in the feed distribution channel in one of the autopsied CMF-S modules, and was believed to be the cause of some, but not all, of the fiber breakage that occurred in the first set of MF modules.

## Addition of Arkal Spin Klin Filter

Fiber breakage occurred many times in the Siemens CMF-S module, even after the 800-μm strainer was placed in front of the booster pump as described in the appendix. Siemens undertook a redesign of their PVDF modules during this test period. The redesigned modules had fewer, thicker fibers in an attempt to make them more robust. In October 2003, these more robust membranes were placed in the Siemens CMF-S system. In addition, the 500-μm strainer located in front of the CMF-S system was replaced by an Arkal Spin Klin 130-μm self-backwashing filter. The Arkal Spin Klin is an innovative all-plastic filter that utilizes diagonally grooved polypropylene discs to create a depth filtration system with intersecting grooves that trap solids. The system uses an air-enhanced backwash process to periodically remove the solids. The following installation problems were experienced:

1. A single compressor was used to feed the air for the Arkal backwash and the Siemens unit. The air demand was too large for the compressor and when the Arkal went into backwash, the Siemens CMF-S system would shut down because of low air pressure.
2. The Arkal discs are color coded according to micron size. The original intent was to have 130-µm discs. The system was sent with 30-µm discs, and the small size of the grooves in combination with the low air pressure resulted in clogging and high differential pressures.

As a result of these challenges, the Arkal filter was bypassed for a period of time and the CMF-S Filter system, incorporating the redesigned modules, was run on water strained only with the 800-µm filter. Fiber breakage events occurred and more modules required replacement. The Arkal filter was finally placed in operation in March 2004 and has proven to be an effective pretreatment method to prevent damage to the hollow fiber membranes.

While the Arkal filter generally provided reliable operation, one operational challenge was biogrowth, which occurred on the discs and inside the housing during times of high biological activity in the feed water. Figure 59 shows one of the most severe biogrowth events experienced with the Arkal filters.

**Figure 59.—Biogrowth in Arkal filter housing.**

This level of biogrowth restricted the flow of ocean water through the discs and caused high differential pressures. A proposed solution to remedy the biofouling issue is to periodically backwash the disc filters with chlorinated water. The presence of chlorine in the backwash water should minimize biogrowth on the discs and in the housing.

## Vibration Issues Associated with Wanner Hydracell High-Pressure RO Pumps

The RO System used for this study has two independent trains. Each train has two pressure vessels operating in series, with three elements in the lead vessel and the four in the trailing vessel, for a total of seven 4-inch-diameter seawater elements. To feed the seven RO membranes in series, the RO pumps produce ~10 gpm at 1,000 psig, and this flow/pressure combination was not readily available in a centrifugal pump. Wanner Engineering offers a positive displacement type pump with super austenitic stainless steel wetted parts that withstand the corrosive ocean water environment. These Hydracell pumps have three pistons that are alternately moved by a wobble plate. The pistons are filled with oil on their return stroke. The oil balances the back side of the diaphragms, causing them to flex forward and back as the wobble plate moves. This provides the pumping action.

These pumps were advertised as having smooth, low pulse output, and the original design of the RO skid had them placed on the frame with the other equipment. Rigid super austenitic stainless steel piping was used to connect the pump discharges with the pressure vessels, as the engineers had experience with flexible hose failures at 1,000 psig. Vibration produced by the Hydracell pumps was accentuated by the combination of having pumps placed on the skid and being rigidly plumbed to the pressure vessels. This caused many problems with the system including:

- The pumps repeatedly lost their alignment and had to be realigned. One of the two pumps had to be rebuilt, as the bearings were destroyed by misalignment.
- Components on the skid vibrated at high frequency, resulting in failures of the Victaulic couplings, fittings, and piping.

After numerous equipment failures on the RO, Wanner was consulted and the following corrections were made:

1. The pumps were removed and placed adjacent to the RO skid, anchored to a concrete base.
2. Variable frequency drives were added to the pumps to lower the motor speed and eliminate the loop that recycled excess water back to the suction of the system.

These changes helped alleviate the vibration on the skid itself, but pipe and Victaulic coupling failures still occurred between the pumps and the pressure vessels. Pulsation dampeners were added to the discharge of the Hydracell pumps, but vibration problems persisted. Additionally, one of the pumps had a diaphragm leak, and the lubrication oil was introduced into the ocean water and ended up irreversibly fouling a set of membranes.

In August of 2006 one of the Hydracell Pumps was replaced with a relatively new pump on the market manufactured by Danfoss. The new pump, model number APP 2.2, is a positive displacement axial piston pump constructed of duplex stainless steel, making it corrosion resistant to ocean water. The pump is lubricated by the ocean water, not oil, so there is no possibility of oil leaking into the ocean water and fouling the RO membranes. The pump produces very little vibration, does not require a pulsation dampener, and is controlled with a variable frequency drive. The second Hydracell pump was replaced with an additional APP 2.2 in May 2007 when the pilot equipment was relocated. Both Danfoss pumps have performed very well since installation.

