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

Submitted by:

West Basin Municipal Water District
17140 Avalon Blvd.
Carson, CA 90746-1296

West Basin Municipal Water District

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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 power plant influent and effluent (post condenser) water sources. Numerous RO membranes were tested and demonstrated that water quality goals were achieved.

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Final Report

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Mission Statements

The mission of the Department of the Interior is to protect and provide access to our Nation's natural and cultural heritage and honor our trust responsibilities to Indian Tribes and our commitments to island communities.

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.

Contents

Glossary

CIP	Clean In Place
CMF-S	Continuous Microfiltration Submerged
gpm	Gallons Per Minute
GFD	Gallons Per Square Foot of Membrane Area per Day
H_2SO_4	Sulfuric Acid
MC	Maintenance Clean
MF	Microfiltration
mg/L	Milligrams per Liter
mgd	Million Gallons per Day
ng/L	Nannograms Per Liter
NaOCL	Sodium Hypochlorite
NH_3	Ammonia
NH_4OH	Ammonium Hydroxide
PDT	Pressure Decay Test
PVDF	Polyvinylidene fluoride
RO	Reverse Osmosis
SBS	Sodium Bisulfite
SCFM	Standard Cubic Feet per Minute
SDI	Silt Density Index
SEM	Scanning Electron Microscope

TDS	Total Dissolved Solids
TOC	Total Organic Carbon
UF	Ultrafiltration
WBMWD	West Basin Municipal Water District
μg/L	Micrograms per Liter
μm	Micrometers or micron
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 power plant intake water as well as 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 time frame 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 power plant 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 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 power plant influent water. These recently developed membranes had the highest boron rejection available.
- Phase B2 evaluated MF and next generation RO membranes on power plant effluent and the Zenon UF System on power plant influent.
- Phase B3 identified two of the four next generation RO membranes for longer-term testing and evaluated all systems on power plant effluent.

Phase B demonstrated that the optimized Phase A MF operating parameters on influent water were unchanged on the warmer effluent water source. Both the MF and UF produced excellent quality filtrate for use as RO feedwater. 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 algal bloom events (Red Tides) during the course of Phase B testing. Periodic testing for algal toxin domoic acid indicated no detection in any RO permeate samples, despite elevated concentrations in feedwater 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 the RO treatment process to be an excellent barrier to this constituent. However, the MF and UF systems did experience loss of permeability during the more severe algal blooms, which temporarily impacted their filtration capacity.

Data was collected on the "next generation" RO membranes which indicated improved performance (lower permeate concentrations of key constituents) over the previous versions tested in Phase A. Each of the "new-generation" RO membranes tested demonstrated the capability of providing permeate water less than 200 mg/L total dissolved solids (TDS) across the power plant influent water temperature range and less than 300 mg/L across the power plant effluent temperature range. Additionally differences were noted in salt rejection performance among the new membranes that 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 executed to evaluate the combination of MF and RO, as well as UF and RO, for the potential application of 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, as well as the ocean water reverse osmosis process operating on microfiltration filtrate. Phase A of this study consisted of microfiltration followed by reverse osmosis. In Phase B of this testing, an ultrafiltration system was added in parallel with the MF system and the results of the reverse osmosis operation were compared operating on feed water from the 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 co-located with an existing ocean water cooled power plant has advantages. One potential advantage is that power may be available at relatively low rates "within the fence" of the power plant. In California, this may result in an energy savings of about $0.05/kwh. 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 power plant and returned to the ocean. Furthermore, the salinity of the RO concentrate is reduced by blending with the power plant discharge water.

Utilization of 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 power plant or water that has already been used in the power plant 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 and exiting the power plant and the warmer return water. A membrane desalination plant operating on the warmer return water would have the advantage of a decreased energy usage associated with a decrease in water viscosity. On the other hand, the warmer water may promote bacteriological growth that may have a higher fouling potential for the membrane treatment processes, and the salinity of the treated water would be slightly higher. Phase A of this work included operation on the cooler power plant influent water. In Phase B, an ultrafiltration membrane process was added and the entire operation switched to the warmer power plant effluent water.

Figure 1 - Pilot Test Equipment at the El Segundo Power Plant

Conclusions

The following conclusions can be drawn from this five year study:

1. The study successfully established the feasibility of utilizing a MF/UF → RO process to produce potable quality water. This was demonstrated on Pacific Ocean water taken from either a power plant intake or the warmer power plant post-condenser effluent source.
2. Each of the "new-generation" RO membranes tested demonstrated the capability of providing permeate water 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. Reverse Osmosis membranes operated effectively at 8 to 12 GFD flux on MF and UF filtrate.
4. Analyses for Domoic Acid in the RO permeate indicated non-detect (less than 0.002 µg/L) results, even when elevated concentrations (2-3 µg/L) existed in the raw feedwater due to substantial algae bloom events.
5. Both the microfiltration backwash and reverse osmosis concentrate waste streams were characterized for disposal options.
6. The Siemens CMF-S microfiltration system:
    a. Confirmed that a flux of 34 GFD was sustainable on the influent feed source (as established in Phase A) and established that this same operating condition was optimum for operation on the effluent source.
    b. Chlorine addition to the backwash was utilized and 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. CIP frequency of every 3 weeks
    d. Required a periodic heated clean-in-place (CIP) to restore membrane permeability. Non-heated CIP's proved to be inadequate to restore the membrane permeability to within 10% of its original level. Successful CIP protocol included:
        i. 2% citric acid recirculation/aeration at 36 - 38ºC followed by
        ii. 400 – 600 mg/L NaOCl recirculation at 20 - 22ºC
    e. Produced filtrate water with turbidity and SDI 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 micron or less.
   g. It was necessary to reduce MF capacity by 25-30% during the most severe algae bloom (Red Tide) events.

7. The Zenon ZW-1000 Ultrafiltration system:
   a. Established a flux of 27.5 GFD was sustainable on the effluent source. While this flux was not demonstrated on the influent source it is expected, based on similarities in UF performance between the two sources at other operating conditions.
   b. Chlorine in the backwash and maintenance clean was utilized and critical to 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. Backwash with 4 mg/L NaOCl 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% 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 approximately 300 µS, fifty percent lower than SWC-4040. The first set of membranes suffered from membrane oxidation with the oxidation much more rapid on the Dow membranes. While not as severe, the second set of Dow membranes experienced a decrease of salt rejecting properties, 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% lower chloride concentration than the other membranes.

10. The continuous chlorination and subsequent ammonia dosing in an attempt to create chloramines proved to be unsuitable for full-scale implementation due to the creation of bromamine and resulting oxidation of the RO membranes. This process was replaced by MF backwash chlorination and continuous sodium bisulfite dosage in front of 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. Flux of 9 GFD may not be optimal for all ocean water sources.

12. Operation on ocean water from the common power plant influent introduced additional challenges for the treatment process. The power plant 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 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 power plant influent (Phase A) and power plant 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 on both power plant cooling loop influent water and warmer power plant cooling loop effluent water.

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 power plant cooling system (≈200 mgd). Existing treatment by the power station consists of a coarse traveling screen (>1 inch) and intermittent chlorination. Standard power plant practice consists of two treatment techniques for controlling organic activity in the cooling loop. Chlorination is manually initiated two times per week for 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) to the pilot plant feed water. Secondly, approximately every two to three months the power plant cooling water is "heat treated" to control biological growth/attachment. Duration of this treatment is one hour at 105 – 120 ^{O}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

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

All values are in mg/L except for pH and temperature

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 operation at other ocean water pilot sites indicated a benefit to the MF performance due to the presence of free chlorine.

The MF system was a Siemens CMF-S system, utilizing 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. Suction is applied to the lumen of the fibers by the MF filtrate pump; 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. At the start of the testing 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 membranes 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 filtrate flow setpoint of 20 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 dram 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 contains filtration time periods segregated by brief backwashes or backpulses. The Zenon ZW1000 system also utilizes a process called a maintenance clean (MC). The MC is a mini CIP where the unit is shut down for approximately thirty minutes and a chemical solution is recirculated.

MF and UF filtrate were directed to cover break tanks, which serve as equalization between the intermittent MF/UF production and the continuous flow RO process. Prior to entry into the break tank, provision was made for chemical addition to the MF filtrate stream. The chemical metering pump was suitable for the addition of either ammonium hydroxide or sodium bisulfite, for chloramine formation or dechlorination, respectively. 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 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 to low concentrations of chloramine, but minimal tolerance to free chlorine.

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 feedwater 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 addition (3mg/L) was injected downstream of the RO booster pump. 20 micron cartridge filters followed the antiscalant addition and provide mixing and a barrier to debris introduced at the break tank. No acid addition to the RO feed stream was used.

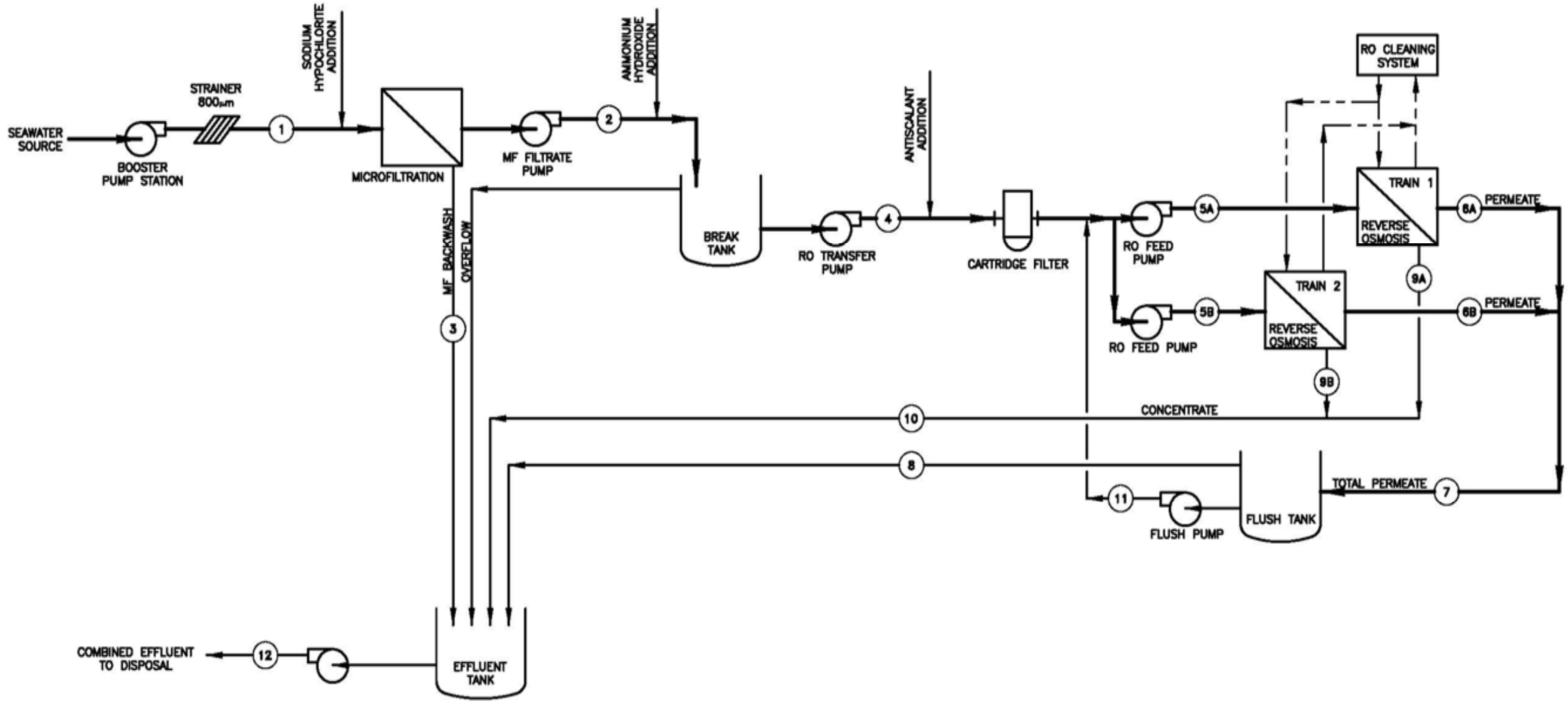
Following cartridge filtration the stream split to feed two identical RO units (Train 1 & Train 2). Each train consisted of a high-pressure pump feeding two, four-inch diameter pressure vessels in series. Each vessel was capable of holding four elements in series. During this study a spacer assembly 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 were experienced over the course of this study. Some of these, as described in the Process and Equipment Challenges section, required modifications to the process flow of the pilot equipment. Figure 3 below contains the revised Phase A testing process flow diagram for the pilot equipment. A discussion of the major issues that required process flow modification is included.

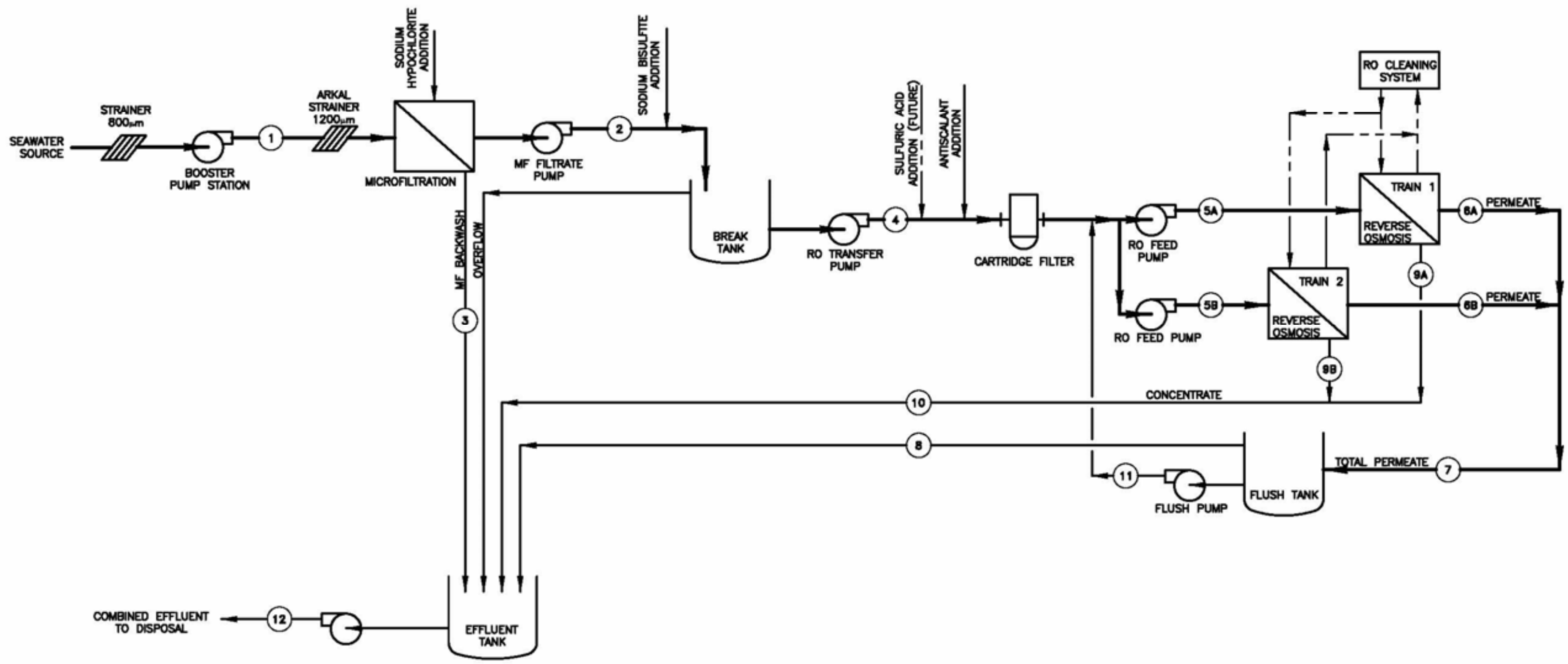
Phase B of the testing introduced the Zenon ZW1000 ultrafiltration system, and the ability to run the equipment on the warmer effluent water source. The phase B testing process flow diagram is included below in Figure 4.

Figure 2 - Initial Process Flow Diagram of the Pilot System



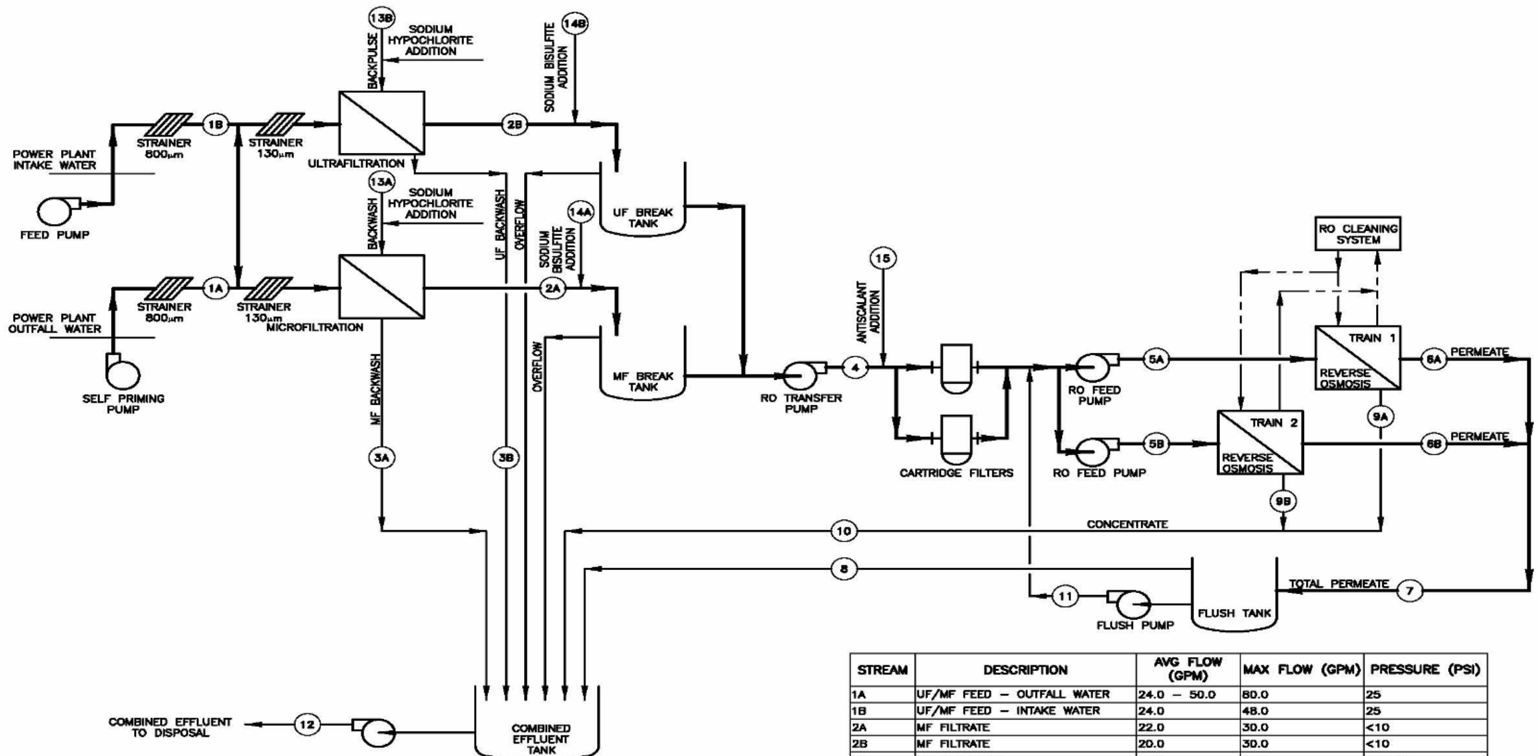
STREAM	DESCRIPTION	AVG FLOW (GPD)	MAX FLOW (GPM)	PRESSURE (PSI)
1	MF FEED	26,250 – 42,600	40.0	25
2	MF FILTRATE	23,000 – 41,000	32.0	<10
3	MF BACKWASH	1,760 – 3,280	40.0	GRAVITY
4	RO LOW PRESSURE FEED	22,400 – 28,600	18.0	40
5	RO HIGH PRESSURE FEED	11,200 – 13,300	9.0	900–1,000
6	RO TRAIN PERMEATE	5,800 – 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 (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 – 26,600	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

Figure 4 - Phase B Process Flow Diagram



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

Microfiltration Optimization and Performance

Phase A Testing

Figure 5 - Siemens CMF-S Microfiltration Pilot System

The Siemens CMF-S system operation was initiated 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 trails are summarized in the following 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.

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 run 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 oring. 1/15 Replaced a second original module that had a crack in the potting. 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	In advertant 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 BW	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 BW	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 BW	20	**Modules Replaced 5/28/04**
	MF 18	6/8/2004-	7314-	26	6.5	34	20 in every BW	20	CIP after very short run-modules reconditioned

Permeability of Original CMF-S Module Design

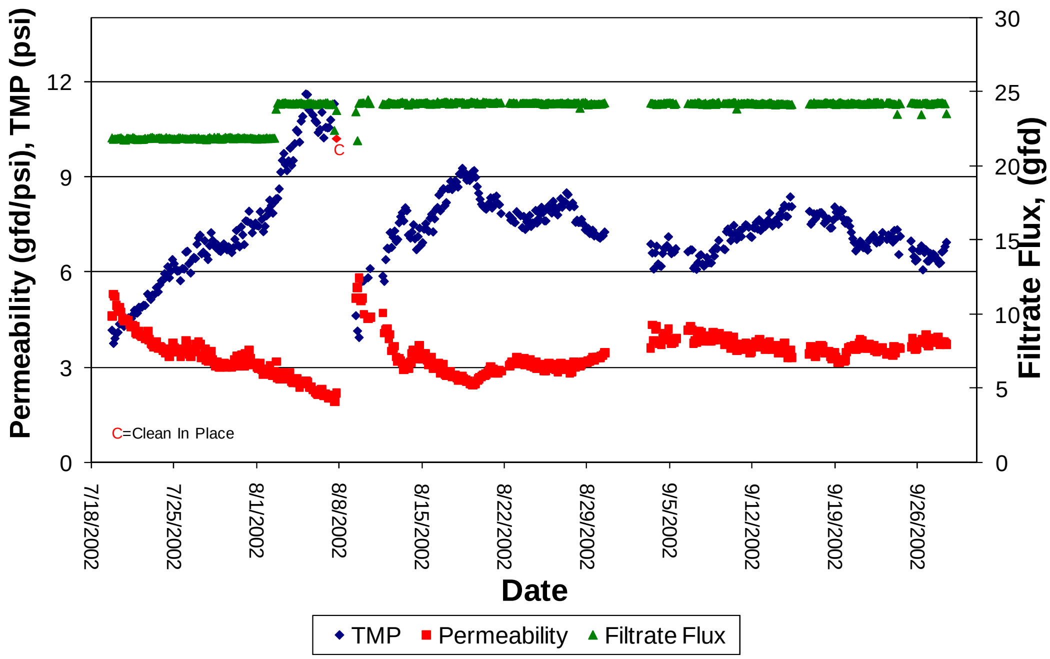
The Siemens CMF-S system runs at constant flux and thus as the membrane fouls, the trans-membrane pressure (TMP) required to maintain throughput rises. However, because transmembrane pressure is also influenced by water temperature and variations in flow, the appropriate method of monitoring membrane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

Permeability is the filtrate flux divided by the temperature corrected transmembrane pressure and is typically reported in units of GFD/psi. The terminal transmembrane pressure (transmembrane pressure where 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 feedwater 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 chlorination followed by MF followed by ammonia dosing process 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 that chloramines and the bromamine damaged the downstream RO membranes. This is discussed further in the Process and Equipment Challenges Section of this document.

