

West Basin Municipal Water District



Ocean Water Desalination Pilot Plant Project

**Investigation of Microfiltration & Reverse Osmosis
Seawater Desalination**

Final Report

July 2005

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1 EXECUTIVE SUMMARY

As California's water supply has become increasingly stressed in recent years by population growth and lack of new sources, advances in desalination technology are making the consideration of large-scale seawater desalination a viable option. West Basin Municipal Water District (WBMWD), in tune to these developments as a result of its extensive desalination facilities for wastewater reclamation, initiated a seawater desalination pilot plant project in 2002. The project investigated the use of microfiltration (MF) membrane processes as pretreatment to seawater reverse osmosis (RO). The objectives of the study include the optimization of MF and RO operational parameters on open-intake seawater, as well as characterization of product and waste streams produced by the membrane processes. The study, operating on an open seawater intake near Los Angeles, produced data which will not only be of value for West Basin's planning and future implementation of desalination, but also those applying these technologies at other locations.

This project, involving the operation of the MF and RO processes on the seawater intake of the El Segundo Generating Station, was completed in 2004. The MF was a PVDF hollow-fiber membrane in a submerged configuration. Two models of RO membranes were tested in parallel systems. Through over ten thousand hours of operation, successful operating parameters for the processes were developed and demonstrated. Highlights of the results include:

- **MF Operation:** An MF operating strategy was developed which maximized operating flux, while maintaining greater than 21 days between chemical cleanings.
- **MF Filtrate Quality:** The MF process was demonstrated to produce excellent filtrate quality, despite variations in feed quality. As a result the RO membranes operated with essentially no measurable fouling at flux rates varying from 8 to 11 GFD.
- **RO Operation:** RO operation of over 10,000 hours (415 days) did not require chemical cleaning. This operation occurred over a complete seasonal cycle. The RO process has demonstrated exceptionally stable performance. This performance is credited to a relatively good quality raw water supply and consistent pretreatment

performance. This performance is considered exceptional compared to many other open intake seawater applications.

- **RO Permeate Quality:** The reverse osmosis systems demonstrated the ability to produce consistently high quality (less than 300 mg/L TDS, pre-stabilization), meeting or exceeding most of the membrane manufacturer's performance predictions. The product water quality produced is excellent and meets all primary and secondary MCL drinking water standards as well as CA-DHS Notification Levels.
- **Boron Rejection:** Data was generated on boron rejection by the RO process, indicating somewhat different performance between the two membranes tested. One membrane producing permeate boron typically 0.8 mg/L, the other 1.1 mg/L.
- **Source Water Quality:** From a desalination process perspective, the source water quality at the test site was very good. Seawater turbidity following 130 micron strainers was generally less than 2 NTU and averaged approximately 1 NTU. Salinity of the feedwater exhibited very little variation. This is a relatively high-quality, consistent feed source in comparison to many existing full-scale and other pilot desalination facilities.^{1,2,3,4}
- **Concentrate Water Quality:** Concentrate water quality (undiluted) was confirmed to meet all California Ocean Plan objectives, with exception of gross beta. Determination of compliance for gross beta requires further analyses, which are underway.

This study has demonstrated that seawater desalination using the MF pretreatment to RO on power plant intake water can be successfully implemented.

¹ *Pilot Testing of MF and UF for Seawater RO Pretreatment*, Lisa Henthorne and Ryan Quigley, 2003 AWWA Membrane Technology Conference Proceedings

² *Trinidad SWRO*, Dr. Kenneth Irwin & John David Thompson, IDA Desalination & Water Reuse Quarterly November/December 2003

³ *Jeddah SWRO Plant Experience, Pretreatment*, Saeed Saad H. Al-Harhi, SWCC, IDA Workshop: Importance of Pretreatment in Reverse Osmosis Desalination March 26, 2004

⁴ *Spanish Desalination Experience*, M. Fariñas, F. J. Bernaola & F. Beltrán, IDA Workshop: Importance of Pretreatment in Reverse Osmosis Desalination March 26, 2004

2 INTRODUCTION

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) has been proposed as an alternative to conventional pretreatment processes for ocean water reverse osmosis (RO). Microfiltration has become a common pretreatment method for RO installations treating municipal wastewater. MF removes colloidal and suspended particulate matter that would pose problems for the reverse osmosis desalination process. A demonstration plant program to evaluate the combination of MF and RO, for the potential application of ocean water desalination in California for the domestic water supply has been executed.

West Basin Municipal Water District's (WBMWD) Ocean Water Desalination Pilot Plant Program tested the capabilities of MF pretreatment in series with a spiral wound RO system. It developed data to determine the optimum operating conditions and cleaning requirements for MF operating on ocean water, as well as the ocean water reverse osmosis process operating on microfiltration filtrate.

The testing occurred at the El Segundo Power Generation Plant (Figure 1). Seawater desalination is energy intensive and a full-scale seawater desalination plant co-located with an existing ocean water cooled power plant has two advantages. The first advantage is that power can be purchased at relatively low rates "within the fence" of the power plant. In California, this can result in an energy savings of $\geq \$0.05/\text{kWh}$. In addition, the seawater desalination plant can also utilize the existing intake and outfall structures that allow seawater to be brought into the power plant and returned to the ocean.



Figure 1 - Test Equipment at the El Segundo Power Plant

3 TEST OBJECTIVES

The objectives of the *Ocean Water Pilot Test Program* were as follows.

1. To determine the optimum membrane operating flux and in-situ membrane cleaning frequency for a MF 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 membranes.
2. To determine the optimum membrane operating flux and in-situ membrane cleaning frequency for an ocean water RO system operating on MF filtrate. Investigate cleaning formulations and techniques for removal of contaminants found in microfiltered ocean water, which foul RO membranes.
3. Characterize the MF backwash and RO concentrate streams to develop data suitable for evaluation of waste stream disposal options.

4 PROCESS FLOW DESCRIPTION

The pilot plant was located on the California coast in the City of El Segundo at the El Segundo Power Generation Plant. Ocean water was brought through an existing open intake to the power plant cooling system (≈ 200 mgd). Existing treatment by the power station consisted of a coarse traveling screen (>1 inch) and intermittent chlorination. Standard power plant practice consisted of two treatment techniques for controlling organic activity in the cooling loop. Chlorination was manually initiated two times per week for a duration of \sim two hours. The addition rate resulted in a total chlorine concentration at the plant outfall (condenser effluent) of approximately 0.06 mg/L. This dosage translated 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 loop was "heat treated" to control biological growth/attachment. Duration of this treatment is one hour at 105 – 120 °F. The pilot equipment was shut down during the heat treat events.

The feed water was Pacific Ocean water with average characteristics as indicated in Table 1.

Table 1 - Ocean Water Quality

Constituent	Value	Constituent	Value
Calcium	407	Bromide	64
Magnesium	1,335	Boron	3.8
Sodium	10,963	Nitrate (as N)	<25
Potassium	404	Fluoride	0.9
Ammonia (as N)	0.05	Silica	<10
Barium	<0.025	Total Dissolved Solids	34,500
Strontium	7.7	pH	8.1
Bicarbonate (as CaCO ₃)	115	TOC	1.2
Sulfate	2,537	Temperature (°C)	15.5 - 24
Chloride	19,080	Temperature (°F)	60 - 75

All values in mg/L except pH and temp

From a desalination process perspective, the source water quality at the test site is very good. Seawater turbidity following 130-micron strainers was generally less than 2 NTU and averaged approximately 1 NTU (figures 16 and 17). Salinity of the feedwater exhibits very little variation. This is a relatively high-quality, consistent feed source in comparison to many existing full-scale and other pilot desalination facilities.^{1,2,3,4}

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 allowed 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 US Filter 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 consisted of four modules submerged in a process tank. Suction was applied to the lumen of the fibers by the MF filtrate pump, drawing water through the walls of the fibers while particulate matter accumulated on the outside surface of the fibers. The CMF-S process included periodic interruption of filtration for backwashing of the fibers. The filtration period was 15 minutes at the start of the testing. Following the filtration period, the fibers were backwashed by reversing the filtrate flow and introducing an air scour across the membrane's outside surface. Subsequently, the process tank was drained and refilled. The entire backwash operation consumed ~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 flux). Details of the MF testing results are discussed in section 5.

MF filtrate was directed to the 150 gallon covered Break Tank, which served as an equalization tank between the intermittent MF 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 at this location. The ammonium hydroxide dose was on a mole ratio of 2:1 $\text{NH}_3:\text{HOCl}$. 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.

A booster pump to the RO system then pumped MF filtrate from the break tank. 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. 3mg/L Permatreat PC-191 Antiscalant addition was injected downstream of the RO booster pump. 20 micron cartridge filters followed the antiscalant addition and provided 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 H.P. 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-Filmtec were selected to provide RO membranes for this study as these two companies had products meeting the treatment requirements and had a substantial share of worldwide reverse osmosis membrane sales. RO performance is discussed below in sections 6 and 9.

Many process and equipment challenges were experienced over the course of this study. Some of these, as described in the appendix, required modifications to the process flow of the pilot equipment. Figure 3 below contains the final process flow diagram for the pilot equipment. See the appendix for a discussion of the major issues that required process flow modifications.

Figure 2 - Initial Process Flow Diagram of the West Basin Ocean Water Desalination Pilot System

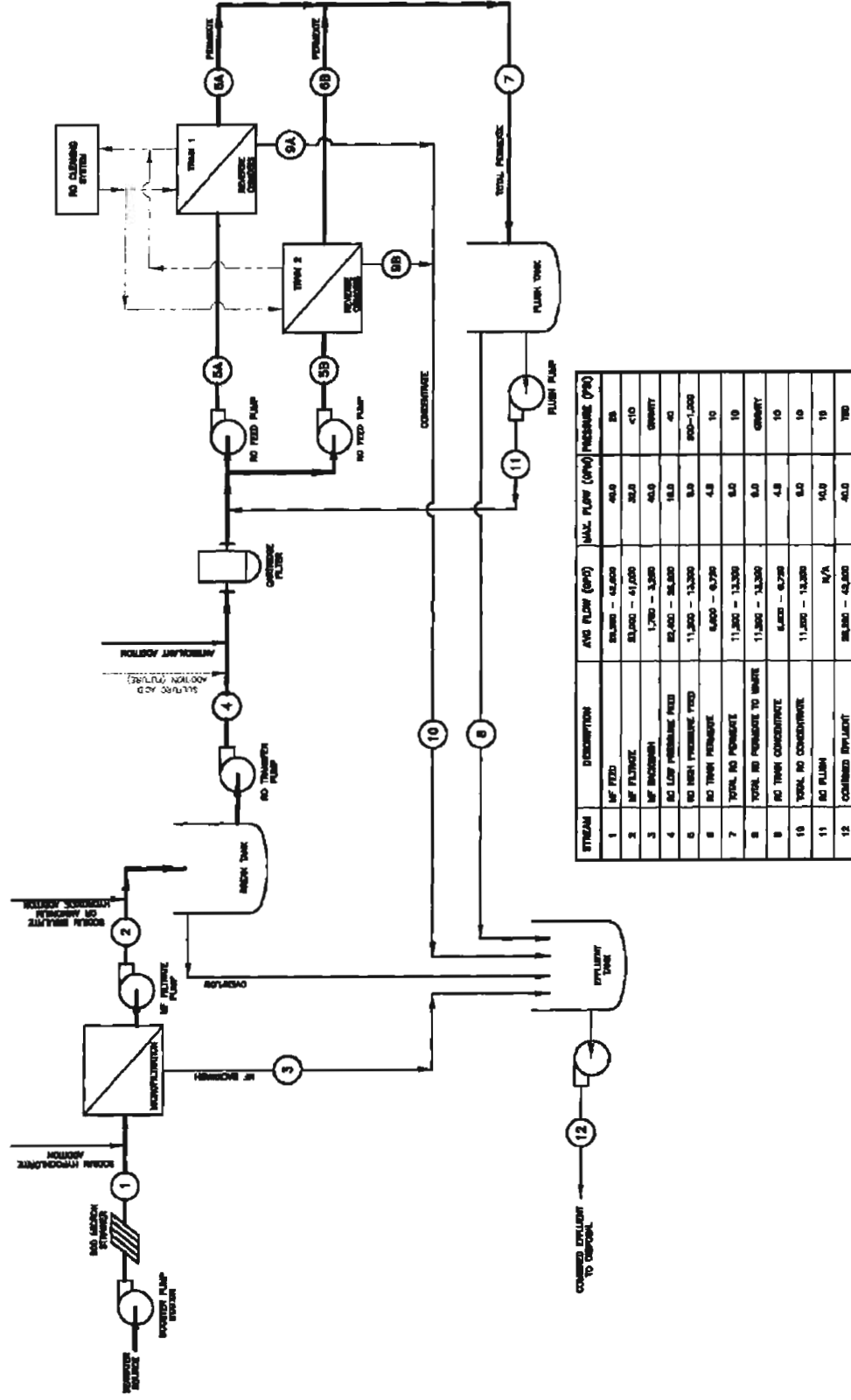
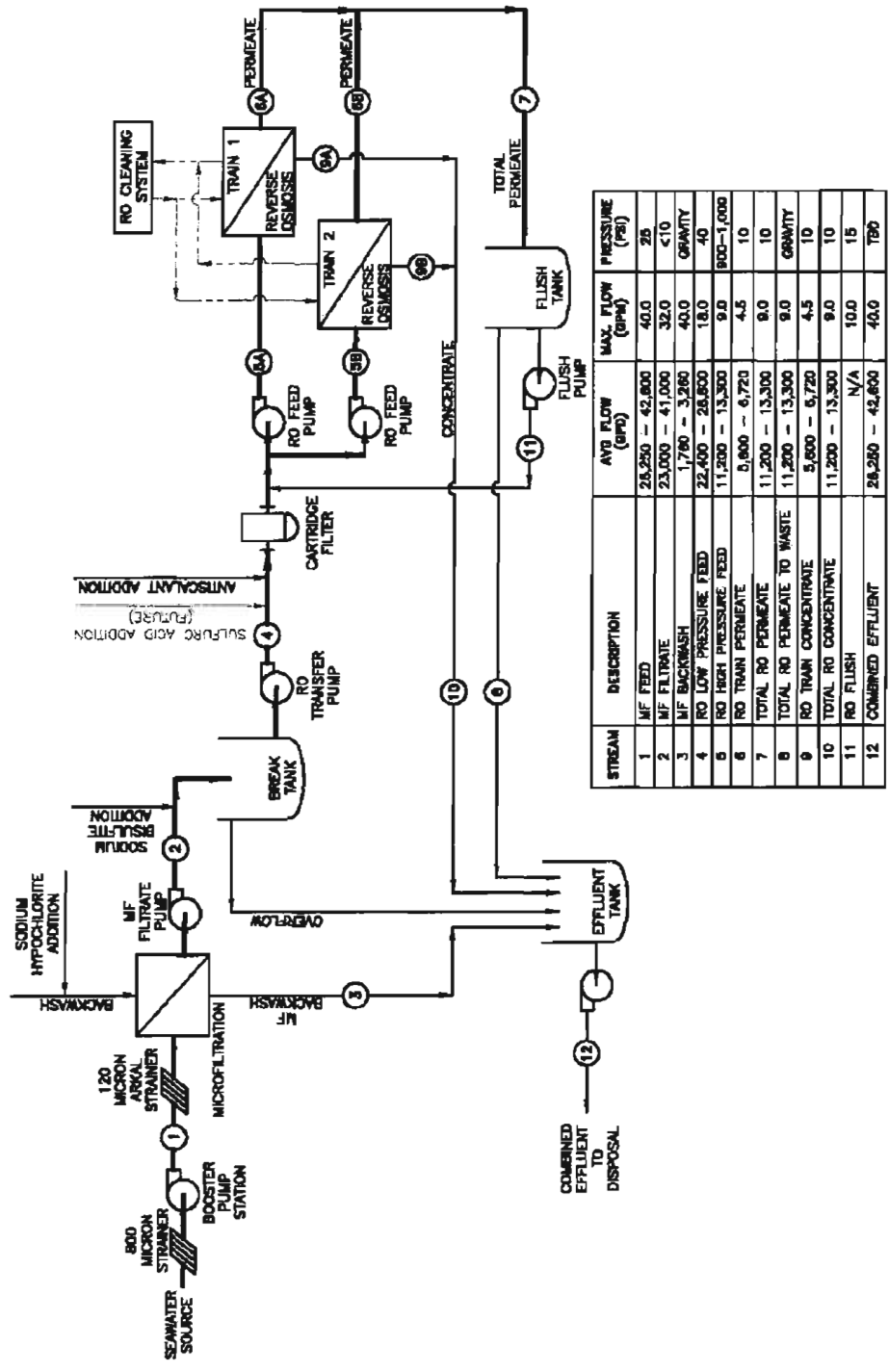
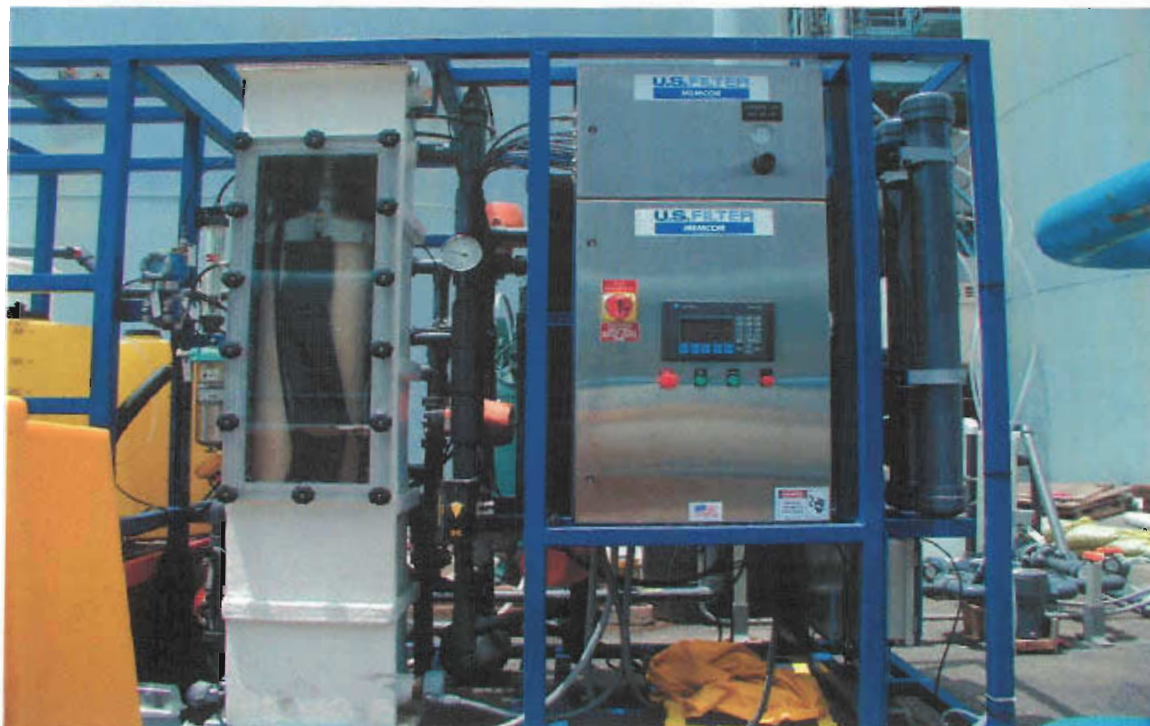