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# Appendix A

## Seawater Desalination Pilot Plant Project
## Microfiltration/Reverse Osmosis Pilot Testing Protocol

West Basin Municipal
Water District

# *West Basin Municipal Water District*

## *Seawater Desalination Pilot Plant Project*

## *Microfiltration/Reverse Osmosis Pilot Testing Protocol*

***May 2002***

***Prepared for the West Basin Municipal Water District by:***

*Separation Processes, Inc.*

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# 1 INTRODUCTION

Seawater desalination will eventually play a significant role in the water supply equation for Southern California. To date, the use of seawater desalination in California has been minimal, primarily due to relatively high cost. Recently, with improved performance and costs, microfiltration (MF) has been proposed as an alternative to conventional pretreatment processes for seawater reverse osmosis (RO). Microfiltration (MF) has become a common pretreatment method for RO installations treating municipal wastewater. However, at the present time there is no pilot plant or demonstration plant program to evaluate the combination of MF and RO for potential application of seawater desalination in California for domestic water supply.

West Basin Municipal Water District's (WBMWD) Seawater Desalination Pilot Plant Program will test the capabilities of MF pretreatment in series with a spiral wound RO system. It will develop data to determine the optimum operating conditions and cleaning requirements for microfiltration operating on seawater, as well as the seawater reverse osmosis process operating on microfitration filtrate. Data will also support development of updated cost assessments of seawater desalination.

The Seawater Desalination Plant will be operated at the El Segundo Power Generation Plant in El Segundo, CA. The research project will take 12 months to complete. This protocol document covers the seven (7) month operational period of the pilot program's total twelve (12) month period.

## 2 OBJECTIVES

The objectives of the *Seawater Pilot Test Program* are:

1. To determine the optimum membrane operating flux and in-situ membrane cleaning frequency for a MF system operating on Southern California coastal seawater. Investigate cleaning formulations and techniques for removal of contaminants found in seawater, which foul the MF membrane.

2. To determine the optimum membrane operating flux and in-situ membrane cleaning frequency for a seawater RO system operating on MF filtrate. Investigate cleaning formulations and techniques for removal of contaminants found in microfiltered seawater, which foul RO membranes.

3. Characterize the MF backwash and RO concentrate streams to develop data suitable for evaluation of waste stream disposal options.

4. Develop design and operating parameters based on the above data to assess cost of seawater desalination for the purpose of production of drinking water by MF/RO.

# 3 PILOT PLANT TREATMENT PROCESS AND EQUIPMENT DESCRIPTION

The pilot plant will be located on the California coast in the City of El Segundo at the El Segundo Power Generation Plant. Seawater is brought through an existing open intake to the power plant cooling system (≈200 mgd). Existing treatment by the power station consists of a course traveling screen (>1 inch) and intermittent chlorination. Chlorination is manually initiated approximately three times per week for a duration of two hours. The addition rate is that which results in a total chlorine concentration at the plant outfall (condenser effluent) of approximately 0.06 mg/L. Approximately every six weeks the power plant cooling loop is "heat treated" to control biological growth/attachment. Duration of this treatment is one hour at $105^{O}$ – $120^{O}$F.

The seawater supply to the pilot desalination process will be a side-stream taken from either the influent or effluent of the power plant Bearing Cooling Water (BCW) heat exchangers. The BCW system's seawater influent is from the main cooling stream which feeds the power plant condensers. Arranging for tie-in to the condenser effluent piping for a warm water source would be physically difficult due to limited exposed piping and logistically difficult due to the need to tie-in at four locations to ensure continuous supply. For this reason the BCW heat exchanger influent and effluent have been selected as the desalination plant feed source. These streams are available at approximately 15 psi at temperatures which fluctuate seasonally (57-$68^{O}$F cool water supply, $86^{O}$-$100^{O}$F heated supply). Operation described in this test plan will use the cool water supply (influent to BCW heat exchanger). A tie-in is also available to provide water from the effluent from the BCW system as an alternate source, should heated seawater be determined to be of interest for subsequent testing. Cool temperature operation of the reverse osmosis system will provide a lower concentration permeate than the warm temperature operation. Based on local project criteria, the benefit of this superior permeate quality is considered to outweigh the cost savings provided by warm water's lower operating pressure. The MF feedwater is Pacific Ocean seawater with approximate characteristics as indicated in Table 1.

