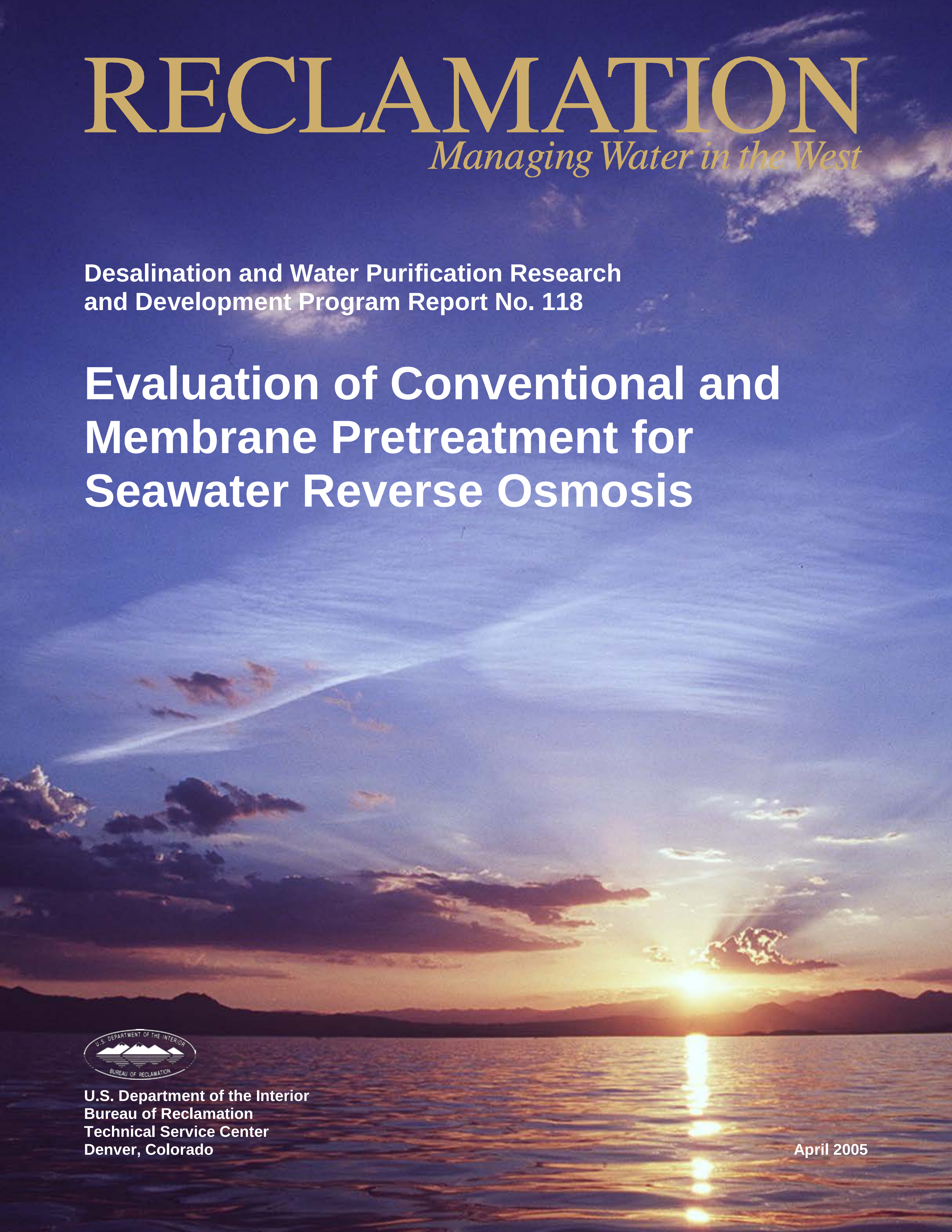

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Managing Water in the West

Desalination and Water Purification Research
and Development Program Report No. 118

Evaluation of Conventional and Membrane Pretreatment for Seawater Reverse Osmosis

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

April 2005

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14. ABSTRACT

A unique series of bench and pilot scale tests were designed and conducted as a part of this project to determine the effect of pretreatment on SWRO performance. The testing program included a literature review, raw water characterization, bench scale testing and two-phase pilot testing. The information collected in the literature review was used provide the study a background on SWRO pretreatment and in the design of the bench and pilot scale experiments.

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EVALUATION OF CONVENTIONAL AND MEMBRANE PRETREATMENT FOR SEAWATER REVERSE OSMOSIS

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Disclaimer

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GLOSSARY OF TERMS

BAIReMT	Batch Internal Recycle Membrane Test
Chl-a	Chlorophyll-a
DO	Dissolved Oxygen
ft^2	square foot
FTIR	Fourier Transform Infra Red Spectroscopy
gal	gallon
gfd	gallons per square foot per day
gpm	gallons per minute
HPC	Heterotrophic Plate Count
hr	hour
in	inch
kg	Kilograms
L	Liter
$L/hr\text{-}m^2$	Liters per hour per square meter
kDa	Kilo Dalton
m^2	square meter
m^3	cubic meter
m^3/min	cubic meter per minute
m^3/day	cubic meter per day
MF	Microfiltration
mg	Milligram
mg/L	Milligrams per Liter
min	minute
mL	Milliliter
MGD	Million Gallons per Day
mm	millimeter
MWH	Montgomery Watson Harza
NaOCl	Sodium Hypochlorite
ND	Non-Detect
NTU	Nephlometric Turbidity Units
O&M	Operations and Maintenance
PDT	Pressure Decay Test
PLWTP	Point Loma Wastewater Treatment Plant
PL	Point Loma
PO_4	Ortho Phosphate
ppm	Parts Per Million
psi	Pounds per Square Inch
RBSMT	Rapid Bench Scale Membrane Testing
RO	Reverse Osmosis

s	seconds
SEM	Scanning Electron Microscopy
SDI	Silt Density Index
SUVA	Specific Ultraviolet Absorbance
SWRO	Seawater Reverse Osmosis
TFC	Thin Film Composite
TOC	Total Organic Carbon
TSS	Total Suspended Solids
TAC	Technical Advisory Committee
UF	Ultrafiltration
UV_{254}	Ultraviolet Absorbance at 254 nanometer
$\Delta\pi$	Net Osmotic Pressure of the Feed and Permeate
°C	Degrees Celsius
μg	Microgram
μmhos	Micromhos
μm	Micron

Calculated Parameters

IAF	Integrated Averaging Factor
J	Membrane Flux (gfd)
J@20°C	Temperature Corrected Membrane Flux (gfd)
J_{SP}	Specific Flux (gfd/psi)
P_{NET}	Net Operating Pressure (psi)
Q_{NET}	Net Permeate Rate (gpm)
R	Salt Rejection (%)
TMP	Transmembrane Pressure (psi)
Y	Recovery

1. EXECUTIVE SUMMARY

There has been increased focus on seawater desalination as a viable alternative water source for coastal communities in recent years. Advances in membrane technology and increased pressure on traditional water supplies have contributed to bringing this almost inexhaustible source of water within reach economically for coastal regions, which are heavily dependent on imported water. In addition to economic limitations a major technical hurdle to application of Seawater Reverse Osmosis (SWRO) desalination for drinking water production has been membrane fouling due to inadequate pretreatment. Recent full-scale experiences have shown that pretreatment is key to the success of SWRO facilities.

Most currently operating SWRO plants use conventional pretreatment. Conventional treatment as used in this report refers to any combination of coagulation, flocculation, sedimentation and granular media filtration. Membrane filtration is now being tested extensively as another pretreatment strategy that could possibly provide superior feedwater quality as compared to conventional pretreatment for SWRO. To study these two major pretreatment alternatives and their impact on SWRO performance, the City of San Diego and their as needed research consultant, MWH, were awarded a cooperative agreement by the Bureau of Reclamation

A unique series of bench and pilot scale tests were designed and conducted as a part of this project to determine the effect of pretreatment on SWRO performance. The testing program included a literature review, raw water characterization, bench scale testing and two-phase pilot testing. The information collected in the literature review was used provide the study a background on SWRO pretreatment and in the design of the bench and pilot scale experiments.

An analysis of the seawater was conducted from three different sites in the San Diego region. These sites included representative sites from open ocean (Scripps Pier), bay location close to open ocean (Shelter Island) and bay location away from open ocean (South Bay Power Plant). It was found that bay water is more susceptible to influence of surface water runoff than open ocean water. Additionally, the water quality in the location inside the bay with poor flushing (South Bay Power Plant) is different from the bay close to open ocean and has a different organic makeup with higher Specific Ultraviolet Absorbance (SUVA) content. Characterization from the influent and the effluent of South Bay Power Plant was also conducted. It was found that the water quality of the effluent from the power plant was of a better quality than the influent in terms of particulate and organic loading. Consequently, the effluent was chosen as a water source for conducting pilot testing. In terms of ion content and makeup the distributions were similar between the influent and the effluent.

Bench scale tests were conducted for optimization of coagulant and polymer dose for conventional pretreatment. The optimum ferric dose for conventional pretreatment of the source water used for the bench scale testing was found to be 4 mg/l with no polymer.

A bench scale setup based on a modification of the Batch Internal Recycle Membrane Test (BAIReMT) developed by DiGiano et al (1999) was designed for screening RO membranes.

SWRO membranes from six manufacturers – Dow, Hydranautics, Koch, Osmonics, Saehan, and Toray were tested. This bench scale method shows great promise for quick evaluation of comparative RO fouling. Based on these tests Hydranautics SWC 4 membrane was selected for pilot testing. This product had not been tested at the pilot scale in previous studies at the time of this study.

Further bench scale testing using seawater fractions (based on size) was conducted using this bench scale setup. These fractions were selected to correlate with size fractions generally associated with existing pretreatment processes. Bench Scale fractionation tests indicated that Microfiltration (MF) and Ultrafiltration (UF) size fractions of seawater caused lower RO flux decline than the conventional filtration size fraction. In addition, these bench scale fractionation tests indicated that there was minimal difference in RO fouling rates between the UF and MF fractions of seawater. Fourier Transform Infra Red Spectroscopy (FTIR) and Scanning Electron Microscopy (SEM techniques) were used to further evaluate fouling magnitude and mechanisms.

The pilot testing was conducted in two phases: Phase I testing was pretreatment evaluation and Phase II was RO evaluation. During Phase I two filter configurations and Microfiltration were evaluated as pretreatment strategies. Water quality data including indicators for particulate, biological, organic and inorganic fractions of the seawater was collected from the pretreatment schemes to determine how these relate to the fouling of the RO membranes. During this phase SDI was between 1 and 3 for the membrane treated water while it was between 3 and 6 for the conventionally treated water. The conventional filters operated at run times between 80 and 100 hours consistently at a loading rate of 6 gpm/sqft. The MF system operated at a flux of 30 gfd for over 600 hours with minimal fouling. The MF system was operated with no use of chlorine and a 30 minute backwash interval was used during the testing.

In the second phase of pilot testing two RO systems were operated simultaneously to evaluate the effect of each pretreatment strategy on RO fouling under different operating conditions and water sources. It was found that the RO performance of the train operating conventional pretreatment is highly dependent on the filter design and water quality, while the RO train operating on MF pretreatment was less sensitive to water quality. In all cases the MF RO train fouled to a lower degree than the conventional RO train.

A discussion of the cost factors influencing full scale RO plants is also presented as a part of this report. This discussion is based on data from current study, literature review and the opinion of the project team. As a result of the analysis conducted it was found that the major cost factors that are influenced by pretreatment include: operational flux, specific flux decline rate, RO influent water quality, and membrane life.

2. INTRODUCTION

The City of San Diego has been promoting the development of alternative water sources to reduce San Diego County's reliance on limited imported water supplies. Seawater desalination presents an attractive opportunity for water scarce regions like San Diego located in coastal areas. It has been traditionally considered the last resort for communities after surface water and groundwater. The primary argument against using available seawater as a water source has been the high cost of seawater desalination compared to the cost of treating water from other sources. However, in recent years two developments have made this option much more viable - increasing pressure on available water sources and the rapid development in seawater desalination technologies like reverse osmosis leading to reduction in costs.

This section presents a background of the state of knowledge on seawater reverse osmosis (SWRO) and the importance of pretreatment to the SWRO process, and lays out the objectives for this study.

2.1 Background

The application of reverse osmosis (RO) membranes to seawater desalination has increased rapidly during the past decade. The industry has recently witnessed the commissioning of numerous full-scale seawater reverse osmosis plants, with capacities upto 45 mgd. A representative list of these plants is shown in **Table 2-1.** Though the majority of these plants are located outside the United States, there are several SWRO plants being proposed in coastal states which are forecasting rapid population growth such as California, Texas and Florida. These proposed plants and their status at the time of writing this report is presented in **Table 2-2**. This interest in SWRO is largely attributed to recent advancements in membrane technology, which have resulted in the production of membranes that operate with significantly lower pressure and higher productivity than those produced in previous generations. These advancements coupled with the decreasing supply of freshwater resources has caused the drinking water industry to take a closer look at SWRO as a sustainable water supply alternative.

While the application of RO membranes for desalination of seawater is not a new concept, there are currently several key issues currently facing the industry. Many of these issues have surfaced during the operation of existing plants and during the planning of future plants. These include proper characterization of seawater quality, evaluation/comparison of pretreatment options (including performance and cost), environmental impacts, types of intake structure, regulatory requirements, meeting multiple water quality objectives and assessment of different suppliers currently offering SWRO membranes. Of these issues one of the most important is the selection of proper pretreatment upstream of the SWRO membranes. Pretreatment is fundamental in ensuring functional efficient operation of the SWRO process. Currently the two most common methods being employed for pretreatment include conventional (e.g. coagulation/ flocculation/sedimentation/filtration) and low-pressure membranes (MF/UF). While both methods have been demonstrated to be successful, adequate comparison has not been made. In order to properly understand the overall effect of pretreatment selection one must look at the impact of pretreatment on RO performance.

2.1.1 Conventional SWRO Pretreatment

SWRO conventional pretreatment, in many respects, is similar to that used in conventional drinking water treatment systems, with the main difference being that the product water is optimized to enhance the efficiency of RO membranes. A typical process flow schematic of deep-bed conventional pretreatment is shown in **Figure 2-1**. As shown, major processes of conventional SWRO pretreatment include coagulation, flocculation, sedimentation and filtration. Chlorine is often dosed upstream of the coagulant addition to reduce biogrowth on the downstream filters. **Table 2-3** provides a list of pilot and full-scale SWRO facilities with conventional pretreatment. Key information shown includes chlorine dose, coagulant type/dose, polymer type/dose (if used) and details on the type of filtration used. As shown, either ferric sulfate or ferric chloride were used for all facilities listed. However, the dose used in each application varied significantly. Ferric coagulant doses listed in **Table 2-3** range from 0.8 to 21 mg/L as Fe. Such data indicates that the dose required for destabilization of colloids varies with location due to differences in raw seawater quality. If the dose is too low there will not be enough positively charged metal ions present to neutralize the negatively charged colloids present in the raw seawater. On the contrary, overdosing may result in charge reversal of the negatively charged particles causing repulsion. The optimal coagulant dose can be determined using bench scale testing (jar tests).

The use of coagulant aids such as inorganic/organic polymers during conventional SWRO pretreatment can aid in destabilization of colloidal matter present in raw seawater. As indicated in **Table 2-3**, four of the eight facilities listed reported using a polymer as part their conventional pretreatment SWRO process. In general, it was found that cationic polymers were generally applied at a low dose (i.e. < 1.0 mg/L) and in several circumstances did not significantly improve filter water quality.

During the next conventional pretreatment step, coagulated flocs are removed by sedimentation and filtration. In general, the filtration process is characterized by media type and filter depth. Filters can be designed using mono, dual or multimedia. Typically, mono media filters use sand, dual media filters use combination of sand and anthracite and multimedia use sand, anthracite and garnet. Characteristics of the different media types typically used in conventional SWRO pretreatment are provided in **Table 2-4**. **Table 2-5** presents the depths of the different media used in seawater pretreatment and compares it to conventional surface water (freshwater) treatment. As shown, typical filter beds range from 1250 to 2250 mm (50-90 inches). Due to increased media amount, deeper bed is considered optimal design for conventional pretreatment. To date the authors have found no studies comparing SWRO performance based on varying bed depth. In general, three types of filters are used in conventional pretreatment including gravity, pressure and continuous backwash.

Of these, the most common type of filters are pressure filters, which offer the advantage of longer filter runs but relatively higher costs (Hagen and Comstock, 2004). Continuous backwash filters are typically applied in situations where there is variability in the water quality of the incoming feed water.

2.1.2 Low Pressure Membrane SWRO Pretreatment

The application of low-pressure membranes such as MF and UF for pretreatment to SWRO is a relatively new concept, which is increasingly being evaluated by the drinking water industry. General characteristics of these processes including consistent product water quality (regardless of feed concentration), absolute barrier to microbial contaminants, reduced footprint, and zero or minimal chemical addition make them an attractive alternative to conventional pretreatment for SWRO applications.