Figure 6 - Performance of Microfiltration System with Continuous Prechlorination (MF Trial I)

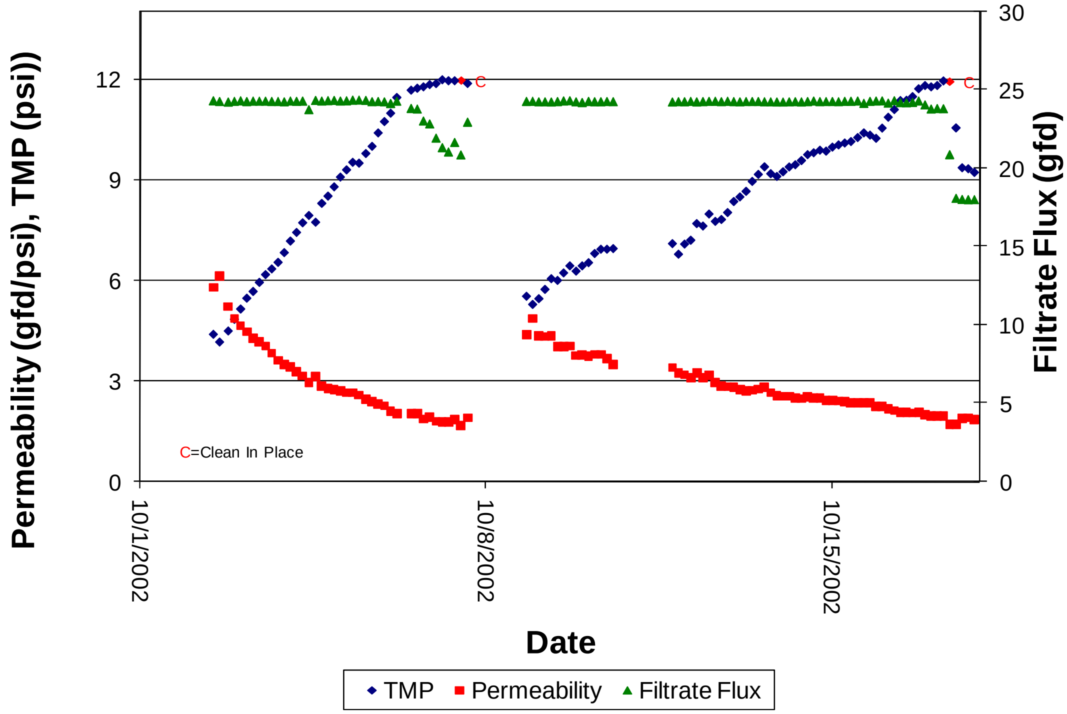


In many other ocean water RO installations on open intakes with conventional filtration pretreatment, a reducing agent, such as sodium bisulfite is added and significant chlorine contact time is allowed to neutralize the oxidant before it contacts the RO membranes. However, as demonstrated in a 1996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

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 ten days before reaching terminal permeability. Operation at 24 GFD was unsuccessful without the chloramination in the feedwater and demonstrated how beneficial the oxidant is to the stable performance of microfiltration membrane process on this feed source.

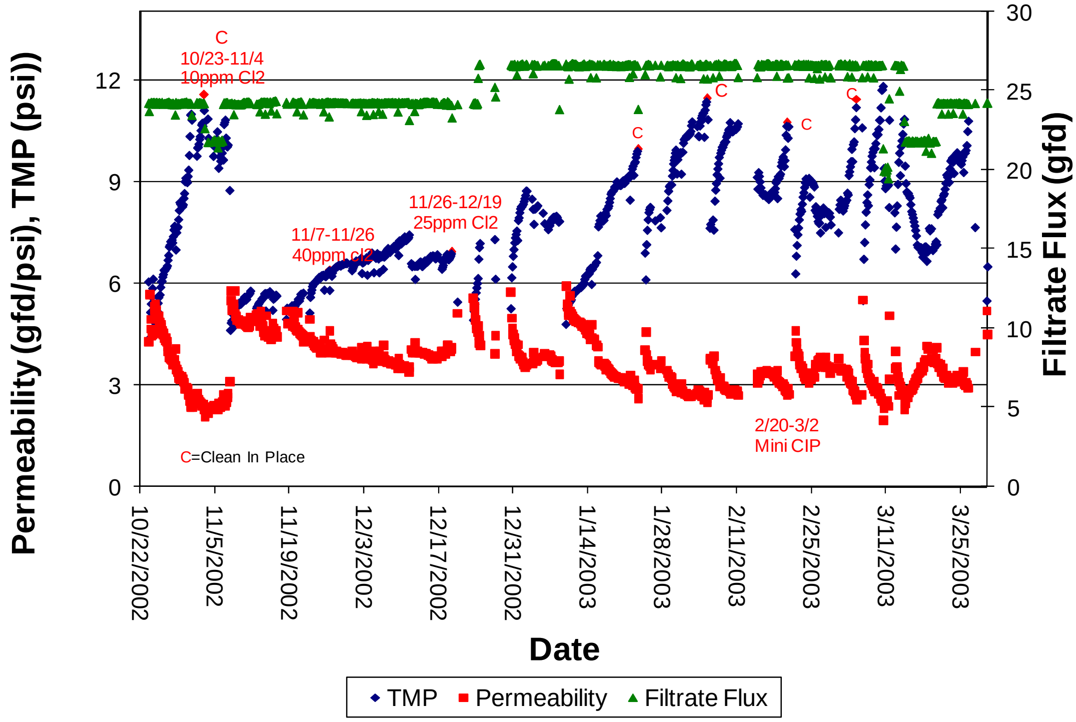
Figure 7 - Performance of Microfiltration System with No Chlorination (MF Trial II)



Trial III-Chlorinated Backwashes

Recognizing the benefit of chlorine to the MF process but accepting that the attempted chloramination of the feedwater (Trial I) presented an adverse impact on the RO membrane, an alternative approach to the use of chlorine was attempted in MF Trial III, chlorinated backwashes. NaOCl (10 mg/L) was attempted in 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 shut down for a chemical clean in place (CIP).

Figure 8 - Performance of Microfiltration System with Chlorinated Backwashes (MF Trial III)



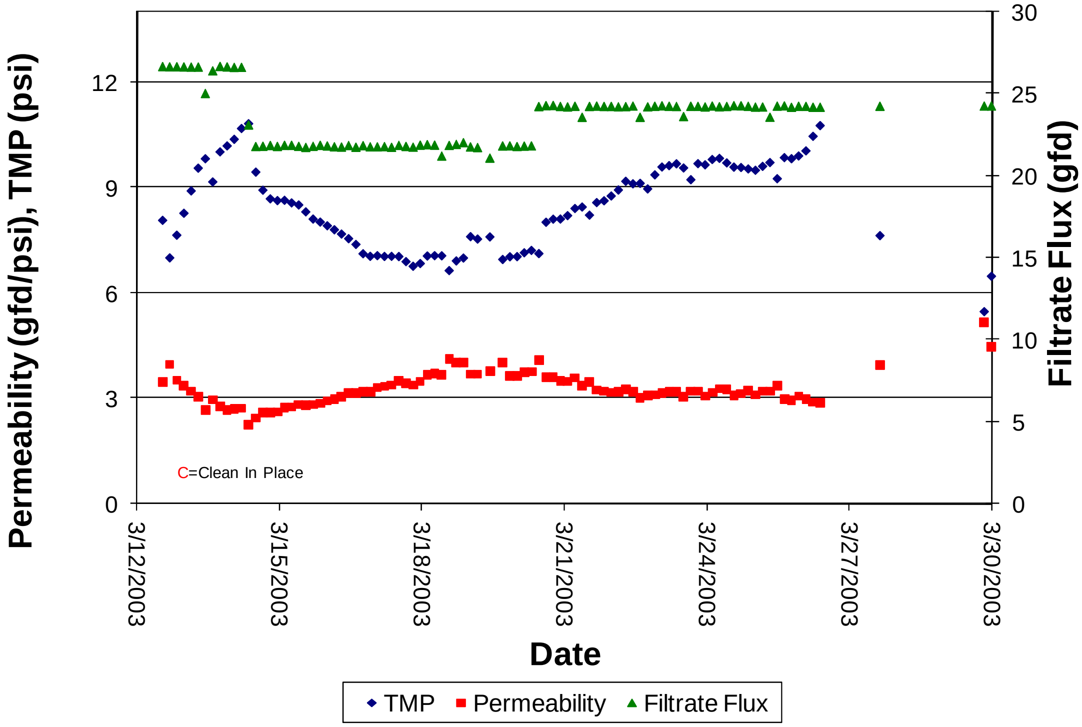
The filtrate flow was then increased from 22 gpm to 24 gpm for run #8, corresponding to a flux increase from 24 to 25.8 GFD. Numerous attempts failed to demonstrate a run time longer than 3 weeks at this flux before a CIP was required. This was compounded by the fact that the CMF-S clean-in-place (CIP) 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 resulted in an improvement in permeability. 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 GFD to 24 GFD with these original CMF-S membranes.

Cleaning Effectiveness

Examination of Figure 8 shows that the "clean" or post "Clean-In-Place" microfiltration permeability's had declined since January 23, 2003. This is a sign of an ineffective CIP procedure. The problem was initiated 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 permeability's are only 4 GFD/psi after the cold water cleanings, whereas with previous heated CIP's, the clean permeability's 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. Hydrochloric acid was utilized 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

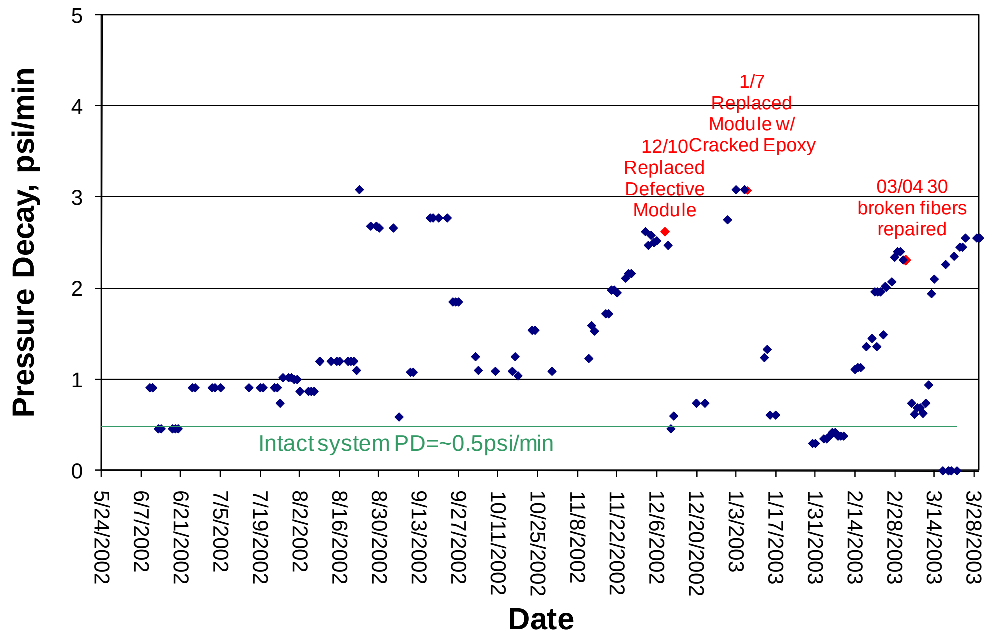
Step	Chemical	Temperature (ºC)	Procedure
1	2% Citric Acid	36 - 38	Perform reverse filtration until membrane cell is filled with MF Filtrate. Add chemicals, heat solution and aerate every 2 minutes. Perform filtrate recirculation for 30 minutes. Repeat 5 minute aeration/5 minute soak cycles 9 times.
2	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 numerous fiber breakage events, and the second on January 7, 2003 after it developed a crack in the epoxy that isolated the feed from the filtrate water. Furthermore, one of the replacement modules demonstrated fiber breakage events as well.

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 lumen at 15 psi, and then a valve on the feed side is opened to atmosphere. 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 drastically 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 quantitate the broken fiber problems observed during this study, on March 4, 2003, a pressure decay was performed on the system resulting in a decay rate of ~2.3psi/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 displays 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

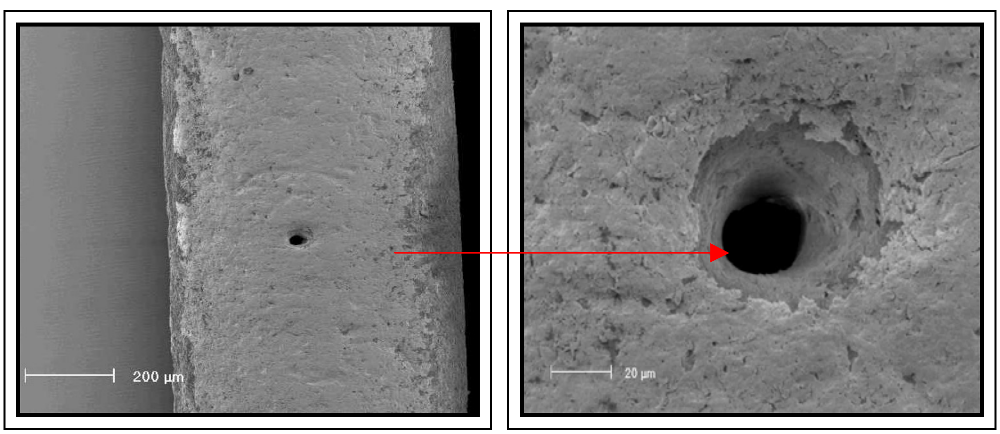


Siemens sent their problematic modules to Australia for autopsy to determine the cause of the fiber breakage and epoxy failures. The results from the analysis of the module with the cracked epoxy can be summed up as:

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. A 130 µm Arkal filter replaced the original 500 µm strainer in front of the MF 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%. SEM photographs showed that other broken fibers that had sheared appeared to have been stretched before failure.

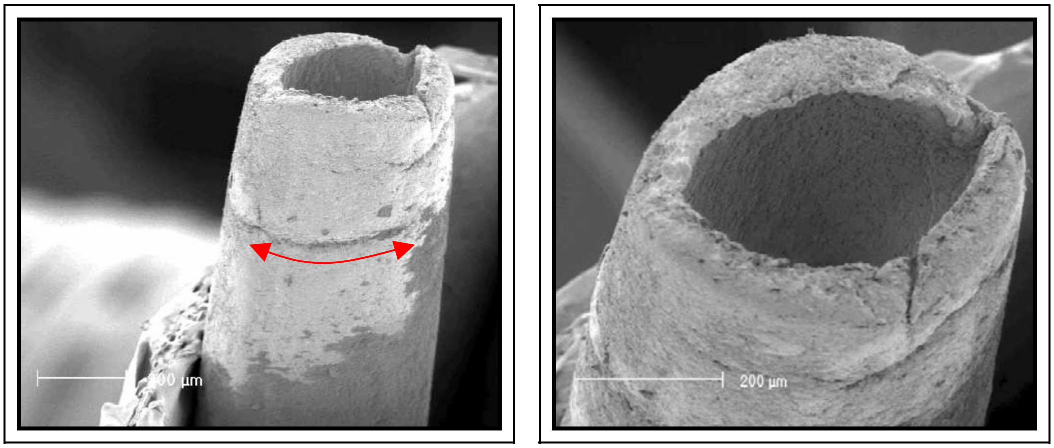
Figure 11 - Air Bubbles Emitted from the Cracked Epoxy During the Pressure Decay Test

Figure 12 - SEM Photographs of a Hole in a CMF-S Module Fiber



A hole in fibre found 490mm from the top. A closer look at the hole shows it appears to have been caused by a sharp object, or by something wearing into the fibre

Figure 13 - Sheared CMF-S Fiber Shows Evidence of Stretch Failure



Broken fibre found 350mm (fibre 2) from the bottom. The fibre has been bent and the surface appears stretched.

The fiber stretching and the fact that three of the six modules displayed no epoxy and very little fiber breakage problems, 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 including:

1. Larger fiber (diameter and wall thickness)
2. Smaller number of fibers in module (different packing density)
3. Reduced fiber area per module

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

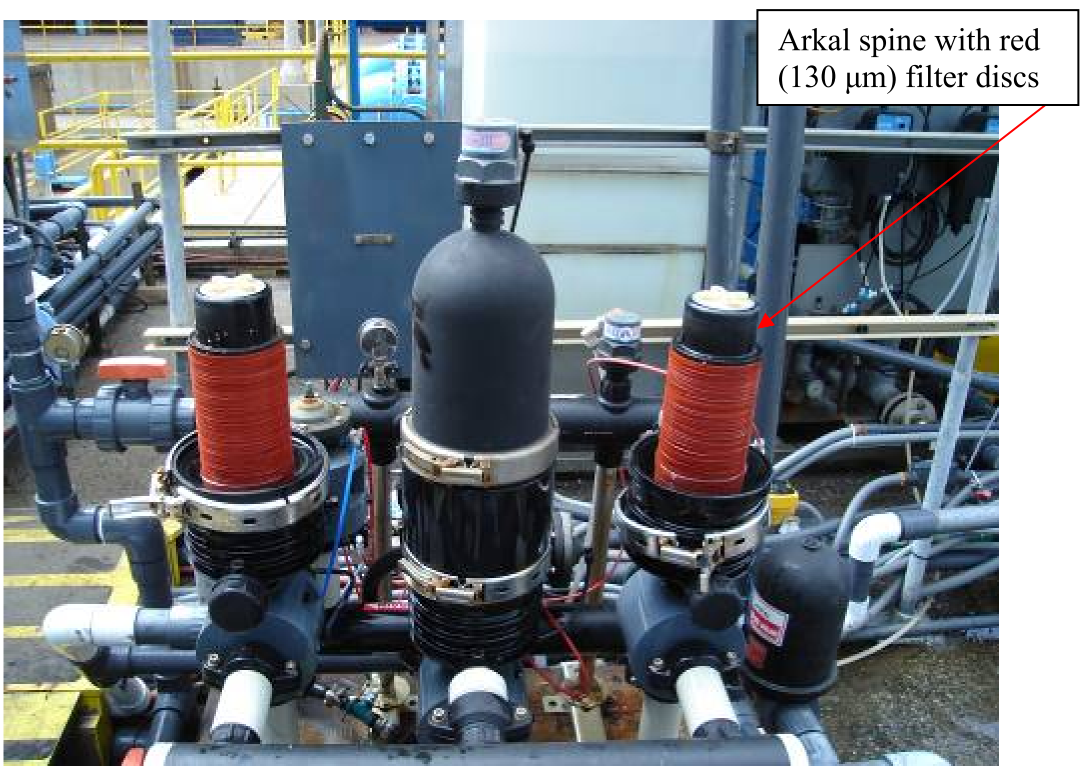
The Arkal filter operates using a specially designed disc filtration technology. Thin, color-coded polypropylene discs are diagonally grooved on both sides to a specific micron size. A series of these discs are then stacked in a column and compressed on a specially designed column or spine. When stacked, the groove on top runs opposite to the groove 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, 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 free and clear, loosening the trapped solids which are flushed out through the drain. Unfiltered seawater is then sent through the clean disc 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

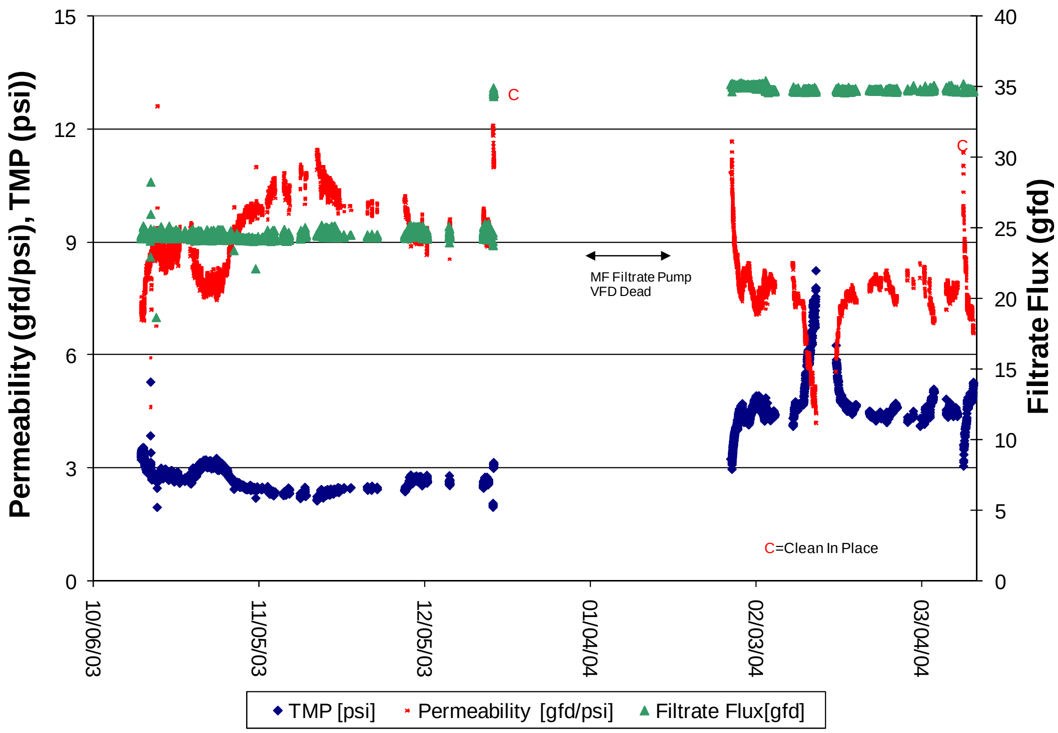


Performance of Newly Designed CMF-S Modules

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 of this document, 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. Per Figure 15 below, this proved to be true. The redesigned modules were first run for eight 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.

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 at these conditions and maintained 3 week run time before a cleaning was required.