Figure 3 – Revised Process Flow Diagram



5 MICROFILTRATION OPTIMIZATION AND PERFORMANCE

Figure 4 - US Filter CMF-S Microfiltration Pilot System



The US Filter CMF-S system was initiated in June 2002, with the first month used as an equipment troubleshooting period. The first stable run started on July 19. The MF trials are summarized in the following Tables 2 and 3. The testing was 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 - 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 130um strainer in front of MF, operation with chlorination of backwash and redesigned MF module

Table 3 - Details of Each Microfiltration Run

<u>Trial</u>	<u>Run #</u>	<u>Dates</u>	<u>MF Run Hours</u>	<u>Total Filtrate Flow, gpm</u>	<u>Per Module Filtrate Flow, gpm</u>	<u>Flux GFD</u>	<u>Target Feed Chlorination (ppm)</u>	<u>Backwash Frequency, min</u>	<u>Comments</u>
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 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 II	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 III	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.
	MF 10	1/24/03-2/5/03	3820?-4028	24	6	25.8	25 in every backwash	15	RO Membranes replaced
	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
	MF 12	2/21/03-3/6/03	4242-4513	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 IV	MF 13	3/6/03-3/11/03	4513-4623	24	6	25.8	25 in every backwash	15	In advertant daily mini CIP with chlorine improved performance
	MF 14	3/12/03-4/3/03	4650-5100		6		40 in every backwash	15	Various flows
	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	6/8/2004-6/23/04	7314-7750	26	6.5	34	20 in every BW	20	CIP after very short run-modules reconditioned

5.1 Permeability of Original CMF-S Module Design

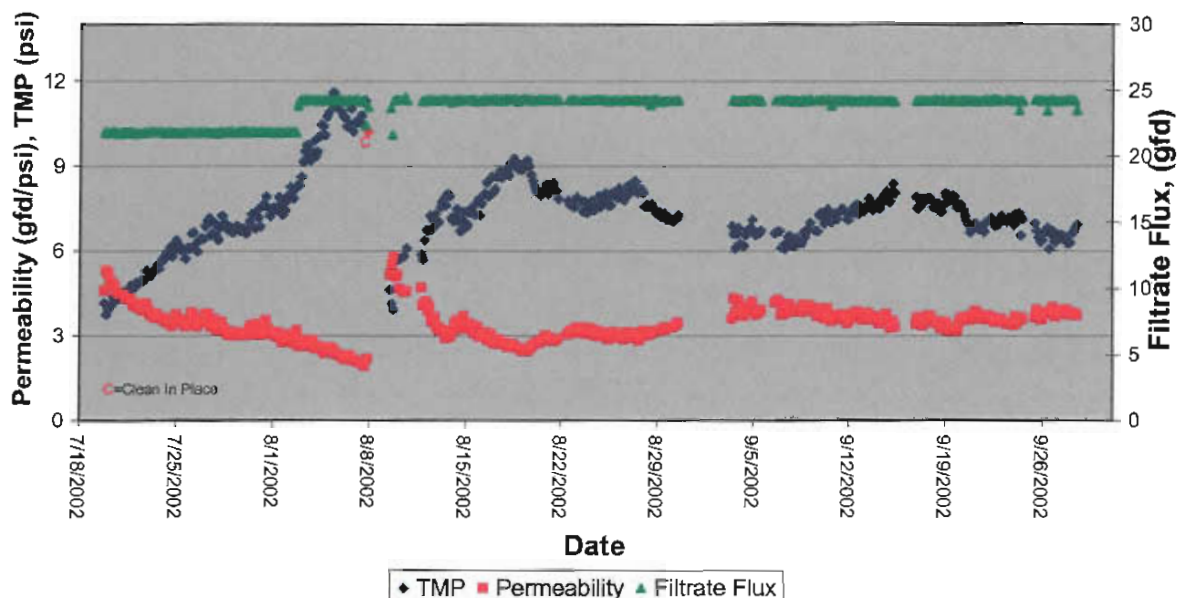
The US Filter CMF-S system runs at constant flux and thus as the membrane fouls, the transmembrane pressure (TMP) required to maintain throughput rises. However, because transmembrane pressure is also influenced by water temperature and variations in flow, the appropriate method of monitoring membrane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

5.1.1 Trial I-Continuous Prechlorination

MF runs 1 and 2 were performed with continuous chloramination at conditions indicated in table 3. The MF demonstrated very stable operation during this period. 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. The continuous chloramination was discontinued following MF Trial I as the NaOCl → MF → NH₄OH → RO 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 than chloramines and the bromamine damaged the downstream RO membranes. See the appendix for further details.

Figure 5 - Permeability of Microfiltration System with Continuous Prechlorination (MF Trial I)

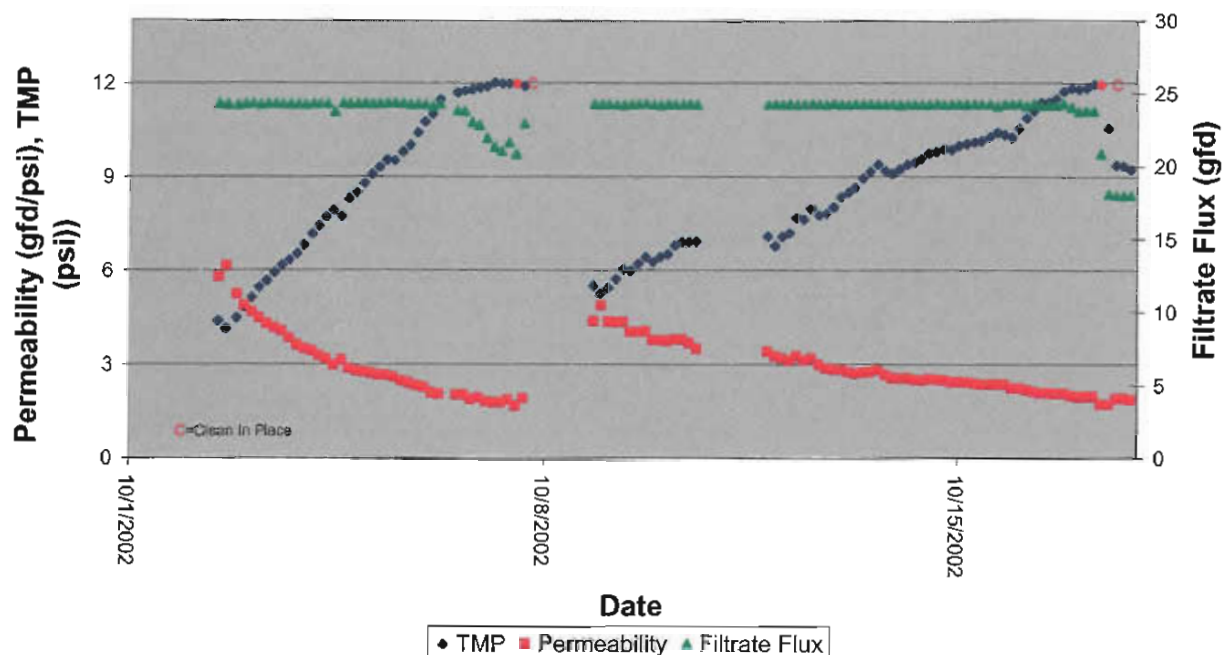


In many past ocean water RO installations on open intakes with conventional filtration pretreatment, a reducing agent, such as sodium bisulfite is added after significant chlorine contact time to neutralize the oxidant before it contacts the RO membranes. However, as demonstrated in a 1996 article in Desalination and Water Reuse, this chlorination/dechlorination process has been shown to enhance the tendency towards biological fouling.⁵ Therefore, this process was not considered a viable option for this study.

5.1.2 Trial II-No Chlorination

Once prechlorination was abandoned, attempts were made to run the US Filter CMF-S system at the same conditions with no chlorination at all. Rapid fouling was observed in two consecutive runs as shown in Figure 6. Note that neither of these runs lasted ten days before reaching terminal permeability. Operation at 24 GFD failed miserably without the chloramination and demonstrated how beneficial the oxidant is to the stable performance of microfiltration membrane process on this feed source.

Figure 6 - Permeability of Microfiltration System with No Chlorination (MF Trial II)

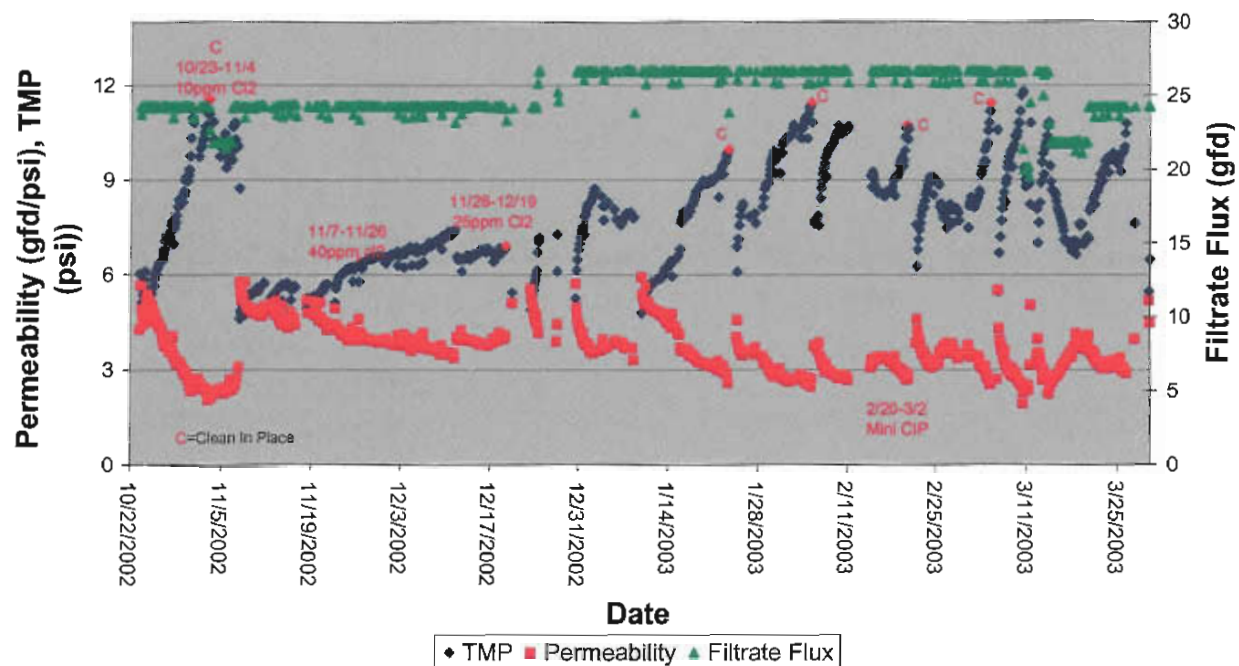


5.1.3 Trial III-Chlorinated Backwashes

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 were implemented. 10 mg/L NaOCl was attempted in every backwash and again rapid fouling was observed as depicted in figure 7. 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 every backwash in run #7, the MF operated for an additional month without requiring a shut down for a chemical clean in place (CIP).

⁵ "Controlling Biological Fouling in Open Sea Intake RO Plants Without Continuous Chlorination," by Hamida and Moch, *Desalination and Water Reuse*, November/December 1996.