A general flow schematic of MF/UF pretreatment system used for SWRO is provided in **Figure 2-2**. As shown, seawater from the intake structure passes through a strainer to remove large particulate matter such as shells and other debris, which can cause physical damage to the low-pressure membranes. As indicated, typical, strainer sizes range from 150 to 800 μm, depending on the quality of the incoming seawater. Sodium hypochlorite may then be dosed to prevent biofouling of the downstream MF/UF membranes. Next, seawater enters into the MF/UF system, which typically consists of membrane holding tanks, membranes, permeate storage tanks, permeate pumps and associated piping. MF/UF membranes are configured to be operated with an inside-out or outside-in flow pattern. In addition, they can be housed in pressure vessels (pressure driven) or submerged in the feed water (vacuum systems). **Figure 2-2** depicts an outside-in submerged system. As shown, a vacuum pump supplies a slight pressure in the membrane holding tank, which draws the water from the outside-in. A portion of the filtrate is then collected in a permeate holding tank and is used to backwash the membranes. During backwash, permeate is pumped from the inside of the fibers to the outside to mitigate the build up of foulants on the membrane surface. Also shown, sodium hypochlorite maybe added to the backwash water. This is typical done instead of adding it continuously upstream of the MF/UF. Lastly, as shown in **Figure 2-2**, sodium bisulfite can be added to the MF/UF permeate prior to RO treatment. These chemicals are added to neutralize the chlorine residual, which can damage the RO membranes.

A list of SWRO pilot and full-scale facilities utilizing MF/UF pretreatment is provided in **Table 2-6**. Pertinent information provided for each facility includes membrane manufacturer/type, prescreen type/micron rating, chemical additions (i.e. type/dose/location), operating flux range, feedwater recovery and backwash frequency (i.e. production time between consecutive backwashes). As shown, the majority of facilities utilized some type of strainer or prescreen prior to MF/UF system. This is necessary to prevent sharp objects such as shell fragments from damaging the hollow fiber membranes. Henthorne et al., 2003, reported it was necessary to replace the typical strainers equipped (500-800 micron) on MF/UF pilot systems with a Arkal Spin Klin unit offering 130 micron filtration using a disk filtration setup. Jew et al., 2003 reported that Pressure Decay Tests (PDT) and Scanning

Electron Microscopy (SEM) tests suggested that hollow fibers may have been cut by sharp objects during pretreatment of seawater using microfiltration. As a result, it was necessary to switch from 800/500 micron strainer to a 800/150 micron strainer.

As shown in **Table 2-6**, several of the facilities dose free chlorine (0.5-1-ppm) on a continuous basis in the feed water to prevent to membrane biofouling. At these facilities, free chlorine was either neutralized using sodium bisulfite (SBS) or converted to chloramines using ammonium hydroxide prior to RO treatment. Jew et al., 2003 show that due to the presence of bromide ion in seawater, attempts to convert free chlorine to chloramine can result in the formation of dibromamines instead of chloramines. The authors reported that dibromamines formed in the MF permeate resulted in subsequent damage to RO membranes as indicated by reduced salt rejection and increase specific flux. To remedy this, continuous chlorination was stopped and chlorine was only added during backwashing and neutralized using SBS prior to RO. However as reported by Henthorne et al., 2004 the use of chlorine only during backwashing is not a fail-safe to potential damaging downstream RO membranes. Specifically the authors reported a reduction in salt rejection of RO membranes from 99.9%-99.3% during UF pretreatment performing Chemically Enhanced Backwashes (CEBs) and neutralized with 1-2 mg/L SBS.

As indicated in **Table 2-6**, several of the facilities report dosing ferric chloride in the raw feed water. The purpose of the coagulant addition would be to neutralize and agglomerate colloidal particles, such as organic matter, prior to membrane filtration. Such particles could then be removed easily during backwashing resulting in reduced membrane fouling. As shown, reported ferric chloride doses (as Fe) are relatively low (0.1-1.5 mg/L) compared to conventional pretreatment (0.8-18 mg/L). To date there has been limited data to quantify the benefit of coagulant addition on the performance of MF/UF during SWRO pretreatment.

As shown, typical operating flux values for MF/UF SWRO pretreatment was reported between 10–70 gfd, feed water recoveries 80-95% and backwashing frequency between 15–30 minutes. Several of the facilities reported using chlorine injection during backwashing to prevent biogrowth on the membrane surface.

2.1.3 Effect of pretreatment on SWRO performance

Table 2-7 provides a list of SWRO projects/facilities utilizing conventional and/or membrane (MF/UF) pretreatment found in recent literature. Based on the information provided, it is possible to compare reported effluent SDI values of the two pretreatment options and downstream performance of RO membranes. As shown, SDI values reported from conventional pretreatment facilities were reported between 1.6-6.0 while the majority of SDI values reported from MF/UF systems were <3.0. These reported SDI values suggest that RO performance following pretreatment by MF/UF would be superior to water treated using conventional pretreatment due to reduction in solid loading to subsequent RO membranes.

As shown in **Table 2-7**, successful operation has been reported at SWRO facilities using both conventional pretreatment and membrane pretreatment. For instance, during the first year of operation, following installation of conventional pretreatment, the Point Lisas SWRO plant, reports consistent SDIs (< 3) from the conventional pretreatment and minimal increase in differential pressure across the pass 1 SWRO membranes. With the implementation of slight modifications to the conventional pretreatment process it is anticipated that cleaning interval of the first pass RO membranes will be 8-9 months (Irwin and Thompson, 2003). Since commissioning in 2002, the United Arab Emirates (UAE) has successfully treated seawater from the Persian Gulf using a 1.3-MGD mobile seawater RO desalination barge, which uses Norit X-Flow UF membranes as pretreatment. To date SDI from the UF membranes has been reported below 3 and RO membranes have been reported to operate at 12-13 gfd and at a feed water recovery of 43%. Though both pretreatment options have been shown successful the proper selection must be based on specific seawater quality which can vary geographically and locally. To fairly compare conventional pretreatment and membrane pretreatment it is imperative to perform a simultaneously evaluation of RO performance following both pretreatment options.

As shown in **Table 2-7**, five of the ten facilities report evaluating RO performance following both conventional pretreatment and MF/UF pretreatment. For example, Henthorne et al., 2003 reports a significant difference in hydraulic performance of SWRO membranes pretreated with membranes as compared to conventional pretreatment. Specifically, the authors reported the MF/UF-RO train operated with no cleaning throughout pilot testing as opposed to the conventional pretreatment-RO train which required cleaning every 6 weeks (based on increase in net operating pressure of 20%). Based on pilot data the authors estimated the cleaning interval of the MF/UF-RO train to be 6 months representing a 75% reduction in cleaning frequency compared to conventional pretreatment-RO train. Brehant et al., 2002 report a less pronounced difference in SWRO performance following membrane pretreatment and conventional pretreatment. While the authors report both trains operated with a steady pressure of 52 bar at 25 deg C during operation at flux of 7.8 gfd and recovery of 30%, the UF train operated for a longer runtime (27 days vs. 21 days) than the conventional pretreatment train. However, it was reported that reduced runtime on conventional pretreatment-RO train was due to operational problems with the conventional unit. Glukstern et al. 2002, reported similar performance of conventional pretreatment-RO and membrane-RO pilot train except during periods of stormy weather which significantly impacted the suspended solids in the feed seawater. During these upsets, it was necessary to stop flow to the RO in the conventional pretreatment train because of break through media filters causing unsafe SDI values.

2.2 Research Needs

Many issues related to the use of RO membranes for seawater desalination still need to be addressed. These include seawater characterization, evaluation of multiple membrane suppliers, evaluation/comparison of pretreatment options, information regarding the effect of foulants present in seawater on RO performance, and the identification of major cost factors related to pretreatment and their impact on overall cost of SWRO facilities.

Characterization of seawater quality is a key factor to consider in operation and design of SWRO facilities. Seawater quality can greatly impact the degree of pretreatment, type and frequency of RO membrane cleaning and the effluent water quality of SWRO facilities. An important aspect of seawater quality is its diverse nature, which is influenced by geographic location and environmental impacts. Geographically seawater can vary in composition due to such affects as currents, evaporation rates, and weather. For example, the TDS content of seawater from the Persian Gulf is typically 49,600 mg/L, while values in the Eastern and Western US have been measured between 18,000-31,000 mg/L (Maxwell, 2004) and 21,000–35,000 mg/L (Lopez, 2004b). Seawater quality can also vary at a given location due to environmental impacts such as storm water runoff, changes in weather/season variability or special flow conditions, and pollution. Such impacts can lead to increased sedimentation, nutrient and pathogenic and salinity loading. An understanding of the nature of the feedwater is required to correlate the extent of the fouling that could be expected on the SWRO membranes. An analysis of seawater with special emphasis on the membrane foulants would help identify the effectiveness of the pretreatment alternatives with regard to membrane fouling.

With the fast paced growth of the SWRO industry there has been an increase in the number of suppliers offering membranes for seawater desalination. To date the major SWRO membrane suppliers include Toyobo, Hydranautics/Nitto Denko, Toray, Dow/Filmtec, Koch/Fluid Systems and Osmonics while a relative new comer to the industry is Saehan Industries. Due to increased application of RO membranes for seawater desalination and market competition, manufacturers are continuously making advancements in membrane technology and subsequently releasing new generation membranes. Over the past several years research of SWRO has primarily focused on increased membrane productivity and system recovery in order to minimize foot print and system capital costs. To meet these goals RO suppliers have re-introduced membranes and vessels that can withstand high pressure hence allowing the systems to be operated at higher recovery and flux (Adham et al., 2003). Due to the recent increase in the number of suppliers and advancements in technology it is prudent to the SWRO industry to perform research that compares operating performance of latest generation membranes. Results from this research could assist municipalities considering SWRO to select qualified suppliers and provide a basis for operating performance to be used in performing preliminary cost assessments.

An effective method for comparing latest generation SWRO membranes from multiple suppliers is to conduct bench scale testing. Such testing is useful for evaluating rejection of organic carbon, dissolved solids and the rate of membrane fouling and scaling. An effective tool for bench scale evaluation is flat-sheet membrane testing module. This unit can serve as a predictive tool as it is capable of simulating some aspects of the flow dynamics of full-scale membrane elements in addition to providing an indication of rejection of organic matter and TDS. Experimental conditions such as feed water source, system recovery, flux, and crossflow velocity can be controlled to allow for direct comparison of performance among various suppliers.

As previously mentioned, one of the key issues related to SWRO is the implementation of proper pretreatment to optimize the performance of the downstream RO process. Currently municipalities and private water suppliers considering SWRO are faced with choosing between the use of conventional pretreatment (e.g. coagulation, flocculation sedimentation and filtration) and low-pressure membranes (e.g. MF/UF). To date a limited number of pilot studies have been performed to evaluate pretreatment options for SWRO. Filteau et al., 2003, evaluated MF as pretreatment to SWRO membranes. The authors reported MF was optimized to operate with a flux of 25-gfd and backwash frequency of 15 min. The authors also reported that chlorine added upstream of the MF system for fouling control, resulted in a significant degradation in integrity of downstream SWRO membranes. Grounds et al., 2000 performed pilot testing of conventional pretreatment to SWRO. During this evaluation coagulant dose, pre-oxidation requirements and filter design was optimized for 26 mgd Point Lisas SWRO plant.

While the above studies provided useful information regarding SWRO pretreatment, only a few studies (Henthorne et al, 2002) have focused on comparing the two options. Such a comparison could be achieved by performing a parallel study of conventional and MF/UF pretreatment at the pilot scale level. During this time, operational and water quality performance data of each pretreatment method could be collected and compared side-by-side. In addition, key operating parameters of each pretreatment method could be optimized to achieve the best effluent water quality and productivity. Results from this analysis would serve as basis performing a cost assessment of the pretreatment methods. The pilot train could then be modified to include RO membranes downstream of each pretreatment option. An evaluation of the difference in RO performance can then serve as a point of comparison when assessing the cost of each pretreatment option.

2.3 Objectives of the Study

The City of San Diego and their as needed research consultant, MWH, was awarded a cooperative agreement by the Bureau of Reclamation to evaluate the effect of pretreatment on Seawater Reverse Osmosis (SWRO) performance. The main purpose on this project was to compare the two most commonly used pretreatment schemes for SWRO - conventional pretreatment (coagulation/filtration) and membrane filtration, in terms of the water quality produced by each treatment train and the effect it has on RO performance. The specific objectives of the study were to:

- Acquire and compile current knowledge on pretreatment for Seawater RO (SWRO) using literature survey.
- Characterize the organic and inorganic constituents of seawater.
- Determine feasible operating conditions for conventional treatment parameters using bench scale testing
- Screen RO membranes for pilot scale testing using bench-scale testing.
- Conduct size fractionation experiments on seawater foulants and conduct RO fouling experiments with these size fractions.

- Perform comparative assessment of the performance of membrane pretreatment with conventional pretreatment by conducting a pilot study.
- Conduct pilot tests on RO membranes using pretreated water from two different pretreatment trains to determine relative rates of fouling
- Determine major cost factors for the two pretreatment schemes.

3. CONCLUSIONS AND RECOMMENDATIONS

3.1 Conclusions from Raw Water Characterization

- Seawater in the San Diego Bay is more susceptible to the influence of surface water runoff than open ocean water.
- The water quality in the location inside the San Diego Bay with poor flushing (South Bay) is different from that in the bay close to open ocean and has a different organic makeup with higher Specific Ultraviolet Absorbance (SUVA) content.
- Water quality of the effluent from the South Bay Power Plant was of a better quality than the influent in terms of particulate and organic loading.

3.2 Conclusions from Bench Scale testing

- The optimum ferric dose for conventional pretreatment of Shelter Island (bay location close to open ocean) water was found to be 4 mg/l $FeCl_3$ with no polymer.
- The bench scale method presented shows promise in quick evaluation of comparative RO fouling.
- Hydranautics SWC 4 membrane was selected for further pilot testing based on the RO screening tests based on comparative flux decline and rejection results.
- Bench Scale size fractionation tests indicated that MF and UF size fractions of seawater foulants can cause lower RO flux decline than conventional filtration size fraction.
- Bench Scale size fractionation tests indicate that there is minimal difference in RO fouling rates between the UF and MF fractions of seawater foulants.
- Bench Scale size fractionation tests indicate that the least amount of flux decline was caused by the seawater size fraction representing tight UF or loose NF membranes (20KDa)
- FTIR and SEM techniques were used to further elucidate fouling magnitude and mechanisms

3.3 Conclusions from Pilot Scale Testing

- MF treated water consistently had lower SDI value (less than 3) when compared to conventionally pretreated water (between 3 and 6)
- MF treated water showed consistently lower RO flux decline than conventionally treated water
- The 90 inch deep filter configuration performed better than 72 inch deep filter in terms of water quality parameters (SDI) and its effect on RO flux decline. For Shelter Island Water 90 inch deep filter RO performance was close to MF RO performance
- For South Bay water the difference between the conventional RO (90 inch deep filter configuration) and MF RO seems more pronounced. This shows that conventional

treatment is highly dependent on water quality as the South bay water can be classified as hard to treat water (high turbidity, high SUVA) by conventional treatment

3.4 Conclusions from Cost Factor Analysis

- There are several cost factors that are impacted by the pretreatment type. The cost factors impacted by the pretreatment type include: operational flux, specific flux decline rate, RO influent water quality, and membrane life.

3.5 Recommended Future Work

- Further SWRO fouling characterization and visualization tests should be conducted using the procedure developed in this study. This will help the seawater desalination community in understanding the scientific basis of organic fouling of RO membranes.
- Additional long term RO fouling tests with continuous feed of seawater and continuous pretreatment operation is recommended to further understand long term fouling.
- Testing under actual operating conditions (in contrast to the aggressive conditions utilized in this study) should be conducted to determine the frequency chemical cleaning under non- aggressive operating conditions is recommended.
- Further optimization of the membrane pretreatment by using coagulation before filtration should be conducted to determine if additional benefits can be accrued by using a coagulation microfiltration process. The effect of such a pretreatment on RO operation should be investigated through bench and pilot scale tests.
- The results from bench scale testing indicate that flux decline of SWRO membranes pretreated with UF or MF pretreatment are similar. Pilot testing should be conducted to compare MF and UF for pretreatment. Also, it should be investigated whether the RO cleaning is more effective if one pretreatment is used versus another.
- A demonstration facility should be constructed and operated to further understand scaleup and water quality issues

4. MATERIALS AND METHODS

4.1 Testing Site

The site used for the desalination pilot evaluation was the Point Loma Wastewater Treatment Plant (PLWTP) in San Diego, California. The PLWTP site overlooks the Pacific Ocean and is close to the San Diego Bay location from which seawater was extracted and transported to the pilot site throughout the study.

Pilot testing was conducted on a concrete slab located at PLWTP. The site had access to sufficient electrical power, and proper drainage lines were provided to meet the needs of all pilot equipment.