Figure 16 - Performance of Redesigned Modules with Arkal Filter (Trial V)

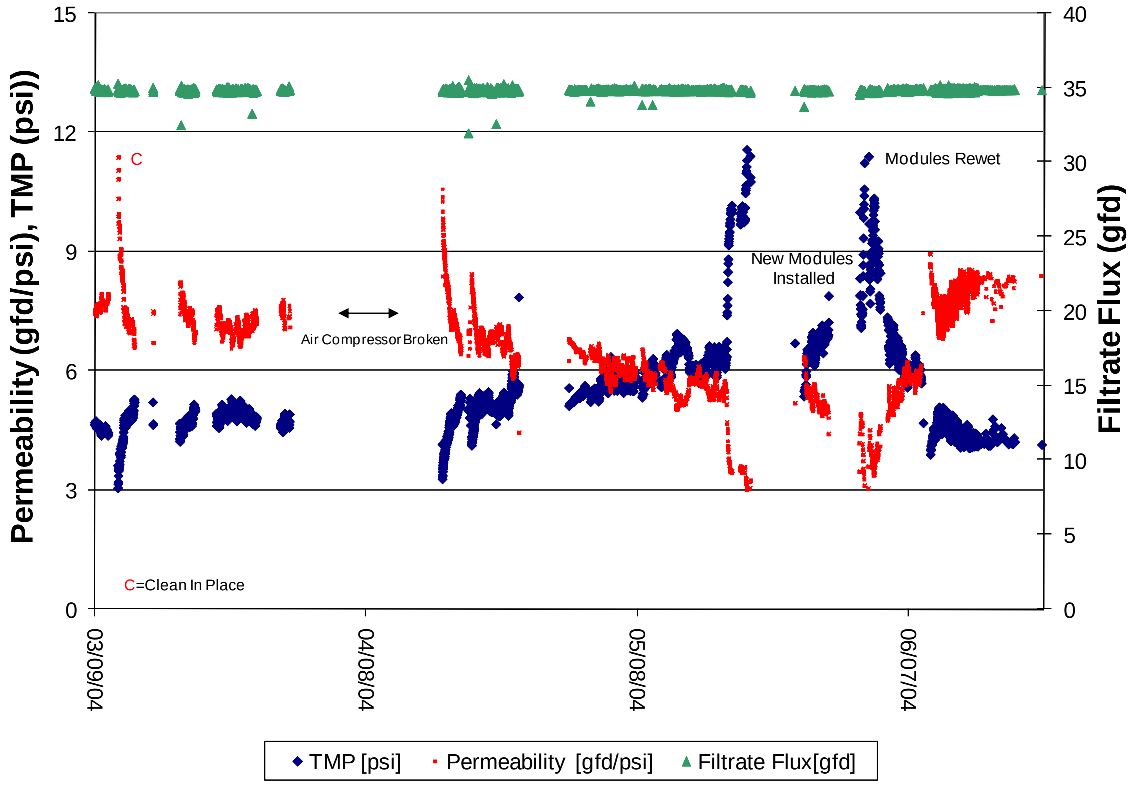


Table 6 - Optimized Siemens CMF-S Microfiltration Run Parameters Phase A

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

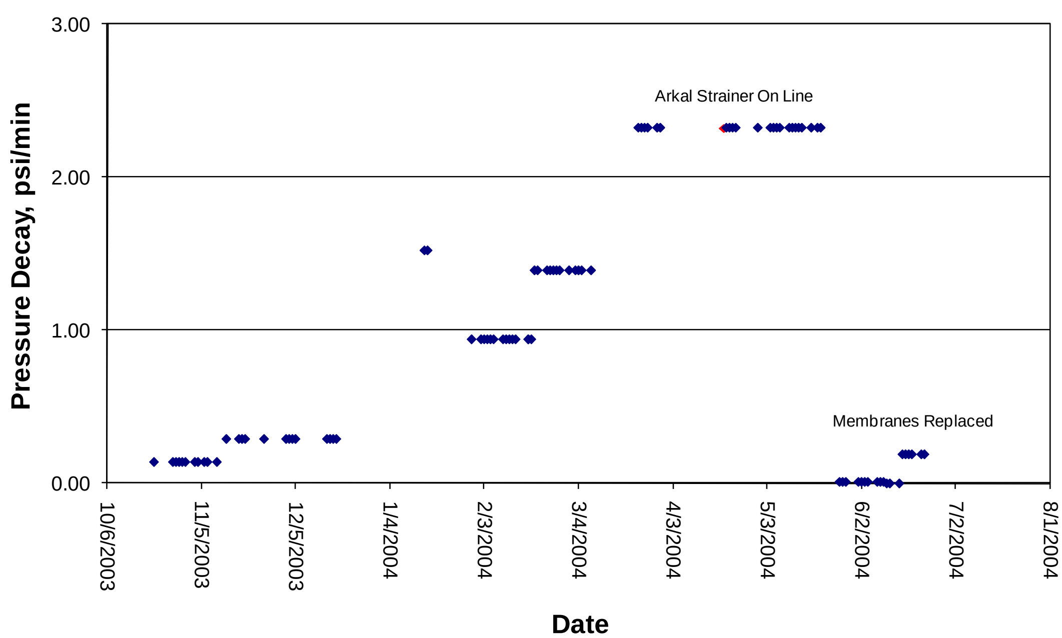
*Optimized Parameters. Non optimized parameters recommended by Siemens.

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. However, it was clear that the new module design may have allowed a significantly higher stable operating flux, but it did not maintain integrity with only the 800 µm strainer as pretreatment per 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 (installed on May 28, 2004). The facts that the damaged module pressure decay did not worsen (Figure 17) and the replacement modules held their integrity with the additional 130µm filter as pretreatment was promising. Phase A concluded with additional run time with the Arkal filter required 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 utilized to condition the raw ocean water such that it is suitable for spiral wound reverse osmosis membranes. This involves particulate matter removal that is best monitored through turbidity measurement and silt density index. Spiral wound reverse osmosis membranes operate best when the RO feed water has turbidity less than 1 NTU and 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 ~1NTU, with peak values of ~5NTU. Per Figures 18 and 19, the MF filtrate turbidity averaged 0.05NTU and typically was <0.1NTU, suitable for RO despite the module and fiber problems.

Figure 18 - Feed Water and Microfiltration Filtrate Turbidity-MF Trials I – III

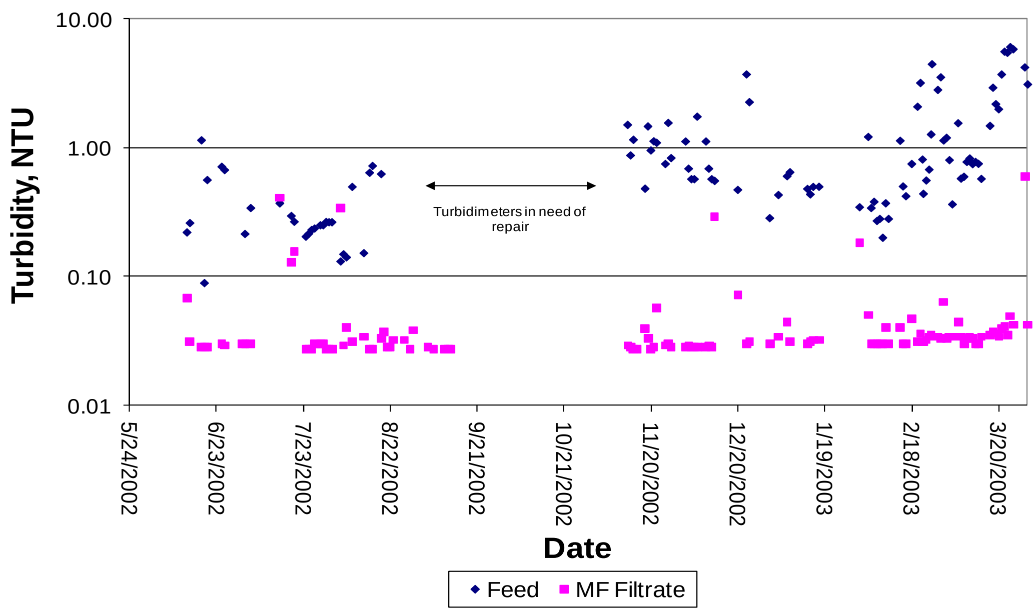
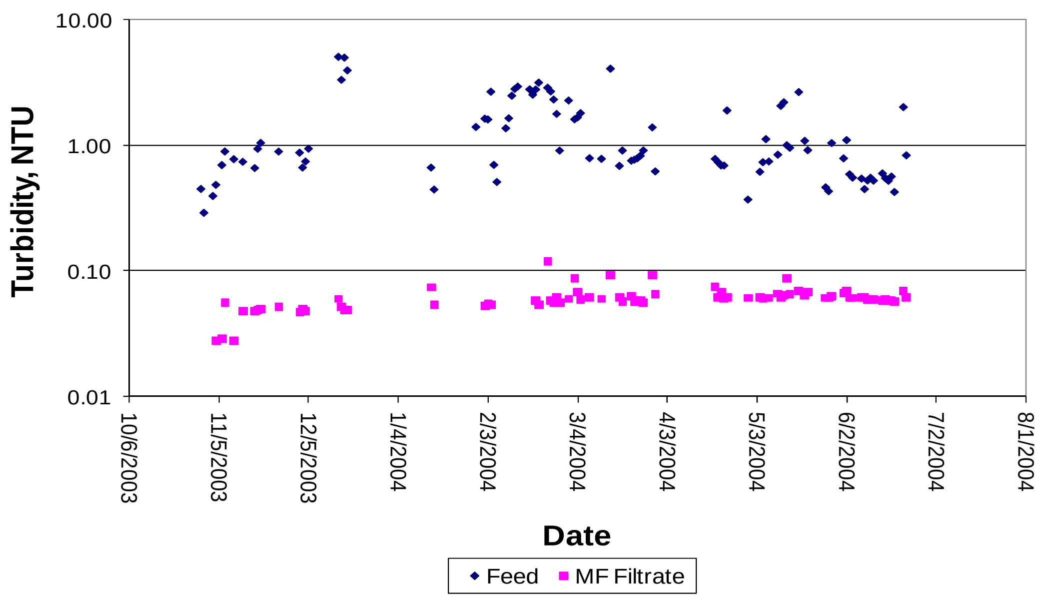


Figure 19 - Feed Water and Microfiltration Filtrate Turbidity-MF Trials IV & V



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 time difference required to filter a volume of water through a 0.45 µm filter pad at a feed pressure of 30 psig, and again after fifteen minutes of continuous filtration. Colloidal and suspended matter clogs the filter pad resulting in increasing SDI_{15} values.

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/permeability decline and differential pressure increase.

The SDI_{15} analysis of the raw ocean water was attempted on a few occasions and was immeasurable, clogging the SDI pad significantly within 5 minutes and almost completely by the fifteen-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 & 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 implemented on May 28, 2004. It was therefore important to find a solution to the fiber breakage events not only because of operating and maintenance efforts required to repair the fiber breaks but to ensure the water quality leaving the MF system is suitable for spiral wound reverse osmosis membranes as well.

Figure 20 – CMF-S System Pressure Decay Results and Filtrate SDI MF Trials I - III

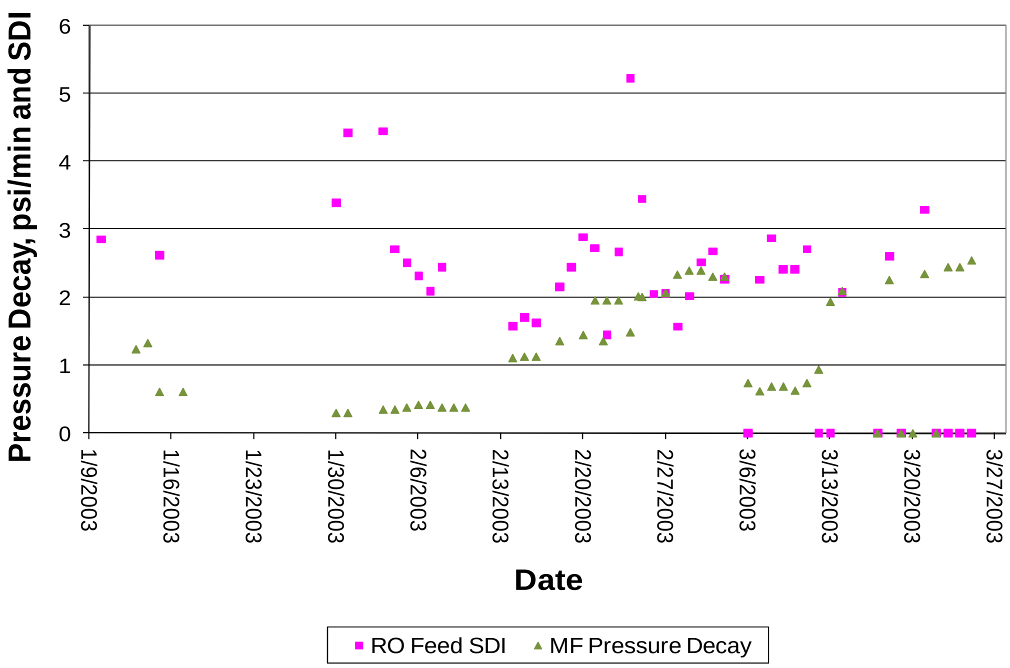
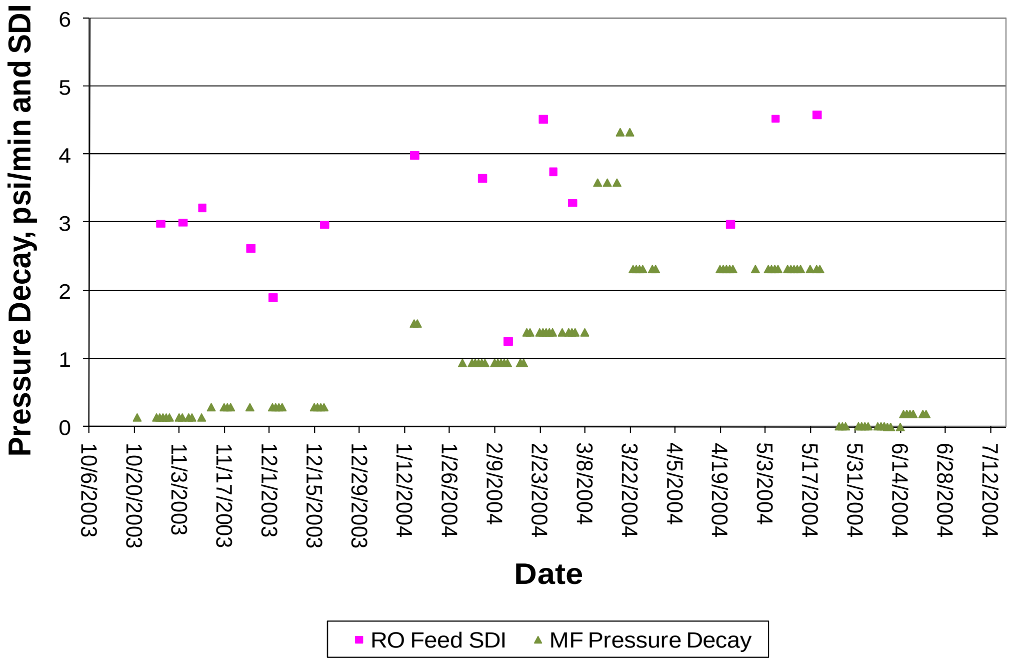


Figure 21 - CMF-S System Pressure Decay Results and Filtrate SDI MF Trials IV & V



MF Filtrate Water Quality Analysis

Weekly water quality analysis demonstrated that the microfiltration system provided a slight removal of TOC (approximately 10% removal). As expected, inorganic constituents were unaffected.

Table 7 - Microfiltration Water Quality Phase A

CMFS FEED

			MF Testing Phases I - III		MF Testing Phases IV & V	
	Units	**DL**	**Ave (15 samples)**	**StdDev**	**Ave (15 samples)**	**StdDev**
Parameter						
UV 254	abs/cm	0.005	0.010	0.003	0.013	0.003
Alkalinity (as CaCO3)	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	1335.3	103.3	1236.0	68.2
Hardness (as CaCO3)	mg/L	200	6514.9	473.7	6060.8	313.1
Sodium	mg/L	25	10963.4	733.2	10285.3	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

	Units	**DL**	**Ave**	**StdDev**	**Ave**	**StdDev**
Parameter						
UV 254	abs/cm	0.005	Typically ND		Typically ND	
Alkalinity (as CaCO3)	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	1338.4	105.0	1256.7	90.2
Hardness (as CaCO3)	mg/L	200	6525.9	490.7	6157.1	409.4
Sodium	mg/L	25	10920.3	808.7	10448.7	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

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

MF BACKWASH -			MF Testing Phases I - III		MF Testing Phases IV & V	
	Units	DL	Ave (15 samples)	StdDev	Ave (15 samples)	StdDev
Parameter						
TOC	mg/l	0.5	1.00	0.37	1.06	0.27
Turbidity	NTU	0.1	7.6	3.5	11.3	8.6

Phase B

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

The following discussion and Tables summarize the MF unit run conditions and events for Phase 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 power plant influent as feedwater. 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 module). However, the operation was impacted by integrity failures during this period, albeit not as severe as experienced in Phase A, prior to the use of the Arkal filter. Table 9A provides a listing of the 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 despite the use of 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 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 where blooms of certain algal species, which contain red-brown pigments, cause the water to appear colored red. This change in feedwater quality required a reduction in operating flux setpoint in order 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 9A - Details of Each Phase B1 MF Run

Feedwater Source: Influent Water

Run #	Dates	Flux (GFD)	Backwash Chlorination (mg/l)	Backwash Frequency	Comments
MF 18	6/8/04 – 9/10/04	34	20 in every backwash tank	~ 20 minutes	New Generation “B” modules, Set 2 Arkal 130 micron
MF 19	9/10/04 – 12/10/04	34	20 in every backwash tank	~ 20 minutes	Several fibers were pinned 9/20/04
MF 20	12/10/04 – 3/10/05	34	20 in every backwash tank	~ 20 minutes	1/15/05 2 pins in one module 2/8/05 Same module replaced due to damage
MF 21a	3/10/05 – 4/27/05	34	20 in every backwash tank	~ 20 minutes	
MF 21b	4/27/05 – 6/6/05	34	20 in every backwash tank	~ 20 minutes	New MF pilot unit installed 4/27/05. Continued operation with previous membrane set
MF 22	6/6/05 – 7/18/05	24.5, 20.5	20 in every backwash tank	~ 20 minutes	Severe Red Tide Event in Late May / Early June

Phase B-2

The Phase B-2 MF operation was defined by the shift of feedwater source from power plant influent to the warmer post-condenser effluent (Table 9B). However, throughout the summer of 2005 severe algae bloom events recurred. This period was marked by operation at reduced flux. Following the subsidence of the algae bloom events, operating flux was increased. During this phase of operation, the replaceable discs in the Arkal pre-filter were changed from 130 micron to 100 micron and subsequently to 40 micron, 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 plugging rate of the Arkal filter. Therefore, a change to 70 micron occurred toward the end of Phase B-2, which was maintained though the balance of Phase B testing. The MF membrane in Phase B-2 experienced a severe fouling event following installation of a set of new Generation “C” membrane (see Table 10), that was not recoverable by CIP. Further discussion is provided below in the MF permeability section.

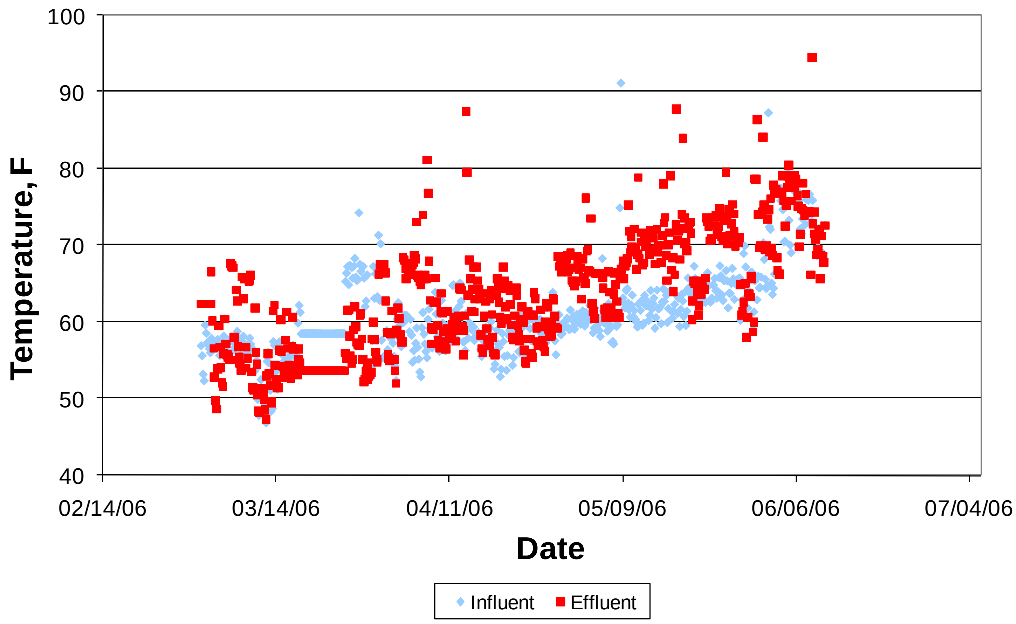
When the power plant was operating, the post-condenser effluent stream which fed the MF process in Phase B-2 was at a higher temperature, compared to the influent stream. The El Segundo Power Plant is a peaking facility and as such does not operate continuously. Therefore, there are periods when the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

Table 9B Details of Each Phase B2 MF Run

Feedwater Source: Post Condenser Effluent

Run #	Dates	Flux (GFD)	Backwash Chlorination (mg/l)	Backwash Frequency (min)	Comments
MF 22	7/18/05 – 9/5/05	20.5	20 in every backwash tank	~ 20 minutes	Reduced flux during algal bloom
MF 23	9/6/05 – 9/16/05	34	20 in every backwash tank	~ 20 minutes	
MF 24	9/18/05 – 9/23/05	27, 34	20 in every backwash tank	~ 20 minutes	9/23 all modules replaced due to fiber integrity issues. Generation "B", Set 3
MF 25	9/26/05 – 10/1/05	34, 27	20 in every backwash tank	~ 20 minutes	Prefiltration tightened from 130 to 100 micron Arkal disc filters prior to run
MF 26	10/19/05 – 11/23/05	27, 32	20 in every backwash tank	~ 20 minutes	November 30th, Prefiltration tightened from 100 micron to 40 micron Arkal disc filters
MF 27	12/9/05 – 12/31/05	31-32	20 in every backwash tank	~ 20 minutes	Generation "C", Set 1, of modules installed
MF 28	1/5/06 – 1/27/06	34	20 in every backwash tank	~ 20 minutes	Irreversible fouling of MF modules on 1/26
MF 29	1/31/06 – 3/6/06	34, 19	20 in every backwash tank	~ 20 minutes	Fouling problems continued – Feb 10th, operation reverted to Influent source until June, due to Effluent feed pump failure
MF 30	N/A	N/A	N/A	N/A	Fouling problems continued
MF 31	4/1/06 – 4/15/06	28	20 in every backwash tank	~ 20 minutes	Set 2 of Generation "C" modules installed due to fouling issues. 40 micron 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	20 in every backwash tank	~ 20 minutes	

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. MF Run #33 in Phase B-3, following correction of backwash chemical dosing, served as final confirmation of performance at the optimized conditions (30-34 gfd). This performance confirmed that the MF can operate for a three to four week period before a CIP is required, in the absence of severe algae bloom conditions.