Figure 7 - Permeability of Microfiltration System with Chlorinated Backwashes (MF Trial III)



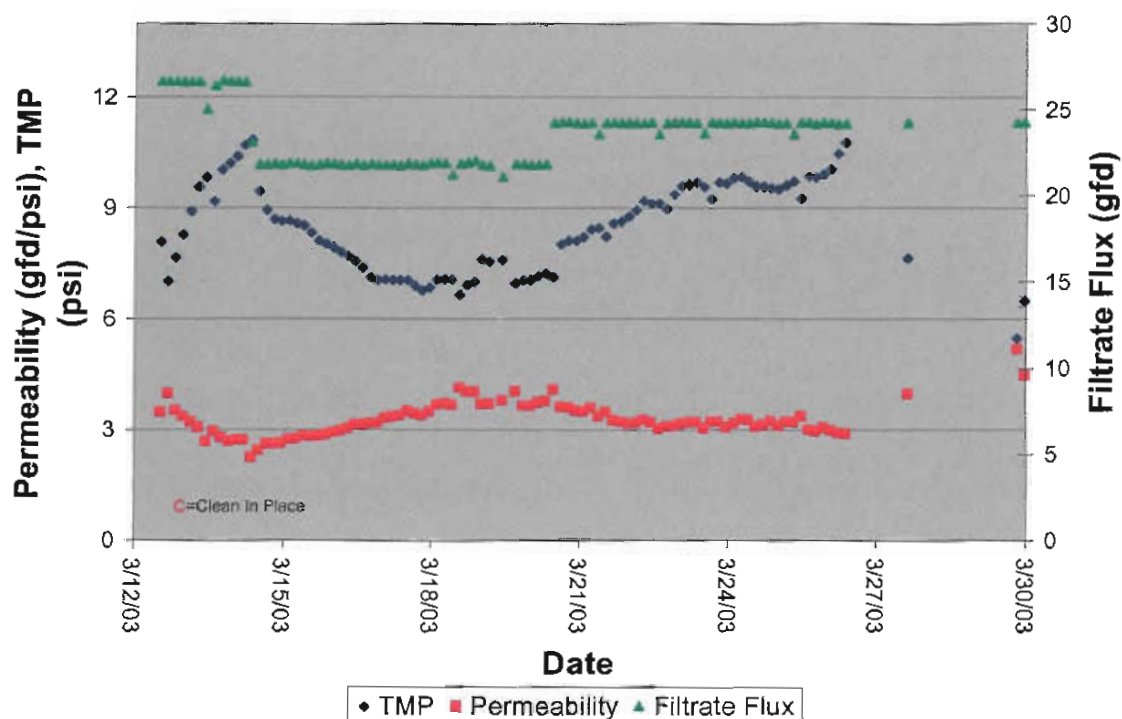
The filtrate flow was then increased from 22 gpm to 24 gpm for run #8, corresponding to a flux increase from 24 to 25.8 GFD. Numerous attempts failed to demonstrate a run time longer than 3 weeks at this flux. This was compounded by the fact that the CMF-S clean-in-place (CIP) heater was disabled for 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 ~6 GFD/psi.

During run 14, the filtrate flow and hence the flux rates were varied as shown in Figure 8.

Figure 8 - Microfiltration Run #14 Specific Flux and Permeate Flux



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.

5.2 Cleaning Effectiveness

Examination of Figure 7 shows that the “clean” or post “Clean-In-Place” microfiltration permeability’s have 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 cleans were performed with cool water on January 23 and February 5, 2003. These cleans were not effective as shown in Figure 7 where the clean permeability’s are only 4 GFD/psi, 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 steps. This enhanced process showed improvement, but failed to fully restore the membranes. Examination of Figures 5 and 6 demonstrates that the following normal heated CIP was quite effective at restoring the membrane permeability. 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. Therefore, each CIP must be heated to be effective.

Table 4 - Effective Microfiltration Cleaning Procedure

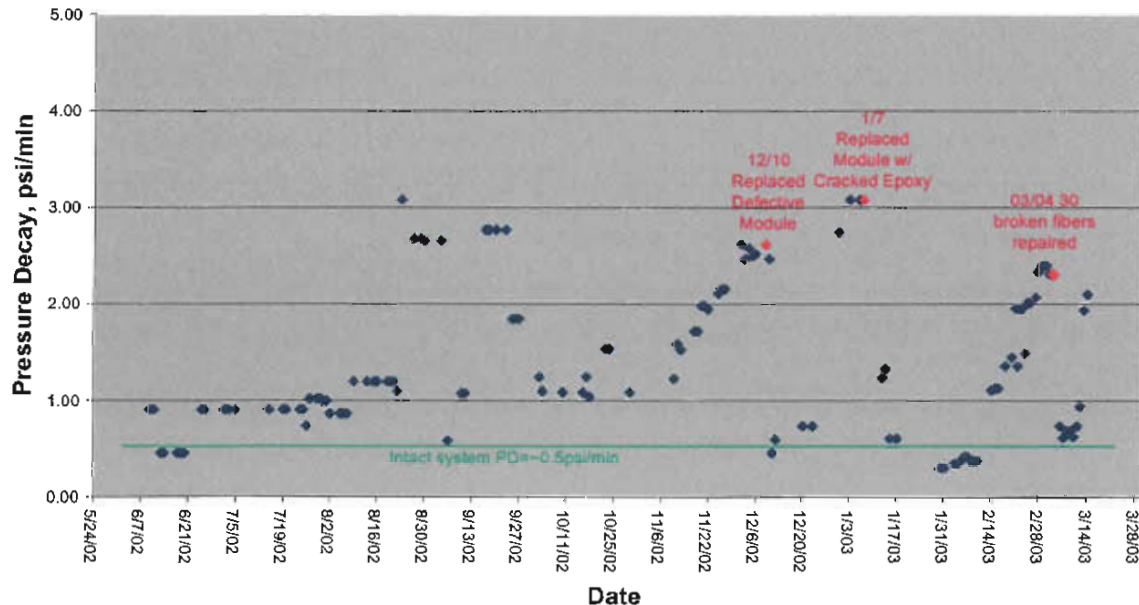
Step	Chemical	Temperature (°C)	Procedure
1	2% Citric Acid	36 - 38	Perform reverse filtration until
2	400 – 600 mg/L chlorine	20	membrane cell is filled with MF Filtrate. Add chemicals, heat solution and aerate every 2 minutes. Perform filtrate recirculation for 30 minutes. Repeat 5 minute aeration/5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

5.3 US Filter PVDF Membrane Module Integrity-Original CMF-S Modules (MF Trials I-III)

The US Filter 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 manually initiated pressure decay test (PDT). During the PDT, the unit was isolated and the lumen (filtrate) side of the modules is drained. Air was then injected to the lumen at 15 psi, and then a valve on the feed side was 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 significant pressure decay. The intact US Filter system with no fiber breaks displays a PDT rate of ~0.5 psi/minute. To quantify the broken fiber problems observed during this study, on March 4, 2003, a pressure decay was performed on the system resulting in a decay rate of ~2.3psi/minute. Thereafter, between 30 and 35 fibers, were repaired on one of the four modules in the system. Each original CMF-S module contained ~14,500 Fibers. Figure 9 below demonstrates that the unit had broken fibers over most of trials I - III of the study. Figure 10 displays air passage during a pressure decay test through the crack that developed in the module epoxy.

Figure 9 - US Filter Microfiltration Unit Pressure Decay Test Results



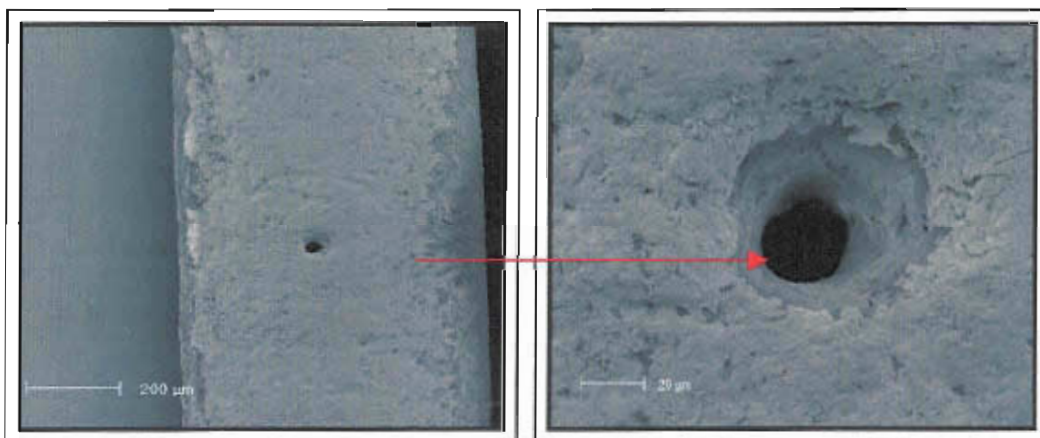
US Filter sent their problematic modules to Australia for autopsy to determine the cause of the fiber breakage and epoxy failures. The results from the analysis of the module with the cracked epoxy can be summed up as:

- 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 um coarse strainer and the standard 500 um 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. Note that a 130um Arkal strainer replaced the original 500um strainer in front of the MF to alleviate this. See the appendix for further details.
- 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 10 - Air Bubbles Emitted from the Cracked Epoxy During the Pressure Decay Test

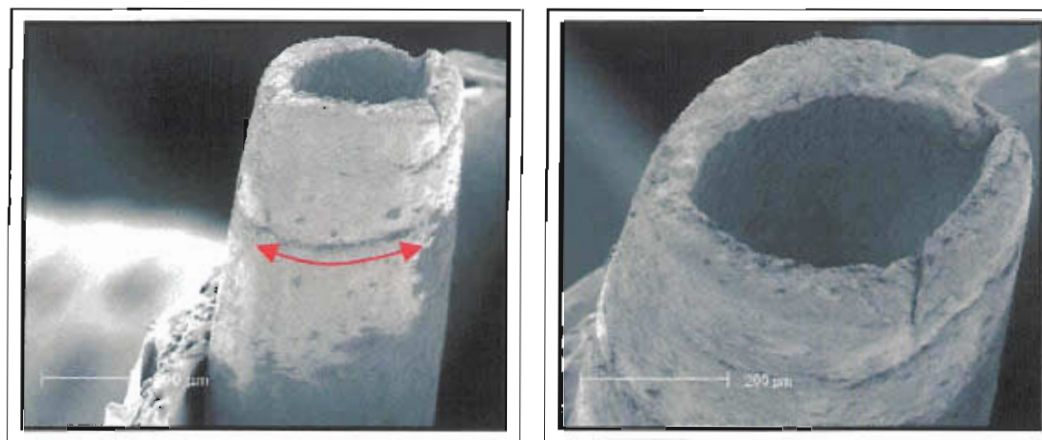


Figure 11 - SEM Photographs of a Hole in a CMF-S Module Fiber



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 12 - 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.

US Filter 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. Different number of fibers in module (different packing density)
3. Reduced fiber area per module

Table 5 - US Filter CMF-S Module Comparison

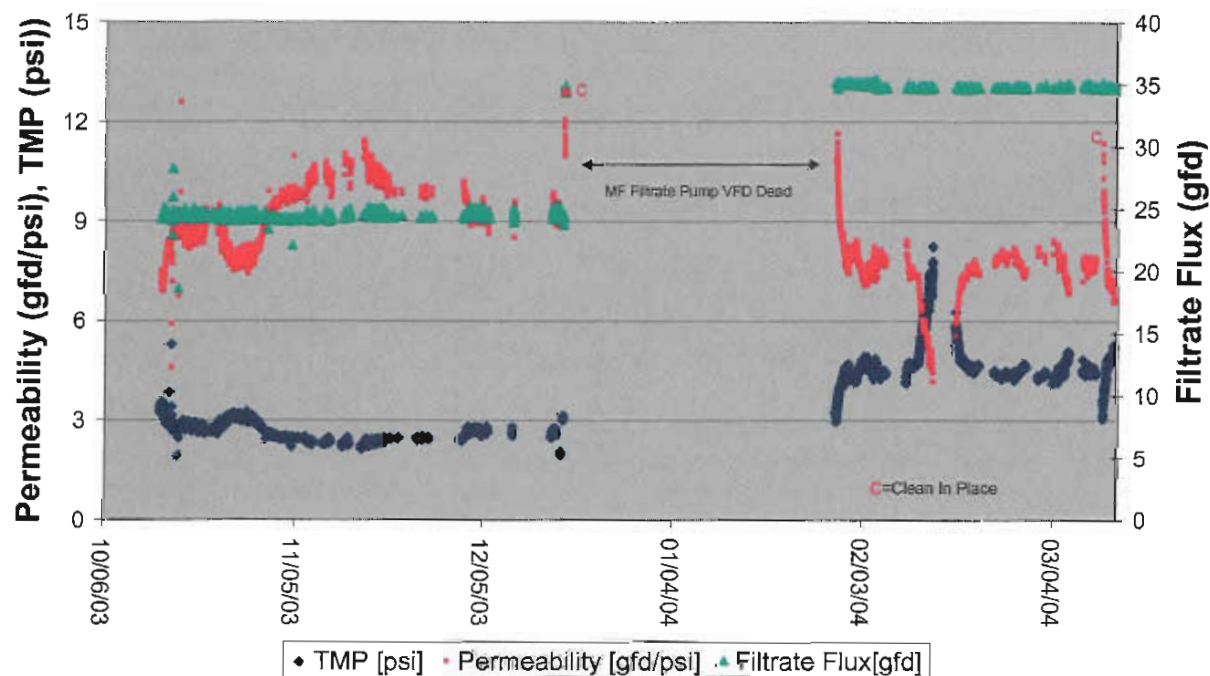
Parameter	Original S10V Module	Redesigned S10V Module
Fiber Outside Diameter, μm	650	800
Fiber Inside Diameter, μm	500	390
Number of Fibers per Module	14,500	9,600
Module Active membrane Area, m^2	31.1	25.3

5.4 Permeability of Newly Designed CMF-S Modules

5.4.1 Trial IV-Redesigned CMF-S Modules Without Arkal Strainer-Permeability

In October 2003, after a delay in testing due to the reconfiguration of the RO feed pumps as discussed in the appendix, the trials commenced with the new, improved US Filter CMF-S module. US Filter 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 13 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” US Filter 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 um strainer was installed for this trial but was bypassed as described in the appendix.

Figure 13 - Permeability of Redesigned MF modules (MF Trial IV)



5.4.2 Trial V-Permeability of New Modules with the Arkal Spin Klin Strainer as Pretreatment

The Arkal Spin Klin 130 um strainer 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 US Filter CMF-S unit was decreased from every 15 to every 20 minutes. Figure 14 shows that one run

was executed at these conditions and maintained 3 week run time before terminal permeability was reached.