The overall pilot treatment process is indicated in the Process Flow Diagram (Figure 1). The first component of the pilot treatment process is a transfer pump which provides sufficient head for delivery of seawater through a 1,000 micron duplex basket strainer to the microfiltration system. The ON/OFF operation of the transfer pump is controlled by the MF system. The strainer design allows cleaning of one basket while the other is in operation, without interruption of the treatment process. Prior to the microfiltration system is a flow paced sodium hypochlorite addition system. The addition system includes a 35 gallon day tank, containing 12.5% NaOCl. Sodium hypochlorite is added to maintain a 1 mg/L chlorine residual through the microfiltration process. Data from MF pilot operation at other seawater pilot sites indicated a benefit to the MF performance due to the presence of free chlorine.

The MF system is a US Filter CMF-S system, utilizing 0.2 micron nominal pore size pvdf hollow fiber technology. The pvdf membrane chemistry has a high tolerance of chlorine and other oxidants, providing wide range of options for control of biological growth within the system and prevention of membrane fouling due to organic matter. The CMF-S process consists of four modules submerged in a process tank. Suction is applied to the lumen of the fibers by the MF filtrate pump, drawing water though the walls of the fibers. Particulate matter is accumulated on the outside surface of the fibers. The CMF-S process includes periodic interruption of filtration for backwashing of the fibers. The filtration period is preselected in the range of 15-30 minutes. Following the filtration period, the fibers are backwashed by reversing the filtrate flow and introducing an air scour across the membrane's outside surface. Subsequently, the process tank is drained to the Combined Effluent Tank and refilled. A critical MF process parameter is the operating flux (filtrate flow per unit area of membrane). Initially the MF will be operated at filtrate flow setpoint of 20 gpm (5 gpm per module; 21.5 gfd instantaneous flux). Adjustments to the flux and backwash frequency will be made per the test plan (section 5 of this document)

MF filtrate is directed to the 150 gallon covered Break Tank, which serves as an equalization tank between the intermittent MF production and the continuous flow RO process. Prior to entry into the Break Tank, provision is made for chemical addition to the MF filtrate stream. The chemical metering pump is suitable for addition of either ammonium hydroxide or sodium bisulfite, for chloramine formation or dechlorination, respectively. Elimination of free chlorine is necessary to protect the polyamide RO membrane, which is subject to damage from exposure to strong oxidants. Initial operation of the pilot will include addition of ammonium hydroxide at this location. The ammonium hydroxide dose will be based on a 3:1 ratio $Cl_2$:$NH_3$-N. This ratio will provide a small excess of ammonia to ensure elimination of free chlorine. The RO membranes have tolerance to low concentrations of chloramine, but minimal tolerance to free chlorine.

MF filtrate will be pumped from the Break Tank, by a booster pump to the RO system. The booster pump discharge at approximately 35 psi, delivering RO feedwater through cartridge prefilters and providing sufficient suction pressure to the RO high pressure pump. Excess MF filtrate (2-8 gpm) will overflow the Break Tank to the Combined Effluent Tank. Operation of the RO booster pump will be controlled by the RO-PLC and require adequate level in the break tank.

Antiscalant and acid addition points are located downstream of the RO booster pump. Antiscalant addition equipment will include a day tank and chemical metering pump. Dose adjustment will be manual as the RO operation is continuous and will be maintained to flow setpoints. The initial dose is anticipated to be 1.5 mg/l. A 10 micron cartridge filter follows acid and antiscalant addition, providing mixing and a barrier to debris introduced at the break tank. Operation in the existing test plan does not require acid addition.

Following cartridge filtration the stream splits to feed two identical RO units (Train 1 & Train 2). Each train consists of a high pressure pump feeding two, four-inch diameter pressure vessels in series. Each vessel is capable of holding four elements in series. During this study a spacer assembly will be used in one vessel to allow operation of seven elements in series. Concentrate flow is manually adjusted to the flow setpoint using the concentrate control valve. The RO units have positive-displacement high-pressure pumps. Therefore, permeate flow is manually adjusted to a setpoint using the H.P. pump recycle control valve. The RO system includes ancillary cleaning and flush systems. Upon shutdown the RO system is automatically flushed with RO permeate.

All process streams including MF backwash, excess MF filtrate, RO permeate and RO brine will be recombined at the Combined Effluent Tank and discharged by a transfer pump to the seawater intake forebay, for eventual discharge to the ocean outfall. MF and RO CIP waste discharges will be collected for proper disposal by El Segundo Power.