Table 9C - Details of Each Phase B3 MF Run

Feedwater Source: Post Condenser Effluent

Run #	Dates	Flux (GFD)	Backwash Chlorination (mg/l)	Backwash Frequency (min)	Comments
MF 33	6/9/06 – 9/20/06	30-34	50 for 2 weeks, then 20 in every backwash tank	~ 20 minutes	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	~ 20 minutes	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 2 summarizes the characteristics of each generation. Membrane material remained PVDF and nominal pore size remained 0.1 micron for each generation.

Table 10 - Summary of Siemens CMF-S Modules Tested

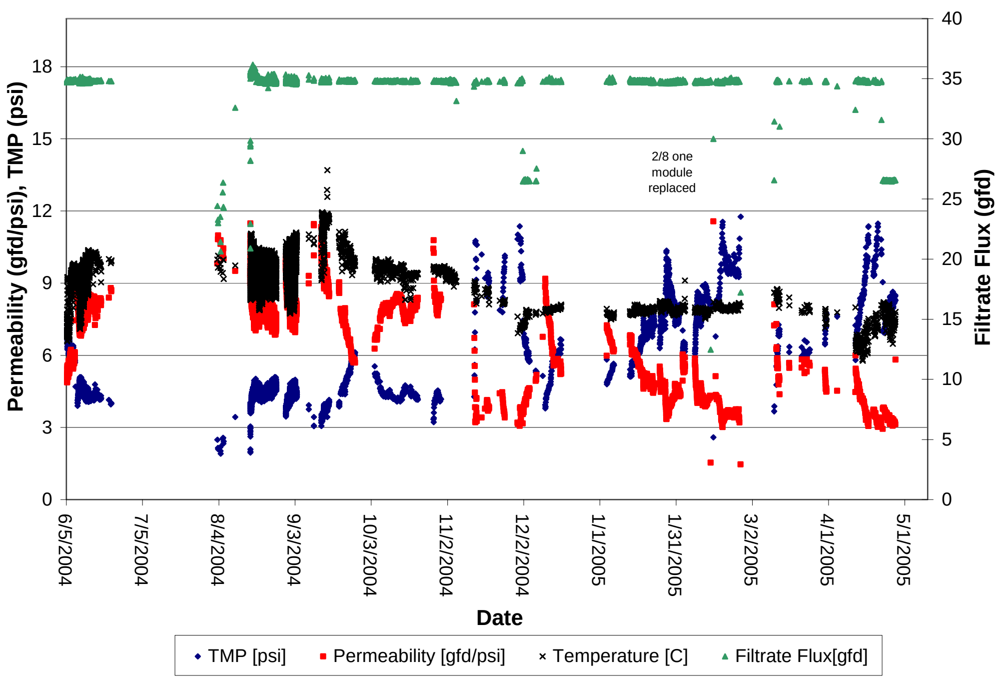
Parameter	Generation A	Generation B	Generation C
Fiber outside diameter, micron	650	800	1000
Fiber inside diameter, micron	390	500	530
Approximate # of fibers per module	14,500	9,600	7,400
Surface area per module, sq. ft.	335	272	262
Achievable flux, GFD	24	34	34
Permeate flow per module, gpd	8040	9248	8908

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 the operating flux of 34 GFD was sustainable for the goal period of 21 days before a CIP was required on influent water 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 the onset of 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 be operated during this event at a reduced flux rate of approximately 20 GFD, approximately 30% less than previous operating flux rates. As the algae bloom conditions subsided in August, the flux rate was able to be increased back to previous values. During this period of testing the feedwater source was switched to the warmer power plant effluent in July.

Figure 24 - CMF-S Performance May 2005 – September 2005

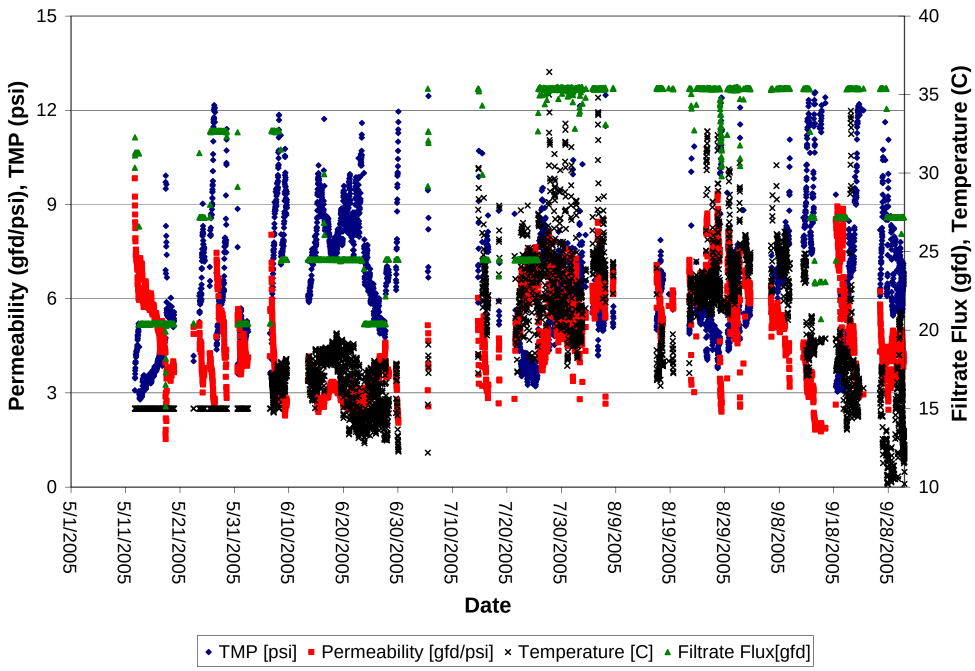


Figure 25 shows the performance of the MF unit from October 2005 to May 2006 (balance of Phase B-2), with feedwater continuing from power plant effluent. The MF unit experienced integrity issues during this time period and several different Arkal prescreen filter disc sizes were utilized in an effort to keep shell particles and other debris from damaging the membrane fibers. MF flux rates varied from 19 to 34 GFD during this period, with one episode of irreversible fouling occurring in January/February resulting in the need to reduce the operating flux to 19 GFD. The irreversible aspect of this fouling was a unique event 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 organics and biological matter on the membrane surface, but 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% sodium percarbonate (40^{O}C) followed by 0.05% H_2SO_4 (40^{O}C). This information was retained for implementation at the pilot, should a similar event occur.

Operational issues with the effluent feed pump resulted in reverting operation back to influent water from February 10th until early June.

Figure 25 - CMF-S Performance October 2005 – May 2006

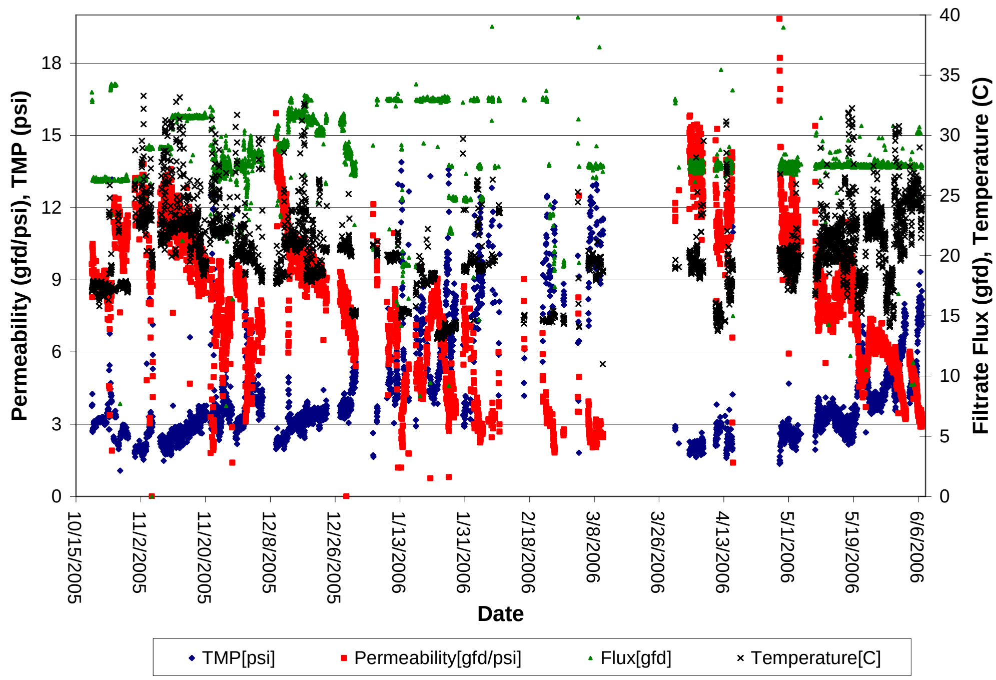
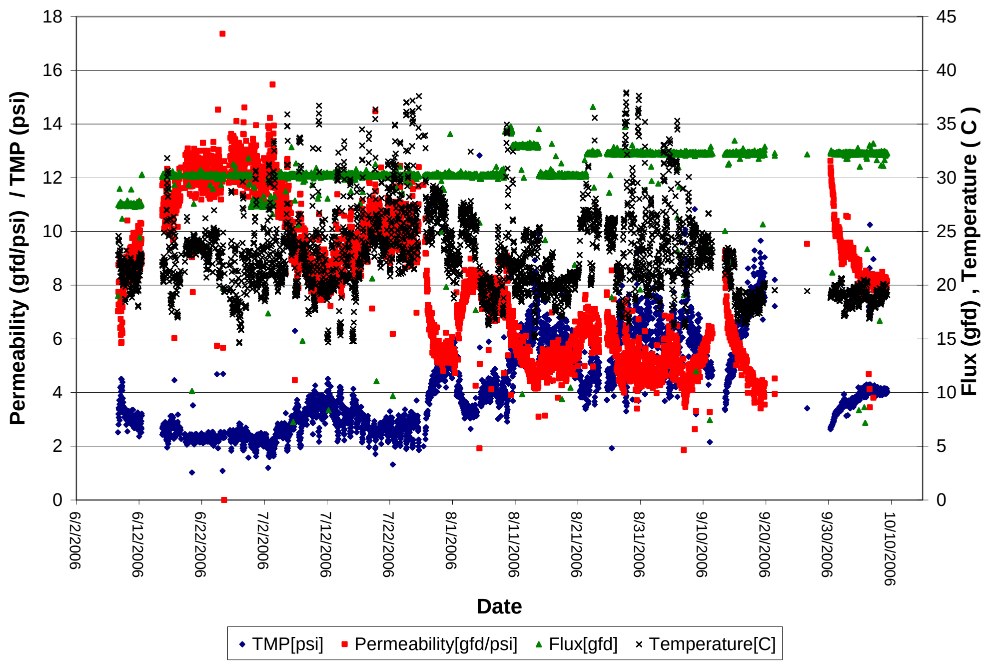


Figure 26 displays the last of the run time for the CMF-S unit. The unit experienced a very long run time during this period of time with flux range of 32-34 GFD. During run #33 the CMF-S maintained 30 GFD for two months without requiring a CIP. Subsequently, in late August 2006, the flux rate was increased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

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

Figure 26 - CMF-S Performance June 2006 – October 2006



MF Filtrate Quality – Phase B

Turbidity

The MF filtrate in general was less than 0.1 NTU as seen in Figure 27. The values greater than 0.1 can generally be attributed to fiber breakage.

Figure 27 - Phase B Siemens CMF-S Turbidity

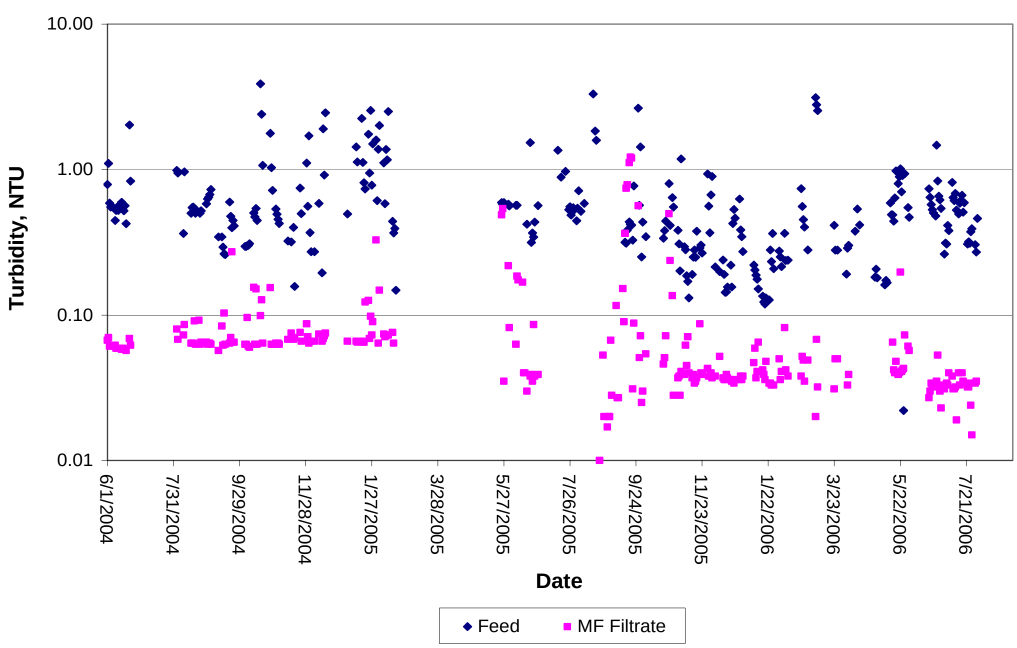


Figure 28 shows the integrity of the MF fibers during phase B with various grades of prescreening. Note that the major fiber integrity issues in August of 2005 correspond with the highest turbidity values as depicted in Figure 27.

Figure 28 - Phase B CMF-S Pressure Decay Test Results

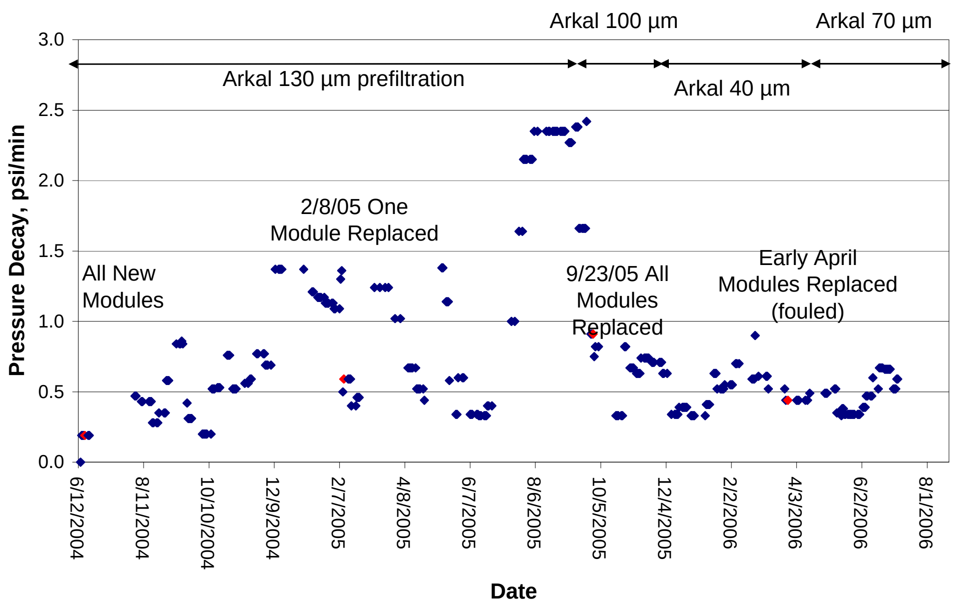
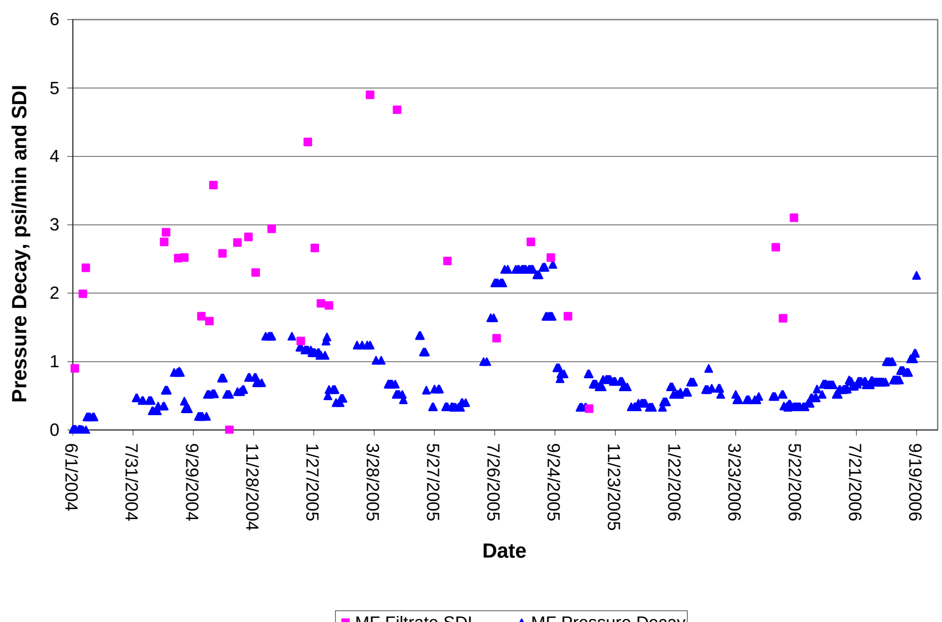


Figure 29 shows the same PDT values plotted with the CMF-S filtrate SDI values. In general, the SDI values for the CMF-S system were below 3, with only three measurements in the 4 to 5 range during this period testing.

Figure 29 - Phase B CMF-S PDT and SDI Values



Tables 11 and 12 show detailed water quality of both the CMF-S feed and filtrate water quality respectively. Like the Phase A testing, the CMF-S system demonstrated approximately 10% removal of TOC, and no removal of inorganic constituents.

Table 11 - 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 CaCO3)	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	1245	52	1254	95	1190	65
Hardness (as CaCO3)	mg/L	200	6089	248	6105	441	5866	294
Sodium	mg/L	25	10237	414	10422	716	9830	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 12 - 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 CaCO3)	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	1249	66	1264	94	1203	80
Hardness (as CaCO3)	mg/L	200	6108	325	6147	432	5930	367
Sodium	mg/L	25	10303	508	10509	683	9941	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. Similar performance with regards to sustainable flux rate and filtrate water quality were observed on both the power plant influent and post condenser effluent water sources. 34 GFD was determined to be the optimum flux for both water sources, and filtrate quality was consistently acceptable as feed to the Reverse Osmosis units. Fiber damage did occur during Phase B testing and pre-filter rating of 70 micron or less was found to be effective at preventing damage. The optimized CMF-S operating parameters are included in Table 13.

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% in order to maintain stable operation and a reasonable period between chemical cleanings.

Three generations of MF modules were trialed during Phase 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% citric acid recirculation/aeration at 36 – 38˚C followed by
- 400 to 600 mg/L NaOCL recirculation at 20 - 22˚C

Table 13 - Optimized CMF-S Parameters

Parameter	Value
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
Refill Duration (seconds)	~35
Backwash chlorination (mg/L)	20

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 power plant 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 Table 14. 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 membrane 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) utilizing a chlorinated backwash operating strategy. This operating strategy used more frequent dosing of chlorine to inhibit and remove foulants of the membrane compared to earlier runs.

Table 14 summarizes the UF unit run conditions during the Phase B testing period:

Table 14 - Details of Each UF Run

Phase B 2 Summary

Feed Source is Power Plant Influent

Run	Dates	Flux (GFD)	Backwash Frequency (min)	# of NaOCl MC per day	NaOCl concentration (mg/l)	# of Citric Acid MC per week	Citric Acid concentration (g/l)	Comments
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.
UF2	5/20/05 – 7/4/05	20.1	28	3	100	1	0.5	Late May/ 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	Power plant 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 power plant effluent during this run.
Feed Source switched to Effluent water Nov 23, 2005								
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 micron to 40 micron.

Run	Dates	Flux (GFD)	Backwash Frequency (min)	# of NaOCl MC* per day	NaOCl concentration (mg/l)	# of Citric Acid MC* per week	Citric Acid concentration (g/l)	Comments
UF 8	1/11/06 – 2/1/06	19	28	2	100	1	0.5	CIP study showed heating CIP solutions to 35-40ºC to be more effective.
Feed Source returned to Influent water Feb. 10, 2006								
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	

Note: Use of citric acid maintenance cleans were stopped after run 10

Phase B 3 Summary

Feed Source is Effluent water

Run	Dates	Flux (GFD)	Backwash Frequency (min)	# of NaOCl MC* per day	NaOCl concentration in MC (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 greater 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.

Run	Dates	Flux (GFD)	Backwash Frequency (min)	# of NaOCl MC* per day	NaOCl concentration in MC (mg/l)	NaOCl used in backwash	NaOCl backwash concentration (mg/L)	Comments
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. .
Equipment relocation, down for 6 months								
UF 16	5/10/07 – 6/19/07	20-25	22-24	1 @ 110ºF	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 @ 110ºF	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 @ 110ºF	100	Yes	2 mg/l In every backwash tank	Demonstration of 27.5 GFD sustainable for 21 days.
UF 19	8/17/07-Through September 2007	27.5	22	1 @ 110ºF	350	Yes	4 mg/l In 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.

UF Permeability

Like the Siemens CMF-S system (and the RO), the UF runs at constant flux and thus as the membrane fouls, the trans-membrane pressure (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 Table 14, early testing in 2005 and 2006 of the Zenon unit on both influent and effluent streams resulted in a sustainable flux rate 16-18 GFD. Figure 31 shows the details of operation between May 2005 and September 2005. The Zenon system was brought online 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. Per figure 31, runs #2 and 3 (May 20, 2005 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 9/14/05.