Figure 14 - Permeability of Redesigned Modules with Arkal Strainer (Trial V)

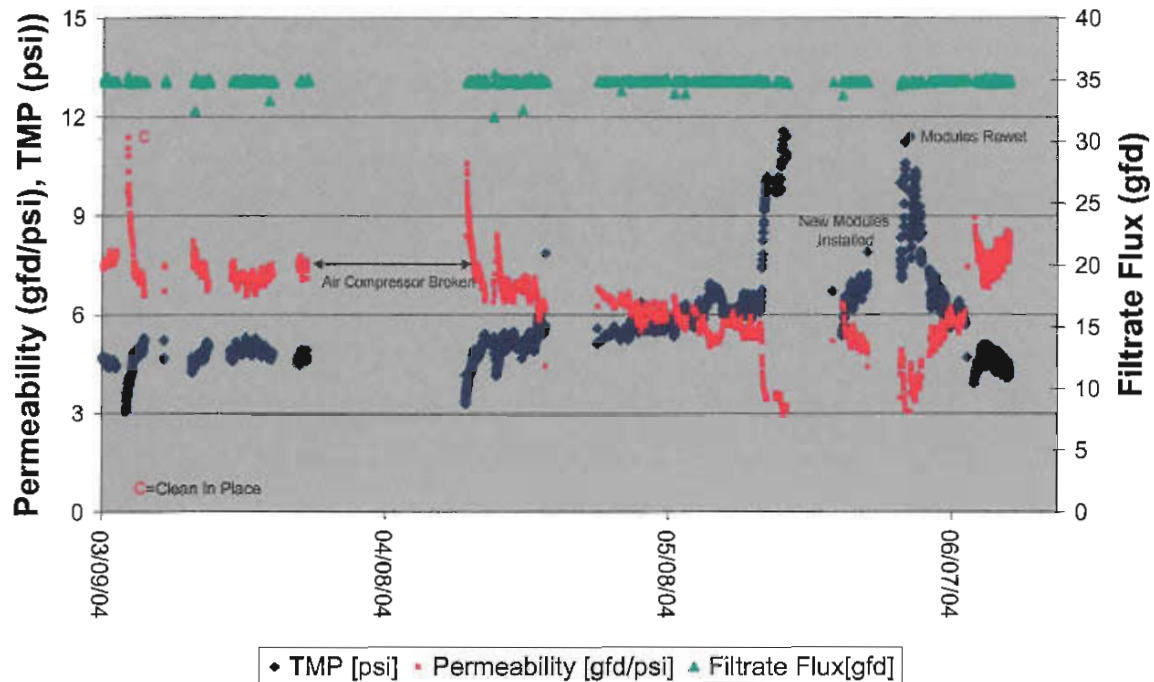


Table 6 - Optimized US Filter CMF-S Microfiltration Run Conditions with Redesigned Modules

Parameter	Value
Filtrate Flow per module (gpm)*	6.5
Filtrate Flux (gfd)*	34
Filtration time between backwashes (min)	20
Recovery	91%
Backwash Parameters	
Air scour Rate (SCFM/module)	7
Air scour Duration (seconds)	30

Parameter	Value
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

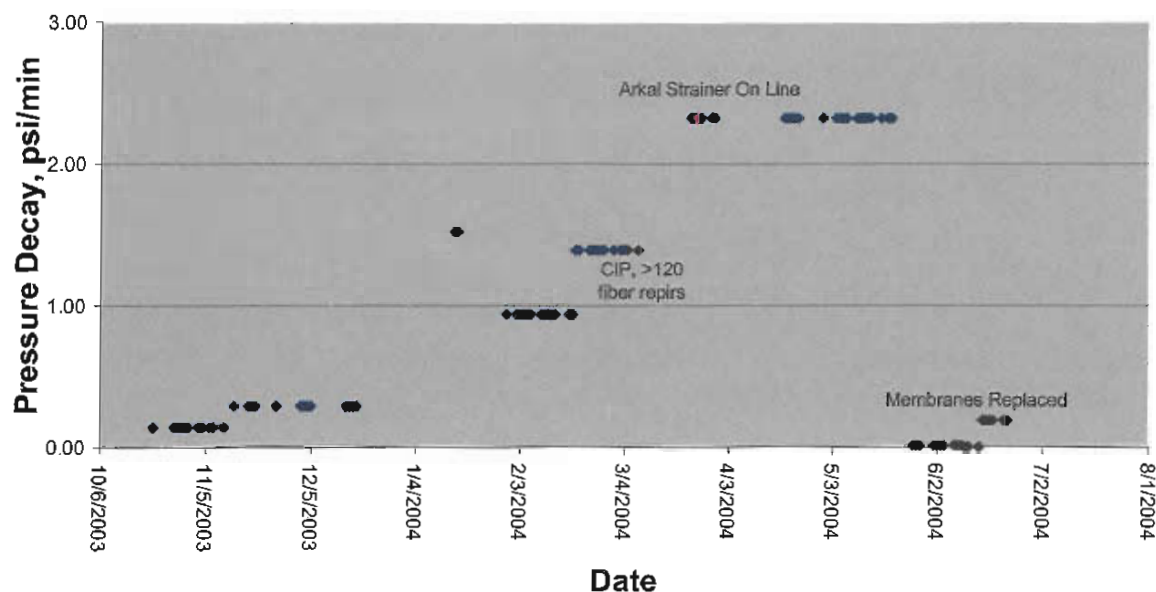
*Optimized Parameters. Non optimized parameters recommended by US Filter.

5.5 New Redesigned US Filter 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 a significant period of time with only the 800 um strainer as pretreatment as the 130 um Arkal Spin Klin strainer was bypassed due to installation problems. However, it is clear the new module design may have allowed a significantly higher stable operating flux, but the more robust modules with only the 800 um strainer as pretreatment did not maintain integrity per figure 15.

The Arkal Spin Klin strainer 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 15) and the replacement modules have held their integrity with the additional 130um strainer as pretreatment is promising. Additional run time with the Arkal strainer would have been required to determine if it would have prevented further MF fiber breakage.

Figure 15 - Integrity of Redesigned CMF-S Modules



5.6 MF Filtrate Quality

The MF pretreatment was utilized to condition the raw ocean water such that it was suitable for spiral wound reverse osmosis membranes. This involved particulate matter removal that was 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.

5.6.1 Turbidity

The presence of suspended material in water causes cloudiness which is known as turbidity.⁶ The raw ocean water and MF Filtrate turbidities were measured once per day at the test site. The incoming ocean water turbidity has averaged ~1NTU, with peak values of ~5NTU. Per figures 16 and 17, the filtrate turbidity averaged 0.05NTU and typically was <0.1NTU, suitable for RO despite the module and fiber problems.

⁶ Water Treatment Plant Operation, 1994 California State University Sacramento.

Figure 16 - Feed Water and Microfiltration Filtrate Turbidity-MF Trials I - III

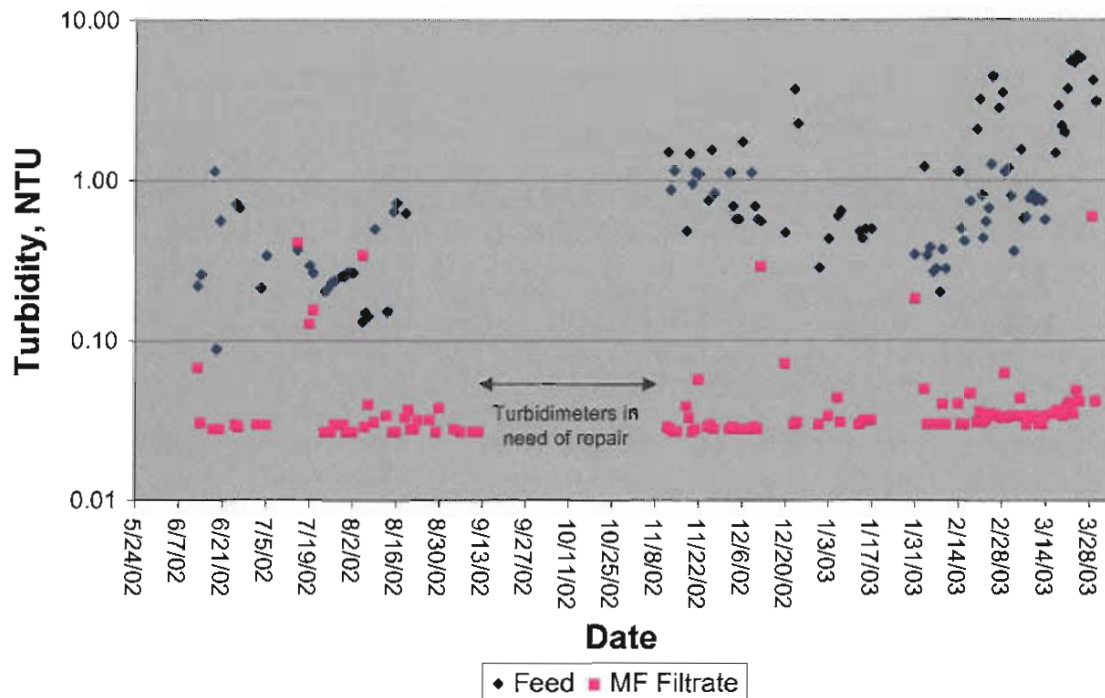
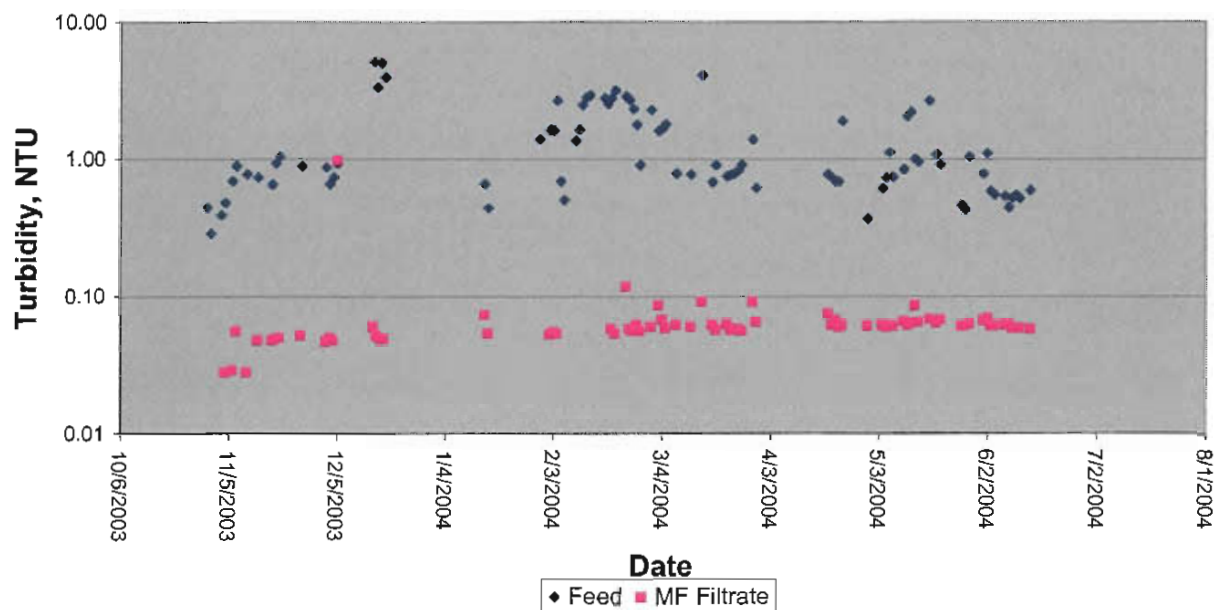


Figure 17 - Feed Water and Microfiltration Filtrate Turbidity-MF Trials IV & V



5.6.2 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 clog 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 4⁷. An SDI_{15} greater than 4 represents water that poses an increased risk to RO membrane fouling and permeability decline.

The SDI_{15} of the raw ocean water was measured on a few occasions and was immeasurable, clogging the SDI pad significantly within 5 minutes and almost completely by the fifteen-minute mark. However, 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 18 and 19 show the RO Feed SDI_{15} and MF pressure decay. A trend between these two parameters is not clear in MF Trials I - III, but 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 2psi/minute. Note that the SDI_{15} reduced to less than 2 after the replacement modules were implemented on May 28, 2004. It is 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.

⁷ Taken from Hydranautics and Dow RO Membrane Data Sheets

Figure 18 - MF System Pressure Decay Results and Filtrate SDI MF Trials I - III

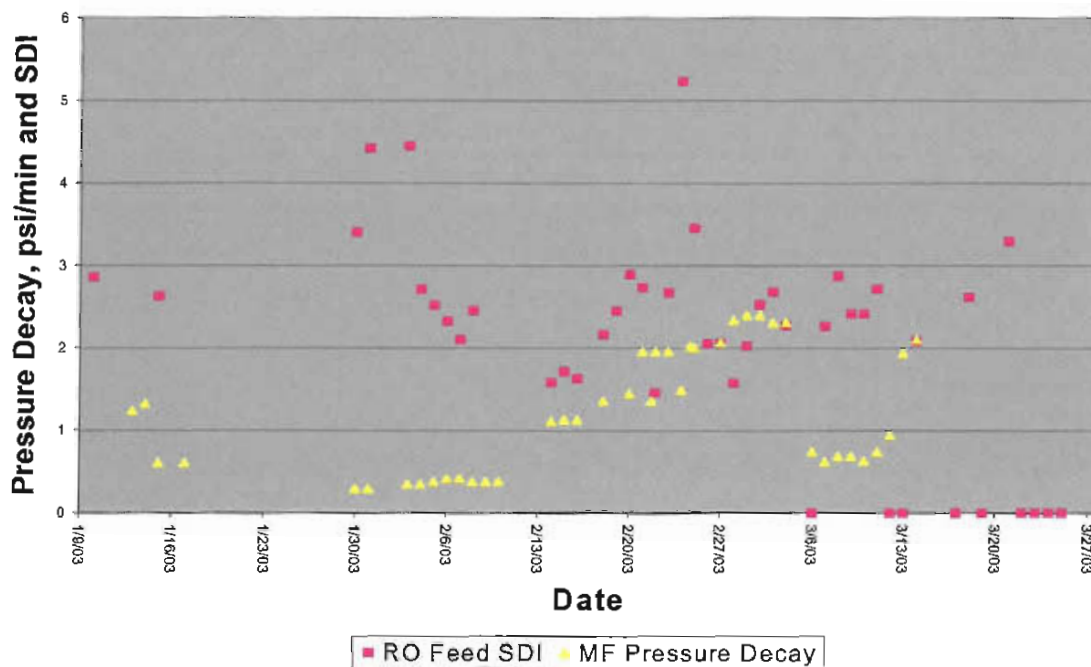
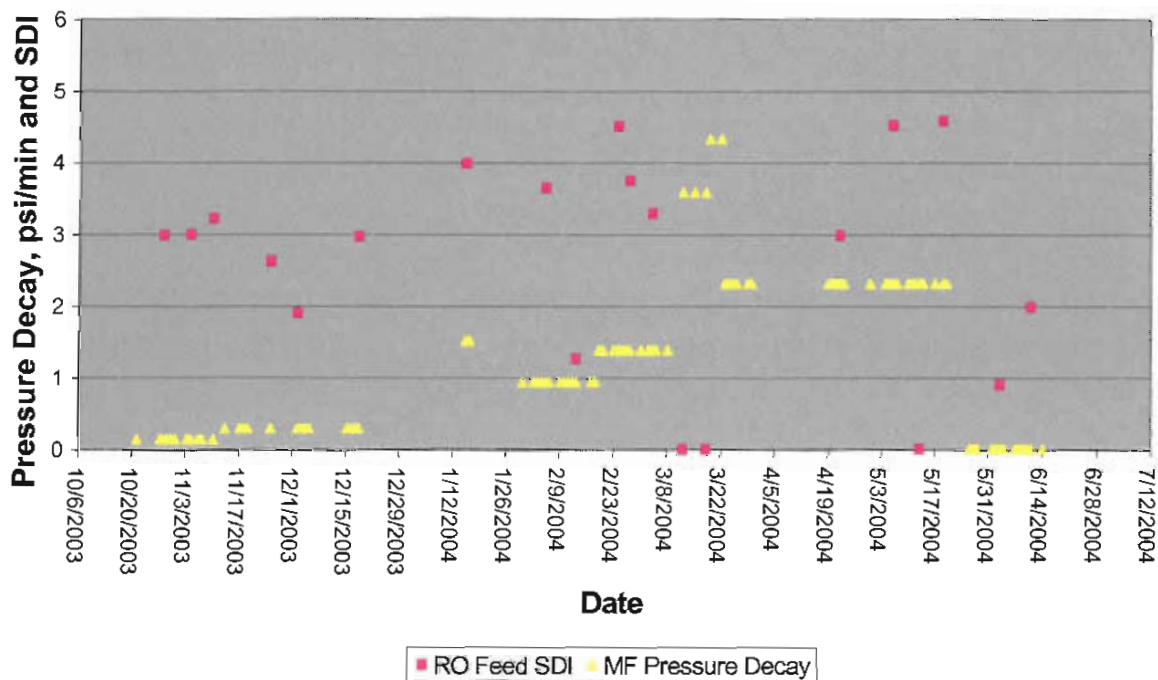