The bench scale evaluation was performed at the MWH Research Center and Fabrication Facility (RCFF) located in Monrovia, CA. Seawater from the San Diego Bay was used as source water for these tests. It was extracted and transported to the bench-scale site in a single 100-gallon (gal) tank. The water was stored at 4°C throughout the study to inhibit biological activity.

4.2 Raw Water Quality Characterization

Seawater from three different sites in the San Diego area was collected and analyzed to understand the difference in water quality between bay and open ocean water.

- Shelter Island (bay close to open ocean - pilot testing water source)
- South Bay (bay away from open ocean - proposed site for full scale plant)
- Scripps Pier (open ocean)

The locations of these three sampling sites and the pilot site are shown on a aerial photograph of the San Diego Bay of the in **Figure 4-1**

The Shelter Island site was used for most of the pilot testing runs and all the bench scale testing. This site is located inside the San Diego bay but is close to the open ocean. A picture of the Shelter Island Boat Ramp where the samples were collected is shown in **Figure 4-2.**

The South Bay site is the site of a power plant. The San Diego County Water Authority is conducting a feasibility study on siting a full scale seawater desalination plant on this site. This site is deep inside the San Diego bay and is poorly flushed. Some pilot testing and raw water characterization was conducted on water from this site. A picture of the South Bay Power Plant site where the samples were collected is shown in **Figure 4-3**.

The Scripps Pier is a research pier owned by the Scripps Institute of Oceanography (SIO) and is on the open ocean. Only raw water characterization was performed on this site. A picture of the Scripps pier site where the samples were collected is shown in **Figure 4-4**.

The seawater sources used for this study were characterized in terms of the inorganic and organic constituents of the water with special emphasis on potential foulants. A list of specific parameters analyzed during the current study was developed from results and recommendations from previous studies (El-Manharawy et. al., 2001, Dalvi et al, 2000)· These include: Temperature, Turbidity, Alkalinity, Sodium, Magnesium, Calcium, Potassium, Barium, Iron, Manganese, Strontium, Silica, Sulfate, Chloride, Phosphate, TDS, Conductivity, Boron, Ultra Violet Absorbance at 254 nm (UV 254), Total Organic Carbon (TOC), and Heterotrpophic Plate Count (HPC). Basic information regarding the sampling time, location and depth; meteorological conditions during sampling; and sample pH, conductivity and temperature was also recorded in the field.

For pilot testing a 5000 gal water truck was used to collect and transport feedwater from these sites to the testing site at the PLWTP. A floating dock strainer screen on the intake of the water truck tank was used to filter large debris from the seawater as it was pumped into the water truck. A schematic of the PLWTP site showing the location of the various equipment used during pilot testing, is provided in **Figure 4-5.**

4.3 Conventional Pretreatment Bench Scale Testing

Jar testing was conducted to determine the optimum coagulant dose. The criteria to assess performance included analysis of TOC, DOC, turbidity and other water quality parameters of the pretreated water, including Silt Density Index (SDI). The jar testing apparatus used for this test is shown in **Figure 4-6**. Initially, the effect of a sole coagulant was investigated. As mentioned in the experimental plan, ferric chloride was the selected coagulant since it is the most commonly used chemical for seawater pretreatment applications, according to the literature. Doses ranging from 1.0 to 16.0 mg/L were evaluated. Afterwards, the combined effect of coagulant and a selected polymer was studied. Three different polymers were recommended by a polymer manufacturer[1] for this particular test. Their preparation and dosing were also established according to recommendations from the manufacturer.

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[1] Nalco Chemical Company, Inc., Naperville, IL

4.3.1 Addition of coagulant only

These jar tests were performed using raw seawater at ambient temperature, no pH correction and ferric chloride as the only coagulant. Selected ferric chloride doses were estimated from literature review of several cases where water quality was similar. The jar tests were performed according to the directions presented on **Table 4-1**, regarding mixing speed, velocity gradient and time. This procedure was the standard for the rest of jar tests presented in this report.

4.3.2 Addition of Coagulant and Polymer

For these experiments, the combined effect of ferric chloride and various concentrations of different polymers were investigated. Optimized ferric chloride concentrations based on the previous tests were (section 4.3.1) were used. As in the previous case raw seawater was used at ambient temperature and without pH correction.

The manufacturer recommended three different polymers for seawater applications, but specific selection required a series of bench tests. Since the charge of particles in seawater is usually negative, the effect of, non-ionic and cationic polymers was tested. The characteristics of each polymer are presented on **Table 4-2**. Instructions for preparing stock solutions of each polymer as well as recommended dosages were also provided by the manufacturer. During preliminary control experiments, polymer concentrations below 0.5 ppm had negligible effect, while at concentrations of 2.0 mg/L and higher the result was adverse and resulted in increased TOC and DOC concentrations. For these reasons, the two polymer doses selected were 0.5 and 1.0 mg/L.

4.3.3 Effect of pH

Additional analyses were performed by including a pH correction step. Hydrochloric acid was added to reduce raw seawater pH to 7.0 and 6.5 respectively. Once this initial pH was attained, additional jar testing was performed under the same conditions used for the experiments described on section 4.3.1(coagulant-only). The purpose of these analyses was to evaluate if a pH-correction step is justifiable based on the results of water quality of the settled water.

4.3.4 Silt Density Index (SDI)

Additional information to ensure a proper selection of the optimized pretreatment conditions was obtained using Silt Density Index (SDI) measurements. SDI is widely known as a suitable indicator of the efficacy of a given pretreatment for RO, since it provides a measurement of the "filterability" of the analyzed water. SDI is a fouling index devised to determine the colloidal particle fouling potential of RO feed water. The test measures the rate of fouling of a 0.45-micron filter membrane.

The volume of water sample needed to obtain a SDI measurement was large (1-2 gallons under the current setup). Hence, it was decided to perform SDI measurements on selected jar testing parameters for selected conditions that showed comparable pretreatment results, in order to define the best conventional pretreatment conditions.

4.3.5 Alternate Coagulant Comparison

To compare ferric chloride to other commonly used coagulants for bench scale optimization, the project team conducted a jar test series to compare the performance of alternative coagulants. The intention of this test series was to benchmark the performance of $FeCl_3$ compared with other commonly used coagulants. The coagulants evaluated included Ferric Sulfate ($Fe_2(SO4)_3$), Aluminum Sulfate ($Al_2(SO4)_3$) and Poly Aluminum Chloride (PACl). This characterization was done with regard to TOC, DOC, Specific UV absorbance (SUVA) and turbidity removal when dosed at metal-molar equivalent rates. Other parameters monitored at the beginning and end of this test series include pH, temperature, alkalinity, conductivity and UV254 absorbance. All jars were dosed, according to **Table 4-3**, with the metal-molar equivalence of 4 mg/l $FeCl_3$, which was found to be the optimum dose in the original coagulant dose optimization, jar testing series.

4.4 RO Membrane Bench Scale Testing

4.4.1 Individual Membrane Testing

These evaluations used a modification of the Rapid Bench-Scale Membrane Test (RBSMT) known as the Batch Internal Recycle Membrane Test (BAIReMT), as proposed as DiGiano et. al. (1999) The BAIReMT requires a much smaller volume of test water than the RBSMT, although a similar experimental apparatus is used in both methods.

The primary piece of experimental equipment is a flat-sheet membrane-testing module called the SEPA® cell[2]. A picture of this unit is provided in **Figure 4-7**. The Sepa® Membrane Cell System is a lab-scale cross-flow membrane filtration unit that can be used to obtain performance data on membrane coupons. It can be used to simulate the flow dynamics of larger, commercially available membrane elements. The operating conditions and fluid dynamics can be varied over broad ranges. The testing module is constructed of stainless steel and is able to accommodate a 24-inch^2 flat sheet of membrane. Water is delivered to

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[2] GE Osmonics, Minnetonka, MN

the test cell using a high-pressure pump to provide the appropriate operating pressure. Permeate flows through the membrane, while concentrate flows across the membrane. Flow rates are controlled by varying the pump speeds and by adjusting the needle valves.

This unit was used to screen a range of seawater RO membranes by testing the performance of rectangular membrane coupons obtained from 2.5 inch by 40 inch or 4 inch by 40 inch elements provided by the manufacturers. The seawater RO elements that were tested were obtained from Hydranautics, Toray, Osmonics, Koch, Dow and Saehan Industries. These manufacturers provided the research team with their most recently developed seawater RO membrane that is commercially available. Membrane characteristics of the membrane elements used are presented in **Table 4-4**.

The BAIReMT configuration requires permeate and concentrate to be returned to the feed tank. In addition, a fraction of concentrate will be recycled to the RO system, in order to increase overall system recovery. A schematic of the experimental setup is shown in **Figure 4-8**.

The membranes were operated at a flux of 12 gfd , a system recovery of 50% and a crossflow velocity of 1 ft/sec.

4.4.2 Seawater Foulant Fractionation RO experiments

The purpose of seawater foulant fractionation was to provide batches of feedwater for RO fouling studies that have been pre-filtered through membranes of different pore sizes. Fouling experiments using RO membranes and these fractions of the foulants present in seawater could provide clues on particle and dissolved matter sizes, in seawater from San Diego Bay, can be expected to contribute most significantly to RO fouling and flux decline.

With the concentrate outlet plugged, the SEPA® cell was operated as a dead-end filtration module during fractionation. Furthermore, a low pressure gear pump was used in place of the high pressure piston pump so that the cell was operated in the low pressure regime of the micro and ultrafiltration membranes that were used for fractionation.

Upon arrival at the lab, all feedwater was filtered at the 1.0μm level by a glass fiber filter. Ten gallons of this feedwater was then filtered through a 0.1μm PTFE cartridge filter to produce the 0.1μm fraction. Another 10 gallons was filtered through a 100 kilo Dalton (kDa) ultrafiltration membrane in the SEPA® cell, in dead end mode, to produce the 100kDa fraction, and so on. All fractionated seawater was collected in ten-gallon batches, and stored in a dark refrigerator at 4°C until use to inhibit biological activity. **Table 4-5** lists the fractions that were produced and the membranes used to create each fraction.

All membranes, and the cartridge filter, were obtained from the same manufacturer[3]. Prior to each fractionation, DI water was filtered through the membrane coupon for 15 minutes to flush the membrane of preservative.

Each of the fractions was sampled for TOC and UV254 and used to estimate Specific UV absorption (SUVA) an indicator of the humic composition of dissolved organics in each fraction.

4.4.3 Fractionated SWRO Fouling Experiments

The purpose of the fouling experiments was to foul RO membranes with each of the seawater fractions prepared in section 4.4.1 and to monitor both the rate and extent of fouling attributable to each fraction. The experimental procedure for operation in the BAIReMT configuration was identical to the procedure described in section 4.4.1. The intent of operating in this configuration is to concentrate potential foulants and dissolved salts to mimic higher recovery RO modules without a large feedwater batch.

Each fouling experiment lasted approximately eight hours; the conductivity in the inner recycle loop was measured every hour. Additionally, the following list of parameters was measured periodically, and used to calculate the specific flux through the RO membrane; the specific flux was, in turn, used to create flux decline curves:

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[3] GE Osmonics, Minnetonka, MN

- Run hours
- Feedwater Pressure
- Permeate Flowrate
- Waste Flowrate
- Recycle Flowrate
- Feed Flowrate
- Recycle Loop Conductivity
- Permeate Conductivity
- Recycle Loop Temperature

The membranes coupons used in the fractionated SWRO experiments were cut from the Hydranautics SWC4-4040 membrane which, in bench scale testing was in the top three performing manufacturers in terms of specific flux, salt rejection and flux decline, among the RO membranes tested. Hydranautics was willing to support the study by providing membranes as in-kind contribution to the project. When not in use the membrane sections were stored in DI water at 4°C to prevent microbiological growth.

4.4.4 Fractionated SWRO Cleaning Experiments

The intention of the cleaning experiments was to predict the reversibility of RO fouling attributable to each seawater fraction. Two fouling runs were conducted for each seawater fraction. The fouled membrane in the second run from each set was used in cleaning experiments. Following the eight hour, fractionated SWRO fouling test, the SEPA cell was flushed with DI water, the permeate port was plugged and the feed and concentrate lines were connected in cycle with a cleaning solution reservoir; this cleaning solution was rinsed over the membrane surface for forty five minutes at very low pressure (driven by a peristaltic pump) to clean the membrane surface.

The cleaning solution for removal of adsorbed organics and colloidal silica, as recommended by Hydranautics was 0.1% w/w NaOH with pH 11.5. Additional cleaning experiments incorporated 2% w/w Citric Acid solutions adjusted to pH 4.0 by NH_4OH for the removal of deposited metal ions.

4.4.5 Fouled Membrane Autopsies

Autopsies of fouled membrane surfaces were conducted to further evaluate the physical and chemical nature of seawater foulants in each fraction. SEM images were taken of foulant layers to better understand cake structure, and FTIR spectral analysis was conducted on fouled membrane surfaces to describe the functional chemistry of important RO membrane foulants in each of the San Diego Bay seawater fractions.

After the completion of each fouling and cleaning experiment, membrane coupons were removed carefully from the SEPA® cell and stored in a Ziploc bag with DI water at or below 4°C until a fouled membrane autopsy including SEM imaging, and FTIR analysis of foulants, could be conducted. All samples were shipped, on ice, to University of New Mexico (UNM) for the autopsies.

4.5 Pilot Testing Setup

The pilot testing was conducted using two treatment trains: one membrane pretreatment train followed by RO and one conventional pretreatment train followed by RO. A schematic illustrating these treatment trains is shown in **Figure 4-9**.

4.5.1 Conventional Pretreatment Train

The conventional pretreatment consisted of a pilot coagulation/sedimentation unit followed by 2 conventional media filters. These pilot units are shown in **Figures 4-10 and 4-11**. The optimum ferric iron and polymer doses were determined from bench testing results. During some testing periods this dose had to be increased. The filter units used during the testing were dual media filters utilizing sand and anthracite. Two different filter configurations were evaluated during the testing. The first configuration evaluated was 72 inches deep with a 24 inch anthracite bed over 48 inch sand bed. The second configuration was 90 inch deep with a 60 inch anthracite bed over a 30 inch sand bed. The second configuration has been tested before in a recent study and has been found to provide consistently good quality pretreatment (MWH, 2000). This configuration was the basis for design of the 35 mgd Point Lisas SWRO plant in Trinidad.

The coagulation/sedimentation pilot unit was operated to provide a flow rate of approximately 0.7 gpm. The loading rate on the media filters was set at 6 gpm/sqft to obtain the best possible water quality out of the conventional treatment train. Twenty five total filter runs were conducted using the conventional setup. The important coagulation and filter design parameters used during these runs is provided in **Table 4-6**.

4.5.2 MF Pretreatment Train

The MF pretreatment train consisted of a disk prefiltration system[4] followed by a MF system. The prefiltration system consists of a set of grooved disks stacked on each other on a spine assembly. These disks are color coded by filtration size. The depth of the grooves determine the nominal filtration size. The spine assembly has a spring compression and internal piston, which are used to alternately compress and release the disks during the filtration and backflush cycles. The backwash operates on a automatic backwash cycle, which was activated once daily during the testing. The filtration is nominally rated at 130 micron and was used to protect the MF system from debris. This system is shown in **Figure 4-12.**

The membrane filtration unit used was a MF unit (CMF –S submerged) supplied by US Filter. The US Filter pilot system included the following components:

- Filtrate / backwash pump
- Filtrate storage tank
- Touch-screen user interface
- Automated operation, filtration, backwash and maintenance clean
- Chemical addition systems for both sodium hypochlorite and acid
- Air compressor for air scour during backwash

The US Filter CMF-S pilot is skid-mounted unit. Photographs of this unit are provided in **Figure 4-13**. The photographs show the square stainless steel tank that contains the 4 membrane modules. In the unit tested, only two of the modules were active. The other two were dummy modules which had the fiber ends sealed. The “clover” of four modules is connected to a common header at the top of the membrane tank. The bottoms of the active modules are connected to aeration caps. With a nominal pore size of 0.1 micron, the US Filter polyvinylidene fluoride (PVDF) membranes remove particulate material, including protozoa, bacteria and some virus. The electrical panel, with LCD display is shown on the photograph on the left. right. The system is completely automated. All system-operating parameters are set using a LCD touch screen user interface with control buttons located at the bottom of the screen. The system flows, pressures and temperatures are displayed on the LCD screen and stored to a database every minute. This data can be downloaded to diskette.

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[4] Spin Klin Filtration System, Arkal Filtration, Israel

The US Filter PVDF membrane is constructed of a hydrophilic, nonionic polyvinylidene fluoride polymer. The membranes utilize outside/in hollow fibers, thus the flow of water is from the outside to the inside of the hollow fiber. In the pilot system tested, two membrane elements were placed in opposite corners of the membrane header within the process tank. Each cylindrical element is approximately 5 inches in diameter and 47 inches long. Each of the two active membrane elements contains 9,600 hollow fibers with an active fiber length of 41.3 inches. The outside-fiber diameter based surface area for each element is 272.3 ft^2. The outer diameter of the hollow fibers is 0.8 mm and the fiber inner diameter is 0.5 mm. The US Filter PVDF membrane is chlorine tolerant. **Table 4-7** summarizes the specification of the US Filter PVDF membrane.