Figure 31 - Zenon Operating Performance May 2005 – September 2005

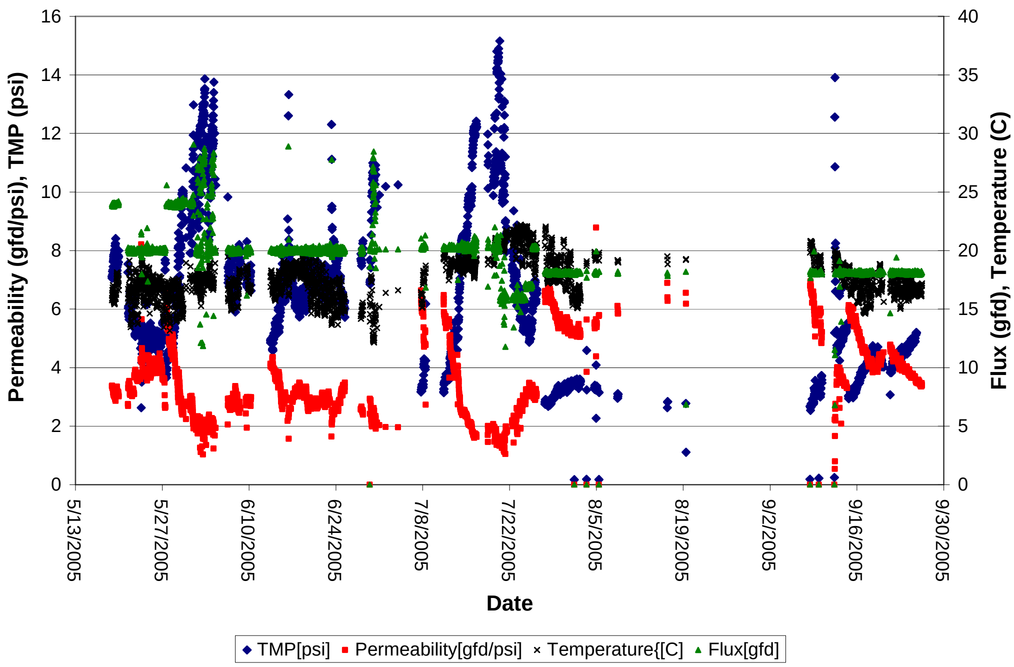
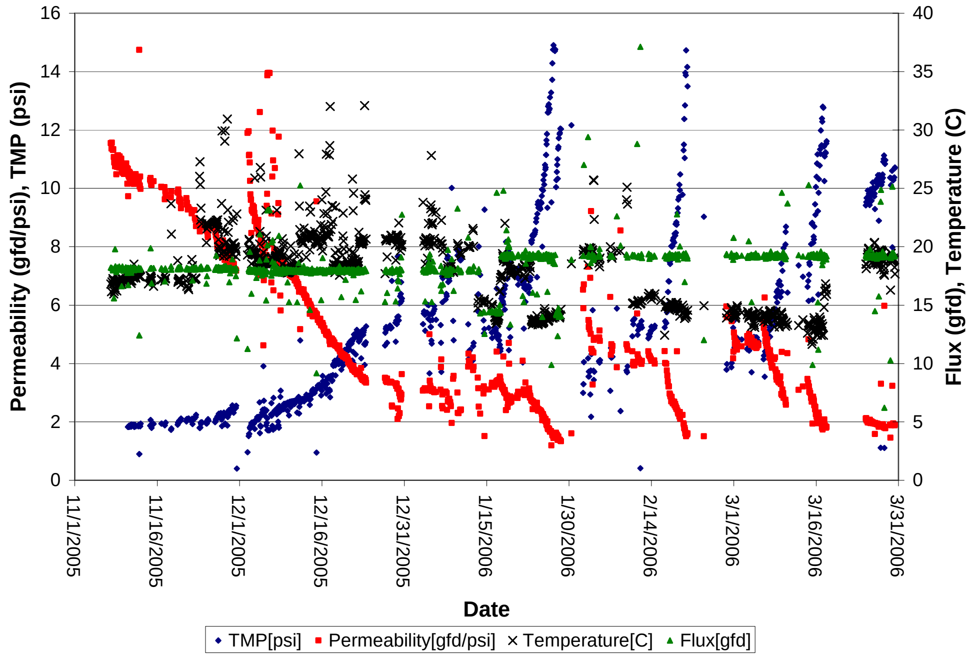


Figure 32 details the continued Phase B-2 operation from November 2005 to March 2006. The unit was switched from power plant influent to power plant effluent during UF Run #6 on November 23, 2005 (site operational requirements). Noteworthy in the data from UF Run #6 is that the rate of permeability loss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

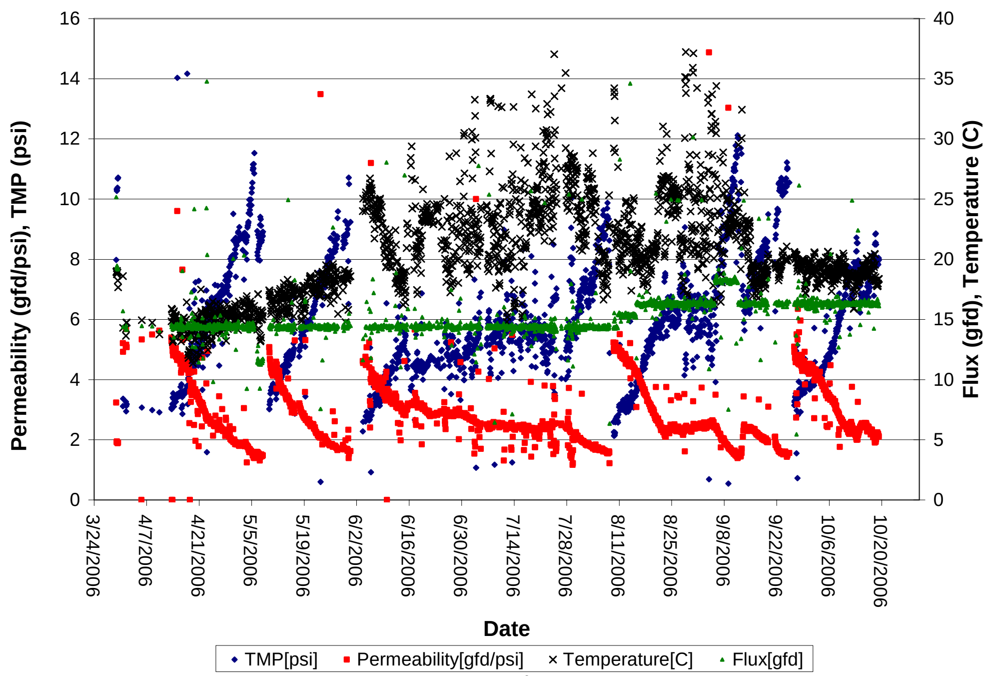
GFD in this period.

Figure 32 - Zenon Operating Performance November 2005 – March 2006



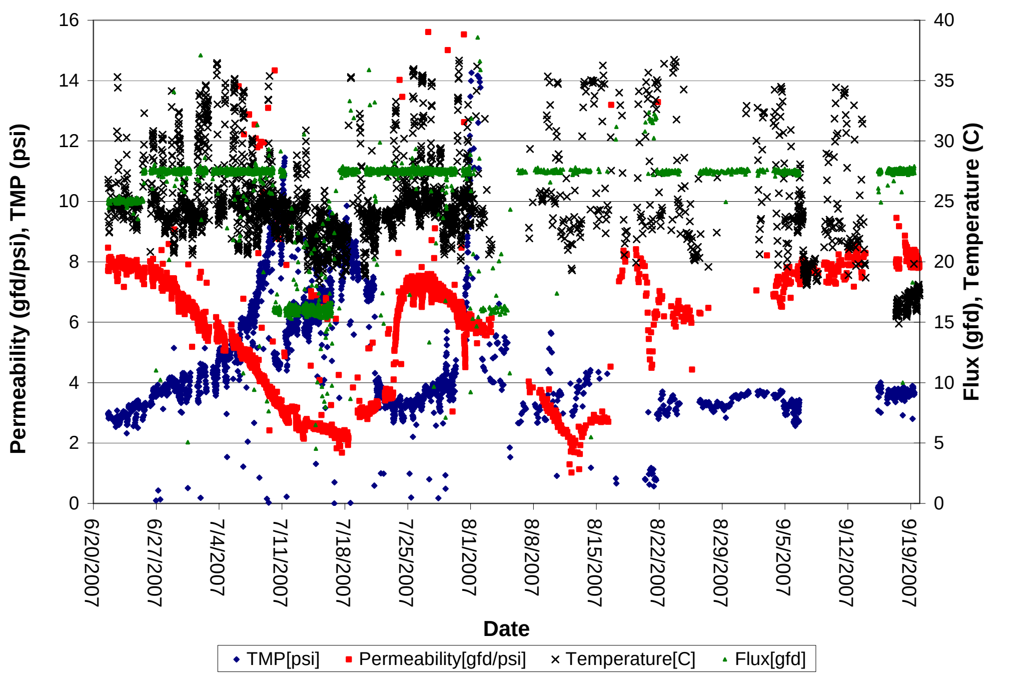
Detailed operating performance for April 2006 to October 2006 is shown in Figure 33. Flux rate was reduced to 14 GFD for a period of this testing, resulting in extended run times between cleanings. Run #13 exceeded 60 days of run time, indicating a flux of 14 GFD was too low as the target CIP frequency 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 utilizes 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 include 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. Per figure 34, the changes provided a drastic improvement in performance as the stable flux rate of 27.5 was achieved.

Figure 34 - Zenon Performance June 2007 – September 2007



The Zenon unit was restarted up in late May 2007, with Run 16 considered a "break-in" period. Run 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%
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 mg/l to 4 mg/l, and in the Maintenance Clean was increased from 100 mg/l 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 15 and 16 show detailed water quality of both the CMF-S feed and filtrate water quality respectively. On average, the Zenon system has also demonstrated approximately 10% removal of TOC.

Table 15 - 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 CaCO3)	mg/L	2	115	6.1	113	1.9	114	1.2
Calcium	mg/L	25	390	32	377	27	391	24
Magnesiu m	mg/L	25	1230	64	1263	111	1206	70
Hardness (as CaCO3)	mg/L	200	6039	329	6142	509	5942	339
Sodium	mg/L	25	10124	447	10407	826	9955	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 16 - 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 CaCO3)	mg/L	2	115	6.1	113	2.1	113	5.2
Calcium	mg/L	25	391	35	381	22	394	23.9
Magnesiu m	mg/L	25	1234	49	1272	97	1213	72.7
Hardness (as CaCO3)	mg/L	200	6059	279	6191	434	5979	350.6

Zenon ZW 1000 Filtrate			Phase B1		Phase B2		Phase B3	
Sodium	mg/L	25	10187	385	10514	700	10042	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 is 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

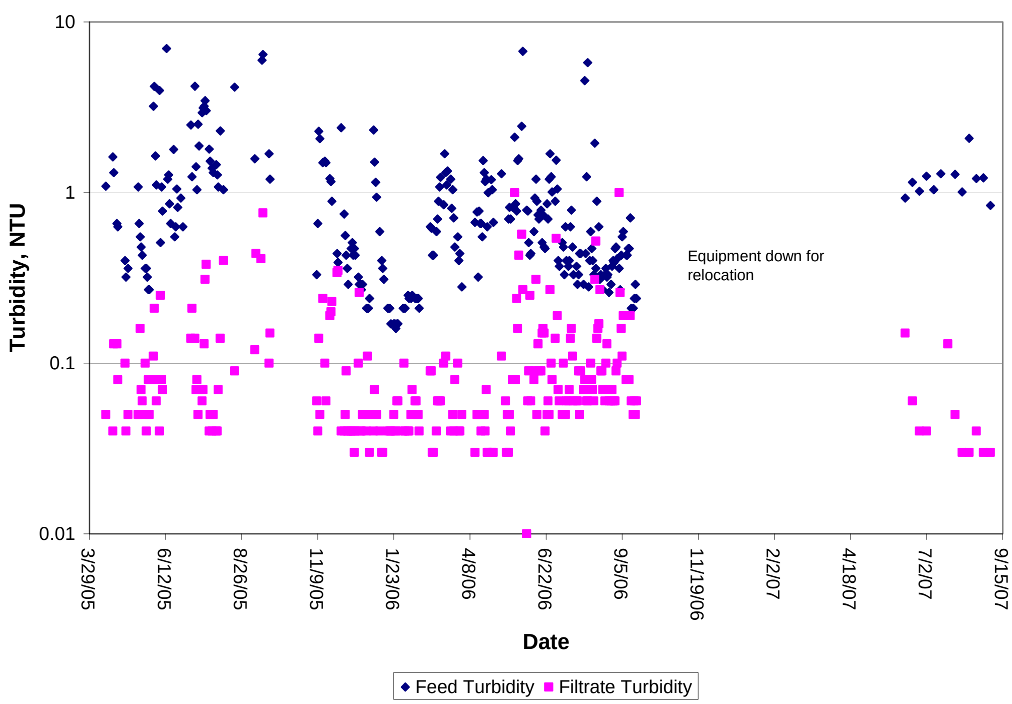
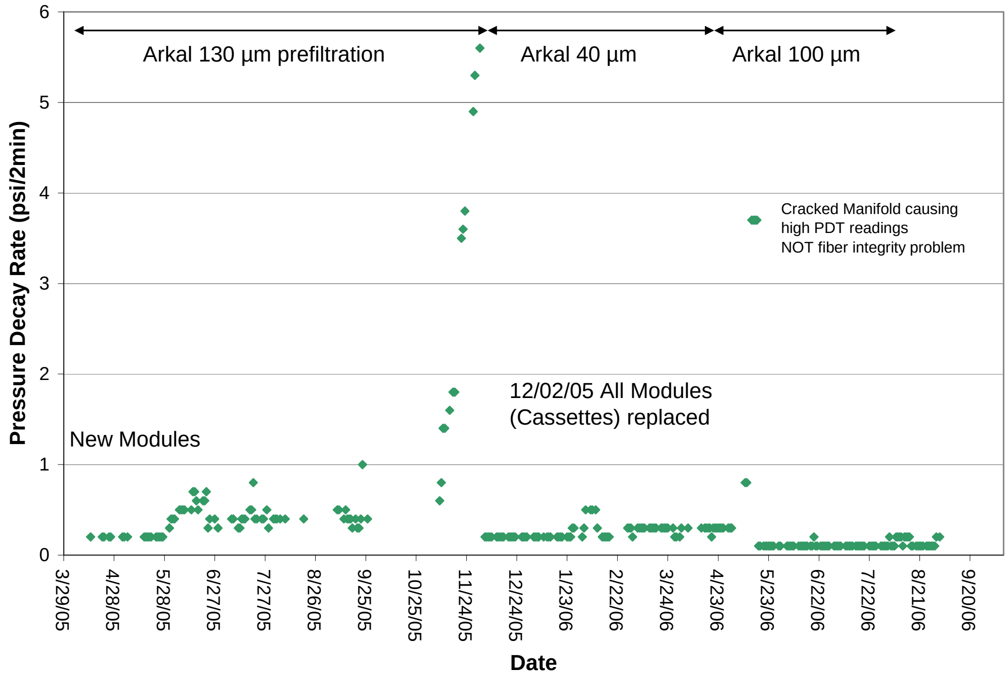


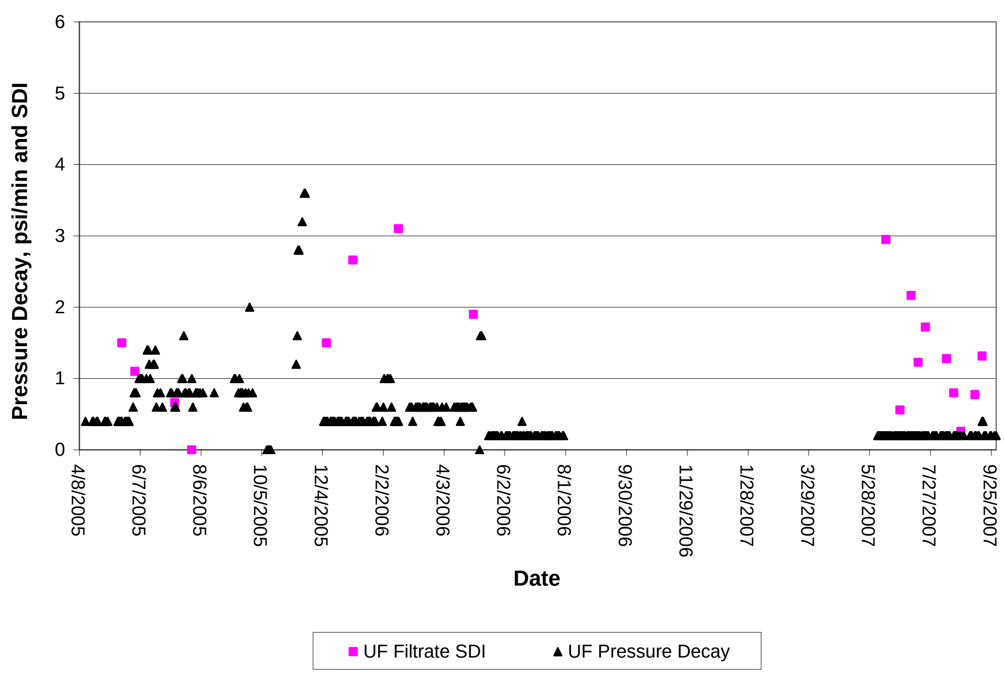
Figure 36 displays the Zenon system membrane integrity from April 2005 to September 2006. The Zenon system had only a single integrity problem where a couple of fibers in one module sheared in half. This event occurred in November of 2005. After membrane autopsy, this event was deemed a membrane manufacturing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

Figure 36 - Zenon ZW1000 Pressure Decay Test Results



The data shown in Figure 37 shows the SDI of the UF filtrate was consistently acceptable as feed to the RO system. Data from June 2007 through September 2007 shows no fiber integrity issues.

Figure 37 - Zenon ZW1000 PDT and SDI Values



UF Summary

The Zenon ZW1000 system was tested on both power plant influent and effluent for a period of approximately two years. Similar performance with regards to sustainable flux rate and filtrate water quality were observed on both the power plant influent and 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 chlorinated backwashes in every backwash combined with daily heated chlorinated maintenance clean has resulted in a sustainable flux rate of 27.5 GFD. Other successful operational parameters are listed in Table 17 below.

Membrane integrity was very good on the Zenon system. The use of a pre-filter rating of 100 micron 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 downstream RO performance.

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

- 2% citric acid recirculation/aeration at 40˚C followed by
- 500 mg/L NaOCL recirculation at 40˚C

Table 17 - 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-inch 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. The smaller 4-inch membranes are a representative smaller version of their 8-inch counterparts and provide equivalent engineering data. However, the standard 4-inch diameter seawater membranes do not have equivalent salt rejection capabilities as 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 as well as the Phase B RO membranes discussed below.

Phase A Testing

Figure 38 - Reverse Osmosis Test Equipment

All phase A testing of the RO utilized a microfiltered feed water source. Phase A of the RO Testing can be grouped into the following trials:

Table 18 - 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 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 19 - Details of Each Phase A Reverse Osmosis Run

Trial	Run #	Dates	MF Filtrate Chemical	RO Feed Antiscalant ppm	Hydranautics Flux, GFD	Hydranautics Recovery	Filmtec Flux, GFD	Filmtec Recovery	Notes
RO I	RO 1	7/15/02-9/6/02	1ppm NH4OH	3	8	50	8	50	RO Membranes show signs of oxidation
	RO 2	9/1/02-9/28/02	1.5ppm NH4OH	3	8	50	8	50	Adjusted NH4OH 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*	*1ppm SBS*	3	*8*	*50*	*8*	*50*	Memcor chlorinated b/w oxidizing RO
	RO 5	11/25/02-12/16/02	2-3ppm SBS	3	8	50	8	50	Increase SBS
	RO 6	12/17/02-1/15/03	2-3ppm SBS	3	8	50	8	50	Both RO pumps repaired, recycle modification
RO II	RO 7	1/15/03-3/9/03	2-3ppm SBS	3	8	50	8	50	1/15-Replaced both HYD and FT RO membranes
RO III	RO 8	3/9/03-4/3/03	3ppm SBS	3	9	50	9	50	Increased RO Flux
	RO 9A	10/21/03 - 11/19/03	3ppm SBS	3	9	50	9	50	Installed RO feed pump VFD
		11/19/03-1/15/04	3ppm SBS	3	9	50	9	50	Infrequent operation to MF/feed flow problems **CIP 12/5**
	RO 9B	1/30/04 - 2/18/04	3ppm SBS	3	9	50	9	50	
RO IV	RO 10	2/18/04 - 6/10	3ppm SBS	3	11	50	11	50	Increased RO Flux

RO Trial I Testing

The reverse osmosis unit consists of two independent trains. Each train has two pressure vessels operating in series, with the lead vessel containing three elements and the tail vessel four elements for a total of seven four-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 subsequently RO process has been used successfully on many wastewater reclamation facilities including the 20 MGD West Basin Water Recycling Plant. 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 is 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 Process and Equipment Challenges section of this document.

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 NH4OH addition rate was increased 50% 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 utilized 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, combined with the addition of sodium bisulfite in front of the RO, was utilized in the remainder of the trials.