Figure 19 - MF System Pressure Decay Results and Filtrate SDI MF Trials IV & V



5.6.3 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 Parameters

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 CaCO ₃)	mg/L	2	115.3	2.1	109.3	1.3
Calcium	mg/L	25	407.0	29.1	388.9	22.2
Magnesium	mg/L	25	1335.3	103.3	1236.0	68.2
Hardness (as CaCO ₃)	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 CaCO ₃)	mg/L	2	115.2	6.3	108.9	4.1
Calcium	mg/L	25	406.2	32.8	393.3	21.6
Magnesium	mg/L	25	1338.4	105.0	1256.7	90.2
Hardness (as CaCO ₃)	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

5.7 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

6 REVERSE OSMOSIS OPTIMIZATION AND PERFORMANCE

Figure 20 - Reverse Osmosis Test Equipment



The RO Testing can be grouped into the following trials:

Table 9 - 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 8GFD
RO III	SBS pretreatment, operation at 9GFD
RO IV	SBS pretreatment, operation at 11GFD

Table 10 - Details of Each Reverse Osmosis Run

Trial	Run #	Dates	MF Filtrate Chemical	RO Feed Antiscalant ppm	Hydранautics Flux, GFD	Hydранautics 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

6.1 RO Trial I Testing

The original pretreatment process, an attempt to create chloramines in seawater, 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→MF→RO process has been used successfully on many wastewater reclamation facilities including the 20MGD West Basin Water Recycling Plant. The ammonia reacts with free chlorine or HOCl to form chloramines.

However, two items complicated the creation of chloramine on this water source. First, ammonia is not present in ocean water and thus must be added. Second, the presence of bromide (Br^-) in ocean water interferes with the reactions. The Pacific Ocean water source used in this study has ~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.

As depicted in Figures 21 and 22, this $\text{NaOCl} \rightarrow \text{MF} \rightarrow \text{NH}_4\text{OH} \rightarrow \text{RO}$ 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 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. (RO Trials II-IV).

Figure 21 - Increasing Permeability of RO Membranes due to Oxidation (RO Trial I)

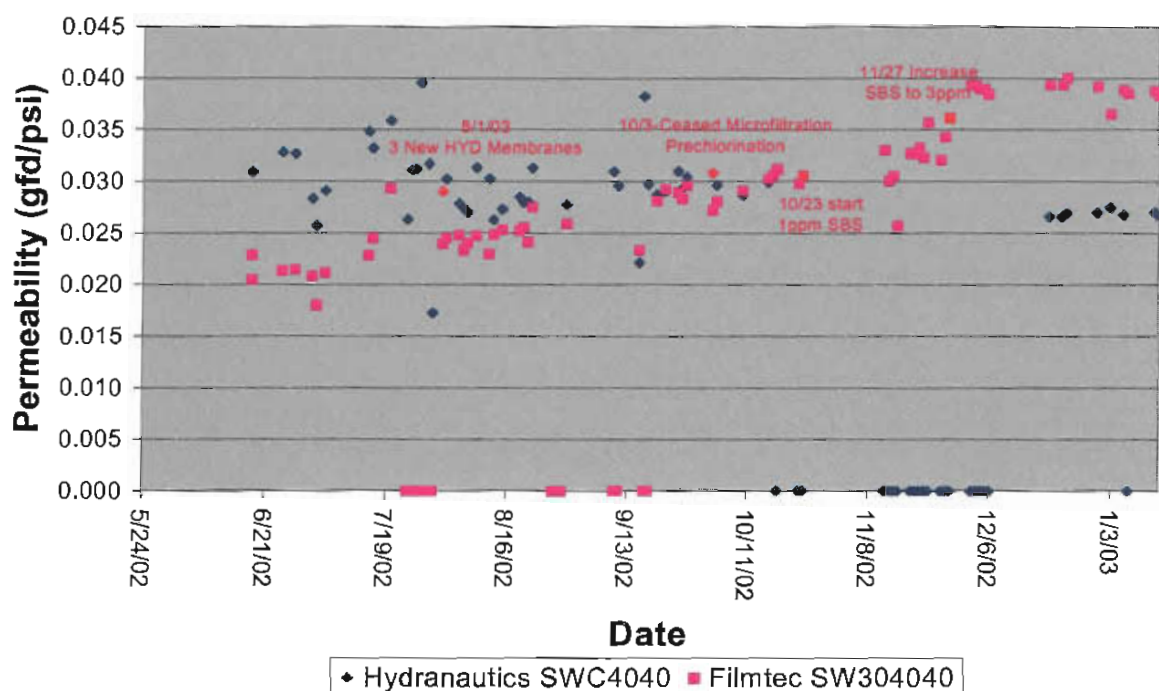
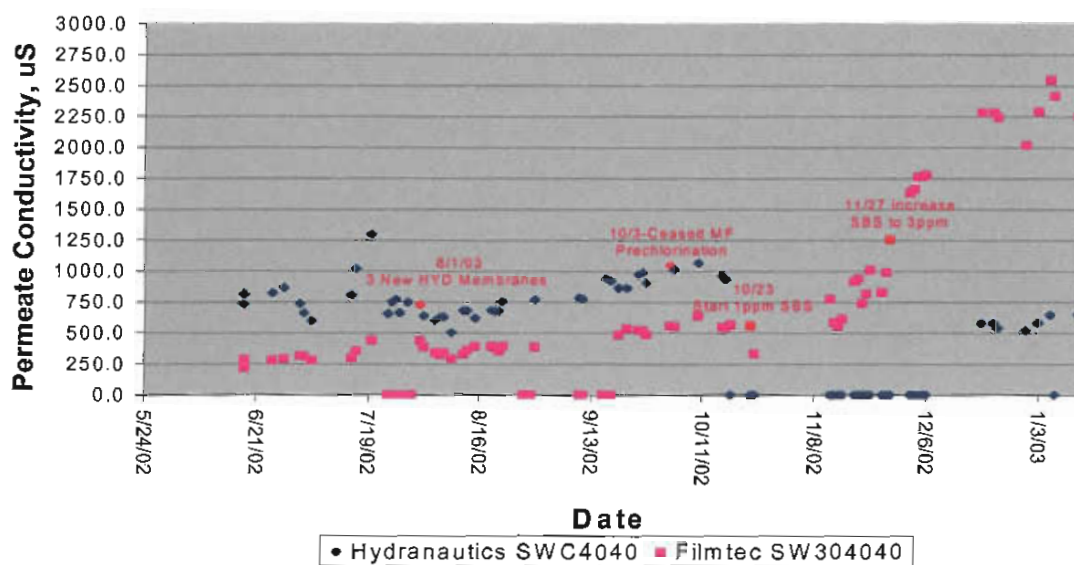


Figure 22 - Increasing Permeate Conductivity of RO Membranes due to Oxidation (RO Trial I)



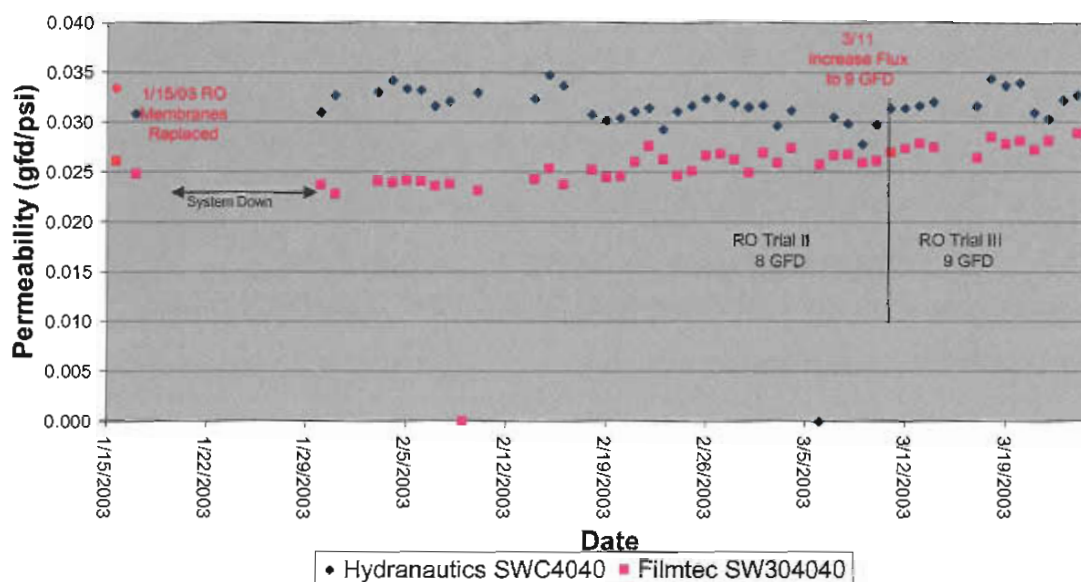
From October through December 2002, the RO was run with the damaged membranes in an attempt to find a pretreatment strategy that would allow the MF to maintain reasonable flux rates and run times without further RO oxidation. The RO membranes were replaced on January 15, 2003 and trial II of the RO testing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

6.2 RO Permeability

Like the MF, 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.)

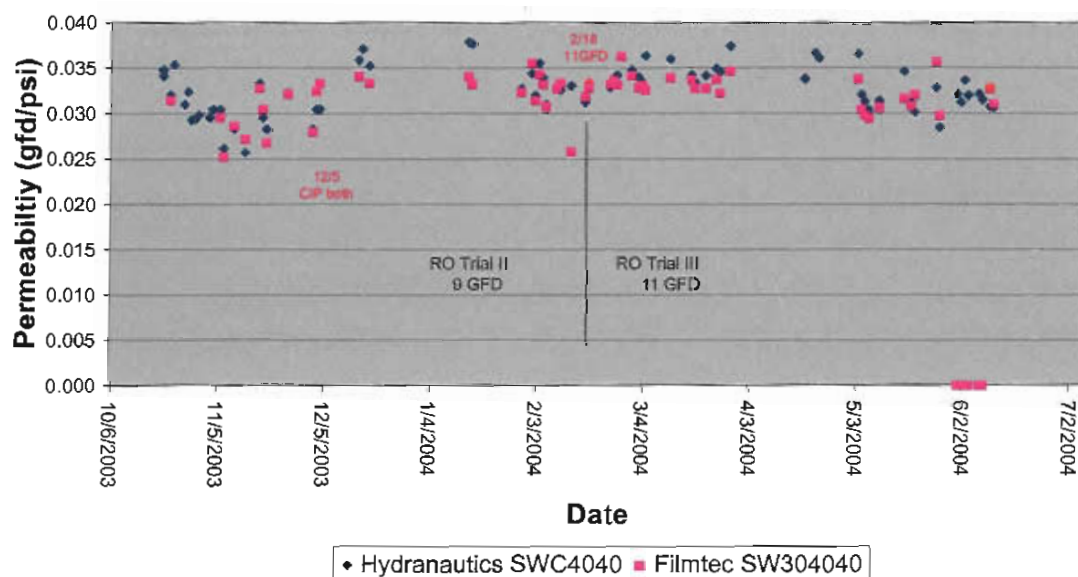
Figure 23 displays that the permeability of the Hydranautics membrane was fairly stable following the replacement of the RO membranes (RO trial II). Filmtec membranes, on the other hand, showed a slight increase in specific flux and as will be discussed in the next section, permeate conductivity as well. These trends are consistent with membrane oxidation. However, the Hydranautics membranes did not show these signs of oxidation. 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 21 and 22 above, which display the results of the Trial I testing of the RO membranes oxidized by the $\text{NaOCl} \rightarrow \text{MF} \rightarrow \text{NH}_4\text{OH} \rightarrow \text{RO}$ process, reveals that the Filmtec membranes experienced deterioration, presumably from oxidation, much faster than the Hydranautics membranes. RO Trial III commenced in March 2003 operating at 9 GFD.

Figure 23 - 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 the appendix. 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 end of figure 23 and the beginning of figure 24) demonstrated that both the Hydranautics and Filmtec membranes did not decrease in permeability over the course of the testing. Thus, no significant fouling was been observed on this set of RO membranes over approximately 3100 hours of testing.

Figure 24 - Reverse Osmosis Membrane Permeability End of Trial III and Trial IV



The optimized RO run parameters are as follows:

Table 11 - Optimized RO Parameters

Parameter	Value
RO Operating Flux (gfd)*	8 - 11
Recovery	50%
Sodium Bisulfite Dose (mg/L)*	3
Antiscalant Dose (mg/L)	3

*Optimized Parameters.

6.3 RO Permeate Quality

Over the course of testing to date, two sets of DOW Filmtec and Hydranautics RO membranes from have been tested, and for each set, the DOW Filmtec SW30-4040 initially produced water of significantly better (lower concentration of most constituents) quality than the Hydranautics SWC-4040. RO Permeate quality was continuously measured via conductivity and biweekly samples were taken for individual analysis.

6.3.1 Conductivity

Figure 25 demonstrates that the conductivity produced by the Filmtec membrane was initially significantly below 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 uS at 11 GFD and 18°C. (Figure 26).