The operating flux of the system was maintained at 30 gallons per square feet per day (gfd) with a backwash frequency of 30 minutes. Two out of the four modules used on the unit were blank modules resulting in a flow of 11 gpm. Blank modules were used to reduce the amount of water that will be needed for the testing and is not expected to impact the results. No Chlorine was used during the testing except for cleaning. The operation parameters that were used for the unit are provided in **Table 4-8.**

4.5.3 RO Trains

The RO pilot was configured in such a way that two totally independent RO trains could be operated on the system. Each RO train consisted of two pressure vessels that hold three 4" by 40" RO elements each. The specifications for this pilot are provided in **Table 4-9** and the pilot is shown in **Figure 4-14**. Each RO train was operated at fluxes of 10 and 14 gfd and recovery values between 25 - 40 %. The RO membrane used for pilot testing was selected based on the bench-scale testing results. This membrane was the SWC 4 membrane manufactured by Hydranautics. Specific membrane characteristics for this membrane are presented in **Table 4-10.**

4.5.4 Operation Schedule

The operation of the pilot was conducted in two phases. In the first phase, which is referred to as the pretreatment evaluation phase, the focus was on operating the pretreatment trains to evaluate the effect of pretreatment on water quality parameters. In the second phase, referred to as RO Operation, the RO systems were operated on the batch of water produced by the pretreatment trains. During data analysis some pretreatment operational and water quality data from the RO operation phase is presented along with the Pretreatment Evaluation Data to present a complete picture of pretreatment process performance.

This project was initially planned with a direct seawater intake. However, during the course of the project planning several difficulties were faced while attempting to secure a site with direct intakes. The most difficult of these was obtaining permits to draw seawater directly from the ocean. In response to this the team decided to conduct the pilot testing under semi batch conditions as elaborated in the sections below.

Pretreatment Evaluation (Phase I)

This initial evaluation phase was conducted during the first six months of pilot testing. During this phase, only the pretreatment trains were operated. This phase was focused on obtaining water quality data from the pretreatment trains.

The MF unit was operated using recycle of backwash and permeate back into to the feed tank. On the other hand, conventional pretreatment train was run without recycle, at a flow rate between 0.7-0.8 gpm, until the batch of water was used up. The feed tank was replenished on a frequent basis to prevent exhaustion, and provided a continuous 2-4 day run. A schematic of this setup is shown in **Figure 4-15.**

RO Operation (Phase II)

This phase of pilot testing was conducted for a period of 3 months. The feed tank for the pretreatment trains were replenished at a minimum of once every two weeks.

During this phase the MF pilot was operated to fill up a 4,900 gal break tank. This tank then feed a pilot scale seawater RO system with six 4"x 40" elements at the rate of 10-12 gpm. The concentrate and permeate from this unit was recycled to the 4,900 gal RO feed tank to enable continuous operation.

The conventional pretreatment train had a pilot scale conventional treatment train operating in the range of 0.6 gpm. This pilot was operated to fill up a 4,900 gal break tank. Water from this tank was then used to feed a pilot scale seawater RO system with six 4"x 40" elements at the rate of 10-12gpm. The concentrate and permeate from this unit was recycled to the break tank to enable continuous operation.

Three runs were completed with each RO train. The parameters for these runs are shown in **Table 4-11.**

4.5.5 Determination of Calculated Parameters

The following section illustrates how common operating parameters were determined throughout the testing period for the RO membrane.

Calculated Feed and Permeate Salinity (TDS)

$TDS_f = C_f/1.52$

$TDS_p = C_p/2.02$

Where,

TDS_f = Feed TDS
TDS_p = Permeate TDS
C_f = Feed Conductivity
C_p = Permeate Conductivity

Here the relationship between TDS and conductivity was determined using data from bench scale testing. Conductivity and TDS data will be collected throughout the testing to refine this relationship.

Feed Osmotic pressure (πf)

$$\pi_f = 0.0115 * (TDS_f) = 0.0115 * (C_f/1.52)$$

This calculation is based on the approximate rule of thumb of 11.5 psi osmotic pressure for 1000 mg/l NaCl. We will use this relationship

Recovery (Y)

Recovery $Y = Q_p/Q_f$
Where, Q_p = Permeate Flow
Q_f = Feed Flow

Integrating Average Factor (I.A.F)

This term denotes the averaging factor on the concentrate/feed side of the membrane depending on recovery.

$$IAF = (\ln(1/(1-Y)))/Y$$

Where, y is recovery expressed as fraction and rejection is assumed to be 100%

Average Osmotic Pressure

On feed/concentrate side (π_{fbl})

$$\pi_{fbl} = IAF * \pi_f$$

On Permeate side (π_p)

$$\pi_p = \pi_{fbl} * C_p/C_f$$

Osmotic pressure difference between feed and permeate ($\Delta\pi$)

$$\Delta\pi = \pi_{fbl} - \pi_p$$

Net Operating pressure (Net Driving Pressure- NDP)

$$NDP = (Pf+Pc)/2 - \Delta\pi - P_p$$

Where,
P_f = Feed Pressure
P_c = Concentrate Pressure
P_p = Permeate Pressure

Trans membrane Pressure (TMP)

$$TMP = (Pf+Pc)/2 - P_p$$

Flux (J)

$J = Q_p/A$
Where, Q_p = Permeate Flow (gpm)
A = Membrane Area

Temperature correction factor (TCF)

With increasing temperature the flux increases because of viscosity changes in the solution and changes in the membrane area. This factor is usually specified by the membrane manufacturer and can be used to normalize flux.

Flux at 25 ^{0}C (J25C)

$$J_{25C} = J \exp(-TCF^* [1/298 + 1/(^0C + 273)])$$

Specific Flux at 25^{0}C

$$J_{sp25c} = J_{25C}/A$$

Rejection

$$R = 1 - (TDSp/TDS_f)$$

4.5.6 Chemical Cleaning of Membranes

All MF and RO chemical cleanings were performed in accordance to the manufacturers recommended protocol. These protocols are provided in **Appendix B**.

US Filter membranes were cleaned in place (CIP) by soaking and circulating citric acid followed by soaking and circulating chlorine.

The RO membranes were cleaned using 0.1 % sodium hydroxide. The chemical solution was mixed using RO permeate in an external cleaning skid which consisted of a 100 gal chemical tank, a heating element and a centrifugal pump. The solution was recycled through the RO concentrate line back to the membrane cleaning tank at a rate of 4-6 gpm for 1 hour. Next, the membranes were allowed to soak for 1 hour. Finally, the cleaning solution was completely drained from the membranes and the system was brought back on-line.

After the first RO run in Phase II manufacturers recommended cleaning procedures did not yield adequate recovery of the RO membrane specific flux. A special cleaning protocol recommended by a speciality cleaning chemical manufacturer[5] was used after Run 1. This procedure is also summarized in **Appendix B.**

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[5] Diamite BFT, King Lee Technologies, San Diego, CA

4.6 Water Quality Monitoring

4.6.1 On-site water quality analyses

Temperature

The temperature of the conventional system feed and filtrate as well as the RO feed was measured using the temperature probe of a conductivity meter[6]. The probe temperature was periodically verified using NIST verified thermometer[7]. The temperature of the MF feed was measured using an inline temperature probe.

Turbidity

The turbidity of the raw water was measured using a lab turbidimeter[8]. The turbidity of the MF feed and filtrate was measured using on-line turbidimeters[9]. The turbidity of the conventional filter effluents were measured using another set of on-line turbidimeters[10]. On-line measurements were periodically verified using the lab turbidimeter.

Conductivity

The raw water, filtered water (conventional and MF) and RO permeate water conductivities were measured using a portable conductivity meter[1]. On-line conductivity of the RO influent and effluent was also monitored using on-line conductivity meter[11]. Measured values were compared with daily conductivity results from the laboratory to ensure continued accuracy.

pH

The pH of the raw water and filtered water (conventional and MF) was measured using a portable pH meter[12]. During chemical cleaning of RO membranes this pH meter was used for adjusting pH.

Dissolved Oxygen

A portable DO probe[13] was used for measuring onsite Dissolved Oxygen during sampling events.

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[6] Hach SensION5, Loveland, CO

[7] ERTCO, Dubuque, IA

[8] Hach Co, Model 2100 N, Loveland, CO

[9] Hach Co., Model 1720D, Loveland, CO

[10] Hach Co., Model 1720 C, Loveland, CO

[11] GLI C53 Contacting Conductivity Analyzer, Hach Co., Loveland, CO

[12] Oakton® Acorn pH 6 pH meter, Vernon Hills, IL

[13] YSI Model 55, Yellow Springs, OH

UV-254 Absorbance

Samples collected for TOC analysis were also analyzed for UV-254 absorbance using a spectrophotmeter[14].

Silt Density Index (SDI)

Silt density index (SDI) analyses were performed on the raw water, filtered water and RO feed using a SDI testing set up. This setup was constructed on site using a pressure regulator and pump. The SDI setup filtered water through a disposable 0.45-µm filter at pressure of 30 psi. The SDI value was determined by continuous monitoring of the flow rate at a constant pressure over a 15-minute period.

Off-site Water Quality Analyses

All off-site water quality analysis were performed at one of the following locations: Point Loma laboratory (PL Lab), MWH Labs, Calscience Environmental Laboratories (CEL Lab) and the Marine Micro Lab. **Table 4-12** summarizes the detection limits and methods used for all of the laboratory analyses that were performed.

Sampling Protocol/Frequency

All water quality samples were collected as grab samples using sample containers provided from the corresponding laboratory. All samples were transported to the lab in a cooler and were processed within the allowable holding period. During sampling, sample ports were allowed to flush before samples were collected. All microbial samples were collected using aseptic technique. The sample ports were flamed and flushed before a sample was collected.

Quality Assurance/Quality Control

Appropriate measures were taken at the pilot site in order to attain the highest amount of quality control and quality assurance. **Appendix C** contains a technical memorandum documenting the QA/QC that was performed throughout the study.

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[14] Hach Co., DR/4000U spectrophotometer, Loveland, CO

5. RESULTS AND DISCUSSION: RAW WATER CHARACTERIZATION AND BENCH SCALE TESTING

5.1 Raw Water Characterization

Tables 5-1 and 5-2 provide a summary of the water quality characterization conducted at the three sites in San Diego. Two sampling events were conducted at these three sites. One sampling event was conducted on 7/21/04 and the other on 11/04/04. The second sampling event was conducted two weeks after a rain event had passed. It can be clearly seen from the water quality data that the greatest variation in water quality was at the South Bay site. This was expected as this site is within the San Diego Bay and is poorly flushed. Hence, the effect of rain events can be clearly seen even after two weeks of a rain event. During each sampling event the TOC and UV 254 are higher at the South Bay site. Also the turbidity at the South Bay site during the second event was very high.

Samples were also collected from the influent and effluent of the South Bay power plant to compare the water quality from these sampling locations. These results presented in **Table 5-3**. This sampling was conducted on 11/04/04 two weeks after a storm event. It can be clearly seen that the water quality coming into the plant is higher in turbidity and organics than the effluent. A high concentration of iron was also measured coming into the South Bay power plant cooling system influent. However, no iron was detected in the effluent. The effluent is higher in temperature as expected as the water is being used for cooling purposes by the South Bay power plant. Considering all water quality parameters it seems like the power plant effluent is more appropriate as a SWRO plant feed source than the influent.

Some of the parameters of interest to SWRO treatment are discussed individually below

5.1.1 TDS

TDS is a direct measure of salinity of the seawater and has an impact on the operational pressure of the process as applied pressure required to maintain a certain production is directly related to the osmotic pressure of the feedwater. It can be seen from **Table 5-1** that the most variation in TDS between the two dates was at the South Bay site. At the other sites the TDS was relatively stable. This was expected as this site is within the San Diego Bay and is poorly flushed. A drastic change in TDS during operation of a SWRO plant could necessitate changes in operating conditions and care during design to be able to accommodate the range of applied pressures needed.

5.1.2 *Iron*

Iron is an important constituent to consider as dissolved iron can pass through the pretreatment process and precipitate on RO membranes causing fouling. It could also promote iron bacteria growth. This was only found in significant quantities in the South Bay Power Plant influent water after the rain event. It is also seen that this was not seen in the South Bay Power Plant effluent indicating that it gets removed while passing through the cooling system of the plant

5.1.3 *Manganese*

Similar to iron, dissolved Manganese can pass through the pretreatment process and precipitate on RO membranes causing fouling. It was only found during one sampling even at the South Bay Influent site.

5.1.4 *Silica*

Silica is important for RO operations as a high concentration of silica in the RO feedwater could lead to its getting concentrated in the concentrate stream and lead to a polymerization reaction on the feed side of the RO membrane causing blinding of the RO membrane. This occurs when silica concentration reaches around 120 mg/L. The silica concentrations at all the three sites were relatively low

5.1.5 *Boron*

Boron is an important consideration while designing SWRO plants. It is a naturally occurring element found in seawater. However, it is currently under discussion for regulation by the EPA. The California Department of Health Services (CDHS) has set an action level for Boron in drinking water at 1.0 mg/L, which is also the standard of the European Union (EU). The range of Boron values found that the three sites ranged from 5 mg/L to below the detection limit (0.5 mg/L).

5.1.6 *Temperature*

The feed water temperature has a substantial impact on RO operation. At higher temperatures membranes can operate at lower pressures for the same flux rate. This is very significant when we consider the seawater intake location at a SWRO plant collocated with a Power Plant. It can be seen from **Table 5-2** that the temperature of the South Bay Power Plant effluent is higher than the influent by approximately 6 deg C. Construction of a SWRO plant that draws its water from the effluent side would hence be a better option at this site. However, the upper limit of temperature for RO membranes (45 deg C for SWC 4 membrane) should also be kept in mind.

5.1.7 UV-254 and TOC

Both these factors represent the organic content and characteristics of the seawater. It can be seen that the TOC was higher at the South Bay site for both the sampling events. Also the UV 254 was substantially higher at the South bay site when compared to the other sites.

5.1.8 Turbidity

Turbidity has a direct impact on the performance of the pretreatment process especially the conventional filtration process shortening runs and leading to early breakthrough of the filters. It was seen that the Scripps Pier site (open ocean) had the lowest turbidity of the three sites while the South Bay site (bay away from open ocean) had the highest. At the South Bay power plant influent the turbidity was very high after the rain event (56.4 NTU) but it went down in the effluent (8.2 NTU). This indicates that substantial settling occurs in the cooling system or the influent or effluent channels of the power plant and hence the power plant effluent would be better raw water source for a SWRO plant than the influent.

5.1.9 HPC

HPC is a indicator of the biological quality of the water. A higher biological load would lead to biological growth in the pretreatment systems necessitating higher maintenance costs. This could also impact the RO system if the biological growth is allowed to proliferate. From **Table 5-2** it can be seen that the open ocean and the bay close to the open ocean had similar biological quality while the bay location away from the open ocean had lower biological water quality as expected. Between the influent and the effluent from the power plant the effluent had slightly higher HPC counts probably because of the higher temperature.

5.2 Conventional Pretreatment Bench Scale Testing Results

5.2.1 Addition of coagulant only

These jar tests were performed using raw seawater at ambient temperature, no pH correction and ferric chloride as the only coagulant. The jar tests were performed according to the directions presented on **Table 4-1**, regarding mixing speed, velocity gradient and time.

Water quality results on the settled water from this first set of jar tests are presented on **Figure 5-1**. The TOC and DOC concentrations are normalized to the initial TOC and DOC concentrations respectively. The decrease in TOC and DOC concentrations with dose is almost identical. The results presented in **Figure 5-1** shows that TOC and DOC decreased continuously with increasing $FeCl_3$ doses. However it can be seen that the point of diminishing returns is reached at about 4-6 mg/L. Doses higher than 4-6 mg/L produce insignificant decreases in TOC and DOC. On the other hand, turbidity decrease is steady up to $FeCl_3$ doses of 2 mg/L. Higher $FeCl_3$ doses produced no significant contribution to turbidity removal and doses higher than 10 mg/L were detrimental to the process. From these results, a ferric chloride dose of 4 mg/L was selected as an appropriate pretreatment. This selection was later confirmed by the results of SDI tests.

5.2.2 Addition of Coagulant and Polymer

A ferric dose of 4 mg/L was used for these tests .The combined effects of ferric chloride and different polymers at various concentrations of were investigated. As in the previous case, raw seawater was used at ambient temperature and without pH correction. The results of these experiments are presented in **Figure 5-2**.

Figure 5-2 shows that addition of a polymer with no charge (8170) slightly increases DOC removal but has a detrimental effect on turbidity removal. In addition, this negative effect increased at higher polymer doses. In all cases turbidity values increased after addition of any amount of this particular kind of polymer. It can be seen from **Figure 5-2** that there was a slight increase in DOC and turbidity removal removal when adding a cationic polymer (7128).