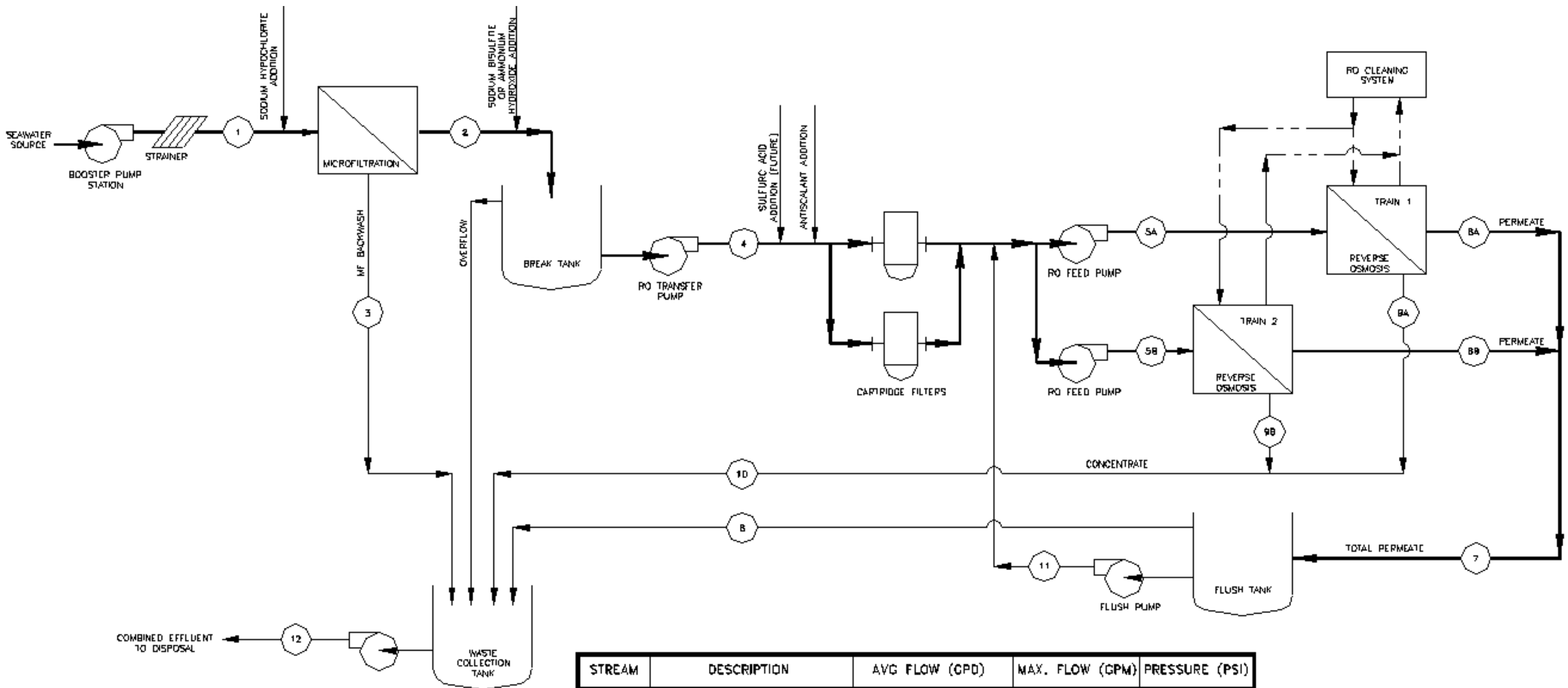
***Table 1.***

***Anticipated Typical Seawater***

| Constituent | Value |
|---|---|
| Calcium | 400 |
| Magnesium | 1,260 |
| Sodium | 10,400 |
| Potassium | 380 |
| Ammonia | <1 |
| Barium | 0.02 |
| Strontium | 8 |
| Bicarbonate | 160 |
| Sulfate | 2,640 |
| Chloride | 18,900 |
| Bromide | 65 |
| Boron | 4.4 |
| Nitrate | <1 |
| Fluoride | 1 |
| Silica | 1 |
| Total Dissolved Solids | ≈34,000 |
| PH | 8.0 |
| TOC | |
| Temperature ($^{O}$F) | 62 (57-68) |

All values in mg/L except pH and temp

## *Figure 1. Process Flow Diagram of the West Basin Seawater Desalination Pilot System*




| STREAM | DESCRIPTION | AVG FLOW (GPD) | MAX. FLOW (GPM) | PRESSURE (PSI) |
|---|---|---|---|---|
| 1 | MF FEED | 26,250 – 42,600 | 40.0 | 25 |
| 2 | MF FILTRATE | 23,000 – 41,000 | 32.0 | <10 |
| 3 | MF BACKWASH | 1,760 – 3,260 | 40.0 | GRAVITY |
| 4 | RO LOW PRESSURE FEED | 22,400 – 28,600 | 18 0 | 40 |
| 5 | RO HIGH PRESSURE FEED | 11,200 – 13,300 | 9.0 | 900–1,000 |
| 6 | RO TRAIN PERMEATE | 5,600 – 6,720 | 4 5 | 10 |
| 7 | TOTAL RO PERMEATE | 11,200 – 13,300 | 9.0 | 10 |
| 8 | TOTAL RO PERMEATE TO WASTE | 11,200 – 13,300 | 9.0 | GRAVITY |
| 9 | RO TRAIN CONCENTRATE | 5,600 – 6,720 | 4 5 | 10 |
| 10 | TOTAL RO CONCENTRATE | 11,200 – 13,300 | 9.0 | 10 |
| 11 | RO FLUSH | N/A | 10.0 | 15 |
| 12 | COMBINED EFFLUENT | 26,250 – 42,600 | 40.0 | TBD |