Figure 25 - Reverse Osmosis Membrane Conductivity Trials II and Beginning of Trial III

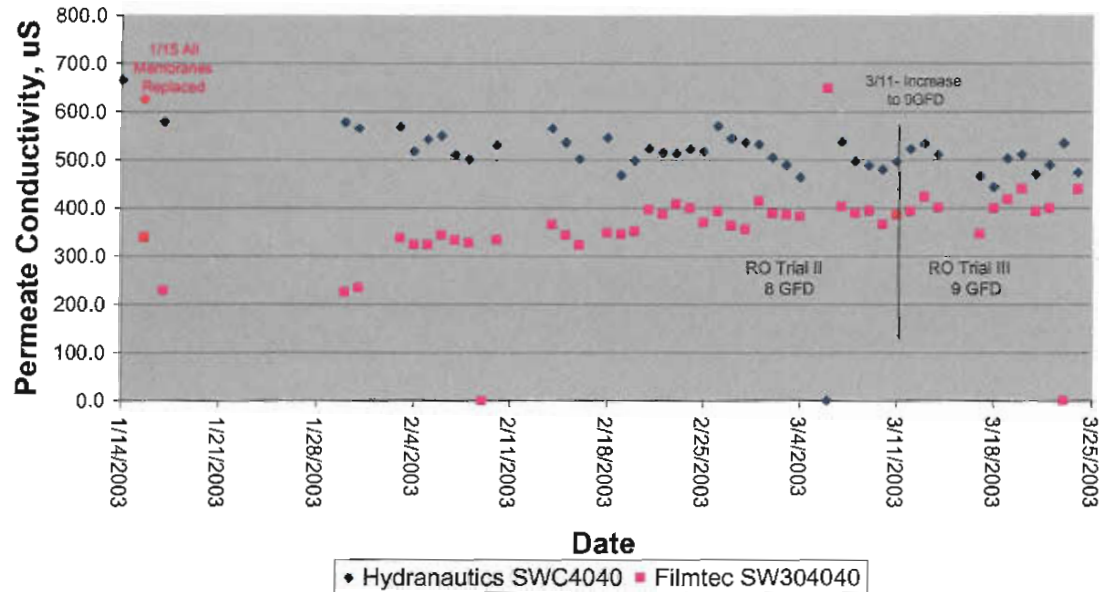
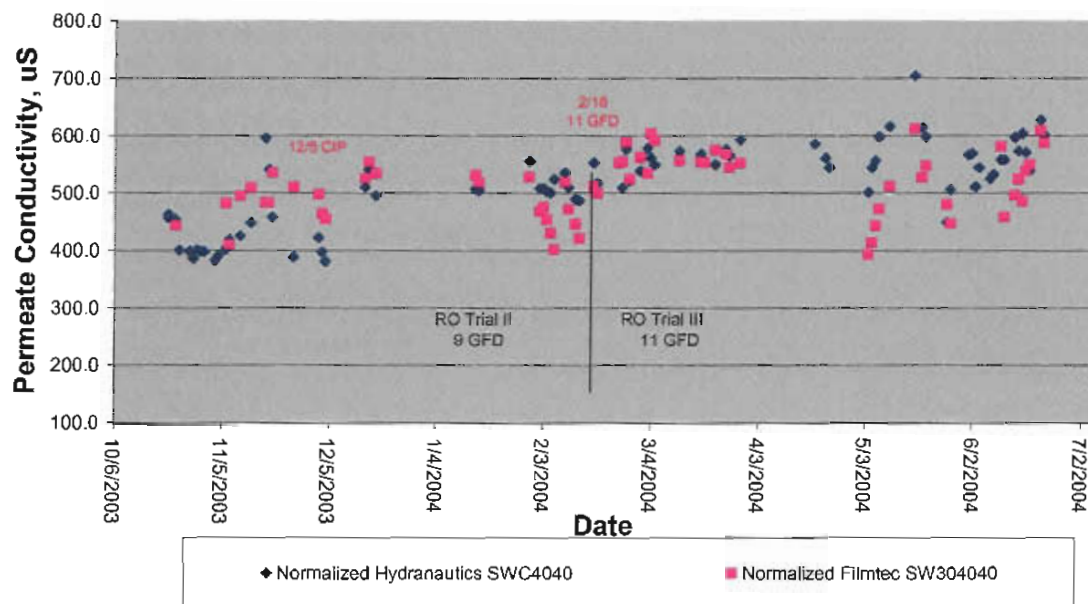


Figure 26 - Reverse Osmosis Membrane Conductivity End of Trials III and Trial IV



6.3.2 Individual Ion Analyses

Tables 11, 12 and 13 summarizes the average results of the laboratory analysis performed on the RO streams for each trial of the testing. The following are evident:

1. At each Trial (flux), each RO membrane produced permeate of TDS <300mg/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 FilmTec (DOW) initially produced water substantially better than Hydranautics. The Filmtec membrane continued to produce better water quality, but the gap between the two membranes reduced as the testing progressed. Boron levels were constantly below 1.5 mg/L and 1.0 mg/L for Hydranautics and Filmtec respectively.

Table 12 - Average RO Membrane Water Quality for Trial II (8 GFD Flux rate)

Parameter	SAMPLE ID					Units
	RO Feed	Permeate		Concentrate		
		Train 1	Train 2	Train 1	Train 2	
		HYD	DOW	HYD	DOW	
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 13 - 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 14 - Average RO Membrane Water Quality for Trial IV (11 GFD Flux rate)

Parameter	SAMPLE ID					Units
	RO Feed	Permeate		Concentrate		
		Train 1	Train 2	Train 1	Train 2	
		HYD	DOW	HYD	DOW	
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

6.4 Reverse Osmosis Membrane Performance Relative to Manufacturers' Predictions

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 15 - RO Performance vs. that Predicted

RO Trial	Membrane	Flux (GFD)	Projected	Actual
			Feed psi/ Permeate TDS / Permeate Boron*	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 does not predict Boron rejection

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 have 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.

6.5 RO Concentrate (Waste) Characterization

The RO concentrate stream was sampled biweekly for the parameters listed above in Tables 12, 13 and 14 in order to characterize the RO waste stream.

6.6 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 as offered by Hydranautics and Dow do not have equivalent salt rejection capabilities as the standard 8-inch products. Therefore, both Hydranautics and Dow were asked to “cherry-pick” their 4-inch diameter inventory and supply membranes that were representative to their 8-inch counterparts in both flow and rejection properties.

7 REGULATORY STANDARDS

7.1 RO permeate quality relative to regulatory standards

The RO permeate produced in this study consistently met drinking water primary and secondary regulatory standards. Permeate from the RO operation was sampled throughout the operation of this study for extensive water quality analyses. A list of the constituents included in these analyses is provided in Table 16. The results of these analyses on both the raw seawater feed and the RO permeate are indicated in table 17. Results indicate that all constituents met primary and secondary drinking water MCL's throughout the course of testing. Likewise, the results indicated compliance with the California DHS Drinking Water Notification Levels (formerly Action Levels). No results indicated concentrations above the Notification Levels.

Overall the testing results demonstrated performance meeting all drinking water regulations throughout the pilot-testing program.

Table 16 - Drinking Water Regulated Constituents

Primary MCL	Primary MCL	Secondary MCL
Constituent - Inorganics	Constituent - SOC's	Aluminum
Aluminum	Alachlor	Color
Antimony	Atrazine	Copper
Arsenic	Bentazon	Corrosivity
Asbestos	Benzo(a)pyrene	MBAS
Barium	Carbofuran	Iron
Beryllium	Chlordane	Manganese
Cadmium	2,4-D	MTBE
Chromium	Dalapon	Odor
Cyanide	Dibromochloropropane	Silver
Fluoride	Di(2-ethylhexyl)adipate	Thiobencarb
Mercury	Di(2-ethylhexyl)phthalate	Turbidity
Nickel	Dinoseb	Zinc
Nitrate (as NO3)	Diquat	TDS
Nitrate + Nitrite (sum as nitrogen)	Endothall	Specific Conductance
Nitrite (as nitrogen)	Endrin	Chloride
Selenium	Ethylene Dibromide	Sulfate
Thallium	Glyphosate	
	Heptachlor	
Constituent - VOC's	Heptachlor Epoxide	
Benzene	Hexachlorobenzene	
Carbon tetrachloride	Hexachlorocyclopentadiene	
1,2-Dichlorobenzene	Lindane	CA-DHS Notification Level*
1,4-Dichlorobenzene	Methoxychlor	Boron
1,1-Dichloroethane	Molinate	n-Butylbenzene
1,2-Dichloroethane	Oxamyl	sec-Butylbenzene
1,1-Dichloroethylene	Pentachlorophenol	tert-Butylbenzene
cis-1,2-Dichloroethylene	Picloram	Carbon Disulfide
trans-1,2-Dichloroethylene	Polychlorinated Biphenyl's	Chlorate
Dichloromethane	Simazine	2-Chlorotoluene
1,2-Dichloropropane	Thiobencarb	4-Chlorotoluene
1,3-Dichloropropene	Toxaphene	Dichlorodifluoromethane
Ethylbenzene	2,3,7,8-TCDD (Dioxin)	1,4-Dioxane
MTBE	2,4,5-TP (Silvex)	Ethylene Glycol
Monochlorobenzene		Formaldehyde
Styrene	Constituent - Radiological	Isopropylbenzene
1,1,2,2-Tetrachloroethane	Combined Radium-226 and 228	Manganese
Tetrachloroethylene	Gross alpha	Methyl iso Butyl Ketone
Toluene	Tritium	Naphthalene
1,2,4-Trichlorobenzene	Strontium-90	NDMA
1,1,1-Trichloroethane	Gross beta	Perchlorate
1,1,2-Trichloroethane	Uranium	n-Propylbenzene
Trichloroethylene		Tertiary Butyl Alcohol
Trichlorofluoromethane		1,2,3-Trichloropropane
1,1,2-Trichloro-1,2,2-Trifluoroethane		1,2,4-Trimethylbenzene
Vinyl Chloride		1,3,5-Trimethylbenzene
Xylenes		Vanadium

*formerly referred to as Action Levels

Table 17 – Regulatory Standard Constituent Test Results (1 of 3)

Compound	Units	Seawater Detection Limit	Seawater Concentration	Drinking Water Detection Limit	Drinking Water Concentration	MCL	AL
1,1,1-Trichloroethane	ug/L		NA	0.5	NA	200	
1,1,2,2-Tetrachloroethane	ug/L		NA	0.5	NA	1	
1,1,2-Trichloro-1,2,2-trifluoroethane	ug/L		NA	10	NA	1200	
1,1,2-Trichloroethane	ug/L		NA	0.5	NA	5	
1,1-Dichloroethane	ug/L		NA	0.5	NA	5	
1,1-Dichloroethylene	ug/L		NA	0.5	NA	6	
1,2,3-TCP	ug/L	0.0050	ND	0.0050	ND		0.005
1,2,4-Trichlorobenzene	ug/L	5.0	ND	5.0	ND	70	
1,2,4-Trimethylbenzene	ug/L		NA	0.5	NA		330
1,2-Dibromo-3-chloropropane	ug/L		NA		NA	0.2	
1,2-Dichlorobenzene	ug/L	5.0	ND	5.0	ND	600	
1,2-Dichloroethane	ug/L		NA	0.5	NA	0.5	
1,2-Dichloropropane	ug/L		NA	0.5	NA	5	
1,2-Diphenylhydrazine	ug/L	5.0	ND	5.0	ND	EPA L2	
1,3,5-Trimethylbenzene	ug/L		NA	0.5	NA		330
1,3-Dichlorobenzene	ug/L	5.0	ND	5.0	ND		
1,3-Dichloropropene	ug/L		NA	0.5	NA	0.5	
1,4-Dichlorobenzene	ug/L	5.0	ND	5.0	ND	5	
1,4-Dioxane	ug/L	0.50	ND	0.50	ND		3
2,3,7,8-TCDD	ng/L		NA		NA	0.03	
2,4,5-TP (Silvex)	ug/L		NA	1	NA	50	
2,4,5-Trichlorophenol	ug/L	5.0	ND	5.0	ND		
2,4,6-Trichlorophenol	ug/L	1.0	ND	1.0	ND	EPA L2	
2,4-D	ug/L		NA	0.5	ND	70	
2,4-Dichlorophenol	ug/L	1.0	ND	1.0	ND	EPA L2	
2,4-Dimethylphenol	ug/L	5.0	ND	5.0	ND		
2,4-Dinitrophenol	ug/L	5.0	ND	5.0	ND	EPA L2	
2,4-Dinitrotoluene	ug/L	5.0	ND	5.0	ND	EPA L1	
2,6-Dinitrotoluene	ug/L	5.0	ND	5.0	ND	EPA L1	
2-Chloronaphthalene	ug/L	5.0	ND	5.0	ND		
2-Chlorophenol	ug/L	5.0	ND	5.0	ND		
2-Chlorotoluene	ug/L		NA	0.5	NA		140
2-Methylnaphthalene	ug/L	5.0	ND	5.0	ND		
2-Methylphenol	ug/L	1.0	ND	1.0	ND	EPA L2	
2-Nitroaniline	ug/L	5.0	ND	5.0	ND		
2-Nitrophenol	ug/L	10	ND	10	ND		
3,3'-Dichlorobenzidine	ug/L	5.0	ND	5.0	ND		
3-Methyl-4-chlorophenol	ug/L	5.0	ND	5.0	ND		
3-Methylphenol	ug/L	5.0	ND	5.0	ND		
3-Nitroaniline	ug/L	5.0	ND	5.0	ND		
4,4'-DDD	ug/L	0.020	ND	0.020	ND		
4,4'-DDE	ug/L	0.010	ND	0.010	ND	EPA L1	
4,4'-DDT	ug/L	0.020	ND	0.020	ND		
4,6-Dinitro-2-methylphenol	ug/L	10	ND	10	ND		
4-Bromophenyl phenyl ether	ug/L	5.0	ND	5.0	ND		
4-Chloroaniline	ug/L	5.0	ND	5.0	ND		
4-Chlorophenyl phenyl ether	ug/L	5.0	ND	5.0	ND		
4-Chlorotoluene	ug/L		NA	0.5	NA		140
4-Methylphenol	ug/L	5.0	ND	5.0	ND		
4-Nitroaniline	ug/L	5.0	ND	5.0	ND		
4-Nitrophenol	ug/L	10	ND	10	ND		
Acenaphthylene	ug/L	5.0	ND	5.0	ND		
Acenaphthene	ug/L	5.0	ND	5.0	ND		
Acetaldehyde	ug/L	30	ND	30	ND		
Alachlor	ug/L	1.0	ND	1.0	ND	2	
Aldicarb	ug/L		NA	4	NA		7
Aldrin	ug/L	0.075	ND	0.075	ND		
Alkalinity	mg/L as CaCO3	2.0	110	2.0	ND		
Alpha-BHC	ug/L	0.050	ND	0.050	ND		0.015
Aluminum	ug/L	0.01	36	0.01	ND	200	
Ammonia	mg/L as N	0.10	ND	0.10	ND		
Aniline	ug/L	5.0	ND	5.0	ND		
Anthracene	ug/L	5.0	ND	5.0	ND		
Antimony	ug/L	0.010	0.14	0.010	0.02		