The results indicated that a neutrally charged polymer (8170) is not beneficial for this particular process. On the other hand, a cationic polymer (7128) slightly improved turbidity removal and DOC removal but at a lower extent. However, the benefits of adding this polymer were not enough to justify implementing a polymer-addition step into the pretreatment process. Therefore polymer addition was not used and $FeCl_3$ was the sole coagulating agent in the pilot testing phase.

5.2.3 Effect of pH

Additional tests were conducted by including a pH correction step. Hydrochloric acid was added to reduce raw seawater pH to 7.0 and 6.5 respectively. The purpose of these analyses was to evaluate if a pH-correction step is justifiable, based on the results of water quality of the settled water.

Results from these experiments are presented in **Figures 5-3 and 5-4,** respectively. When these results were plotted comparatively as a function of pH, the results showed that pH-correction steps did provide slight improvement in turbidity removal and DOC removal especially when the pH was reduced from a pH of about 8.0 to 7.0. At the selected dose of 4 mg/L the turbidity removal increase is about 6 % and the DOC removal improvement is about 5 %. However, there was no additional benefit when pH was further reduced to 6.5. pH adjustment was however, not used during pilot testing as the improvement is relatively small. pH adjustment could be considered in future testing to further optimize conventional pretreatment.

5.2.4 Silt Density Index (SDI)

Additional information to ensure a proper selection of the optimized pretreatment conditions was obtained using Silt Density Index (SDI) measurements. SDI is widely known as a suitable indicator of the efficiency of a given pretreatment for RO, since it provides a measurement of the "filterability" of the analyzed water.

Under the current experimental setup, SDI tests required water to be prepared in advance according to the selected pretreatment conditions. For this reason, this analysis was performed only for selected conditions that showed comparable pretreatment results, in order to define the best treatment option. These results are presented in **Table 5-4** and **Figure 5-5.** This clearly shows that a ferric dose of 4 mg/L with no polymer addition will produce the best water for RO treatment. It has to be noted that the SDI results might be affected by polymer carryover for the results with the polymer however this could be possible in actual plant operation where polymer could carryover to the RO membrane.

5.2.5 Alternate coagulant testing

Prior to jar testing the feedwater was sampled for pertinent water quality parameters. After rapid injection of concentrated coagulant solutions in each of the jars and agitation according to the speeds and intervals shown in **Table 4-1**, grab samples were taken from each jar and re-tested for the quality parameters.

The general water quality results for the raw water dosed with the coagulants are provided in **Table 5-5**. These general water quality results are all within expectations, the addition of coagulants in every case lowered alkalinity slightly, lowered pH and slightly increased conductivity. The most significant result here is the turbidity reduction between the jar with no coagulants and the jars with chemical coagulants; all coagulants performed similarly on this basis.

Table 5-6 compares the final organic content in each jar dosed with chemical coagulants. TOC and DOC analysis were conducted by Calscience Environmental Labs, and UV254 absorption was conducted. The source water for jar testing was San Diego Bay water collected at Shelter Island; no filtration was employed. All coagulants were dosed at the metal molar equivalent of *4mg/l* FeCl3.

According to these results we might expect the addition of any of these chemical coagulants to perform similarly for overall organics removal. In terms of actual TOC and DOC removal $FeCl_3$ did not perform as well as the other coagulants. However in terms of SUVA reduction

$FeCl_3$ performed slightly better than the other coagulants. $FeCl_3$ coagulated water had the lowest value of SUVA which is said to be a relative measure of aromatic content of the dissolved organics in the water and higher values of SUVA have been generally associated with greater tendency for membrane fouling (EPA Membrane Filtration Guidance Manual, 2003).

5.3 RO membrane bench scale testing

5.3.1 Individual Membrane Testing

Results from the experiments using the six membranes are presented in **Figures 5-6** through **5-8**, and **Table 5-7**. **Figure 5-6** shows the Net Operating Pressure increase during the eight hours of testing. The initial range of net operating pressures is between 253 and 294 psi. This figure cannot be used for comparison purposes as the impact of small variations in flux and temperature has not been considered even though it was attempted to maintain the flux at 14 gfd. It shows the range of variations in net operating pressure increases for each membrane.

Figure 5-7 shows the specific flux decline normalized at 25°C for the membranes tested. The normalized specific flux values for each membrane are identified using the first letter of each manufacturer's name at the right hand side of the Figure. A similar presentation of the rejection values obtained for each membrane is shown in **Figure 5-8**. It can be seen that the top three membranes in terms of the lowest specific flux decline and highest rejection include Hydranautics, Koch and Dow.

Rejection data of specific constituents from samples collected at the end of each run are shown in **Table 5-7**. Rejection data on Dow and Koch membranes were not available for the specific ions but they are expected to be similar to the Hydranutics membrane based on salt rejection data. It can be seen from the data in bold that the Hydranautics SWC 4 membrane had the highest values of rejection for most constituents. The high flux decline rates seen are expected considering that the system is working at high flux conditions (12 gfd). As a result of this testing Hydranautics SWC 4 was selected as the membrane to be tested at the pilot scale.

5.3.2 Seawater Foulant Fractionation Experiments

5.3.2.1 Fractionated SWRO Fouling Experiments

Table 5-8 presents TOC and UV 254 results on each seawater fraction. Based on these parameters, it seems that fractionation to the levels used in this study does not have a significant effect on the organic matter composition.

All the SWRO fouling experiments on the fractionated seawater foulants were conducted using different sections of the SWC 4 membrane obtained from Hydranautics. **Figure 5-9** presents normalized specific flux decline results for the fractionated RO fouling experiments. It plots the normalized flux decline at each fractionation level, all data is normalized to the initial specific flux. The initial specific flux ranges from 0.043 gfd/psi to 0.053 gfd/psi, these variations are within expectations. Manufacturing inconsistencies yield different clean flux values for different sections of a membrane. These results show that the tighter the fractionation level the lower the net flux declines. According to this data the 0.1μm (representing MF membrane) prefiltration shows an improvement over the 1 μm prefiltration (representing conventional filtration). There appears to no measurable improvement between the 0.1um (representing MF) and 100kDa fractions (representing UF membrane). However, when the prefiltration is further reduces to 20kDa (representing tight UF or NF) further gains in reducing flux decline are obtained.

Based on these results we can expect that MF pretreatment to be more effective in reducing RO fouling than conventional filtration. However, UF pretreatment (of 100 kDa cutoff) may not provide any further benefit. The tight UF/loose NF membranes might be the most effective in reducing fouling but will operate at higher pressures. Coagulation before the microfiltration or ultrafiltration membrane was not a part of the current study. However such a strategy could also target organic matter not removed by any specific prefiltration strategy and could provide additional flux decline mitigation.

5.3.2.2 *Fractionated SWRO Cleaning Results*

Figure 5-10 depicts typical results of the cleaning experiments for each fractionated source. The first bar in each pair is the normalized specific flux of the RO membrane (SWC 4, Hydranautics) at the end of the eight hour fouling experiment with the associated source water fraction. The second bar in each pair is the flux of the same membrane coupon after being cleaned by NaOH solution.

The flux recoverability is highest for the 1.0um fraction. At the 0.1um fraction there was less flux recovery. The flux recoverability is similar for both the 100kDa and 20kDa fractions which suggests there is some fouling component present in both fractions which is not removed by the NaOH cleaning performed as a part of these tests. Only one set of cleaning experiments ~~were~~ was conducted for each fraction.

5.3.2.3 *FTIR Analysis of Fouled and Cleaned membranes*

After the fouling or cleaning experiments were completed, membrane coupons were removed from the SEPA cell and stored, refrigerated in deionized water until analysis. For preparing the membrane for FTIR analysis one inch square sections of clean membrane were cut out of a fresh coupon and transferred into a petri dish for drying. The membrane section was dried in a 150 degree F oven for four hours prior to FTIR reading.

The FTIR spectrum from an unused thin-film composite (TFC) membrane is shown in **Figure 5-11**. The spectrum shows a broad region of absorption between 3600 and 3000 cm-1, which is due to stretching of the O-H bond in hydroxyl functional groups, and sharper peaks at 2916 and 2848 cm-1, which are due to stretching of C-H bonds. A region of strong absorption exists between 1800 and 650 cm-1, containing many sharp peaks. The number of peaks in this area suggests that the TFC membrane contains a variety of functional groups. Carboxylates, amides, and carbonyl functional groups from aldehydes or ketones can be identified from these peaks.

The FTIR spectrum of the cleaning solution is shown in **Figure 5-12**. The strong absorption of the O-H stretching from water is evident in the spectrum. The presence of some organic compounds in the cleaning solution is also evident because of the C-H bond stretching shown at 2916 and 2848 cm-1.

A fouled membrane that had been used to filter seawater that was prefiltered through a 1.0 μm filter is shown in **Figure 5-13**. To facilitate comparison with the spectrum of the unused TFC membrane, the TFC spectrum has been adjusted to have the same baseline absorption between 2800 and 2000 cm-1, and scaled vertically so that the 1242 cm-1 peak is nearly identical height in the two spectra. A number of similarities are noted between the spectra. Many of the sharp absorption peaks between 1700 and 650 cm-1 in the unused membrane are also present on the fouled membrane. Most notable are the peaks at 1585, 1489, 1240, 1107, and 1012 cm-1; the doublet at 1319 and 1292 cm-1; the doublet at 1166 and 1151 cm-1; the doublet at 715 and 690 cm-1; and the triplet at 873, 852, and 833 cm-1. The strong correspondence between these peaks suggests that the TFC membrane material can be detected through the fouled layer. As the infrared light wave can only penetrate about 1 μm into the sample, the fouled layer appears to be less than about 1 μm thick (after the membrane has been dried). However, a number of distinct differences can also be observed in these spectra, including several regions of absorption on the fouled membrane that are not present on the unused membrane. These regions of stronger absorption include a region of sharp absorption between 4000 and 3500 cm^{-1} and a doublet at 2360 and 2343 cm^{-1}. The sharp peaks in the region of 4000 and 3500 cm^{-1} are characteristic of water vapor. Nitrogen- and silica-containing functional groups absorb infrared light in this region, and may be masked by the spectra of water vapor. The doublet at 2360 and 2343 cm^{-1} is due to carbon dioxide from the atmosphere adsorbed onto the sample surface. In addition, several peaks of higher absorption are observed between 1558 and 1488 cm^{-1}. These peaks may be due to ring stretching of aromatic organics indicating fouling by these organics.

Regions of weaker absorption on the fouled membrane are also observed. The O-H band between 3600 and 3000 cm^{-1} is absent on the fouled membrane. In addition, a number of bands in lower wavenumbers, most notably between 1454 and 1262 cm^{-1}, between 1093 and 883 cm^{-1}, and between 823 and 696 cm^{-1} have lower absorption than the unused TFC membrane. The TFC membrane material had weaker absorption in these areas, and the foulant covering the membrane prevents the absorption in these weaker absorbing areas, causing a reduction of absorption. Overall, it appears that the membrane is covered with material that does not absorb light strongly in the infrared region.

The spectra produced by prefiltration experiments at 0.1 μm, 100 kDa, and 20 kDa are shown in **Figure 5-14**. When compared to **Figure 5-13**, absorption in several areas is subdued relative to other peaks when compared to the membrane fouled with 1.0 μm prefiltered water. This comparison of relative peak heights suggests that the 0.1 μm experiment deposited less material than the 1.0 μm experiment. The relative peak heights appear identical between the spectra for the 0.1 micron and the 100 KDa fraction fouled membranes, suggesting that the smaller prefiltration experiments did not remove any foulants that were not removed by the 0.1 μm experiment. This conclusion is corroborated by the flux performance in the 0.1 μm and 100 kDa, fractions which are nearly identical (**Figure 5-9**). However, the 20 kDa fraction has the lowest flux decline of all the fractions and corresponds with the lowest peak heights corresponding to this fraction in the comparative FTIR spectra.

The FTIR spectra were also used to attempt determine the ability of cleaning solutions to remove foulants from the membrane surface. A comparison between a membrane fouled by the 1 micron fraction of seawater and a membrane similarly fouled and cleaned with NaOH is shown in **Figure 5-15**. In the cleaned membrane, the doublet at 2360 and 2343 cm^{-1} and the region of sharp absorption between 4000 and 3500 cm^{-1} are not as intense relative the peak at 1242 cm^{-1} as it is in the fouled membrane, but these are due to carbon dioxide and water vapor, respectively, and are not indicative of the ability of the cleaning solution to remove material from the membrane. Overall, the cleaned membrane is more similar to the unused membrane than the fouled membrane. These results suggest that the NaOH cleaning solution was able to remove many of the foulants that decreased the absorption of the unused membrane spectrum. It should be noted that the membrane cleaned with NaOH shows a spectrum very similar to the spectrum pretreated at 0.1 μm, suggesting that the NaOH can remove the same compounds removed by prefiltration at smaller pore sizes.

5.3.2.4 SEM Imaging of Fouled and Cleaned membranes

After the fouling or cleaning experiments were completed, membrane coupons were removed from the SEPA cell and stored, refrigerated in deionized water until analysis. For preparing the membrane for SEM analysis one inch square sections of clean membrane were cut out of a fresh coupon, transferred onto a SEM stub and attached with double sided carbon tape. The edges of each mounted specimen was then heavily coated with colloidal graphite paint.

Figure 5-16 represents a clean RO membrane. **Figures 5-17** and **5-18** represent the SEM images of the fouled and cleaned RO membranes after two different levels of prefiltration. All of them are magnified to the same scale. The only visible difference in the fouled and cleaned membranes is at the 1 micron prefiltration level, where the amount of material after cleaning with NaOH is much reduced compared to the fouled membrane indicate substantial flux recovery after this cleaning possibly because of the ease of removal or larger size debris from the membrane surface. For the 0.1 micron prefiltration there seems to be no difference between the fouled and the cleaned membrane. Both these membrane samples infact look identical to the clean membrane. This is possibly because the resolution of the SEM images is not large enough to go beyond the 1 micron level. Similar results (to the 0.1micron prefiltration) were seen for the smaller size prefiltration RO fouling experiments in terms of SEM images. However, good quality images for these smaller fractions could not be obtained.

6. RESULTS AND DISCUSSION: PILOT TESTING RESULTS

The operation of the pilot was conducted in two phases. In the pretreatment evaluation phase, the focus was on operating the pretreatment trains to evaluate the effect of pretreatment on water quality parameters. In the RO Operation phase the RO systems were operated on the batch of water produced by the pretreatment trains.

Two different water sources were used during the pilot study. Raw Water was collected from the Shelter Island Boat Ramp for most of the study. This location provided a representative water which is characteristic of seawater bays close to open ocean. However, the South Bay (San Diego Bay) Power Plant effluent was also used as a source. This source is deep inside the San Diego Bay where the bay "dead ends". The water here is poorly flushed was found to be influenced by surface run off and had high turbidity and organic content. Raw water quality data for these two sites are summarized in Section 4: Raw Water Characterization and Bench Scale Testing Results.

6.1 Pretreatment Evaluation

During this phase of testing the two pretreatment trains were operated continuously to obtain reliable water quality information on pretreated water as well as operational data from these two trains. This testing was conducted over a period of six months.

6.1.1 Conventional Pretreatment

As described in Section 4 the conventional pretreatment train consisted of a pilot coagulation/sedimentation unit followed by 2 conventional media filters. The optimum ferric doses were determined from bench testing results and for Shelter Island water was set at 4 ppm $FeCl_3$ while for South Bay water it was maintained between 7 and 13 ppm. No polymer was added during the testing. The coagulation/sedimentation pilot unit was operated to provide a flow rate of approximately 0.7 gpm. The loading rate on the media filters was set at 6 gpm/sqft. Two different dual media filter configurations were evaluated. The first configuration evaluated was 72 inches deep with a 24 inch anthracite bed over 48 inch sand bed. The second configuration consisted of a 30 inch sand bed over a 60 inch anthracite bed.

Tables 6-1 and **6-2** present the range of values of feed and settled water onsite lab measurements for the two different water sources evaluated during this testing. The tables show that there is little difference in the water quality parameters between the feed and the settled water. From the turbidity data for raw and settled water it seems that the settling process for this pretreatment train was not optimum. In light of this fact the conventional pretreatment train could be considered a direct filtration train. This does not seem to affect the operation of the conventional filters in terms of the water quality produced. However, further optimization is possible with longer settling times.

Tables 6-1 and **6-2** clearly show the difference between the two water sources in terms of turbidity and UV 254 measurements, both of which are higher for the South Bay water. This data is in agreement with the discussion of water quality differences present at the different sites presented in Section 5.

Table 6-3 presents a summary of the water quality analysis conducted by the MWWD Labs (Point Loma and Marine Micro Labs) and Calscience Labs on the feedwater on the feed water. These analyses were only conducted for the Shelter Island Water source.

The operational sequence of pilot testing with conventional pretreatment is summarized in **Table 6-4**. The operational results from operation of the conventional filters are summarized in **Table 6-4** and **Figures 6-1** through **6-3**. Twenty-five filter runs were completed with the conventional pretreatment train. Runs 1 through 20 were conducted with the 72 inch deep filter configuration while Runs 21 through 25 utilized the 90 inch deep filter configuration. Runs 1 through 23 were on Shelter Island water while Runs 24 and 25 were conducted using South Bay Effluent Water.