### *Table 2a. Chemical Addition Dose*

| Chemical | Expected Dose | Controlling Parameter |
|---|---|---|
| Sodium Hypochlorite | 1-3 mg/L | Free Chlorine concentration in MF Feed and Flow Paced |
| Ammonium Hydroxide (i) | 1-3 mg/L | Ratio Based on MF Filtrate Free Chlorine Concentration and Flow Paced |
| Sodium Bisulfite (i) | 3-9 mg/L | " |
| RO Antiscalant (ii) | 1.5 mg/L | RO Feed Flow |

Footnotes:
(i) Either Ammonium Hydroxide or Sodium Bisulfite will be used, not both.
(ii) Permatreat 191 by Nalco Chemical Corporation

### *Table 2b. Chemical Dilution and Addition Rate*

| Chemical | Bulk Chemical Concentration as Purchased | Day Tank Dilution | Pump Output at Typical Flow of Receiving Stream |
|---|---|---|---|
| Sodium Hypochlorite | 10% | 1:4 (2%) | 12 ml/min at 20 gpm |
| Ammonium Hydroxide (i) | 5% | 1:4 (1%) | 6 ml/min at 20 gpm |
| RO Antiscalant (ii) | 100% | 1:199 (0.5%) | 11 ml/min at 12 gpm |

# 4 PROGRAM SCHEDULE

The Seawater Pilot Testing Program is a twelve-month program. The operational period of the program is anticipated to last approximately seven months. The sequence of events are as follows:

**Phase 1** – Design & Equipment Procurement (January 1, 2002 – February 28, 2002)

During Phase 1 the treatment equipment will be specified including detailed design of the RO pilot system. Contracts will be established for procurement and lease of all equipment. Additionally the pilot test protocol is prepared.

**Phase 2** – Site construction & equipment installation (March 1, 2002 – May 31, 2002)

Phase 2 includes preparation of the site and installation of the equipment. Startup and commissioning of the equipment will occur during this period.

**Phase 3** **-** Pilot testing **-** (June 1, 2002 – December 31, 2002)

Phase 3 is the program's period of structured testing, which is the subject of this protocol document. If MF or RO equipment installation is completed early, Phase 3 will begin as soon as possible.

# 5 TEST PLAN

## 5.1 Responsibilities

### 5.1.1 General

This Pilot Testing Program is under the management of West Basin Municipal Water District (West Basin). The following entities are participants, with responsibilities as noted below. Separation Processes, Inc. (SPI) will be the test engineer. United Water Services (UWS) will the testing site operator. During the testing program, SPI will direct changes to the system operating conditions in accordance with the guidelines of the test protocol. Changes to operating conditions will be communicated in writing to the UWS operations supervisor. The UWS supervisor or appropriate staff will maintain the pilot operating conditions, equipment, etc., in accordance with the operating guidelines established by the Test Protocol and SPI. These procedures will ensure integrity of the data. Documentation of sampling, chain of custody, and QA/QC will be maintained by UWS for all water quality samples.

The West Basin will retain control of all data from the test program. Operating data, water quality data and other testing results will not be disclosed by the project participants to other outside parties without approval of West Basin.

### 5.1.2 West Basin Responsibilities

West Basin is the owner and/or lease holder of the treatment equipment providing overall management of the study.

### 5.1.3 United Water Services Responsibilities

UWS has the following responsisbities

1. Provide operating labor for day-to-day operations, including data collection, recharge of chemical day tanks and normal adjustment of controls.
2. Maintain study's primary record set and distribute as instructed by the protocol and West Basin.
3. Collect samples and provide laboratory analyses per this protocol.
4. Be the primary contact with NRG during the execution of this study.
5. Procure expendable items during the study, including chemicals for addition and CIP, cartridge filters, etc.
6. Remove used CIP solutions for off-site disposal

### 5.1.4 Separation Processes Responsibilities

Separation Processes, Inc. is providing engineering services including test protocol development, equipment specification, data monitoring to ensuring compliance with testing protocol and decisions regarding changes to process operating conditions, data analysis and reporting.