Compound	Units	Seawater Detection Limit	Seawater Concentration	Drinking Water Detection Limit	Drinking Water Concentration	MCL	AL
Arochlor-1016	ug/L		NA	0.5	NA	0.1	
Arochlor-1221	ug/L		NA	0.5	NA	0.1	
Arochlor-1232	ug/L		NA	0.5	NA	0.1	
Arochlor-1242	ug/L		NA	0.5	NA	0.1	
Arochlor-1248	ug/L		NA	0.5	NA	0.1	
Arochlor-1254	ug/L		NA	0.5	NA	0.1	
Arochlor-1260	ug/L		NA	0.5	NA	0.1	
Arsenic	ug/L	0.010	1.14	0.010	ND	10	
Asbestos	MFL	0.200	ND		NA	7	
Atrazine	ug/L	0.50	ND	0.50	ND	3	
Azobenzene	ug/L	5.0	ND	5.0	ND		
Barium	mg/L	0.5	ND	0.5	ND	1000	
Bentazon	ug/L		NA	2	NA	18	
Benzene	ug/L		NA	0.5	NA	1	
Benzidene	ug/L	10	ND	10	ND		
Benzo(a)anthracene	ug/L	5.0	ND	5.0	ND		
Benzo(a)pyrene	ug/L	5.0	ND	5.0	ND	0.2	
Benzo(b)fluoranthene	ug/L	5.0	ND	5.0	ND		
Benzo(g,h,i)perylene	ug/L	5.0	ND	5.0	ND		
Benzo(k)fluoranthene	ug/L	5.0	ND	5.0	ND		
Benzoic Acid	ug/L	100	ND	100	ND		
Benzyl Alcohol	ug/L	5.0	ND	5.0	ND		
Beryllium	ug/L	0.005	ND	0.005	ND	4	
Beta-BHC	ug/L	0.050	ND	0.050	ND		0.025
bis(2-chloroethoxy)methane	ug/L	5.0	ND	5.0	ND		
bis(2-chloroethyl)ether	ug/L	5.0	ND	5.0	ND		
bis(2-chloroisopropyl)ether	ug/L	5.0	ND	5.0	ND		
bis(2-ethylhexyl)phthalate	ug/L	5.0	ND	5.0	ND		
Boron	ug/L	100	NA	10	NA		1
Bromacil	ug/L	10	ND	10	ND		
Bromide	mg/L	0.50	65	0.50	0.73		
Butachlor	ug/L	0.38	ND	0.38	ND		
Butylbenzylphthalate	ug/L	5.0	ND	5.0	ND		
Cadmium	ug/L	0.005	ND	0.005	ND	5	
Calcium	mg/L	0.05	237	0.05	0.863		
Carbazole	ug/L	5.0	ND	5.0	ND		
Carbofuran	ug/L		NA	4	NA	18	
Carbon Disulfide	ug/L		NA		NA		160
Carbon Tetrachloride	ug/L		NA	0.5	NA	0.5	
Chlorate	ug/L		NA		NA		800
Chlordane	ug/L	0.10	ND	0.10	ND	0.1	
Chloride	mg/L	1000	20000	5.0	170	500	
Chlorobenzene	ug/L		NA	0.5	NA	70	
Chlorothalonil	ug/L	5.0	ND	5.0	ND		
Chrysene	ug/L	5.0	ND	5.0	ND		
cis-1,2-Dichloroethylene	ug/L		NA	0.5	NA	6	
Color	PCU		NA	1	NA	15	
Copper	ug/L	0.005	0.393	0.005	0.467	1000	
Cr(VI)*	ug/L	0.30	ND	0.30	ND		
Cyanide	mg/L	0.010	ND	0.010	ND	200	
Dalapon	ug/L		NA	0.5	NA	200	
Delta-BHC	ug/L	0.50	ND	0.50	ND		
Di (2-ethylhexyl)adipate	ug/L		NA	5	NA	400	
Di (2-ethylhexyl)phthalate	ug/L		NA	3	NA	4	
Diazinon	ug/L	0.25	ND	0.25	ND	EPA L2	6
Dibenz(a,h)anthracene	ug/L	5.0	ND	5.0	ND		
Dibenzofuran	ug/L	5.0	ND	5.0	ND		
Dichlorodifluoromethane	ug/L		NA	0.5	NA		1
Dichlorofluoromethane	ug/L		NA	0.5	NA		1000
Dichloromethane	ug/L		NA	0.5	NA	5	
Dieldrin	ug/L	0.020	ND	0.020	ND		
Diethylphthalate	ug/L	5.0	ND	5.0	ND		
Dimethoate	ug/L	1.0	ND	1.0	ND		
Dimethylphthalate	ug/L	5.0	ND	5.0	ND		
Di-n-butylphthalate	ug/L	5.0	ND	5.0	ND		
Di-n-octylphthalate	ug/L	5.0	ND	5.0	ND		
Dinoseb	ug/L		NA	2	NA	7	
Diquat	ug/L		NA	4	NA	20	

Compound	Units	Seawater Detection Limit	Seawater Concentration	Drinking Water Detection Limit	Drinking Water Concentration	MCL	AL
Endosulfan I	ug/L	0.020	ND	0.020	ND		
Endosulfan II	ug/L	0.010	ND	0.010	ND		
Endosulfan Sulfate	ug/L	0.050	ND	0.050	ND		
Endothal	ug/L		NA	45	NA	100	
Endrin	ug/L	0.10	ND	0.10	ND	2	
Endrin Aldehyde	ug/L	0.050	ND	0.050	ND		
EPTC	ug/L		NA	1	NA	EPA L1	
Ethylbenzene	ug/L		NA	0.5	NA	700	
Ethylene dibromide	ug/L		NA	0.02	NA	0.05	
Fluoranthene	ug/L	5.0	ND	5.0	ND		
Fluorene	ug/L	5.0	ND	5.0	ND		
Fluoride	mg/L	2	NA	0.02	NA	2	
Formaldehyde	ug/L	30	ND	30	ND		100
Gamma-BHC (Lindane)	ug/L	0.20	ND	0.20	ND	0.2	
Glyphosate	ug/L		NA	25	NA	700	
Heptachlor	ug/L	0.010	ND	0.010	ND	0.01	
Heptachlor epoxide	ug/L	0.010	ND	0.010	ND	0.01	
Hexachlorobenzene	ug/L	0.50	ND	0.50	ND	1	
Hexachlorobutadiene	ug/L	5.0	ND	5.0	ND		
Hexachlorocyclopentadiene	ug/L	1.0	ND	1.0	ND	50	
Hexachloroethane	ug/L	5.0	ND	5.0	ND		
Indeno(1,2,3-cd)pyrene	ug/L	5.0	ND	5.0	ND		
Iron	ug/L	0.01	41.7	0.01	0.077	300	
Isophorone	ug/L	5.0	ND	5.0	ND		
Isopropylbenzene	ug/L		NA	0.5	NA		770
Lead	ug/L	0.005	0.188	0.005	0.86	15	
Magnesium	mg/L	0.05	1060	0.05	4.28		
Manganese	ug/L	0.005	1.19	0.005	ND	50	
MBAS	mg/L	0.050	ND	0.050	ND	0.5	
Mercury	ug/L	0.005	0.008	0.005	0.005	2	
Methoxychlor	ug/L	10	ND	10	ND	40	
Methyl isobutyl ketone	ug/L		NA	5	NA		120
Metolachlor	ug/L	0.50	ND	0.50	ND		
Metribuzin	ug/L	0.50	ND	0.50	ND		
Molinate	ug/L	0.50	ND	0.50	ND	20	
MTBE*	ug/L		NA	0.5	NA	5	
Naphthalene	ug/L	5.0	ND	5.0	ND		170
N-butylbenzene	ug/L		NA	0.5	NA		70
Nickel	ug/L	0.005	0.239	0.005	0.037	100	
Nitrate	mg/L as NO3	10	ND	0.50	ND	45	
Nitrite	mg/L as NO2	10	ND	0.50	ND	1	
Nitrobenzene	ug/L	5.0	ND	5.0	ND	EPA L1	
N-nitrodimethylamine	ug/L	5.0	ND	5.0	ND		0.010
N-nitrosodi-n-propylamine	ug/L	5.0	ND	5.0	ND		
N-nitrosodiphenylamine	ug/L	5.0	ND	5.0	ND		
N-Propylbenzene	ug/L		NA	0.5	NA		260
Oxamyl	ug/L		NA	20	NA	200	
PCBs	ug/L	0.10	ND	0.10	ND	0.5	
Pentachlorophenol	ug/L	5.0	ND	5.0	ND	1	
Perchlorate	ug/L	20	NA	4	NA	EPA L1	4
Phenanthrene	ug/L	5.0	ND	5.0	ND		
Phenol	ug/L	5.0	ND	5.0	ND		4200
Picloram	ug/L		NA	1	NA	500	
Prometon	ug/L	1.0	ND	1.0	ND	EPA L2	
Prometryn	ug/L	0.50	ND	0.50	ND		
Propachlor	ug/L	0.50	ND	0.50	ND		
Pyrene	ug/L	5.0	ND	5.0	ND		
Pyridene	ug/L	5.0	ND	5.0	ND		
Sec-butylbenzene	ug/L		NA	0.5	NA		260
Selenium	ug/L	0.01	ND	0.01	ND	50	
Silica	mg/L	1.0	ND	0.10	ND		
Silver	ug/L	0.005	ND	0.005	ND	100	
Simazine	ug/L	0.50	ND	0.50	ND	4	
Sodium	mg/L	5.00	9060	5.00	126		
Strontium	mg/L	0.01	8.49	0.01	0.025		
Styrene	ug/L		NA	0.5	NA	100	
Sulfate	mg/L	1500	2800	0.75	4.2	500	
TBA	ug/L		NA	2	NA		12
TDS	mg/L	100	32000	10	300	1000	
Tert-butylbenzene	ug/L		NA	0.5	NA		260

7.2 RO Concentrate Characterization

The RO concentrate stream was sampled biweekly for the parameters listed above in Table 12, for example, in order to characterize the RO waste stream. Tables 12, 13 and 14 provide average values for the operating periods.

Regular sampling of the RO streams was performed throughout the study. Table 18 provides a listing of the constituents identified in the California Ocean Plan. The project sampling included analysis of the RO concentrate stream for these constituents as indicated in Table 18. The results were confirmed to be below the Ocean Plan limits with the exception of Gross Beta and copper. In the case of Gross Beta further analysis is needed to make the determination, which is underway. Additionally, the concentrate gross beta is compliant with Code of Federal Regulations, Title 10, and California Code of Regulations Title 17 and its presence is due to levels in the raw seawater feed. Copper levels were observed to be elevated in the RO concentrate and identified to be introduced in the treatment process, not simply concentration of feedwater levels. The primary source of copper was traced to inappropriate material selection for a transfer pump in the treatment process. Replacement of this pump dropped the copper level in the RO concentrate substantially, to below the Ocean Plan limit. The initial elevated copper levels are not considered indicative of a condition which would occur in full-scale application of this treatment process with proper equipment selection.

Table 18 - California Ocean Plan Regulated Constituents

Arsenic	Acrolein*	DDT
Cadmium	Antimony	1,4-Dichlorobenzene
Hex Chromium	bis(2-chloroethoxy) methane	3,3-dichlorobenzidine
Copper	bis(2-chloroisopropyl) ether	1,2-Dichloroethane
Lead	Chlorobenzene	1,1-Dichloroethylene
Mercury	Chromium (III)	Dichlorobromomethane
Nickel	di-n-butyl phthalate	Dichloromethane*
Selenium	Dichlorobenzenes	1,3-dichloropropene
Silver	Diethyl phthalate	Dieldrin
Zinc	Dimethyl phthalate	2,4-Dinitrotoluene
Cyanide	4,6-dinitro-2-methylphenol	1,2-Diphenylhydrazine
Total Chlorine Residual	2,4-dinitrophenol	Halomethanes
Ammonia (as N)	Ethylbenzene	Heptachlor
Acute Toxicity	Fluoranthene	Heptachlor Epoxide
Chronic Toxicity	Hexachlorocyclopentadiene	Hexachlorobenzene
Phenolic Compounds	Nitrobenzene	Hexachlorobutadiene
Chlorinated Phenolics	Thallium	Hexachloroethane
Endosulfan	Toluene	Isophorone
Endrin	Tributyltin*	NDMA
HCH	1,1,1-Trichloroethane	N-nitrosodi-N-propylamine
Gross Alpha	Acrylonitrile	N-nitrosodiphenylamine
Gross Beta	Aldrin	PAHs
Radium 226	Benzene	PCBs
Radium 228	Benzidine	TCDD Equivalents*
Tritium	Beryllium	1,1,2,2-Tetrachloroethane
Total Uranium	bis(2-chloroethyl) ether	Tetrachloroethylene
	bis(2-ethylhexyl) phthalate	Toxaphene
	Carbon Tetrachloride	Trichloroethylene
	Chlordane	1,1,2-Trichloroethane
	Chlorodibromomethane	2,4,6-Trichlorophenol
	Chloroform	Vinyl Chloride
		* not analyzed

8 CONCLUSIONS

1. The US Filter CMF-S microfiltration system with the redesigned S10V PVDF membranes:

- a. Demonstrated that a flux of 34 GFD was sustainable if oxidant addition to the backwash was utilized. 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

For further details see Table 6 above.

- b. 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 - 38°C followed by
- ii. 400 – 600 mg/L NaOCl recirculation at 20 - 22°C

- c. Produce filtrate water with turbidity and SDI suitable for spiral RO membranes when the MF system maintained integrity. For the US Filter CMF-S system this translates to pressure decay test results of less than 2 psi/min.
- d. While fiber breakage during this testing has been at an unacceptable rate, it is believed that this is related to shell fragments in the feed water. The addition of the Arkal filter in front of the redesigned CMF-S module did show signs of alleviating this issue, but more data is required.

2. Reverse Osmosis membranes operated effectively for 8 to 11 GFD flux on microfiltered permeate. No noticeable fouling was observed during continuous operation. RO parameters were found to be:

- a. Flux = 8 to 11 GFD
- b. Recovery = 50%

-
- c. 3 mg/L sodium bisulfite dosage in feed to prevent oxidation by chlorine added in the microfiltration process
 - d. 3mg/L antiscalant dosage in feed to prevent the precipitation of sparingly soluble salts.
 - e. Typical feed pressure of:
 - i. 810 to 850 psig for 8 GFD
 - ii. 840 to 870 psig for 9 GFD
 - iii. 870 to 920 psig for 11 GFD
3. Both the microfiltration backwash and reverse osmosis concentrate waste streams were characterized for disposal options. The characteristics of each of these streams are listed in Tables 8 and 12, 13 and 14 respectively.
4. The continuous $\text{NaOCl} \rightarrow \text{MF} \rightarrow \text{NH}_4\text{OH} \rightarrow \text{RO}$ process proved to be unsuitable for full-scale implementation due to the suspected mechanism of the creation of bromamine. This process was replaced by MF backwash chlorination and continuous sodium bisulfite dosage in front of the RO. See the appendix for a detailed discussion. Test results indicate that this is a viable approach. However, increases in RO membrane salt passage indicate that some oxidant may be reaching the RO, requiring some refinement of the process.
5. Operation on ocean water from the common power plant intake introduced additional challenges for the treatment process. The power plant heat treatment cycles, which clear the intake 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 um in series with a 500 um strainer proved to be ineffective at preventing sand and crushed shell fragments from reaching the MF system and puncturing fibers. Additional testing showed that the 800 um strainer in series with the 130 um Arkal Spin Klin filter showed signs of alleviating this problem, but further testing is required.

-
6. Each of the RO membranes tested demonstrated the capability of providing permeate water less than 300 mg/L TDS. Typical RO permeate as produced included:
 - a. 150 to 230 mg/L of TDS and 0.6 to 1.2 mg/L of Boron at 8 GFD
 - b. 178 to 185 mg/L of TDS and 0.8 to 1.1 mg/L of Boron at 9 GFD
 - c. 160 to 200 mg/L of TDS and 0.8 to 1.1 mg/L of Boron at 11 GFD
 7. Two sets of Hydranautics and Dow membranes were tested, and in each set, Filmtec membranes initially produced significantly lower permeate concentration. The first set of membrane results suffered from membrane oxidation with the oxidation much more rapid on the Filmtec membranes. While not as severe, the second set of Filmtec membranes experienced a decrease of salt rejecting properties, whereas the Hydranautics water quality was more stable.