Figure 39 - Increasing Permeability of RO Membranes due to Oxidation (RO Trial I)

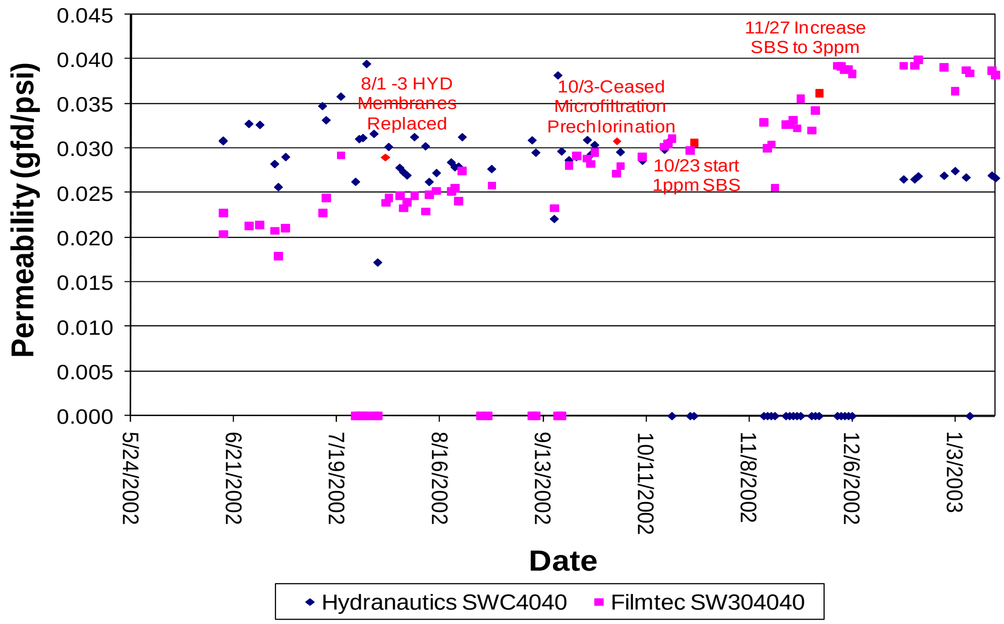
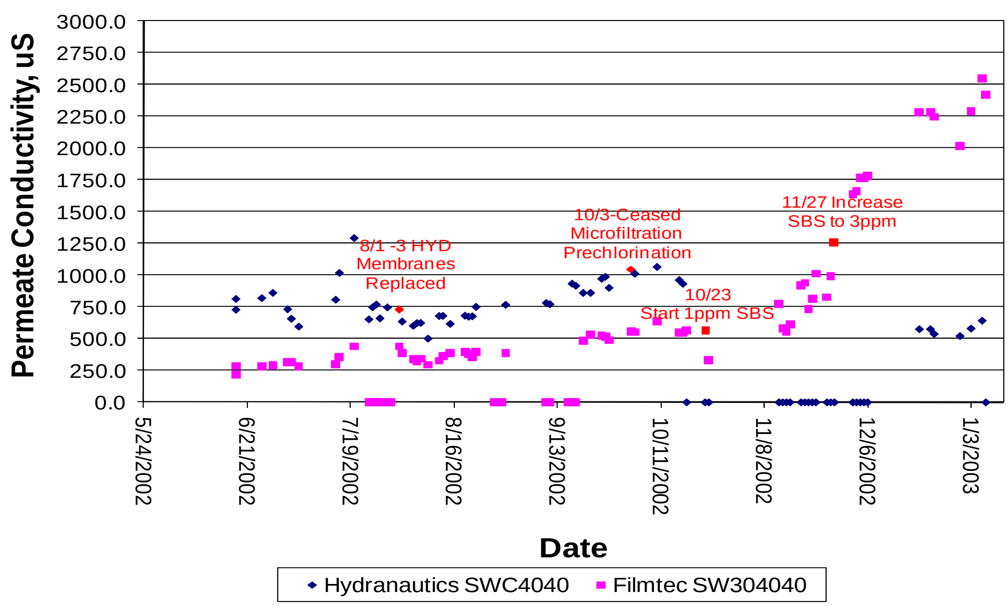


Figure 40 - Increasing Permeate Conductivity of RO Membranes due to Oxidation (RO Trial I)



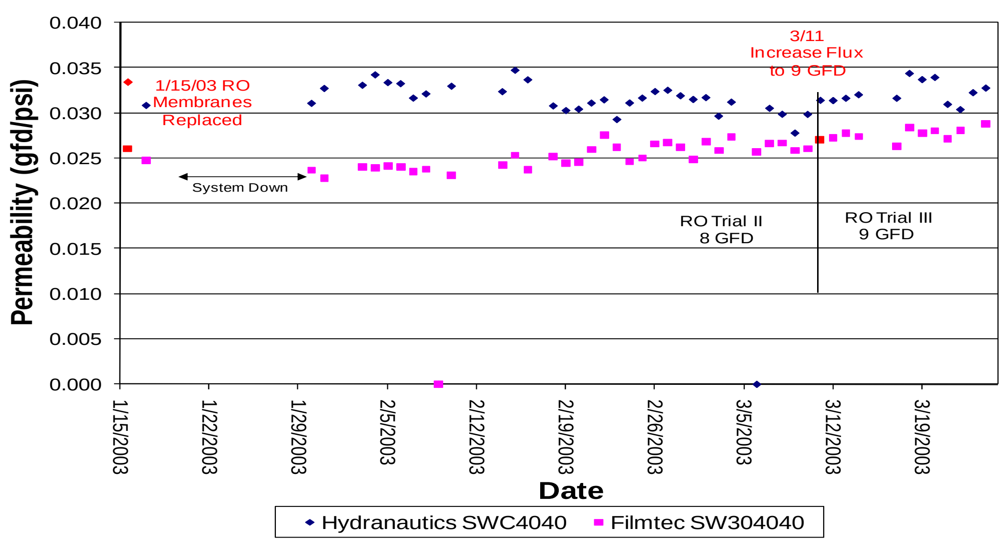
From October through December 2002, the RO was run with the damaged membranes in an attempt to find a pretreatment strategy that would allow the MF to maintain reasonable flux rates and run times without further RO oxidation. The RO membranes were replaced on January 15, 2003 and trial II of the RO testing 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 a distinctly different approach to 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.)

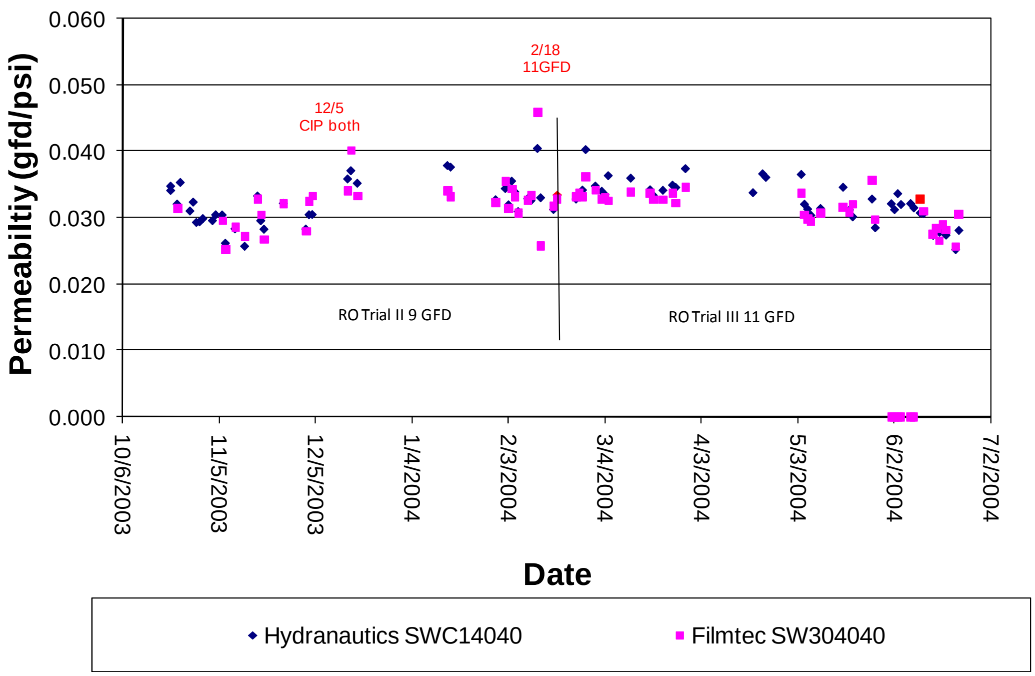
Figure 41 displays that the permeability of the Hydranautics membrane was fairly stable following the replacement of the RO membranes (RO trial II). Dow membranes, on the other hand, showed a slight increase in specific flux and as will be discussed in the next section, permeate conductivity as well. These trends are consistent with membrane oxidation. However, the Hydranautics membranes did not show these signs of oxidation, and these 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, examination of Figures 39 and 40 above, which display the results of the Trial I testing of the RO membranes oxidized by the chlorine followed by ammonia addition (failed chloramination) process, reveal that the Dow membranes experienced deterioration, presumably from oxidation, much faster than the Hydranautics 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. This is discussed further in Process and Equipment Challenges. Testing was resumed in October 2003. A drop in permeability was immediately observed and 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 operated at 9 GFD flux. The flux was increased to 11 GFD on February 18, 2004. Comparison of 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 3100 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 20 - 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 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 feedwater temperature. (Figure 44).

Figure 43 - Reverse Osmosis Membrane Conductivity Trials II and Beginning of Trial III

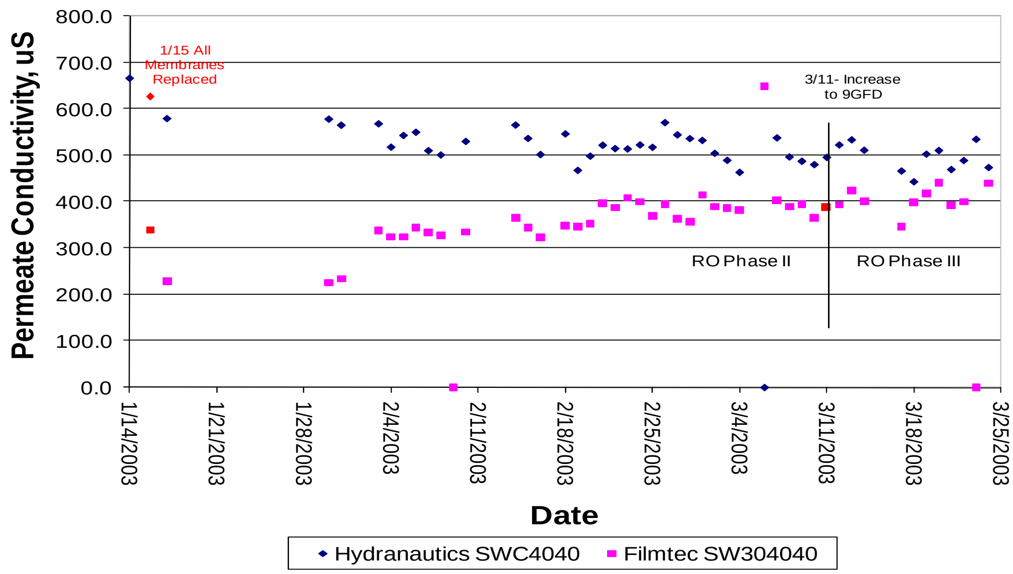
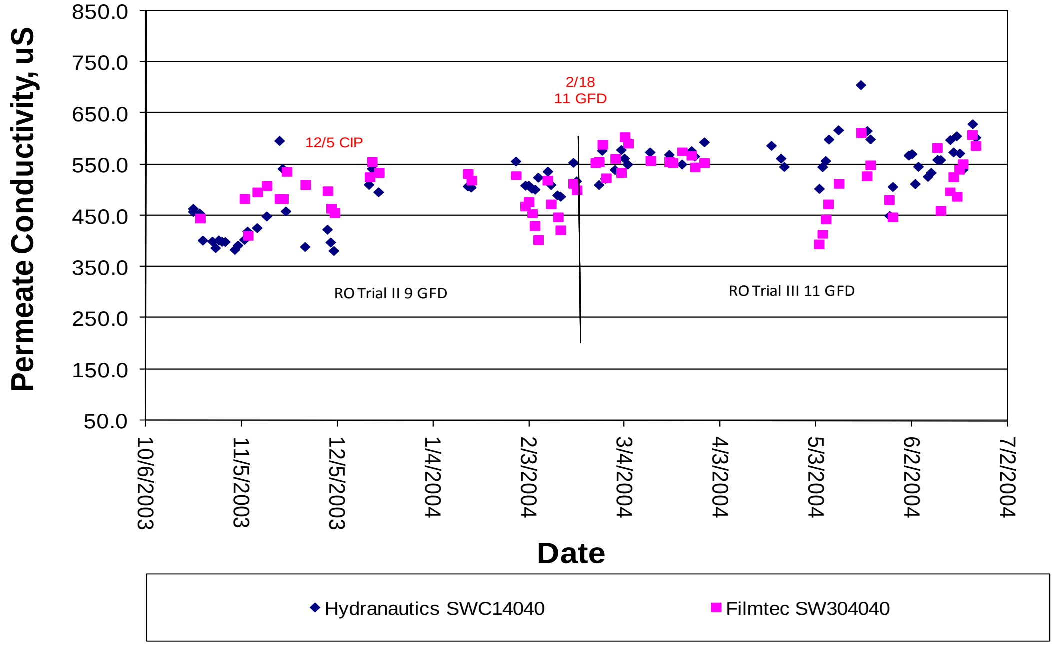


Figure 44 - Reverse Osmosis Membrane Conductivity End of Trials III and Trial IV



Individual Ion Analyses

Tables 20, 21 and 22 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 of 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. For both Boron and TDS, the Dow membrane initially produced water substantially lower concentration than the Hydranautics membrane. The Dow membrane continued to produce lower concentration, but the gap between the two membranes lessened as the testing progressed. Boron levels were constantly below 1.5 mg/L and 1.0 mg/L for Hydranautics and Dow, respectively.

Table 21 - Average RO Membrane Water Quality for Trial II (8 GFD Flux rate)

	SAMPLE ID					
	RO Feed	Permeate		Concentrate		
		Train 1	Train 2	Train 1	Train 2	
Parameter		**HYD**	**DOW**	**HYD**	**DOW**	**Units**
TDS	34750	230	150	69000	67000	mg/L
Lab pH*	8.1	6.9	6.5	7.9	7.9	UNITS
Alkalinity (as CaCO3)	115	<2	<2	212	214	mg/L
Bicarbonate (as CaCO3)	114	<2	<2	210	212	mg/L
Carbonate (as CaCO3)	1.3	<0.1	<0.1	1.5	1.6	mg/L
Hydroxide (as CaCO3)	0.06	<0.01	<0.01	0.04	0.04	mg/L
Sulfate	2533	<10	<10	5538	5463	mg/L
Chloride	18875	111	70	35325	34975	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	1360	2.0	2.6	2504	2460	mg/L
Hardness (as CaCO3)	6586	9.4	13.1	12156	11937	mg/L
Ca Hardness (as CaCO3)	986	1.5	2.8	1846	1807	mg/L
Sodium	11175	77	46	20600	20400	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: Ave temperature 22C, Four samples
Maximum TDS: 290 HYD, 160 Dow
Maximum Boron: 1.3 HYD, 0.7 Dow

Table 22 - Average RO Membrane Water Quality for Trial III (9 GFD Flux rate)

	SAMPLE ID					
	RO Feed	Permeate		Concentrate		
		Train 1	Train 2	Train 1	Train 2	
Parameter		**HYD**	**DOW**	**HYD**	**DOW**	**Units**
TDS	34167	185	178	64667	64667	mg/L
Lab pH*	8.0	6.6	6.6	7.8	7.8	UNITS
Alkalinity (as CaCO3)	112	<2	<2	205	205	mg/L
Bicarbonate (as CaCO3)	111	<2	<2	204	204	mg/L
Carbonate (as CaCO3)	1.1	<0.1	<0.1	1.2	1.3	mg/L
Hydroxide (as CaCO3)	0.05	<0.01	<0.01	0.03	0.03	mg/L
Sulfate	2538	<10	<10	5265	5160	mg/L
Chloride	18967	100	95	35050	33950	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	1260	1.5	2.4	2410	2457	mg/L
Hardness (as CaCO3)	6133	7.1	11.2	11716	11925	mg/L
Ca Hardness (as CaCO3)	944	1.4	2.2	1792	1808	mg/L
Sodium	10383	68	63	19867	20133	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: Ave temperature 22C, Five samples
Maximum TDS: 240 HYD, 230 Dow
Maximum Boron: 1.2 HYD, 1.0 Dow

Table 23 - Average RO Membrane Water Quality for Trial IV (11 GFD Flux rate)

	SAMPLE ID					
	RO Feed	Permeate		Concentrate		
		Train 1	Train 2	Train 1	Train 2	
Parameter		**HYD**	**DOW**	**HYD**	**DOW**	**Units**
TDS	34800	200	160	71400	68600	mg/L
Lab pH*	8.0	7.1	6.8	7.7	7.8	UNITS
Alkalinity (as CaCO3)	108	<2	<2	205	205	mg/L
Bicarbonate (as CaCO3)	107	<2	<2	204	204	mg/L
Carbonate (as CaCO3)	1.0	<0.1	<0.1	1	1	mg/L
Hydroxide (as CaCO3)	0.0	<0.01	<0.01	0	0	mg/L
Sulfate	2492	<10	<10	5370	5276	mg/L
Chloride	18580	112.8	93.1	35000	34460	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	1304	1.0	1.3	2514	2498	mg/L
Hardness (as CaCO3)	6392	4.3	6.4	12326	12231	mg/L
Ca Hardness (as CaCO3)	1021	<1.2	1.5	1974	1945	mg/L
Sodium	10480	75.2	57.3	20240	20040	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: Ave temperature 21C, 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. A comparison of the performance of each membrane versus that predicted by the projection software programs is listed below:

Table 24 - RO Performance vs. Predicted

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

* *Hydranautics software did not predict Boron rejection at that time*

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

Overall permeate concentration for both membranes operating in Trials II – IV experienced increases which are considered abnormal. Both the steady increase over a period of operation as observed with 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 was planned (8 GFD) for Phase B of the testing.

RO Concentrate (Waste) Characterization

The RO concentrate stream was sampled biweekly for the parameters listed above in Tables 21, 22 and 23 in order to characterize the RO waste stream. The recovery of the RO was 50% 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 power plant influent and effluent, as well as during seasonal water quality events such as Red Tides and biofouling episodes.

While 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 power plant 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, as the SWC4+ membrane had demonstrated the highest chloride rejection of all membranes previously tested. This selection provided Phase B-3 with the membrane which most efficiently removed boron (Dow) and the one which achieved the lowest chloride concentration (Hydranautics).

Table 25 lists the operating parameters of the RO membranes during the Phase B period of testing:

Table 25 - Details of Each Phase B RO Run
Phase B1 Summary
Feed Source is Power Plant Influent

Run #	Dates	Pretreatment Chemical	Antiscalant MG/L	Membrane A	Membrane A Flux (GFD) / % Recovery	Membrane B	Membrane B Flux (GFD) / % Recovery	Comments
RO11	6/10/04 to 11/16/04	3 mg/L SBS	3	Hydranautics SWC1-4040 Set B	12 GFD / 50%	Dow SW30-4040 Set B	12 GFD / 50%	Flux increased from 11 to 12 GFD to investigate performance at higher flux.
RO12	11/17/04 to 12/10/04	3 mg/L SBS	3	Hydranautics SWC1-4040 Set B	8 GFD / 50%	Dow SW30-4040 Set B	8 GFD / 50%	Flux reduced back to 8 GFD to compare performance vs. previous runs.
RO13	12/17/04 to 2/24/05	3 mg/L SBS	3	None	NA	Toray TM810	10, 12 GFD / 50%	Begin testing of next generation RO membranes
RO14	2/25/05 to 4/27/05	3 mg/L SBS	3	None	NA	Koch 1820SS	10, 12 GFD / 50%	
RO15	5/15/05 to 7/17/05	3 mg/L SBS	3	Dow SW30HRL E-4040	10, 12 GFD / 50%	Hydranautics SWC4+ 4040	10, 12 GFD / 50%	Red Tide event started in late May/ Early June. RO membranes experienced fouling

Phase B2 Summary -*Feed Source is Power Plant Effluent*

Run #	**Dates**	**Pretreatment Chemical**	**Antiscalant MG/L**	**Membrane A**	**Membrane A Flux (GFD) / % Recovery**	**Membrane B**	**Membrane B Flux (GFD) / % Recovery**	**Notes**
RO16	7/18/05 to 12/5/05	3 mg/L SBS	3	Dow SW30HRL E-4040	10,12 GFD / 50%	Hydranautics SWC4+ 4040	10,12 GFD / 50%	
RO17	12/06/05 to 5/20/06	3 mg/L SBS	3	Toray TM810	12 GFD / 50%	Koch 1820SS	10,12 GFD / 50%	Operation reverted to Influent water Feb 10th due to feed pump issues. RO Fouling occurred in mid March, coinciding with another algae bloom.

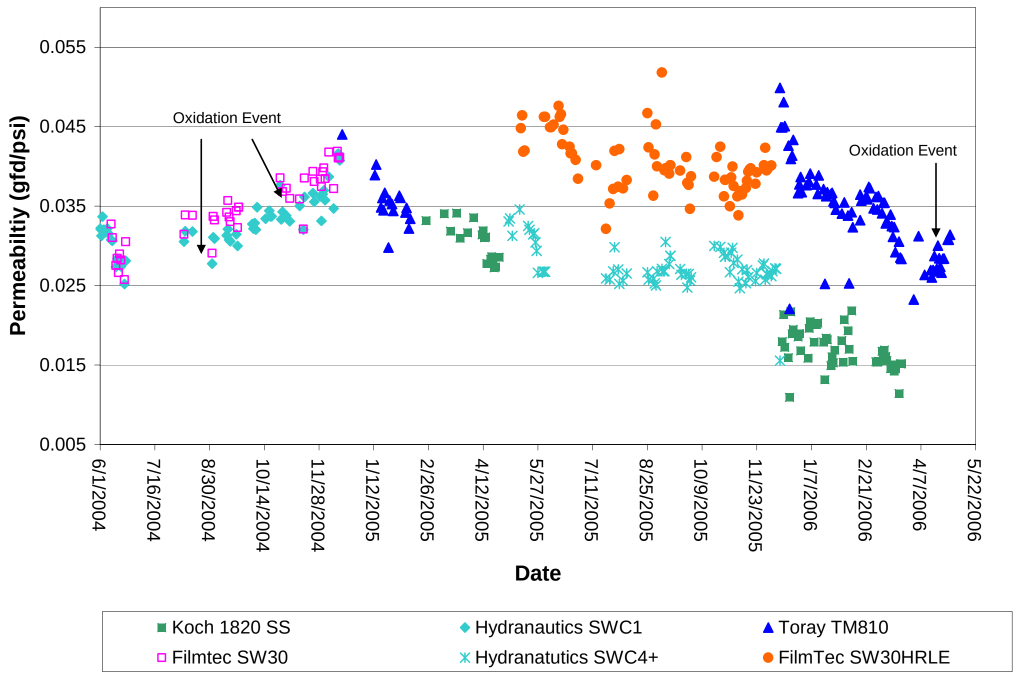
Phase B3 Summary - *Feed Source is Power Plant Effluent*

Run #	Dates	Pretreatment Chemical	Antiscalant MG/L	Membrane A	Membrane A Flux (GFD) / % Recovery	Membrane B	Membrane B Flux (GFD) / % Recovery	Notes
RO18	5/23/06 to 8/1/06	3 mg/L SBS	3	Dow SW30HRL E-4040	12 GFD / 50%	Toray TM810 Set B	12 GFD / 50%	Dow SW30HRLE -4040 and Toray TM810 selected for further testing
RO19	8-1-06 to 10-15-06	3 mg/L SBS	3	Dow SW30HRL E-4040 Set B	12 GFD / 50%	Toray TM810 Set B	12 GFD / 50%	RO HP Pump failure required new set of Dow membranes to be installed. Biofouling of Toray membranes experienced, CIP restored performance
RO20	6/11/07 Through September 2007	3 mg/L SBS	3	Dow SW30HRL E-4040 Set B	12 GFD / 50%	Hydranautics SWC4+ 4040	12 GFD / 50%	Hydranautics installed for further evaluation based on possible need for higher chloride and boron removal. Biofouling experienced for both trains.

RO Permeability

Figure 45 displays the permeability of all membranes tested in Phases B1 and B2. June 2004 through November 2004 consisted of further evaluation of the Dow SW30-4040 and Hydranautics SWC1-4040 membrane at a flux rate of 12 GFD to compare performance at previous flux rates of 8, 9 and 11. Unfortunately, an operational upset with the sodium bisulfite pump caused 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 power plant influent water. The Toray membrane showed strong performance with respect to both permeability and permeate quality.

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

In May – 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 experienced permeability loss during this time frame, and it is possible that dissolved organics present as the result of the algae bloom passed through the MF membrane and fouled the RO membrane.