### 5.1.5 El Segundo Power Responsibilities

El Segundo Power, LLC is the owner of the test site at the El Segundo Power Plant. They will provide space, electrical power, feed water and discharge stream disposal.

### 5.1.6 Equipment Contractor Responsibilities

An Equipment Contractor will be retained to provide and install all treatment process equipment. The exception being the MF equipment is provided by the West Basin through lease from US Filter. The Equipment contractor will be responsible for installation of the MF system in coordination with US Filter personnel and in accordance with US Filter instructions.

## *5.2 Operating Plan*

Operation of both the MF and RO processes will occur throughout Phase 3 of this study. While the operation of the RO requires operation of the MF to supply feedwater, the operating plans for each process are otherwise independent.

### 5.2.1 MF process operating plan

The CMF-S system will be operated initially at the conditions indicated in table 3. Over the course of operation the trans-membrane pressure (TMP) is expected to rise gradually, due to deposition of matter on the fibers, which are not completely removed by normal backwash. The system will be taken out of service when the TMP has risen to the maximum limit value. At that time a Clean in Place (CIP) procedure will be performed to remove the accumulated solids from the fiber surface, which normal backwashing has been unable to remove. The period of operation from initial operation until maximum TMP is reached is the CIP interval. A goal of this pilot test is to determine the maximum flux and/or backwash interval, which will allow a CIP interval of at least 21 days. Based on the CIP period achieved at the initial operating conditions, adjustments will be made to the operating conditions of subsequent runs. These changes will be directed by SPI and communicated in writing to UWS. Likewise the effectiveness of the CIP procedure will be assessed by observation of the post-CIP TMP value. Modifications of the CIP procedure will be made to improve CIP performance or eliminate steps that are unnecessary. The system will be manually taken off line when the Power Plant performs the periodic high temperature treatment of the plant intake piping. UWS will be responsible to coordinate with the plant personnel to schedule pilot equipment shutdowns as needed.

***Table 3. MF Operating Conditions***

| | |
|---|---|
| Instantaneous Flux | 21.5 gfd |
| Instantaneous filtrate flow | 20 gpm |
| Backwash interval | 15 minutes |
| Maximum TMP | 12 psi |
| Backwash chemical addition | None |
| NaOCl addition | Dose sufficient to maintain 1 mg/L free chlorine in the MF feed stream |
| Clean-in-place (CIP) cleaning procedure | US Filter standard two step Acid/Hypochlorite procedure as published in their operation literature. |

### 5.2.2 RO operating plan

As previously discussed, a primary objective of the study is the determination of the maximum operating flux, which allows operation for 30 days or more between chemical cleanings. The RO system will be operated initially at the conditions indicated in table 4. As shown, the initial operating flux will be 8 gfd. This level will be maintained for 1,000 hours of operation. Chemical cleaning of the membrane will be performed if the Specific Flux declines from its initial value by 20%, if the Normalized Differential Pressure increases by 25% or upon 1,000 hours of operation, which ever occurs first. If the period of operation before cleaning exceeds 30 days, the operating conditions after cleaning will be adjusted by increasing the operating flux by 1 gfd. This adjustment of operating flux will continue for subsequent runs up to 12 gfd. If the Specific Flux is extremely stable during a run, the Engineer may elect to increase the operating flux of the subsequent run by more than 1 gfd.

If the period of operation between cleanings is less than 30 days, the same operating flux will be repeated for the following run. A cleaning period of less than 30 days for the second run will result in a decrease of operating flux by 1 gfd for the subsequent run, but not less than 8 gfd.

The initial cleaning formulation and technique will be the generic procedure recommended by the membrane manufacturer for seawater desalination applications. Should this not be successful, variation in the formulation will be tried. A successful cleaning if defined as one that recovers the Specific Flux to 95% of its initial stabilized value.

If RO membrane performance is so stable that the engineer elects to increase the operating flux by more than 1 gfd in a single increment, assessment of performance at 12 gfd may occur well before the end of the test period. At the engineer's discretion the RO membrane may be replaced by an alternate RO membrane.

***Table 4. Initial RO Operating Conditions***

| | Train 1 | Train 2 |
|---|---|---|
| Membrane manufacturer | Dow FilmTec | Hydranautics |
| Membrane Element Model | SW30-4040 | SWC1-4040 |
| Quantity of elements | 7 | 7 |
| Element active membrane area ($ft^2$) | 80 | 70 |
| Total active membrane area ($ft^2$) | 560 | 490 |
| Initial Permeate flow (gpm) | 3.1 | 2.7 |
| Initial Flux (gfd) | 8 | |
| RO recovery | 50% | |
| Ammonium Hydroxide addition | Dose sufficient to maintain 3:1 ratio of MF filtrate free chlorine concentration to ammonia-N addition. | |
| Antiscalant Addition | 1.5 mg/L Nalco PermaTreat 191 | |
| Clean-in-place (CIP) criteria | 20% loss of initial Specific Flux or 25% increase in normalized Differential Pressure | |
| Clean-in-place (CIP) procedure | Membrane manufacturer's generic formulation and procedure | |

## *5.3 Standard Sampling Methods*

To ensure the accuracy of all collected data, consistent sampling methods with respect to location, timing, and the technique must be maintained. Additionally, for samples analyzed at off-site laboratories, consistency in sample preservation, packaging and shipping is required. Membrane operational parameters such as flow, pressure, and time since last backwash will be recorded at the time of sampling. All analyses will be performed according to Standard Methods[1].