Highlights of the results include:

- **MF Operation:** An MF operating strategy was developed which maximized operating flux, while maintaining greater than 21 days between chemical cleanings.
- **MF Filtrate Quality:** The MF process was demonstrated to produce excellent filtrate quality, despite variations in feed quality. As a result the RO membranes operated with essentially no measurable fouling at flux rates varying from 8 to 11 GFD.
- **RO Permeate Quality:** The reverse osmosis systems demonstrated the ability to produce consistently high quality (less than 300 mg/L TDS, pre-stabilization), meeting or exceeding most of the membrane manufacturer's performance predictions.
- **Boron Rejection:** Data was generated on boron rejection by the RO process, indicating somewhat different performance between the two membranes tested. One membrane producing permeate boron typically 0.8 mg/L, the other 1.1 mg/L.

9 FULL SCALE COST ANALYSIS

A conceptual level cost estimate has been prepared for full-scale implementation of the desalination process developed in this project. Capital and O&M costs for each component of the treatment process and facility were estimated. Equipment and construction costs can certainly be site specific and subject to change. These values assumed construction of the facility in Southern California and 2005 costs.

The basis of the estimate is as follows:

- Two cases are indicated. 20 and 40 mgd nominal product capacity. A 95% on-line factor was used resulting in annual production rate values which are used for the unit cost calculations (\$/kgal & \$/ac.ft.).
- Seawater intake and outfall structures were not included. This estimate assumes connection to existing power plant cooling loop. Cost of a connection to the cooling loop was included.
- Feed and discharge pipelines were included between the seawater source and the treatment facility, assumed to be a distance of 1,400 feet.
- The treatment process included the use of the 130 micron disk filters described in this report.
- All treatment equipment costs include installation.
- Chlorine costs include that used in MF backwash and product disinfection. Ammonia was included assuming chloramination of product.
- RO energy recovery device was assumed to be pelton wheel device at 88% efficiency.
- Electricity cost at \$0.05/kWh
- Capital amortization over 25 years at 6%.
- All treatment equipment was placed within a building, at \$250/sq.ft.
- No cost for land was included.
- Product Storage, pump station and distribution pipelines are not included.
- No cost for permitting, feedwater assessment, sanitary survey, etc, were included.

COST ESTIMATE - SEAWATER DESALINATION
WITH MICROFILTRATION AND REVERSE OSMOSIS

TABLE 19 - Conceptual Level Capital Cost Estimate

Nominal design capacity (mgd)	20.0	40.0
1.1 Process Equipment		
Pre-strainers	\$ 1,050,000	\$ 2,150,000
Microfiltration System	\$ 11,600,000	\$ 21,050,000
MF Basins	\$ 250,000	\$ 500,000
Reverse Osmosis System	\$ 37,000,000	\$ 70,000,000
Interconnecting Piping	\$ 2,495,000	\$ 4,685,000
Chemical Addition (Lime, CO ₂ , NaOCl etc.)	\$ 1,100,000	\$ 1,400,000
Electrical (to sites, transformers, MCC, etc.)	\$ 9,272,000	\$ 16,314,000
Instrumentation & Controls	\$ 3,000,000	\$ 4,500,000
Miscellaneous (5% of above)	\$ 3,277,000	\$ 6,010,000
subtotal	\$ 69,044,000	\$ 126,609,000
1.2 Pipelines & Pump Stations		
Feed Pump Station at Power Plant		
-Tie-in to seawater cooling loop, pumps, foundations etc.	\$ 2,870,000	\$ 4,875,000
Feed Pipeline from Feed Pump Station to Desal Facility		
36"/50" Inlet pipeline - 1,400 ft.	\$ 605,000	\$ 840,000
Discharge Pipeline from Desal Facility to Power Plant Outfall		
30"/42" Discharge pipeline - 1,400 ft.	\$ 505,000	\$ 705,000
Product Pipeline		
- 30"/42" Desal Facility to Product Storage Tank - 300 ft.	\$ 110,000	\$ 155,000
subtotal	\$ 4,090,000	\$ 6,575,000
1.3 Treatment Plant Building (@ \$250/sq. ft.)	\$ 17,375,000	\$ 26,625,000
1.4 Site development	\$ 2,900,000	\$ 4,100,000
1.5 Engineering (10%)	\$ 9,500,000	\$ 16,400,000
1.6 Project Management (5%)	\$ 4,750,000	\$ 8,200,000
Subtotal	\$ 107,659,000	\$ 188,509,000
1.7 Contingency (30%)	\$ 32,825,000	\$ 56,645,000
Total Estimated Cost	\$ 140,484,000	\$ 245,154,000

COST ESTIMATE - SEAWATER DESALINATION
WITH MICROFILTRATION AND REVERSE OSMOSIS

TABLE 20 - PERTINENT OPERATING PARAMETERS

Nominal Plant Product Capacity, mgd	20.0	40.0
2.1 Reverse Osmosis System		
Recovery	50%	50%
Permeate Flow, gpm	13,889	27,778
Feed Flow, gpm	27,778	55,556
Concentrate Flow, gpm	13,889	27,778
RO Feed Pressure, psig		
Average	975	975
Pump Efficiency	0.83	0.83
Motor Efficiency	0.93	0.93
Motor/VFD Efficiency	0.91	0.91
Energy Recovery Device Efficiency	0.88	0.88
# of RO Membrane Elements	5,845	11,690
2.2 Microfiltration System		
Recovery	90%	90%
Filtrate Flow, gpm	27,777.8	55,555.6
Feed Flow, gpm	30,864.2	61,728.4
Backwash Flow, gpm	3,086.4	6,172.8
MF Filtrate pump TDH (feet)	39.0	39.0
# of MF Membrane Modules	5,400	10,800
Power Plant Pump Station Pump TDH	69	69
2.4 Product Quality, TDS mg/L		
RO Permeate	<325	<325
Stabilized Product	<400	<400

COST ESTIMATE - SEAWATER DESALINATION
WITH MICROFILTRATION AND REVERSE OSMOSIS

TABLE 21 - DAILY COST OF CHEMICALS

Nominal Plant Product Capacity, mgd	20.0	40.0
3.1 Threshold Inhibitor (TI)		
Pounds of TI per Day	1,000.8	2,001.6
Cost per Pound*	\$ 1.00	\$ 1.00
Cost of TI per Day	\$ 1,000.80	\$ 2,001.60
3.2 Sodium Bisulfite (SBS)		
Pounds of SBS per Day	166.8	333.6
Cost per Pound*	\$ 0.23	\$ 0.23
Cost of SBS per Day	\$ 37.53	\$ 75.06
3.3 Lime		
Pounds of Lime per Day	6,136.8	12,273.6
Cost per Pound	\$ 0.053	\$ 0.053
Cost of Lime per Day	\$ 325.25	\$ 650.50
3.4 Carbon Dioxide (CO₂)		
Pounds of CO ₂ per Day	8,558.0	17,116.1
Cost per Pound	\$ 0.05	\$ 0.05
Cost of CO ₂ per Day	\$ 427.90	\$ 855.80
3.5 Chlorine (Cl₂)		
Pounds of Cl ₂ per Day	166.80	333.60
Cost per Pound*	\$ 0.67	\$ 0.67
Cost of Cl ₂ per Day	\$ 111.76	\$ 223.51
3.6 Ammonia (NH₃)		
Pounds of NH ₃ per Day	16.7	33.4
Cost per Pound*	\$ 1.27	\$ 1.27
Cost of NH ₃ per Day	\$ 21.18	\$ 42.37
Total Cost of Chemicals / Day	\$ 1,886.89	\$ 3,773.78

*cost per pound based on 100% strength

COST ESTIMATE - SEAWATER DESALINATION
WITH MICROFILTRATION AND REVERSE OSMOSIS

TABLE 22 - DAILY ENERGY REQUIREMENTS

Nominal Plant Product Capacity, mgd	20.0	40.0
4.1 Seawater Intake Pump Station		
Power Plant Pump Station Pump Brake HP	648	1296
Power Plant Pump Station Pump Motor In HP	712	1424
Power Plant Pump Station Pump Motor In kW	531	1063
4.2 MF Filtrate pumping		
MF Filtrate Pump Station Pump Brake HP	330	659
MF Filtrate Pump Station Pump Motor In HP	362	725
MF Filtrate Pump Station Pump Motor In kW	270	540
4.4 RO Transfer Pump		
RO Transfer Pump Brake HP	781	1,562
RO Transfer Pump Motor In HP	858	1,717
RO Transfer Pump Motor In kW	640	1,281
4.3 RO High Pressure Pump		
RO Pump Brake HP	18,452	36,904
Energy Recovery Device HP	6,810	13,620
RO Pump Motor Brake HP	11,642	23,284
RO Pump Motor Input HP	12,793	25,587
RO Pump Motor Input kW	9,544	19,088
Other RO Power, kW	477	954
Total RO Power, kW	10,021	20,042
Total Power Required, kW	11,463	22,926

COST ESTIMATE - SEAWATER DESALINATION
WITH MICROFILTRATION AND REVERSE OSMOSIS

TABLE 23 - O&M COSTS SUMMARY

Nominal Plant Product Capacity, mgd	20.0	40.0
5.1 Cost Factors		
Capacity Factor	100%	100%
On-Line Factor	95%	95%
Operating Days per Year	347	347
Water Production, kgal/year	6,935,000	13,870,000
Water Production, ac.ft./year	21,283	42,566
Operators, Hours/Day	40	60
Maintenance, Hours/Day	12	24
Labor Cost per Hour	\$ 40	\$ 40
MF Module Cost	\$ 600	\$ 600
RO Element Cost	\$ 500	\$ 500
RO&MF Memb. Life, Years	5.0	5.0
Energy Cost per kWhr	\$ 0.05	\$ 0.05
5.2 Annual O&M Cost		
Chemicals	\$ 654,280	\$ 1,308,559
Energy	\$ 4,769,669	\$ 9,539,338
Operating Labor	\$ 554,800	\$ 832,200
Maintenance Labor	\$ 175,200	\$ 350,400
Maintenance Materials	\$ 702,420	\$ 1,225,770
RO Membrane Replacement	\$ 584,500	\$ 1,169,000
MF Membrane Replacement	\$ 648,000	\$ 1,296,000
Miscellaneous (5%)	\$ 404,443	\$ 786,063
Total O&M Cost	\$ 8,493,312	\$ 16,507,331

COST ESTIMATE - SEAWATER DESALINATION
WITH MICROFILTRATION AND REVERSE OSMOSIS

TABLE 24 - TOTAL COST OF WATER

Nominal Plant Product Capacity, mgd	20.0	40.0
6.1 Capital Amotization		
Project Capital Cost	\$ 140,484,000	\$ 245,154,000
Amortization Period, Years	25	25
Interest Rate	6%	6%
Capital Recovery Factor	0.07823	0.07823
Annual Capital Recovery	\$ 10,989,602	\$ 19,177,593
6.2 Annual & Volumetric Costs		
Total Annual Cost of Water	\$ 19,482,914	\$ 35,684,923
Cost of Water, \$/ac.ft.	\$ 915	\$ 838
Cost of Water, \$/kgal	\$ 2.81	\$ 2.57

10 APPENDIX -PROCESS AND EQUIPMENT CHALLENGES

10.1 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.⁸ 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 was 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 inactivated 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 was added after significant chlorine contact time to neutralize the oxidant before it contacts the RO membranes. However, this chlorination/dechlorination process was shown to actually enhance the tendency towards biological fouling of the RO.⁹

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:

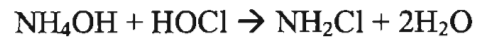
⁸ Reverse Osmosis and Nanofiltration, 1999 American Water Works Association.

⁹ “Controlling Biological Fouling in Open Sea Intake RO Plants Without Continuous Chlorination,” by Hamida and Moch, Desalination and Water Reuse, November/December 1996.

Reaction 1, Addition of sodium hypochlorite:



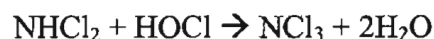
Reaction 2, Formation of monochloramine



Reaction 3, Formation of dichloramine



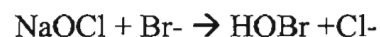
Reaction 4, Formation of trichloramine⁵



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 creation of chloramine on this water source. First, ammonia is not present in ocean water and thus must be added. Second, the presence of bromide (Br⁻) in ocean water interferes with the reactions above. The Pacific Ocean water source used in this study has ~64 mg/L of Br⁻. Br⁻ substitutes for Cl⁻ in reactions 1 - 4 listed above such that the chlorine addition to ocean water actually produces hypobromous acid (HOBr) instead of HOCl. Furthermore, subsequent ammonia addition creates bromamines instead of chloramines. The following reactions apply:

Reaction 5, Addition of NaOCl to ocean water



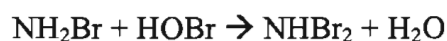
Reaction 6, side reaction with chloramines



Reaction 7, subsequent Ammonia addition



Reaction 8, dibromamine formation¹⁰



¹⁰ Handbook of Chlorination and Alternative Disinfectants, 1999, White.

To protect the RO membranes from oxidation by HOBr, the mol ratio of $\text{NH}_3:\text{HOCl}$ addition should be ~2:1 or greater. A 1 mg/L NaOCl addition and subsequent 1mg/L NH_4OH addition utilized in this pilot study represented an $\text{NH}_3:\text{HOCl}$ mol ratio of 2.1:1. However, HOBr and bromamines are suspected to be stronger oxidants than their chlorine equivalents, HOCl and chloramines. There was little information or data on the exposure of thin film 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 21 and 22, this $\text{NaOCl} \rightarrow \text{MF} \rightarrow \text{NH}_4\text{OH} \rightarrow \text{RO}$ 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 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 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.

10.2 Power Plant Heat Treatment Cycles

The pilot trials were started in June 2002. Soon thereafter, the power plant performed a heat treat cycle. Approximately every one to three months the power plant intake 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 °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 um 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 um strainer, the booster pump, and the 500 um strainer on the US Filter 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 as discussed in section 5.3 above.

10.3 Addition of Arkal Spin Klin Strainer

The study experienced numerous US Filter CMF-S module fiber breakage events, even after the 800 um strainer was placed in front of the booster pump as described in the appendix. US Filter 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 US Filter CMF-S system. In addition, the 500 um strainer directly in front of the CMF-S system was replaced by an Arkal Spin Klin 130 um self backwashing strainer. 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 US Filter unit. The air demand was too large for the compressor and when the Arkal

went into backwash, the US Filter 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 um discs. The system was sent with 30 um discs and this in combination with the low air pressure resulted in clogging and high dP's.

As a result of these challenges, the Arkal strainer 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 um. Fiber breakage events occurred and more modules required replacement. The Arkal strainer was finally placed in operation in March 2004 and did show signs of alleviating the fiber breakage problem.

10.4 Vibration issues associated with Wanner Hydracell High Pressure RO Pumps

The RO System utilized for this study has two independent trains of 4M, 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 superaustenitic stainless steel wetted parts that withstands the corrosive seawater 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:

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- 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, swagelok 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. The pumps were given variable frequency drives to lower the motor speed and eliminate the loop that recycled excess water back to the suction of the system.