The conventional filter runs were terminated when the head reached approximately 100 inches of water. Based on this criterion it was found that the filter run lengths were between 57 and 120 hours (barring coagulation upsets and runs beyond the target head loss of 100 inches of water) for all these cases **(Table 6-4)**. The coagulation upsets when the coagulant feed pump failed correspond to the shortened run lengths. The spikes in the turbidity correspond to operation of the system immediately after a backwash before filter maturation. The operating turbidity was consistently below 0.1 NTU after filter maturation. The backwashes were initiated manually and were between 10 and 15 minutes in length.

All SDI measurements were conducted after filter maturation when the filtrate turbidity had stabilized. The SDI measurements varied widely for the 72 inch deep filter and were between 2 and 5.3 for Shelter Island Water (**Table 6-4**). For the same water using the 90 inch deep filter the SDI values were more consistent and were between 2.8 and 3.2. The SDI values for the 90 inch deep filter was between 3.4 and 3.8 for the South Bay Water. The RO manufacturers recommend that the RO feed water SDI be below 5 and preferably below 3. These SDI measurements for the two different filter configurations and the two different water sources are shown in **Figures 6-2** and **6-3.**

The run lengths for Shelter Island water with the two filter configurations are essentially identical but in terms of water quality (turbidity and SDI) the 90 inch deep filter configuration produces more consistent data. Also the relatively lower water quality from South Bay tends to impact filtered water quality in terms of turbidity and SDI.

6.1.2 MF Pretreatment

As described in Section 3 the membrane pretreatment train consisted of a disk filtration system followed by a US Filter CMF –S MF unit. The operating flux was maintained at 30 gfd for most of the testing. The backwash frequency was set at 30 minutes and no chemicals were used during actual operation of the unit.

The operational results from operation of the MF unit are summarized in **Figures 6-4** and **6-5**. The system ran reliably at 30 gfd for a period of 670 hours with minimal fouling. A cleaning was conducted at the end of this period and the system put back into service at 35 gfd. The system operated for approximately 200 hours with minimal fouling. It was decided to raise the flux to 40 gfd after this. However, it was found that the raw water feed pump could not hydraulically provide this flow rate. Hence this run was terminated.

The turbidity values for the MF treated seawater was consistently below 0.1 NTU with most values between 0.03 and 0.04 NTU. **Figure 6-6** provides the SDI values. Unlike the conventional filters no variation in the SDI values can be seen as a result of switching to South Bay Water. The SDI values for the MF treated seawater were consistently below 3 for all cases and was well within RO manufacturers' recommendations for feedwater RO SDI.

Table 6-5 provides a comparison of particulate, organic and biological quality of the conventionally pretreated seawater to the MF permeate. The TOC and TSS values are similar for these two pretreatment streams. However, the HPC values for the MF permeate (between 1 and 98)was much lower than the conventional filtrate (between 8 mad 1600).

6.2 RO operation

This phase of pilot testing was conducted for a period of 3 months. Three distinct runs were conducted on the two RO trains to directly compare the flux decline rate of two complete RO trains utilizing two different pretreatment strategies operating on the same batch of water. The objective of this phase to draw conclusions regarding the effect of feedwater quality and pretreatment strategy on RO flux decline.

Both RO trains were operated under the same operating conditions. The RO operating conditions for this phase are summarized in **Table 6-6**. The overall specific flux decline curves for the two trains are shown in **Figure 6-7** and comparative rejection values are illustrated by means of the feed and permeate conductivity for each train in **Figure 6-8**. An attempt was made to maintain the temperature between at 25°C ± 5°C during the runs by using a large feed tank for equalization of temperatures . Most of the operating data was obtained within this range. There were few excursions for the first run with the conventionally pretreatment RO treatment train. It was determined that the feed tank volume was lower than the other train used and in subsequent runs the tank volume was increased resulting in temperatures within the desired range.

To understand the flux decline effects, flux was normalized to 25 deg C and specific flux calculated using the procedures described in section 4.5.4. These specific flux curves are shown in **Figure 6-7**. To directly compare the flux decline rates during each run and to eliminate the effect of initial specific flux, specific flux curves for each run were normalized to initial specific flux. These specific flux curves are presented in Figures in **Figures 6-9** through **6-11**. A figure showing the calculated specific flux decline rate for each run is shown in **Figure 6-12**.

These figures show that in Run 1 the specific flux declined more rapidly for the conventional RO train compared to the MF RO train. At the end of this run the conventional RO train had declined to 67 % of its initial flux while the MF RO train had declined to 73 % of its initial flux.

During Run 2 the specific flux decline rate was similar for the conventional RO train and the MF RO train during the initial part of the run. At the end of this run, however, the specific flux was lower for the conventional RO train (82% of initial specific flux) was slightly lower than the MF RO train (88% of initial specific flux). The 90 inch deep filter configuration was used for Run 2 while the 72 inch deep filter configuration was used for Run 1. The small difference in flux decline indicates that with proper design of filters it might be possible to use conventional filters to pretreat seawater to a degree such that the flux decline in RO is similar to membrane pretreated seawater. Shelter Island seawater was used for both Runs 1 and 2. From the water quality analysis (Section 5) it can be seen that this is a relatively good water source with a low level of particulates and organic matter especially when compared to seawater from South Bay which has a relatively high turbidity as well as organic content.

The specific flux decline trend for Run 3 is shown in **Figure 6-11**. This run was terminated when the specific flux of one of the trains reached 0.023 gfd/psi. South Bay seawater was used for this run. It can be seen that even though the same conventional filter configuration was used as Run 2 the flux decline rate was slightly higher for the conventional RO train in Run 3. Since South Bay water collected during this run was higher in organics and particulates to Shelter Island water used in Run 2 it was expected that the pretreatment trains might not perform as well as in Run 2. This was indeed the case for the conventional RO train where the flux decline rate is larger than in Run 2. However, the MF RO train flux decline was similar to the flux decline in Run 2 indicating that the MF treated water quality was comparatively independent of the variation in source water quality.

Several water quality measurements were conducted during the RO operation part of the testing. These are summarized in **Tables 6-7** through **6-11**. **Tables 6-7** and **6-8** show the feed water quality for the two RO membranes. It can be seen that in terms of the parameters measured including the conductivity, TDS, hardness and alkalinity, there was no significant difference between the two treatment train. This is expected as neither of the two treatment trains are designed to remove dissolved substances. For the other parameters measured including TOC, HPC and chlorophyll–a, there was no significant difference measured either. A small difference could be expected from the different pretreatment strategies but this not quantifiable in these parameters. For example, some TOC measurement could be expected from the coagulation process if a sufficiently high dose of coagulant is used. Also, the biological quality of MF permeate has been reported to be superior to conventional filtrate so the HPC values of this stream should in principle be lower. Not much data exists on chlorophyll-a which is an indicator of algae growth and was used as a quality control measurement to make sure the tanks used on sites were not growing algae. Very small quantities were found indicating good management practices.

The RO permeate values from the two treatment trains are shown in **Tables 6-9** and **6-10**. These values are similar as expected. The amount of TDS, alkalinity and hardness removal expected from two RO membranes operating on two differently pretreated waters is not expected to be different.

A complete sweep of ions was conducted during Run 2 of the testing. This data is provided in **Table 6-11.** These values are similar, as pretreatment strategies used are not expected to impact removal of ions.

7. RESULTS AND DISCUSSION: COST FACTOR EVALUATION

The major objectives of the cost factor evaluation effort were to

- Identify major cost factors related to the SWRO process and differentiate them based on pretreatment;
- Use information from current pilot testing, literature review and discussions with the project group and advisors to evaluate cost factors and relate them to operating conditions;

A recent study funded by the USBR has presented in depth data on full scale treatment trains incorporating different pretreatment scenarios (Henthorne and Quigley, 2005). Hence to avoid duplication, the current evaluation focused on the major factors influencing costs and the effect of pretreatment rather than a full-blown cost analysis. The following discussion is a qualitative discussion based on data from current study, literature review and discussions within the project team. Whenever applicable, interpretive charts and tables correlating data from the current study to cost factors ~~are~~ referenced. It should be kept in mind that most of these factors are interrelated and a separate discussion of each factor is presented to illuminate aspects that are pertinent to each cost factor.

7.1 Cost Factors for SWRO

A list of the cost components for Capital and O&M costs for the SWRO process is presented in **Table 7-1**. These components are used in the ensuing discussion to identify how the cost factors selected for discussion affect the Capital and O&M costs. Codes have been assigned to each cost component to make these correlations easier to follow in the next table.

Table 7-2 lists the cost factors associated with the SWRO process and the cost components associated with each cost factor. This table also indicates where the cost factor is impacted by pretreatment. Each cost factor is discussed individually below.

7.1.1 Membrane Properties

These are intrinsic properties of the membrane that have an impact on the Capital and Operational Cost of a SWRO treatment train.

7.1.1.1 Initial Specific Flux

Initial Specific Flux is defined as the ratio of the temperature corrected flux of new membrane to the net driving pressure. It is also referred to as the water mass transfer coefficient and is an intrinsic membrane property. Since net driving pressure is calculated by accounting for the applied pressure, pressure drop across the membrane, permeate pressure and the osmotic pressure on the feed side of the membrane, it is independent of feed water TDS and operating conditions. Specific flux has a great impact on capital and operating costs. Lower specific flux equates to more membrane area needed if the applied pressure is required to be kept below a certain value. Conversely, if flux is set at a certain value then larger pumps will be needed to maintain higher pressure. Larger pumps mean extra equipment costs and higher power consumption. The bench scale results from the current study indicated the initial specific flux of the RO membranes tested ranged from 0.049 - 0.056 gfd/psi. Since this is an intrinsic RO membrane property it is not affected by the pretreatment type.

7.1.1.2 Salt rejection

Salt rejection of SWRO membranes has to be relatively high to meet drinking water quality criteria. The USEPA secondary MCL for TDS is 500 mg/L and most seawater RO plants target permeate TDS values less than this number. Some plants target other constituents (like Sodium or Boron) and set permeate water quality goals. Low rejection SWRO membranes could necessitate addition of a second stage to meet specific water quality goals. This impacts capital costs by making extra membranes, vessels and equipment necessary and increases the power costs because of the additional pumping required. Pretreatment is usually not a factor affecting the salt rejection of RO membranes unless chlorine is used in the process and is not completely quenched before reaching the SWRO process, which can compromise the integrity of the RO membranes.

Bench scale results from this study showed that salt rejection varied among the RO membranes tested and ranged from 97.8 - 98.7%. However, as expected no difference in salt rejection was observed during pilot testing which compared MF RO to conventional RO. During pilot testing both trains utilized identical SWRO membranes (Hydranautics SWC4) and achieved 99.66% salt rejection. Such results demonstrate salt rejection by SWRO membranes is not impacted by type of pretreatment.

7.1.1.3 Effective Membrane area

Different membrane manufacturers have different amounts of membrane area per element. This could impact the overall capital cost as less number of elements need to be used for membranes with higher effective membrane area per element to obtain the same total production. For example, the 4 inch by 40 inch SWC 4 RO element used during pilot testing had an effective membrane area of 70 ft^2 while a similar membrane product of the same dimensions obtained from another manufacturer (DOW SW 30 4040) has an active area of 80 ft^2. However it should be kept in mind that the actual cost of the membrane elements might also vary and could contribute by itself to a difference in capital costs. Since membrane area is a property of the membrane element, it is not affected by pretreatment method.

7.1.2 Operational Flux

Operational Flux is a principal capital cost factor in that it determines how many membrane elements need to be used for a particular size of plant. Thus it also determines the number of vessels and plant footprint. Here, pretreatment could play an important role in determining what flux the RO membranes can run at without excessive flux decline. Operation at high flux rates (beyond the critical flux) can lead to excessive flux decline as demonstrated during the current pilot testing as can be seen during Run 1 of RO operation phase of pilot testing for both treatment trains). For example, the normalized specific flux decline rate of the MF RO train during operation at 14 gfd and 10 gfd was 11.2 % and 1.2 % per day, respectively. A similar trend was also observed for the conventional pretreatment train; however, it should be noted that noted that the MF treated water showed a lower flux decline than the conventionally treated water under all conditions. This indicates that MF treated water can be run at higher fluxes than the conventionally treated water for the same run lengths. Also, since flux decline leads to shortened cleaning intervals a higher Operational flux would lead to more frequent cleaning and contribute to operational costs.

7.1.3 Feed Water Recovery

Feed Water Recovery is defined as the amount of permeate produced per unit feed flow. Higher recovery has a positive impact on all the capital cost factors as the feed flow required to be processed through the treatment train is less for the same amount of permeate production. Wilf et al, 2001 emphasized that the size of all pretreatment equipment (including storage tank, booster pumps, filtration equipment and chemical dosing) along with concentrate piping and the out fall facility is directly impacted by feed water recovery. However the authors also reported that changing recovery rate and flux from 45% and 8 gfd flux to 55% recovery and 11 gfd increased power consumption from 4.2 – 4.6 kWh/m3. Recovery is usually limited by the higher pressure requirements and the potential for membrane scaling due to precipitation of inorganics on the membrane surface.

Pretreatment is not expected to have an impact on the recovery of the RO process unless some TDS or hardness removal pretreatment strategy is used as a part of the treatment train. However, one consideration would be the fact that salt rejection increases with increased recovery. To achieve the same level of salt rejection as observed at a lower recovery it would be necessary to increase permeate flux. From this regard, the type of pretreatment could impact sustainable recovery of SWRO process. Lastly, while conventional or MF/UF pretreatment would not provide removal of these constituents while NF could contribute to overall RO feed TDS reduction.

7.1.4 Specific Flux Decline Rate (Fouling Rate)

The Specific Flux Decline rate is representative of the decline in productivity of the membrane with time. It is a function of the membrane properties, operating conditions and the feed water quality. The rate of membrane fouling affects most of the operational cost components of the SWRO process. For instance, more power is needed for pushing the water through the membrane as it fouls. In addition, the membranes require more frequently cleaning resulting in increased chemical and labor costs. Furthermore, increased cleaning can cause membrane deterioration and ultimately shorten membrane life. It is expected that membrane pretreatment, as compared to conventional, will result in lower specific flux decline rate. This is primarily due to the ability of membranes to consistently achieve high levels of particulate matter regardless of feed water concentration. Results from the current pilot testing showed the normalized specific flux decline rate for Runs 1, 2 and 3 for the MF RO and conventional RO trains were 11.2 %, 1.2% and 1.1% per day respectively, and 13.8%, 1.7 % and 2.3 % per day, respectively.

7.1.5 RO Feed Water Quality

Seawater characteristics have a great effect on the Capital and Operational Cost of a SWRO treatment train. A description of how common seawater constituents impact SWRO costs is provided below.

7.1.5.1 Salinity

Salinity has an impact on the operational pressure of the process as applied pressure required to maintain a certain production is directly related to the osmotic pressure of the feedwater. Also important to consider is the variation in salinity. Salinity impacts the power usage and chemical consumption (antiscalant use). It is a primary consideration while designing a plant and affects all Capital cost factors except pretreatment processes, which usually are not impacted by salinity. The most effect the salinity has on Capital Cost factors is on the selection of the feedwater pumps as these would have to provide a higher head for a certain production rate when compared to a sourcewater with more salinity. Salinity also greatly impact the power consumption of the SWRO plant and hence controls O&M costs as power costs are in most cases the major contributor to overall operational costs. If there are large differences in salinity with time, specific changes will have to be made into the RO operation to account for the different pressure required at different time. Salinity is not impacted by pretreatment unless NF pretreatment is used. A recent USBR study on seawater desalination of seawater under the influence of surface water runoff demonstrates the impact of salinity changes on the operational performance of the RO membranes (Reiss et al., 2004).

7.1.5.2 Temperature

Temperature has two direct impacts on the RO process. First, increased feed water temperature allows more water to diffuse across the membrane allowing for decreased operating pressure under constant flux conditions. Secondly increased feed water temperature results in increased salt passage across the membrane. Wilf et al., estimates the impact of temperature and flux on salt passage at around 3% per °C. As a result, feed water temperature can impact all the capital cost factors except pretreatment as the RO plant can be sized smaller compared to when the feed temperatures are lower. Temperature also increases salt passage so the advantage of higher production has to be balanced against the water quality goals. Temperature is not impacted by pretreatment unless there is considerable hold time between the pretreatment and the RO process. Temperature has a beneficial effect on power consumption as less pressure is needed for the same production if the temperatures are higher.

7.1.5.3 Turbidity (particulate load)

Seawater with high particulate matter can impact the performance of pretreatment processes leading to poorer quality water reaching RO membranes. This would effect all operational cost components due to fouling of RO membranes. This was seen during pilot testing during Runs 2 and 3. The pretreatment configurations were the same for these two runs. However, the water was from two different sources. For

Run 2 Shelter Island water with lower turbidity was used while highly turbid South Bay water was used for Run 3. The UV 254 values for South Bay water were also higher than Shelter Island water. From **Figure 7-1** it can be seen that the impact on the RO flux decline rate is much higher for the conventionally pretreated water compared to the MF pretreated water indicating that conventional treatment is more sensitive to water quality changes then MF treatment.