## 5.4 Data Handling Protocol [2]

Successful implementation of the performance testing will require coordination between all testing participants. All performance testing activities will be thoroughly documented. Documentation will include field logbooks, photographs, data sheets, electronic databases and chain-of-custody forms.

Original field sheets and chain-of-custody forms will accompany all samples shipped to the analytical laboratory. Copies of field sheets and chain-of-custody forms for all samples will be maintained in the project files. The data management system used in the pilot testing program will involve the use of computer spreadsheets and manual recording of operational parameters for the membrane equipment on a daily basis.

Where applicable, the MF and RO electronic data loggers will be used for automatic entry of testing data into computer databases. Specific portions of the MF computer database for operational and water quality parameters will be downloaded by manual importation into Microsoft Excel weekly. In spreadsheet form, the data will be manipulated into a convenient framework to allow analysis of membrane equipment operation. At a minimum, backup of the computer databases to diskette will be performed on a weekly basis.

Daily measurements of all values, including those electronically data logged, will be recorded on specially-prepared data log sheets. An operating logbook will include a record of events (equipment starts, stops, maintenance, instrument calibrations) and description of any problems or issues Photocopies will be made of each data-log and operating logbook page. The original sheets will be stored on-site; one photocopy will be forwarded to the engineer (SPI) at least once per week during each testing phase; and a second photocopy will be placed in the project file. This protocol will not only facilitate referencing the original data, but offer protection of the original record of results. .

Each membrane test run will be assigned a run number, which will then be tied to the data from that experiment through each step of data entry and analysis. As samples are collected and sent to the laboratories, the data will be tracked by the same system of run numbers. The run number designation will indicate both MF and RO run (e.g. "Run MF2/RO4").

### 5.4.1 Data Collection

The data indicated in Tables 5, 6 & 7 will be recorded daily by the operator on data sheets (Appendix A). Values indicated with an asterisk are also recorded in the MF or RO data loggers. Water analyses schedule is indicated in Table 8.

***Table 5. MF Pretreatment Operating Data Requirements.***

| |
|---|
| Strainer inlet pressure (psi) |
| Strainer outlet pressure (psi) |
| NaOCl day tank level |

***Table 6. MF Operating Data Requirements.***

| |
|---|
| Filtrate Flow (gpm)* |
| TMP (psi)* |
| Temperature ($^{O}$C) * |
| Conductivity (μmho/cm) * |
| pH* |
| Run Time (hours) * |
| Backwash frequency setpoint* |
| Backwash flow (gpm) * |
| Backwash flow duration (sec) * |
| Backwash pressure (psi) * |
| Backwash chemical requirements |
| Air flow (cfm) * |
| MF feed turbidity (NTU) * |
| MF Filtrate turbidity (NTU) * |
| $NH_4OH$ day tank level |
| Pressure decay test start pressure (psi) * |
| Pressure decay test end pressure (psi) * |
| Pressure decay duration (sec) * |
| Pressure decay test result (psi/min) * |

## *Table 7. RO Operating Data Requirements.*

| Data requirements common for both trains |
|---|
| Run time (hours)* |
| Cartridge Filter inlet pressure (psi) |
| Cartridge Filter outlet pressure (psi) |
| Temperature ($^{O}$C) * |
| Feed Conductivity (μmho/cm) * |
| Feed pH* |
| Antiscalant day tank level |

| Data requirements for each train |
|---|
| Feed Pressure (psi)* |
| Interstage pressure (psi) |
| Concentrate Pressure (psi)* |
| Total Permeate flow (gpm) * |
| Bank 2 Permeate Flow (gpm)* |
| Concentrate flow (gpm) * |
| Permeate Conductivity (μmho/cm) * |
| Individual vessel permeate conductivity profile (μmho/cm) (weekly) |