In December of 2005, the Toray TM810 and Koch 1820SS membranes were reinserted into the system for continued testing on Phase B-2 power plant effluent. The Toray membranes started up with higher permeability and higher conductivity than when operated in Phase B-1, and after substantial troubleshooting, 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 occurring in the membranes as they were in storage for 6 months. On February 10th, operation reverted back to influent water operation due to a malfunction of the effluent water supply pump. In mid March both sets of membranes experienced a loss in permeability. This event coincided with an algae bloom, confirmed by elevated levels of domoic acid present in the feedwater as well as by satellite imagery of the Santa Monica Bay source water.

Light energy utilized in photosynthesis by higher plants and algae cells 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 46 - Chlorophyll-a levels off the coast of southern California September 2006

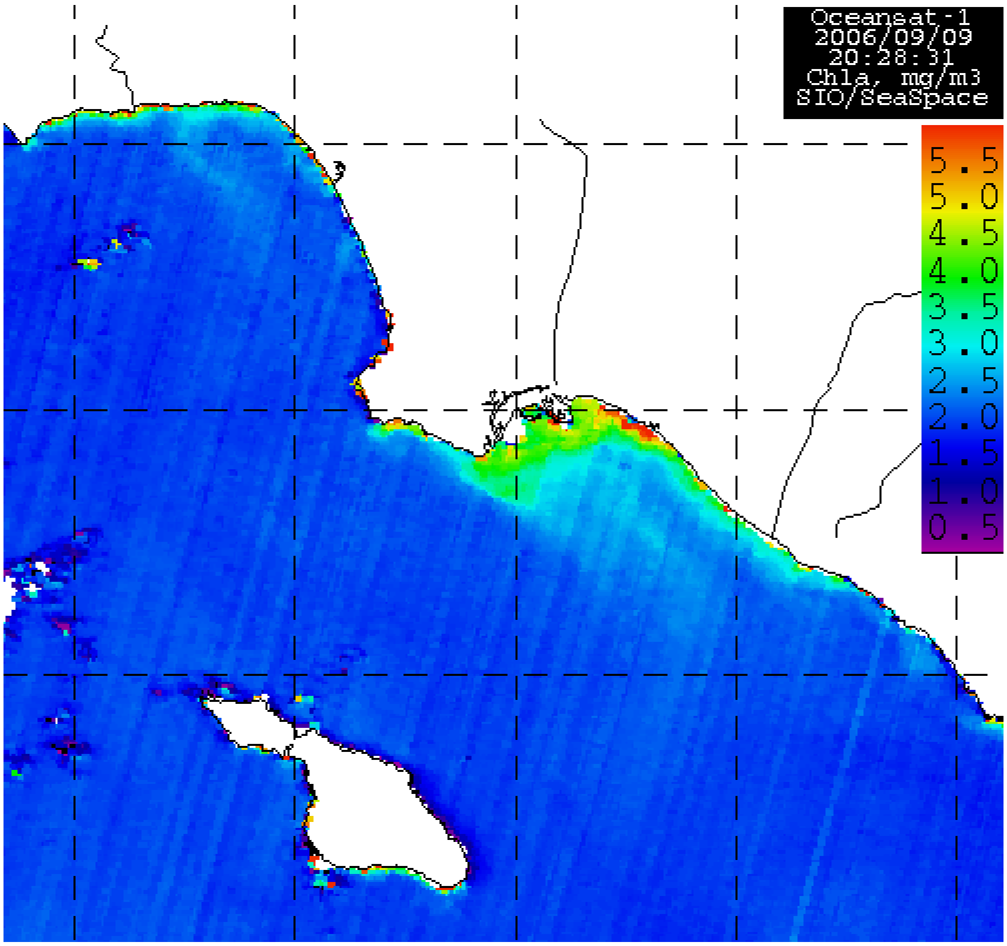
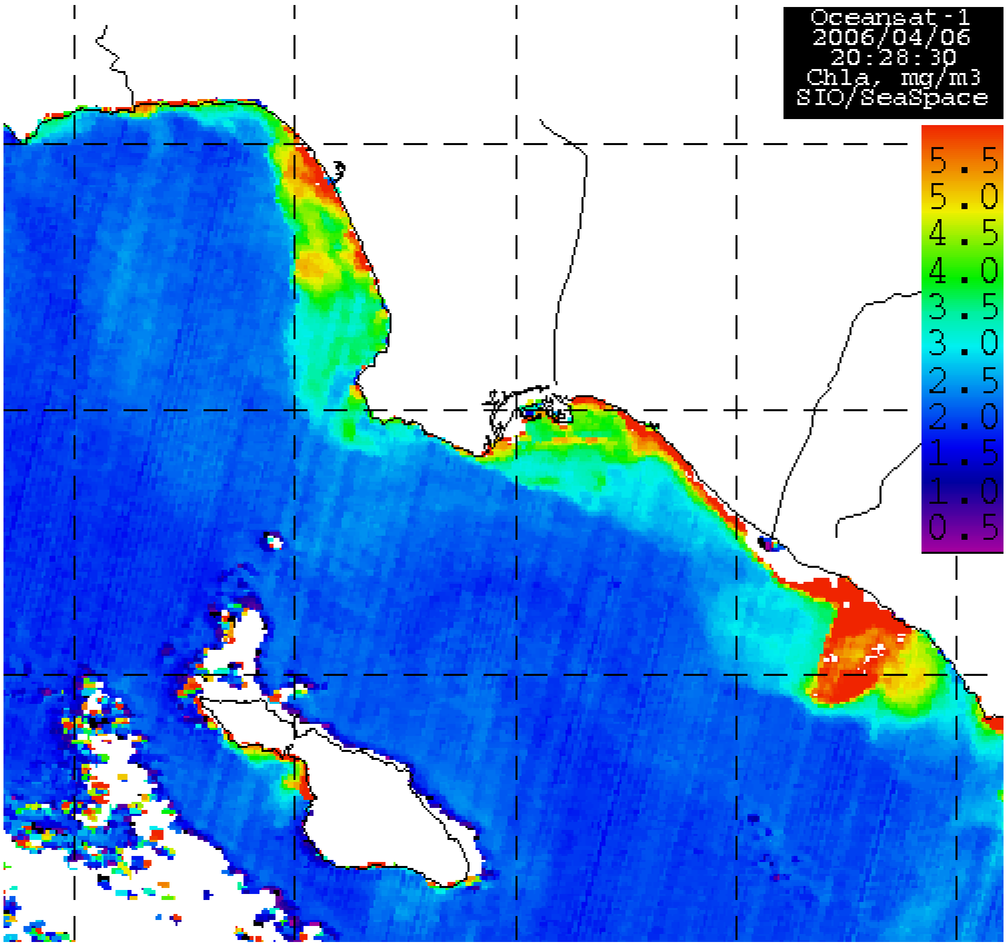


Figure 47 depicts the chlorophyll-a levels during the algal bloom in April 2006.

Figure 47 - Elevated chlorophyll-a levels off the coast of southern California April 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 feedwater, 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.

The Toray TM810 and Dow SW30HRLE membranes were operated from June 2006 to October 2006 on power plant effluent. A high pressure feed pump seal failure leaked oil into the feed water 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 2006 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% citric acid cleaning solution (pH ~2) heated to 35 – 38 ºC followed by a caustic cleaning solution with 2% Avista P111 membrane cleaner (pH ~ 10.5) heated to 35 – 38 ºC was successful in restoring performance.

Figure 48 - Toray TM810 Permeability August 2006 – October 2006

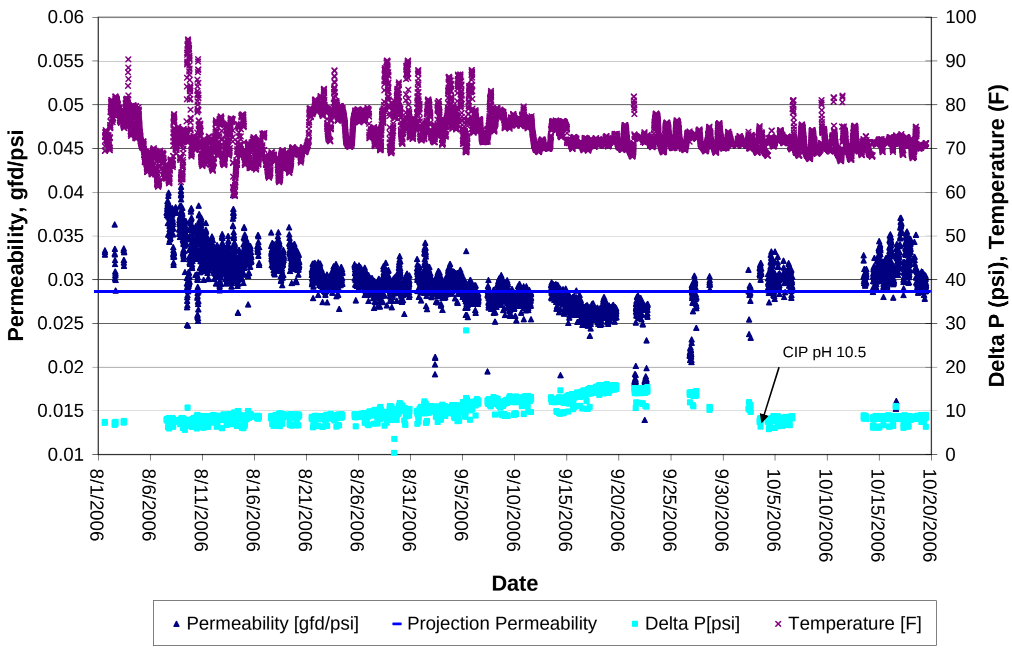
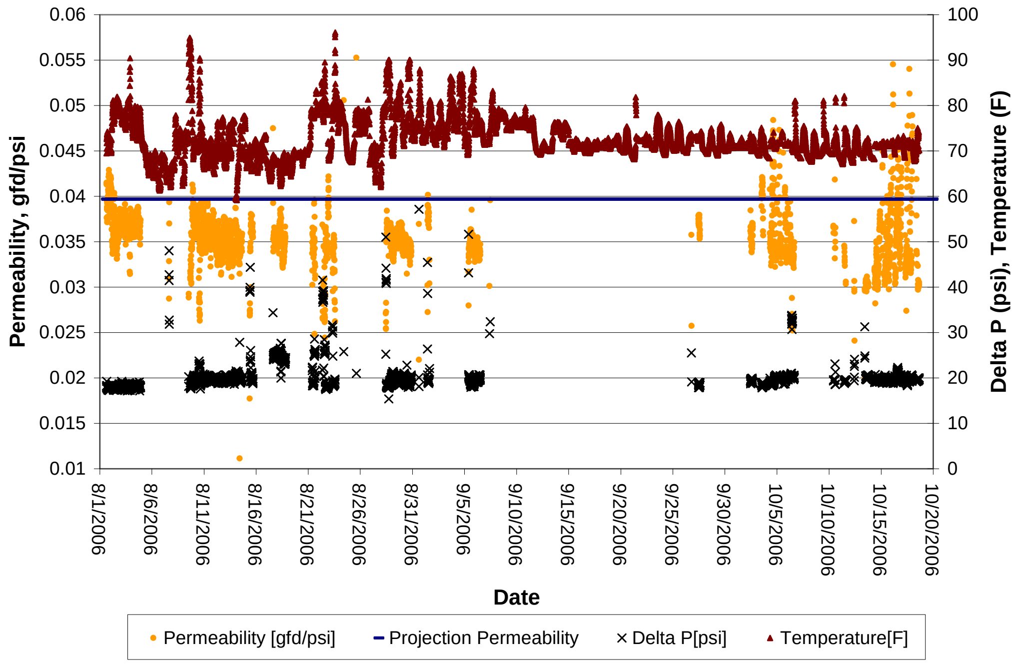


Figure 49 illustrates the Dow SW30HRLE membrane operation from August to October 2006. Mechanical issues as discussed in the Process and Equipment Challenges Section of this document limited the run time during this period, but a loss in permeability was witnessed for the Dow membranes.

Figure 49 - Dow SW30HRLE Permeability August 2006 – October 2006



Phase B3 restarted in June 2007 with a new set of Hydranautics SWC4+ membrane to further evaluate the low TDS permeate quality seen in previous testing, along with the previous set of Dow SW30HRLE membrane. When the Dow RO membranes were brought back on line in June 07 the permeability declined. In early September 2007, a CIP was performed consisting of a 2% citric acid cleaning step (pH ~2) heated to 35 – 38 ºC followed by a caustic cleaning step with 2% Avista P111 membrane cleaner (pH ~ 10.5) heated to 35 – 38 ºC. This is the same cleaning procedure that was used successfully on the Toray membranes in September 2006, however it had no effect on restoring 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 trialed 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% 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 30-35ºC (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 membranes, but initial data for the Dow membranes suggests that more foulant may be able to be removed with another cleaning step.

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

Figure 50 - Dow SW30HRLE Permeability June 2007 – September 2007

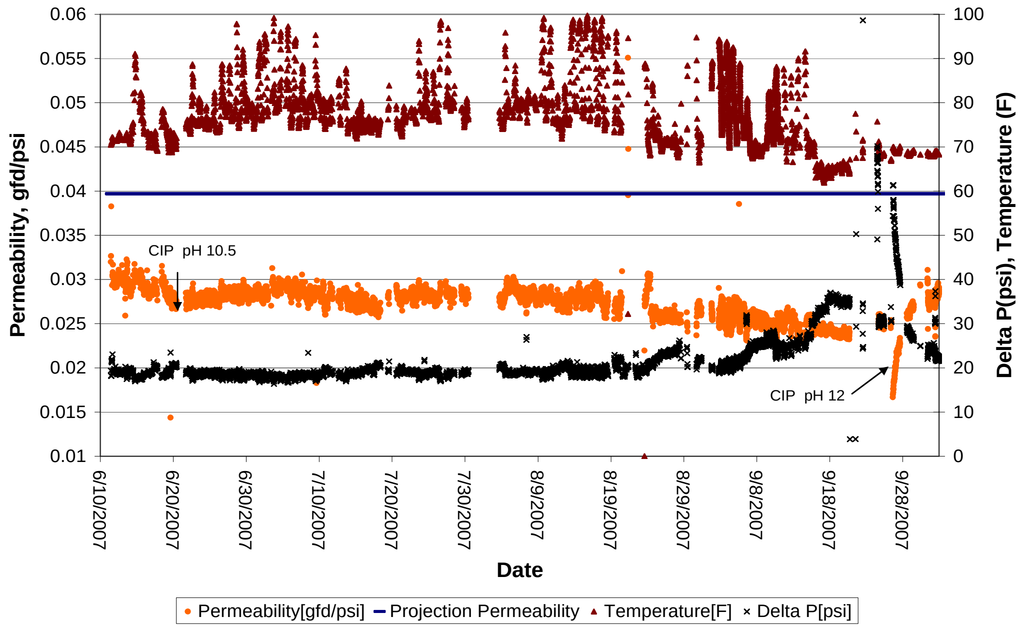
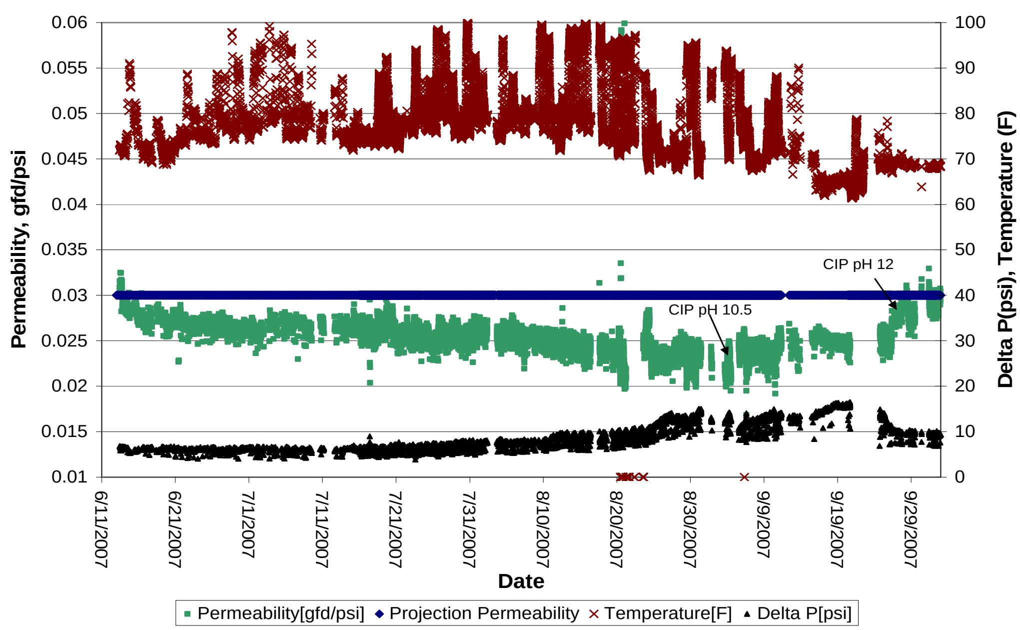


Figure 51 - Hydranautics SWC4+ Permeability June 2007 – September 2007



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

Table 26 - Startup feed pressure requirements

	Feed Pressure Normalized to 25ºC, psi	
Membrane	**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 power plant influent water at a temperature of approximately 65ºF and the other on influent water with an average temperature range of 60-65ºF. 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 power plant effluent water at 72-78ºF, 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 power plant effluent water, with an elevated temperature range of 75-90ºF. There was a continuous presence 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 as well as 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+ membrane. This fouling coincided with a severe algae bloom present in the ocean water where the pilot plant is located. The feed water source was influent water, with an average temperature of approximately 65ºF. A two step cleaning procedure was used for this first fouling event. Step 1 was a 2% citric acid (pH ~2) heated to 35 – 38ºC. Step 2 was a high pH with a generic formulation of:

- 1% sodium tripolyphosphate,
- 1% tetrasodium EDTA
- 1% trisodium phosphate
- The pH was adjusted to 11 and heated to 35 – 38ºC.

This generic formulation is commonly used for cleaning RO membrane; however, it had no effect on restoring 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 feedwater and by satellite imagery. The feed water source was influent water with an average temperature range of 60-65ºF. 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% solution, pH 11) and P112 (1% solution, pH 12), both heated to 35ºC. The P111 cleaner improved #4042 permeability by 23%, and the P112 cleaner improved # 4010 permeability by 27% bringing the flow within 16% of its original factory flow data. 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 feedwater source was effluent water, and water temperature was elevated to an approximate range of 72-78ºF. Since there was no evidence of an algae bloom during this time frame, and biogrowth was found in the break tanks, a cleaning with 2% citric acid (pH ~2) and Avista P111 (pH~ 10.5), both heated to 35 – 38ºC, was performed on the Toray membrane. This cleaning proved to be successful in restoring permeability. This procedure was not able to be performed on the Dow membranes due to the timing of pilot plant relocation effort.

The fourth and final fouling event occurred in August and September of 2007 on the same set of Dow membranes mentioned in the above paragraph, and a new set of Hydranautics SWC4+ membranes. The feedwater source was effluent water, with an average temperature range of approximately 75-90ºF. The power plant was running consistently during the summer of 2007, and temperature spikes occasional reached 100ºF. There was also a persistent presence 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 RO system piping. A cleaning with 2% citric acid (pH ~2) and Avista P111 (pH~ 10.5), both heated to 35 – 38ºC, was performed on the Hydranautics membrane with no effect on restoring performance. Biogrowth continued in the system until another CIP was implemented on the Hydranautics membrane approximately two weeks later. This CIP procedure utilized a 2% citric acid (pH ~2) heated to 35º C and 1% Avista P112 (pH 12) heated to only 30º 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 utilized on the Dow membranes, with the only difference being heating the P112 solution to 35º C, per Dow specifications. This procedure did have some effect on restoring permeability, but the operating data after the cleaning suggests that more foulant may be able to be removed.

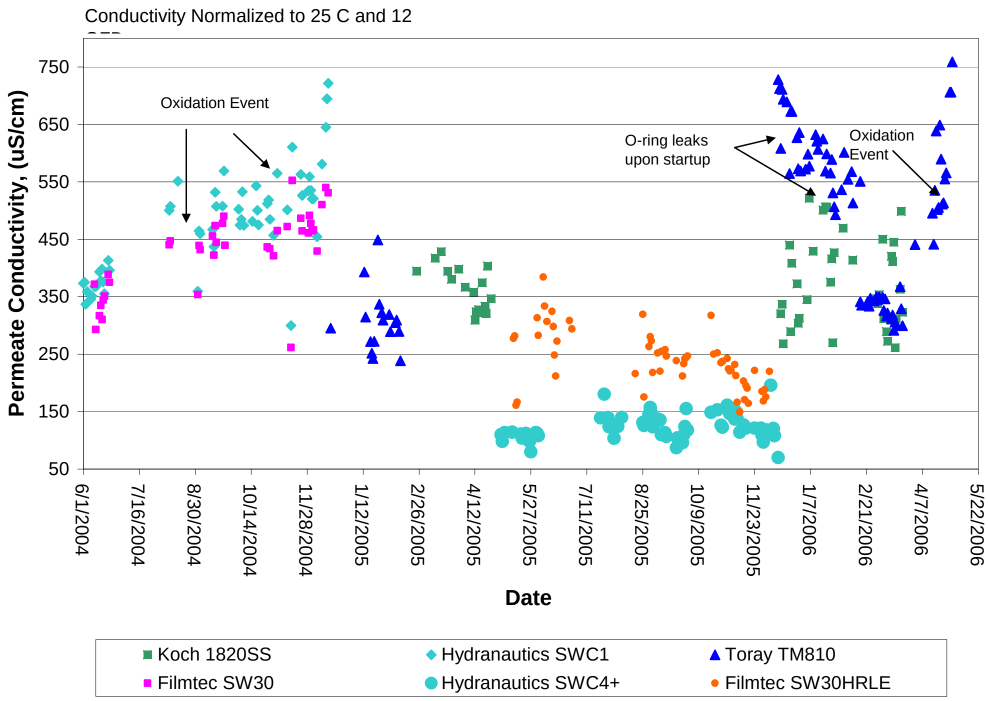
RO Permeate Water Quality

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 (lowest permeate conductivity) of all membranes tested, followed by the Dow (Filmtec) SW30HRLE and Toray TM810 respectively. The Koch 1820SS membrane showed the lowest rejection of the next generation RO membranes.

It should be noted that there were two operational upsets previously mentioned that 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 for each of these membranes (greater than 450 µS/cm) can be seen in Figure 52.