Also, the membrane pretreatment processes themselves are generally more robust and the permeate quality is not impacted by the changes in water quality, including turbidity. This was demonstrated in Phase I : Pretreatment evaluation part of this study where the MF permeate turbidity was between 0.02 and 0.04 NTU irrespective of influent water quality while the conventional filtrate turbidity was between 0.02 and 0.07 NTU (after maturation) depending on the feedwater source. Galloway et al., 2003, emphasize that from an operational stand-point membrane pretreatment is less labor intensive than conventional during periods of varying feed water quality.

7.1.5.4 SDI

The SDI of seawater is considered an indicator of the propensity of the water to foul RO membranes. In recent years there has been some discussion of the suitability of this parameter as an indicator of fouling since it is not found to be a reliable indicator of fouling in several cases. If considered an indicator of fouling, high feedwater SDI's could impact all the components of operational costs. Raw water SDI's could not be obtained on any for the water sources as the filters became completely plugged in less than 5 minutes. During the pretreatment phase of testing, it was found that MF SDI's were consistently lower than the conventional pretreatment. The MF permeate SDI's were between 1 and 3 while for the conventional process it was dependent on feedwater source and filter configuration and varied widely from 2 to 5.3.

RO feedwater SDI's are presented for Runs 2 and 3 of Phase II in **Figure 7-1**. It shows some correlation with the RO flux decline rate. However, the magnitude of the difference in SDI does not directly correlate with the observed flux decline.

7.1.5.5 Biological Quality

Biofouling is one of the mechanisms of fouling in RO membranes. Poor biological water quality of RO feed could cause higher flux decline affecting all the operational cost components. This is definitely impacted by pretreatment strategy as a better biological water quality can be produced from an well-operated pretreatment process. In this study HPC was used as a biological indicator. A well operated MF is a complete barrier for most bacteria and hence HPC numbers should be relatively low. During the pretreatment evaluation part of this study it was found that the HPC values for the MF permeate (between 1 and 98) were much lower than those of conventional filtrate (between 8 and 1600) indicating better biological quality. However, during the RO testing phase of the study such a difference was not observed. The average HPC of the MF RO feed water was 329 cfu/ml while that of the conventional RO feed was 360 cfu/ml. This is probably due to contamination and sample handling errors.

7.1.5.6 Organic Matter

High level of organic matter in the water could cause greater organic fouling of membranes. This will impact all the operational cost components of the RO process. This is clearly seen from **Figure 7-1** where the two RO systems were operated during Runs 2 and 3 under the same operating conditions but with widely varying UV 254 which is sometimes used as a surrogate for aromatic organic matter content. Run 3 with higher organic matter showed higher fouling for both the conventional RO system and the MF RO system.

Again, there was not much difference between the TOC values before and after the conventional or MF pretreatment processes. The MF process is not expected to remove any organic matter but due to the use of coagulant some organic matter removal could be expected from the conventional pretreatment process. This was probably not observed during our testing due to the low doses utilized (4 mg/L as $FeCl_3$). Eventhough it was not a part of this study a further investigation on the impacts of coagulation before MF on organic matter and turbidity removal and its benefit of coagulation into the robust MF process.

7.1.6 Permeate Water Quality Goals

Permeate Water quality goals determine the operating conditions of the RO process. Since the operating conditions dictate the design as well as operation of the RO system. This will impact all operating cost components and capital cost components (except pretreatment). Pretreatment type has minimal effect on the RO permeate water quality. This can be seen from the detailed water quality results from Run 2 of the RO testing phase presented in **Table 6-11**. There is minimal difference between the Water Quality parameters

7.1.7 Membrane Life

Membrane life is directly impacted by the quality of the membrane element construction materials and the exposure of the membrane to oxidants and/or cleaning chemicals. Improper quenching of any oxidants used for biofouling control in the pretreatment train could cause rapid deterioration of the RO membrane shortening its life. This factor is directly related to the Specific Flux Decline Rate as membranes that show large flux decline will have to be cleaned more frequently. Cleaning brings the membrane into contact with high and low pH solutions, which will slowly degrade the membranes over the long run decreasing their rejection capabilities. This factor is impacted indirectly by pretreatment as pretreatment determines if the membrane is being cleaned more regularly. Several authors have suggested MF pretreatment may result in longer life of SWRO membranes due to less cleaning (Ebrahim et al., 1997, Galloway et al., 2003, Henthorne 2005).

7.1.8 Chemical Costs

A variety of chemicals are used throughout the SWRO process and therefore chemical costs represent a major operational cost factor. The chemicals that may be used in a SWRO plant include biocides (or oxidants) for biofouling control, chlorine, antiscalant, cleaning chemicals, sodium bisulfite for quenching free chlorine, coagulant and acid for pH adjustment. Of these chemicals, the cost of cleaning chemicals are most impacted by pretreatment as pretreatment can determine how frequently these chemicals are utilized.

In terms of pretreatment chemical requirements, in general, membrane pretreatment requires fewer chemicals than conventional. Typical chemicals required for MF pretreatment process include chlorine during routine operations in backwash and chemicals like chlorine, caustic and citric acid for chemical cleaning. For the conventional process coagulants such as Ferric, Alum or PACL are used in addition to chlorine that is used for biofouling control. Since the coagulant feed for the conventional pretreatment process is continuous the amount of chemical consumption for convention pretreatment is in general higher than membrane processes. Burney et al., estimated chemical costs associated with pretreatment (based on conventional and MF) for a 27,276 m^3 SWRO plant. The authors reported the chemicals costs for the conventional pretreatment process was approximately 2.8 times higher (i.e. 4.662 vs. 1.637 fils / m^3) than that required for MF.

7.1.9 Post Treatment

Post treatment processes include alkalinity recovery, disinfection and corrosion control measures and are affected by specific water quality parameters of the RO permeate like alkalinity, hardness and pH. A low alkalinity and hardness permeate will need larger size post treatment equipment and more chemicals. This is not expected to be impacted by pretreatment as the water quality parameters of the RO permeate following different pretreatment strategies are similar.

7.1.10 Membrane Configuration

Membrane configuration and staging of vessels has a major impact on the capital costs as it effects the number of elements used and consequently the number of vessels and the plant footprint. Membrane configuration is primarily dictated by feed water quality as well as the desired permeate water quality and is not impacted by pretreatment

7.1.11 Type of Energy Recovery Devices

Use of Energy Recovery Devices ~~are~~ is very prevalent in SWRO plants. Additional equipment is required but significant cost savings using energy recovery devices have been reported. The selection and use of these devices are not impacted by pretreatment.

7.1.12 Seawater Intake and Outfall Location

The distance of the intake and out fall locations from the plant adds to the capital cost and to operational costs due to the extra pumping power required. This is not impacted by pretreatment.

7.1.13 Characteristics of Product Conveyance Line

The length of the product conveyance line adds to the capital cost and to operational costs due to the extra pumping power required. This is not impacted by pretreatment.

7.1.14 Plant Utilization Factor

This does not affect any of the capital or operational cost components listed but plant down time effects the cost per unit production of water by reducing average production. It has been reported (Wilf et al, 2001) that the offline time for membrane pretreatment process (i.e. backwashing) is lower than filter backwashing time.

7.1.15 Product water Blending

Blending reduces the amount of water that has to be treated and hence affects all the capital and operational cost components. This decision is not based on the type of pretreatment used.

7.1.16 Seawater Intake Type

Seawater intake type greatly affects the water quality coming into the pretreatment process. Consequently it affects all the operational costs including extra labor to keep prescreens clean. It affects capital cost by requiring extra pretreatment steps. However, the intake type itself is not impacted by the pretreatment selected.

There are three distinct type of intakes: beach well intake, surface water intake and submerged intake. Where it can be implemented, beachwell intake will provide the best quality feed in terms of suspended solids, dissolved organics, intake temperature and in some cases feed pH. Feedwater quality from a beachwell intake is so consistent in terms of SDI and turbidity, that beachwells alone have been considered as pretreatment process for side by side evaluation with conventional pretreatment methods. If a beachwell intake is not suitable, the depth of the intake directly affects intake water quality. It has been reported that that surface intake from the first 35m of the ocean (surface intake) is heavily loader with marine organisms as well as algae, and that intake from below 35m deep (submerged) has 1/ 20 the debris loading found in shallow waters (Gille, 2003). Ocean depths exceed 35m anywhere from 10m to several kilometers from the shoreline in most cases.

REFERENCE LIST

Abdullah, H., 2004, East Bay Municipal Utility District, EBMUD, personal communication

Al-Sheikh A. H. H. 1997. SWRO Pretreatment with an Emphasis on the Jeddah Plant Operation Experience. *Desalination*, 110: 183-192.

Antrim, B., 2004, Koch Membrane Systems, personal communication

Baig, M.B., Al Kutibi, A., A., 1998, "Design features of a 20 migd SWRO desalination plant, Al Jubail, Saudia Arabia" Desalination 118 (1998) 5-12

Bell, R., 2004, Municipal Water District of Orange County (MDOWC), personal communication.

Bou-Hamad et al, 1997, Performance evaluation of three different pretreatment systems for seawater reverse osmosis technique, Desalination, 110, 1997, 85-92.

Boyle Engineering Corporation, 1989, Marin Municipal Water District (MMWD) Feasibility Study Report

Brehant, A., Bonnelye, V., Perez, M., "Comparison of MF/UF pretreatment with conventional filtration prior to RO Membranes for surface seawater desalination." Desalination 144 (2002) 353-360.

Cleasby J. L. 1985. Unconventional Filtration Rates, Media and Backwashing Techniques. *Innovations in the Water and Wastewater Fields*. E. A. Glysson *et. al.* eds., Butterworths: 1-23.

Dalvi A.G.I., Al-Rasheed, R, Javeed M.A., Studies on organic foluants in seawater feed of reverse osmosis planets of SWCC

Diana Mourato, M. Singh, C. Painchaud, R. Arviv, Immersed membranes for desalination pretreatment, Proceedings of the IDA World Congress on Desalination and Water Reuse, Bahamas, 2003.

DiGiano, F. A., S. Arweiler, and J. A. Riddick. 1999. Alternative Tests for Evaluating NF Fouling. *JAWWA* 92:2:103-115

Dow, 2004, Website: www.dow.com

Ebrahim S.H., et al. "Conventional Pretreatment System for the Doha RO Plant: Technical and Economic Assessment." Desalination 102 (1995) 179-187.

Ebrahim et al., 1997 "Microfiltration system as pretreatment for RO units: Technical and economic assessment".

El Azizi et. al, 2002, Design Criteria of 10,000 m3/day SWRO desalination plant of Tajura, Libya, Desalination (2002), vol. 153, 273-279

El-Manhawary S., A. Hafez. "Technical Management of RO System." Desalination 131 (2000) 173-188.

Everest, W., 2004, Orange County Water District, OCWD, personal communication

Franks,R., 2004, Hydranautics, personal communication

Furukawa, 2004 "An Introduction to Desalination", AMBAG Desalination Workshop Monterey CA May 20th 2004.

Galloway et al, "UF for Seawater RO Pretreatment", 2003a, Proceedings of the IDA World Congress on Desalination and Water Reuse, Bahamas, 2003.

Galloway M., Von Gottberg A., Mahoney J., 2003b, " UF for Seawater RO Pretreatment"
2003 AWWA Membrane Technology Conference Paper Proceedings

Gille, Detlef. "Seawater Intakes for Desalination Plants." Desalination 156 (2003) 249-156.

Gluckstern, P., M. Priel, Mark Wilf. "Field Evaluation of Capillary UF Technology as Pretreatment for Large Seawater RO Systems." Desalination 147 (2002) 55-62.

Hagen, D and Comstock, D., "Granular Media Filtration Pretreatment" International Desalination Association Conference, RO Pretreatment Workshops, San Diego March 2004

Hassan, A.M. et al. "A Demonstration Plant based on the new NF-SWRO Process." Desalination 131 (2000) 157-171.

Henthorne, L., R. Quigley. "Evaluation of MF, UF and Conventional Pretreatment for SWRO Applications." International Desalination Association Conference Proceedings, 2003.

Henthorne, L., R. Quigley. "Evaluation of Membrane Pretreatment for Seawater RO Desalination." Aqua Resources International, Desalination and Water Purification Research and Development Report No. 106. Spring 2005.

Hyflux, 2004, website: www.hyflux.com

IDA, 2004, IDA Website, Featured Plants list: www.idadesal.org

Irwin, K.J., J.D. Thompson. "First year Operation of 26.25mgd (119,000m3/day) Point Lisas Desalination Plant." Proceedings IDA World Congress 2003.

Jew V., Shoengerger P., Filteau, G., Alt, S., Desalination of Pacific Ocean Water Via Microfiltration and Reverse Osmosis. American Water Works Annual Conference, 2003

Kurihara, M., 2003, "Focus on Membrane Technology for Water Treatment" Presented by Dr. Masaru Kurihara, Toray Industries, Sept 2003, Toray website: http://www.toray.com/ir/library/pdf/lib_a101.pdf

Lopez, C., 2004a, SDCWA Planning and Environmental Committee Desal Project Update, dated 7/14/04

Lopez, C., 2004b, San Diego County Water Authority (SDCWA), personal communication

Maxwell, J., 2004, IDA RO Pretreatment Workshop, San Diego, CA, March 2004

Miller, W., 2004, IDA RO Pretreatment Workshop, San Diego , CA, March 2004

Montgomery Watson, 2000, Point Lisas Reverse Osmosis Pretreatment Pilot Study: Trinidad and Tobago, Nov. 2000

Nadeem A. B., Ebrahim S., Abdel-Jaw AD M., Bou-Hamed S., "Seawater Pretreatment by Microfiltration: An Economic Assessment" Water Resources Development Vol. 15., No. 3, 323-331, 1999

Pentell, S.E., Seawater Desalination in California, California Coastal Commission Report, October 1993.

Reahl, G., 2004, Ionics, personal communication

Reiss, R., "Evaluation of desalination on waters under the influence of surface water runoff for pretreatment, water quality, and pathogen removal performance." Final Report Submitted to USBR Agreement no. 02 fc 81 0829, Sept 2004.

Sacksteder, C., 2004, The Dow Chemical Company, personal communication.
Wilf,M., and Schierach,M.,K., 2001, Improved performance and cost reduction of RO seawater systems using UF pretreatment, Desalination, 135, (2001) 61-68

Toray, 2004, Seawater Application list (Rev 9/04) provided by Toray Membrane America, Inc

TWDB, 2003, Biennial Report on Seawater Desalination DRAFT Report dated 10/19/2003

Viesmann, W., Jr., and Hammer, M.J., Water Supply and Pollution Control, 1998, Addision Wesley Longman Inc., 1998

Voutchkov, N., 2004, "Demonstrating seawater desalination feasibility", Water 21 Magazine, February 2004

Yoon, S.K., 2004, Saehan Industries, personal communication

Zidouri H. 2000. Desalination in Morocco and Presentation of Design and Operation of the Laayoune SWRO Plant. *Desalination*, 131: 137-145.