***Table 8. Water Quality Parameters***

| Laboratory Analysis Frequency | | | | | | | | |
|---|---|---|---|---|---|---|---|---|
| **Parameter** | **Raw Feed** | **MF Feed** | **MF Filtrate** | **MF Backwash** | **Break Tank Influent** | **RO Feed** | **RO Permeate** | **RO Conc.** |
| Sampling location | S010 | S020 | S050 | S030 | S035 | S135 | S185-1&-2 | S175-1&-2 |
| pH | | Daily | -NA- | -NA- | | | | |
| Turbidity (NTU) | Daily | Daily | Weekly | Weekly | | Weekly | | |
| TOC (mg/L) | Weekly | Weekly | Weekly | Weekly | | | | |
| DOC (mg/L) | | Weekly | -NA- | -NA- | | | | |
| $UV_{254}$ ($cm^{-1}$) | Weekly | Weekly | Weekly | -NA- | | | | |
| Total Alkalinity (mg/L as $CaCO_3$) | | Weekly | Weekly | -NA- | | | | |
| Total hardness (mg/L as $CaCO_3$) | | Weekly | Weekly | -NA- | | | | |
| Calcium hardness (mg/L as $CaCO_3$) | | Weekly | Weekly | -NA- | | | | |
| Manganese (mg/L) | | Monthly | -NA- | -NA- | | | | |
| TDS (mg/L) | | Monthly | Monthly | Monthly | | | | |
| Free chlorine residual (mg/L) | | Daily | Daily | Daily | Daily | Daily | Weekly | Weekly |
| Total chlorine residual (mg/L) | | | | | Daily | Daily | Weekly | Weekly |
| Complete mineral analysis (constituents listed in Table 1) | | | | | | Biweekly | Biweekly | Biweekly |
| Silt Density Index (15 min) | | | | | | Weekly | | |
| Total Heterotrophic Plate Count | Daily | Daily | Daily | | | Weekly | | Weekly |

### *5.5 Membrane Integrity Verification*

In addition to on-line filtrate turbidity measurement, the CMF-S microfiltration system will automatically perform a pressure decay test (PDT) to assess the integrity of the membrane barrier. This test will be performed daily, requiring less than 5 minutes down time. Results of the PDT are recorded in the CMF-S data logger. Out of range results may indicatre a broken fiber and require troubleshooting efforts to identify and plug the damaged fiber.

RO membrane integrity is monitored by on-line conductivity measurement of RO permeate. Additionally, conductivity profiles are performed on the individual RO vessel permeate streams weekly.

# 6 PERFORMANCE EVALUATION

## 6.1 Parameters for Evaluation of Performance

The following calculated parameters, together with the data collection specified in the previous section, will be used for evaluation of performance of the membrane systems.

Microfiltration System

- Transmembrane pressure (psi)
- Specific Flux (gfd/psi)
- Period between chemical cleanings
- Filtrate Turbidity (NTU)

Reverse Osmosis System

- Specific Flux (gfd/psi)
- Normalized Differential Pressure (psi)
- Normalized Permeate Conductivity (μmho/cm)
- Period between chemical cleanings

# 7 REPORTING

## 7.1 Monthly Progress Reporting

During the course of pilot plant operation (Phase 3), SPI will prepare a brief monthly progress report of pilot activities and results.

## 7.2 Final Report

At the conclusion of the Phase 3 testing, SPI will compile the entire data set and prepare a final report of results.

# 8 QUALITY ASSURANCE/QUALITY CONTROL

Quality assurance and quality control of the operation of the membrane equipment and the measured water quality parameters will be maintained during the Performance Testing Program.

When specific items of equipment or instruments are used, the objective is to maintain the operation of the equipment or instructions within the ranges specified by the Manufacturer or by Standard Methods. Maintenance of strict QA/QC procedures is important, in that if a question arises when analyzing or interpreting data collected for a given experiment, it will be possible to verify exact conditions at the time of testing.

Equipment flowrates and associated signals will be documented and recorded on a routine basis. A routine daily walk through during testing will be established to verify that each piece of equipment or instrumentation is operating properly. Particular care will be taken to confirm that any chemicals are being fed at the defined flowrate into a flowstream that is operating at the expected flowrate, such that the chemical concentrations are correct. This will be accomplished through chemical drawdown measurements. In-line monitoring equipment such as flowmeters, etc. will be checked to confirm that the readout matches with the actual measurement (i.e. flowrate) and that the signal being recorded is correct. Flow measurement accuracy will be confirmed monthly by stopwatch-and-bucket techniques. The accuracy of on-line water quality instruments will be verified monthly by comparison to grab sample results using West Basin laboratory bench instrumentation. The items listed are in addition to any specified checks outlined in the analytical methods.

**REFERENCES:**

[1] American Public Health Association, American Water Works Association, Water Environment Federation. Standard Methods for the Examination of Water and Wastewater, 20th Ed. 1999.

[2] Adapted with permission from NSF International. Protocol for Equipment Verification Testing for Physical Removal of Microbiological and Particulate Contaminants. May 14, 1999.