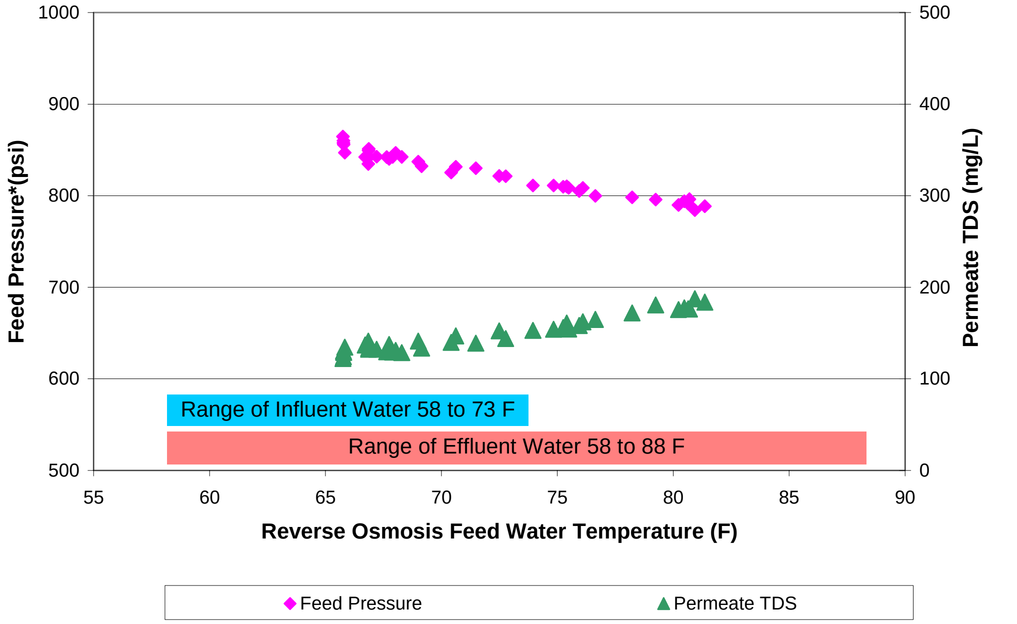
Another noteworthy point relates to the re-installation of the Toray and Koch membranes in December 2005. After substantial troubleshooting involving o-ring leaks with the Toray membrane, two new elements were installed on February 16, 2006 and permeate conductivity returned to the values seen in previous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

Figure 52 - Summary of RO Conductivity



One important aspect of RO membranes is their response to changes in feed water temperature. When the temperature of the feedwater 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 feedwater 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 to a temperature band of 65-80ºF. 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 RO Membrane



*Feed Pressure Normalized to 10 GFD, Not Normalized for Temperature, SW30HRLE Membrane

Figure 54 - Permeate Boron Concentration vs. Temperature at 12 gfd

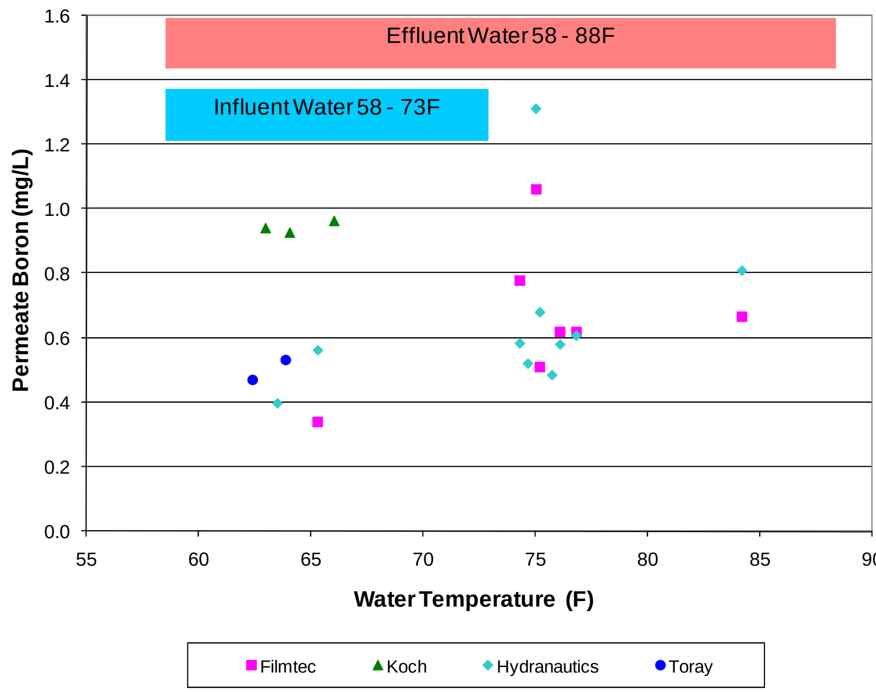


Figure 55 - Permeate Chloride Concentration vs. Temperature at 12 gfd

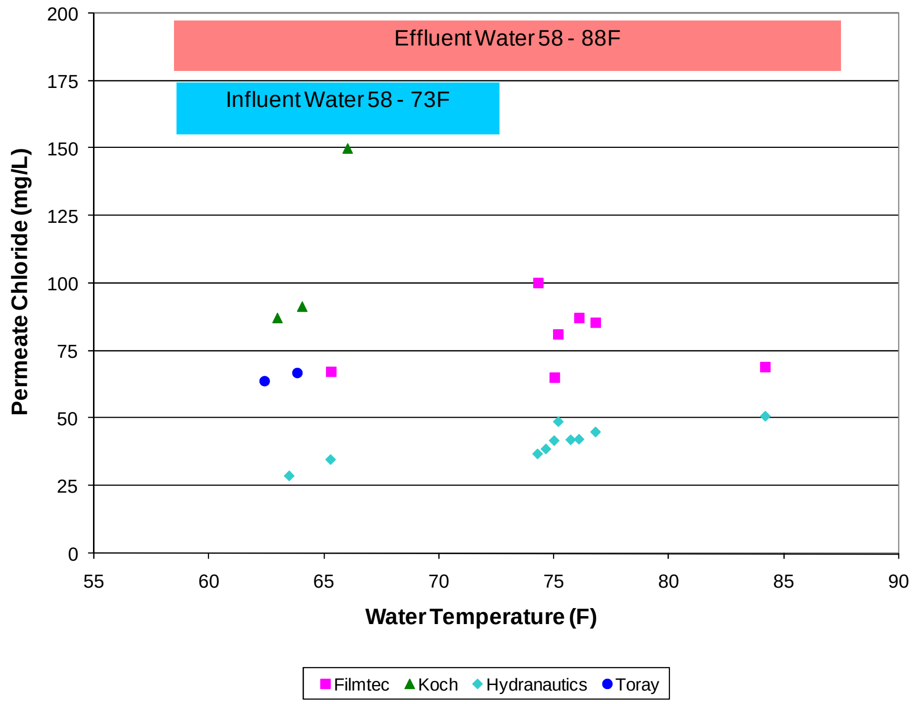


Table 27 (27a and 27b) displays the average feed, permeate and concentrate water quality from the operation of the membranes. Of particular interest in the RO Permeate are TDS, boron, and chloride. The current notification level from CA Department of Health Services for boron is 1 mg/l. Chloride levels 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 27a - Average Water Quality June 2004 – July 2006, Hydranautics and Dow

	Hydranautics SWC4+ 12 GFD Average Temp 23.4 °C			Filmtec SW30HRLE 12 GFD Average Temp 24.1 °C			
Parameter	**RO Feed**	**RO Permeate**	**RO Concentrate**	**RO Feed**	**RO Permeate**	**RO Concentrate**	**Units**
TDS	33889	69.8	68400	33857	128.9	57714	mg/L
Lab pH*	8	6.4	7.8	8	6.6	7.8	UNITS
Alkalinity (as CaCO3)	111	<2	217	111	2.5	187	mg/L
Bicarbonate (as CaCO3)	110	<2	216	110	2.4	186	mg/L
Carbonate (as CaCO3)	1	<0.1	1.26	1.06	<0.1	1.11	mg/L
Hydroxide (as CaCO3)	0.05	<0.01	0.03	0.05	<0.01	0	mg/L
Sulfate	2629	<2	5752	2636	4.5	5120	mg/L
Chloride	19944	41.3	39200	20057	79	33557	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	1252	0.1	2493	1240	1.7	2087	mg/L
Hardness (as CaCO3)	6111	0.5	12172	6053	8.2	10190	mg/L
Ca Hardness (as CaCO3)	954	<0.25	1905	947	1.3	1595	mg/L
Sodium	10667	25.3	21340	10671	47.8	18171	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 27b - Average Water Quality June 2004 – July 2006, Toray and Koch

	Toray TM810 12 GFD Average Temp 20.2 °C			Koch 1820SS 12 GFD Average Temp 17.8 °C			
Parameter	**RO Feed**	**RO Permeate**	**RO Concentrate**	**RO Feed**	**RO Permeate**	**RO Concentrate**	**Units**
TDS	34000	127.5	64000	32500	140	62500	mg/L
Lab pH*	8	6.9	7.7	8	6.7	7.8	UNITS
Alkalinity (as CaCO3)	111	2.5	206	104	<2	205	mg/L
Bicarbonate (as CaCO3)	109	2.5	205	103	<2	203	mg/L
Carbonate (as CaCO3)	0.98	<0.1	0.98	0.97	<0.1	1.21	mg/L
Hydroxide (as CaCO3)	0.05	<0.01	0.03	0.05	<0.01	0.03	mg/L
Sulfate	2570	5.1	5433	2535	11.8	5595	mg/L
Chloride	19025	75	35500	18450	89	33900	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	1225	1.9	2250	1280	0.5	2420	mg/L
Hardness (as CaCO3)	6007	8.8	11068	6232	2.6	11801	mg/L
Ca Hardness (as CaCO3)	962	1.5	1803	961	0.4	1835	mg/L
Sodium	10020	43.7	18333	10350	60.9	19100	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 power plant was operating, with temperatures reaching 100ºF at times. When the temperature of the feedwater is elevated, salt passage through the membrane increased resulting in an 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

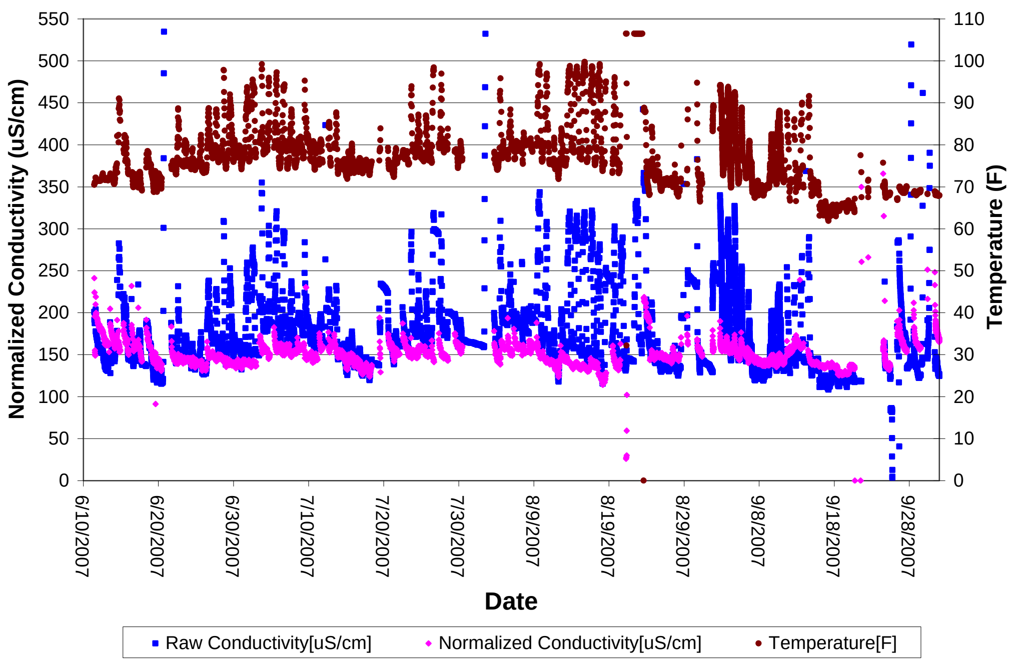
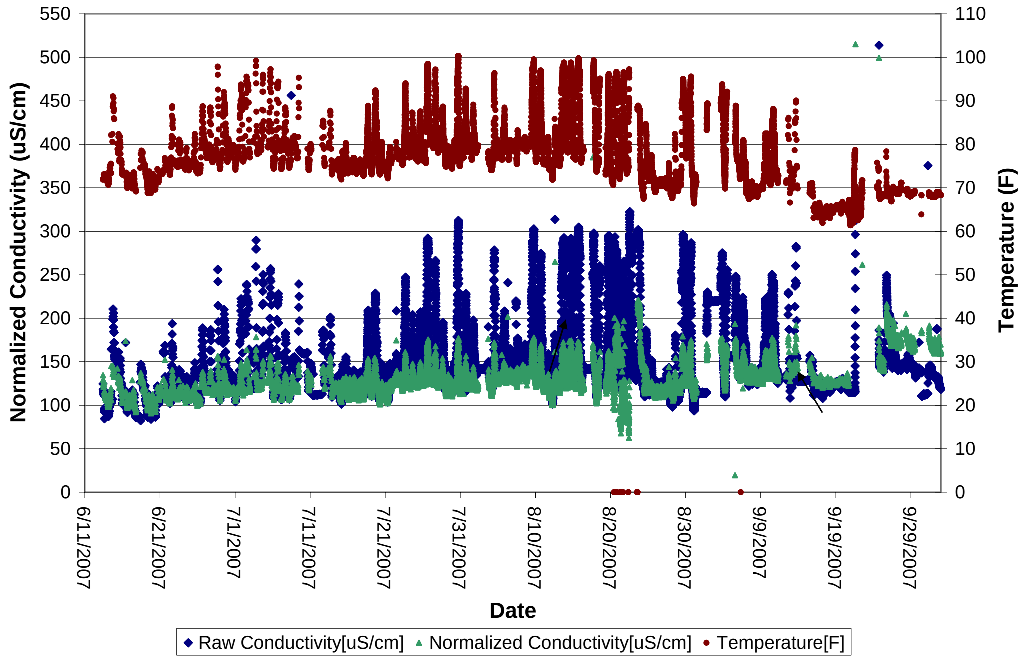


Figure 57 - Hydranautics SWC4+ Permeate Conductivity June 2007 – September 2007



Water samples were collected throughout the period of testing for detailed analyses. The flux rate of the RO membranes were 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 28 and 29 below. The TDS, chloride, and boron concentrations are also compared to the manufacturers' projected performance at those conditions. Both the Dow and Hydranautics membranes experienced permeability loss since 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.

Table 28 - Dow Average Water Quality June 2007 – August 2007

Parameter	Filmtec 8 GFD Ave Temp 25.2°C RO Feed	Filmtec 8 GFD RO Permeate	Filmtec 8 GFD Projected Permeate	Filmtec 10 GFD Ave Temp 28.3°C RO Feed	Filmtec 10 GFD RO Permeate	Filmtec 10 GFD Projected Permeate	Filmtec 12 GFD Ave Temp 22.2ºC RO Feed	Filmtec 12 GFD RO Permeate	Filmtec 12 GFD Projected Permeate	Units
TDS	37000	107	262	38500	105	260	36000	64	139	mg/L
Lab pH*	8.1	7.1		8.2	7.1		8.2	7.3		UNITS
Alkalinity (as CaCO3)	113	<2		115	<2		116	<2		mg/L
Bicarbonate (as CaCO3)	112	<2		113	<2		114	<2		mg/L
Carbonate (as CaCO3)	1.3	<0.1		1.5	<0.1		1.7	<0.1		mg/L
Hydroxide (as CaCO3)	0.06	<0.01		0.071	<0.01		0.08	<0.01		mg/L
Sulfate	2580	2.5		2590	2.5		2630	2.5		mg/L
Chloride	19450	60.2	153	19100	61	152	19350	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	1335	0.94		1355	0.83		1240	0.6		mg/L
Hardness (as CaCO3)	6551	4.6		6626	4		6144	3.2		mg/L
Ca Hardness (as CaCO3)	1054	0.7		1046	0.6		1038	0.5		mg/L
Sodium	11000	38.7		11100	38		10300	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 29 - Hydranautics Average Water Quality June 2007 – Aug 2007

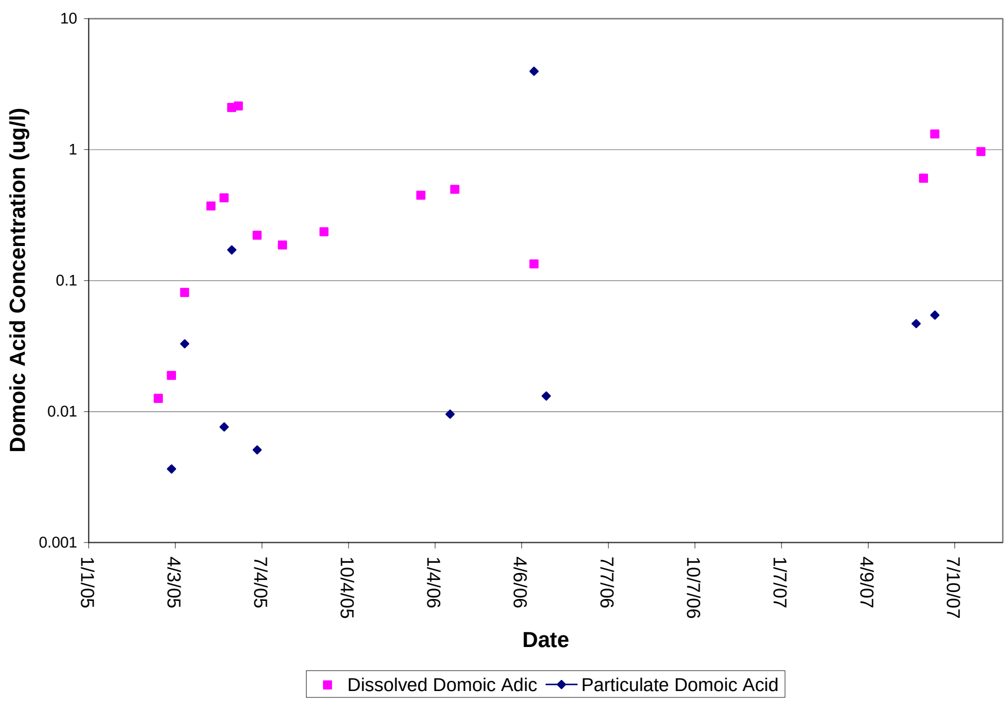
Parameter	Hydranautics 8 GFD Ave Temp 25.2ºC RO Feed	RO Permeate	Projected Permeate	Hydranautics 10 GFD Ave Temp 28.3ºC RO Feed	RO Permeate	Projected Permeate	Hydranautics 12 GFD Ave Temp 22.2ºC RO Feed	RO Permeate	Projected Permeate	Units
TDS	37000	91	194	38500	91	169	36000	58	111	mg/L
Lab pH*	8.1	6.3		8.2	6.3		8.2	6.4		UNITS
Alkalinity (as CaCO3)	113	<2		115	<2		116	<2		mg/L
Bicarbonate (as CaCO3)	112	<2		113	<2		114	<2		mg/L
Carbonate (as CaCO3)	1.3	<0.1		1.5	<0.1		1.7	<0.1		mg/L
Hydroxide (as CaCO3)	0.06	<0.01		0.071	<0.01		0.08	<0.01		mg/L
Sulfate	2580	<2		2590	<2		2630	<2		mg/L
Chloride	19450	51	113	19100	49	99	19350	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	1335	0.39		1355	0.32		1240	0.3		mg/L
Hardness (as CaCO3)	6551	1.9		6626	1.5		6144	1.6		mg/L
Ca Hardness (as CaCO3)	1054	0.3		1046	0.3		1038	0.7		mg/L
Sodium	11000	32		11100	31		10300	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

Algal Toxins

Another important water quality aspect of ocean water desalination has to do with the presence of algal toxins in the ocean 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, that 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 molecular weight of domoic acid (311) is large enough to be rejected by the RO membrane.

Figure 58 - Domoic Acid Levels in Ocean water 2005 - 2007



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 less than 300 mg/L TDS on the influent water throughout its temperature range.

Phase B testing provided valuable information on four next generation RO membranes. Of the four membranes tested, Dow SW30HRLE, Hydranautics SWC4+, Toray TM810, and Koch 1820SS, all but the Koch 1820SS product warranted consideration for further testing. The Koch 1820SS lower boron rejection and lower permeability 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.

The difference in chloride rejection versus boron rejection among the membranes tested was unexpected and noteworthy to those developing full-scale implementation of ocean water RO. The Hydranautics SWC4+ achieved similar permeate boron concentrations as Dow and Toray membranes, but demonstrated substantially lower chloride concentration, 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 power plant influent and the warmer power plant 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 can be seen in Figure 53.

Operation of the power plant during the summer months coincided with an increase in algal biomass in the ocean. This increased presence of marine microorganisms together with the elevated temperatures of the post condenser effluent seems to exacerbate RO biofouling. Cleaning trials over the course of Phase B testing indicate 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 differences between Phase A and B is 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. All permeate samples tested resulted in Non Detect, even when levels of domoic acid in the feedwater were considered high when compared to average concentrations. The lower detection limit in the test for presence of domoic acid is 0.002 μg/liter.

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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. (American Water Works Assoc., 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 past ocean water RO installations on 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. (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→MF→RO process has been used successfully on many wastewater reclamation facilities including the 20 MGD West Basin Water Recycling Plant. 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[5] $NHCl_2 + HOCl \rightarrow NCl_3 + 2H_2O$

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

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% to 1.5 mg/L. 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. Chlorine in the 20 - 40 mg/L range was then utilized 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, combined with the addition of sodium bisulfite in front of the RO was utilized in the remainder of the trials.

Power Plant Heat Treatment Cycles

The pilot trails were started in June 2002. Soon thereafter, the power plant performed a heat treatment, or "heat treat" cycle. Approximately every one to three months the power plant influent which feeds the pilot equipment was "heat treated" to control biological growth/attachment. The heat treat consists of recirculation of ocean water at 105 – 120 ^{O}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 where shells and other particulate matter were 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 experienced with the first set of MF modules.

Addition of Arkal Spin Klin Filter

The study experienced numerous Siemens CMF-S module fiber breakage events; even after the 800 µm strainer was placed in front of the booster pump as described in the appendix. Siemens underwent 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 utilizes 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 shutdown on 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 this 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. 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 feedwater. 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 housing.

Vibration Issues Associated with Wanner Hydracell High Pressure RO Pumps

The RO System utilized for this study has two independent trains of 4 membranes, four-inch pressure vessels in a 1:1 array. To feed the seven 4" RO membranes in series, the RO pumps produce ~10 gpm at 1000 psig, and this flow/pressure combination was not readily available in a centrifugal pump. Wanner Engineering offers a positive displacement type pump with superaustinitic 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 1000 psig. Vibration produced by the Hydracell pumps was accentuated by the combination of having pumps placed on the skid and 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 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 failures still occurred between the pumps and the pressure vessels. Pulsation dampeners were added to the discharge of the Hydracell pumps, but vibration problems still persisted. Additionally, one of the pumps experienced 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 vibrations and 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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