APPENDIX A
Tables and Figures

Table 2-1: Full-Scale SWRO Facilities Worldwide

Plant Name	*Location*	*Capacity (MGD)*	*RO Membrane MFG*	*Status*	*Year Completed*	*Reference*
Ashkelon	*Israel*	*72.0*	*Dow Filmtec*	*In Construction*	*2005*	*Dow, 2004*
Baughers Bay	*Grand Cayman*	*1.6*	*Dow Filmtec*	*In Operation*	*1990*	*Sacksteder, 2004*
Rambla Morales	*Spain*	*13.2*	*Dow Filmtec*	*In Operation*	*2004*	*Dow, 2004*
Al Dur	*Bahrain*	*11.9*	*Dupont*	*In Operation*	*1989*	*Kurihara, 2003*
Pembroke	*Malta*	*14.3*	*Dupont*	*In Operation*	*1994*	*Kurihara, 2003*
Al-Zour South	*Kuwait*	*28.5*	*Dupont*	*In Operation*	*1999*	*Kurihara, 2003*
Marbella	*Spain*	*14.9*	*Dupont*	*In Operation*	*1999*	*Kurihara, 2003*
Al-Jubail	*Saudi Arabia*	*20.0*	*Dupont / Toray*	*In Operation*	*2000*	*Kurihara, 2003*
Palma de Mallorca	*Spain*	*16.6*	*Dupont, Toray*	*In Operation*	*1999*	*Kurihara, 2003*
Tajura	*Tripoli, Libya*	*10.0*	*Koch/Fluid Systems*	*In Operation*	*1993*	*El Azizi et al., 2002*
Kindasah	*Saudia Arabia*	*3.3*	*Hydranautics*	*In Operation*	*2000*	*Hydranautics, 2004*
Larnaca	*Cyprus*	*14.3*	*Hydranautics*	*In Operation*	*2001*	*Hydranautics, 2004*
Cartagena	*Spain*	*17.0*	*Hydranautics*	*In Operation*	*2002*	*Hydranautics, 2004*
Tampa Bay	*Tampa Bay, Florida*	*25.0*	*Hydranautics*	[1]*In Operation*	*2003*	*Hydranautics, 2004*
Antofagasta	*Chile*	*13.7*	*Hydranautics*	*In Operation*	*2003*	*Hydranautics, 2004*
Carboneras	*Almeria, Spain*	*31.7*	*Hydranautics*	*In Operation*	*2003*	*Hydranautics, 2004*
Fujairah	*Fujairah, UAE*	*45.0*	*Hydranautics*	*In Operation*	*2004*	*Hydranautics, 2004*
Changjing Chemical	*China*	*1.1*	*Saehan*	*In Operation*	*2004*	*Yoon, 2004*
Campo de Mauricia	*Spain*	*37.0*	*to be decided*	*In Construction*	*2006*	*Furukawa, 2004*
Tianjin	*China*	*26.4*	*to be decided*	*In Construction*	*2006*	*Hyflux, 2004*
Taweelah	*Abu Dhabi, UAE*	*60.0*	*to be decided*	*In Construction*	*2006*	*IDA, 2004*
Ashdod	*Israel*	*38.0*	*Toray*	*In Operation*	*2004*	*Toray, 2004*
Tuas	*Singapore*	*36.0*	*Toray*	*In Construction*	*2005*	*Toray, 2004*
Hamma	*Algeria*	*30.0*	*Toray*	*In Construction*	*2006*	*Toray, 2004*
Algiers	*Algeria*	*53.0*	*Toray*	*In Construction*	*2006*	*Toray, 2004*
Trinidad	*Trinidad & Tobago*	*28.8*	*Toray*	*In Operation*	*2002*	*Toray, 2004*
Okinawa	*Japan*	*10.5*	*Toray, Hydranautics*	*In Operation*	*1996*	*Hydranautics, 2004*
Jeddah RO1	*Saudi Arabia*	*15.0*	*Toyobo*	*In Operation*	*1989*	*Kurihara, 2003*
Jeddah RO2	*Saudi Arabia*	*15.0*	*Toyobo*	*In Operation*	*1994*	*Kurihara, 2003*

Table 2-2: Proposed SWRO Facilities (United States)

Project Sponsor(s)	*Project Name*	*Year of Study*	*Proposed Location(s)*	*Proposed Capacity*	*Status*	*Proposed Start Up Date*	*Reference*
East Bay Muncipal Utility District (EBMUD, San Francisco Utilities Commission (SFPUC), Santa Clara Valley Water District (SCVWD), Contra Costa Water District (CCWD)	*Bay Regional Desalination Project*	*2003*	*1. Mirant Power Plant 2. Near San Francisco Bay Bridge 3. Oceanside*	*20 MGD increments (up to 120 MGD ultimate)*	*on going*	*unknown*	*Abdullah, 2004*
Los Angeles Department of Water and Power (LADWP)	*LADWP Seawater Desalination Project*	*1993*	*Scattergood Power Station*	*12 MGD*	*Currently studying 1) max. capacity 2) conc. disposal issues*	*2015*	*Miller,2004*
Orange County Water District (OCWD) / Municipal Water District of Orange County (MWDOC)	*OCWD/MWDOC*	*2003-2004*	*AES, Huntington Beach, CA*	*50 MGD*	*Filed Expression of interest for consortium*	*2008*	*Everest, 2004*
West Basin Municipal Water District	*West Basin Municipal Water District Desal Project*	*2002*	*to be determined*	*20 MGD*	*Completed pilot testing. Will build 0.5 mgd demo*	*2008*	*Miller, 2004*
Long Beach Water Department (LBWD)	*Long Beach Seawater Desalination Project*	*2001*	*Long Beach, CA*	*9-10 MGD*	*Completed pilot testing. Will build 0.3 mgd demo*	*demo:2005;full sclae:2010*	*Lopez, 2004a*
Marin Muncipal Water District	*Marin Muncipal Water District Seawater Desalination Project*	*1989*	*Corte Madera, CA*	*15 MGD*	*RFP issued 2004 additional pilot testing*	*unknown*	*Boyle Engineering, 2004*
California American (Cal-am) Water Company; Monterey County Water Resources Agency	*Coastal Water Project*	*2004*	*Moss Landing, CA*	*20 MGD*	*Completing Env. Assessment to be submitted to CPUC summer 05'. Planning on 60 gpd pilot plant*	*mid 2009*	*Lopez, 2004a*
Municipal Water District of Orange County (MWDOC)	*MWDOC Ocean Water Desalination Project*	*2004*	*Site 1: Dana Point Site 2: San Onofre Nuclear Generating Station (SONGS)*	*Site 1: 25 MGD*	*Current*	*unknown*	*Bell, 2004*
Texas Water Development Board (TWDB)	*Lower Rio Grande Valley-Brownsville Feasibility Study*	*2003*	*Brownsville Port Authority Complex (Brownsville, TX)*	*25-100 MGD*	*TWDB requesting funding from the Texas Legislature to perform pilot testing*	*2010-2040*	*TWDB,2003*
Texas Water Development Board (TWDB)	*Corpus Christi Feasibility Study*	*2003*	*Barney Davis Power Plant (Corpus Christi, TX)*	*25 MGD*		*2010*	*TWDB, 2003*
Texas Water Development Board (TWDB)	*Freeport Feasibility Study*	*2003*	*Dow Chemical Complex (Freeport, TX)*	*10 MGD*		*2010*	*TWDB, 2003*
South Florida Water Management District, Lee County Utilities, Florida Power and Light	*Feasibility Study on Seawater Desalination*	*2001*	*Ft. Myers Power Plant (Caloosahatchee River)*	*30 MGD*	*Phase I complete. Phase II beginning.*	*unknown*	*TWDB, 2003*
San Diego County Water Authority (SDCWA)	*South County Binational desalination*	*2003*	*South Bay Power Plant*	*25 MGD*	*on going*	*2015*	*Lopez, 2004b*

Table 2-3: Conventional SWRO Pretreatment Case Studies

Project / Facility Name	Chlorine Dose	Coagulant		Polymer		Filter Specifications	Notes	Reference
		Type	Dose	(Y/N)	Dose			
Tampa Bay Water Seawater Desalination Plant	NA	$Fe_2(SO_4)_3$	NA	YES	NA	Dual Sand filtration. 9 mm Stage 1; 6 mm Stage 2	Use prechlorination; dechlorination; cartridge filters	Maxwell, 2004
Evaluation of Membrane Pretreatment for SWRO Desalination	0 mg/L	$FeCl_3$	6-7 mg/L	NO	NA	Multi Media (sand/anthracite/garnet-equal volumes). Single Stage. 3 ft diameter beds. Total Filter Depth 48 inches. 96% recovery, BW filter 1 (once/12 hours), BW Filter 2 (once/24 hours). Filter loading rate of 2 gpm/ft2		Henthorne & Quigley, 2003
Evaluation of desalination of water under the influence of surface water runoff f	NA	$Fe_2(SO_4)_3$	NA	NA	NA	Two Stage multi media filtration		Reiss et. al, 2004
Point Lisas SWRO Pretreatment Pilot Study	0-0.5 mg/L	$FeCl_3$	7-21 mg/L	Yes	1 mg/L	Compared Single and Two Stage Filtration; Dual Media (sand/anthracite); Media size, mm (1/0.5; 0.8/0.4; 1.2/0.6); Filter Depth, in (60/30) Filtration Rate gpm/ft2 (4.0, 6.0 and 9.0).	Optimal Ferric Dose 16 mg/L as Fe. Tested Cationic Polymer stopped after 1 year -no significant improvement	Montgomery Watson, 2000
Carlsbad Desalination Project	1.0 mg/L	$Fe_2(SO_4)_3$	3.5 mg/L	Yes	0.5 ppm	Two Parkson Dynasand backwash filters in series. Filter #1 coarse (9 mm) sand media bed. Filter #2 finer (5 mm) sand media. Continuous Backwash	Cationic polymer tested for short time but stopped as no significant improvement in water quality observed	Voutchkov, 2004
Al Jubail Pretreatment Pilot Study	1-1.2 mg/L	$FeCl_3$	0.8 mg/L as Fe	YES	0.2-0.4	Dual Media Anthracite: 600-1200 mm (1.4-2.5 mm); Sand 800-1400 mm (0.63-1.0 mm); Support Gravel 1: 150 mm (2.0- 3.15); Support Gravel 2: 150 mm (3.15-5.60).	Filter Run Time 24+ hours; sulfuric acid dosed to reduce feed pH 8.3-6.7; sodium bisulfite used for dechlor upstream of RO.	Baig et, al .,1998
Doha Reverse Osmosis	1.94	$Fe_2(SO_4)_3$	4-5 mg/L	Yes	0.3-0.5	Dual Media: support layer (0.3 m); Silica Sand 0.7-12 mm (1 m); Hydroanthracite 1.4-2.5 mm (0.7 m)	Cationionic polymer addition + acid addtion (ph 6) reduced SDI and coagulant dose.	Ebrahim et al, 1995
Marin Desal Pilot Plant Study	0 mg/L	$FeCl_3$ (see notes)	14-20 mg/L	Yes	0.3-0.4	Three 24 inch dual media filters; 10 inches of #20 sand and 20 inches of 0.65-0.85 mm anthracite. Loading rate 3.7 gpm ft2	Jar test were performed using ferric chloride and ferric sulfate prior to testing. Polymer used in two locations coagulant aid and filter aid	Boyle Engineering, 1989
Mekrot Water Company	1.2 mg/L	$FeCl_3$	0.3-0.7 mg/L	No	NA	Two Stage Single and Dual Media; Velocity 6.5-7.2 m/h; Backwash every 100 h with air scouring		Gluckstern et. al 2002

Table 2-4: Specifications of media used in conventional SWRO pretreatment (Hagen and Comstock, 2004)

Media Type	Specific Gravity	Effective Size (mm)	Uniformity Coefficient	Specific Surface Area
Anthracite	1.6	0.9-1.0	1.5	1000
Sand	2.6	0.42-0.55	1.5	1600
Garnet	4.0	0.20-0.28	1.6	5300

Table 2-5: Media depths used in conventional SWRO pretreatment and freshwater applications

Seawater Utility and Pilot Plants in Operation			
Single/ Dual	Media types	Effective Media Size	Media Depth
Single[1]	Gained reinforced plastic bed	various	9 m
Dual[2]	Anthracite	NA	600 mm
	Fine sand	NA	600 mm
	Gravel	NA	400 mm
Dual[3]	Hydroanthracite	1.40-2.50 mm	700 mm
	Fine sand	0.70-1.20 mm	1000mm
	Support	various	300 mm
Dual[4]	Anthracite	1.00 mm	1500 mm
	Sand	0.50 mm	750 mm
Freshwater Recommendations			
Dual[5]	Anthracite	0.90-1.10 mm	600-750 mm
	Sand	0.45-0.55 mm	600-750 mm
	Coarse sand	0.80-2.00 mm	50-75 mm
Dual[6]	Coarse anthracite	1.50 mm	NA
	Underlying sand	0.75-0.90 mm	NA

Sources: [1]Zidouri (2000)
[2]Al-Shiekh (1997)
[3]Ebrahim *et. al.* (1995)
[4]Irwin and Thompson (2003)
[5]Great Lakes Upper Mississippi Board of State Public Health & Environmental Managers
(As referenced in Viessman and Hammer 1998)
[6]Cleasby (1985)

Table 2-6: MF/UF SWRO Pretreatment Case Studies

Project / Facility Name	Plant Type	MF/UF Membranes	Pre screen	Chemical Addition	Flux (gfd)	Feed Water Recovery (%)	Backwash Frequency	Reference
Evaluation of Membrane Pretreatment for SWRO Desalination	Pilot	Zenon (UF), Norit (UF), Hydranautics (UF), Memcor (MF), Pall (MF)	Replaced strainers with Arkal Spin Klin (130 micron discs)	Feed :no chlorine. 0.1-1.5 ppm ferric chloride (Hydranautics only). Product, 1-2 ppm SBS (Norit Only)	50-60 gfd	93-94	Norit used CEB	Henthorne & Quigley, 2003
Evaluation of desalination of water under the influence of surface water runoff	Pilot	Norit X-Flow (UF)	200 micron basket strainer	Ferric Sulfate	36 gfd	80%	N/A	Reiss et. al, 2004
West Basin Seawater Desalination Pilot Plant Project	Pilot	US Filter CMFS (MF)	Phase I&II: 800/500 micron strainer; Phase III: 800/150 micron strainer	Phase 1: 1 ppm NaOCL in feed followed by NH4OH. Phase II NaOCl in BW followed by SBS.	24 gfd	90+	15 with 25-40 mg/L Cl2	Jew et. al, 2003
Carlsbad Desalination Project	Pilot	Hydranautics Submerged (UF)	N/A	NO	10 gfd	N/A	N/A	Voutchkov, 2003
Marin Desal Pilot Plant Study	Pilot	Aqua Chem (UF), EPOC EXXFLOW (MF)	UF: 50 micron bag filter	none	25-35 gfd	unknown	15 min	Boyle Engineering, 1989
Mobile Seawater Desalination Plant	Full scale	Norit X-Flow (UF)	unknown	none	45 gfd	80+	CEB done every 24-72 hours. Alternate NaOCl and HCl. SBS and NaOH used in product	Reahl, 2004
Port Hueneme	Pilot	Koch (UF)	200 micron self backwashing	0.75-1 ppm ferric chloride	40 gfd	85%	tested with no chemicals. No chlorine.	Antrim, 2004
MEW and KISR	Pilot	Memcor CMF	Coarse Strainer	0.5 ppm chlorine to feed; removed by SBS in product	63 gfd	94.7	5-25min	Ebrahim et al, 1997
ONEDEO Services in Gibraltar	Pilot	Aquasource (UF)	200 micron prefilter	1 mg/L ferric chloride	35-88 gfd	not reported	30 min with 5 mg/L free chlorine	Brehant et al, 2002
Mekrot Water Company	Pilot	Hydranautics capillary hollow Fiber (UF)	50 micron screen filter	0.3 ppm ferric; no continuous chlorine	35-70 gfd	not reported	15-30 min with 20 ppm chlorine	Gluckstern et al, 2002

Table 2-7: Pretreatment Comparison

Project / Facility Name	Pretreatment Type	Effluent SDI	RO Membranes	RO Performance	Reference
Evaluation of Membrane Pretreatment for SWRO Desalination	MF/UF	< 3	Koch Toray	No cleaning required during pilot testing. Assumed 6-month cleaning interval	Henthorne & Quigley, 2003
	Conv.	3.4-5.5	Koch Toray	Operated 12.5 gfd, six week cleaning interval	Henthorne & Quigley, 2003
Evaluation of desalination of water under the influence of surface water runoff	UF	N/A	Toray TM810	N/A	Reiss et. al, 2004
	Conv.	N/A		N/A	Reiss et. al, 2004
Point Lisas SWRO Plant	Conv.	3	Pass 1 Train: Toray SU-820; Pass 2 Train: Toray SUL-G20F	Pass 1RO some rise in differential pressure, recoverable by low pH cleaning. With slight modifications to CP, RO cleaning interval expected 8-9 months. Pass II as expected to due lower salinity, stable performance for the first year.	Montgomery Watson, 2000 &Irwin and Thompson, 2003
NF-SWRO Desalination System	dual media filtration + NF^2	</= 1 (after NF)	Toyobo HM 8255 HFF & Toyobo HM 8355 HFF	not reported	Hassan et al , 2000
Gibraltar	UF	0.8	FilmTec SW30-2540	Phase I - flux 13 LMH; recovery 30 %. Feed pressure 52 bar @ 25 deg C for 27 days. Phase II flux 20 LMH; recovery 40%. Feed pressure 66 bar@ 20 deg C 20 days	Gluckstern et al, 2002
	Conv.	2.7-3.4		Phase I: flux 13 LMH; recovery 30 %. Feed pressure 52 bar @ 25 deg C for 21 days. Shorter runtime due to various operational problems. Phase II only ran 3 days due to operational problems.	Gluckstern et al, 2002
West Basin Seawater Desalination Pilot Project	MF	2-3	Hydranautics SWC 4040 Dow Filmtec SWC30-4040	Operated approx 60 days no fouling on Hydra @ 8-9 gfd, 50 % recovery. Filmtec showed increasing permeability.	Jew et al, 2003
Carlsbad Desalination	Conv.	unknown	Hydranautics SWC 4040	not reported	Voutchkov, 2004
	UF	unknown		not reported	
Marin Desal Pilot Study	MF	0.7-5.3	Hydranautics 8040- SWC1	not reported	Boyle Engineering, 1989
	[1]Conv.	5.9-6	Hydranautics 8040- -SWC1	not reported	
Mobile Seawater Desalination Plant	UF	unknown	Toray SU-820 FA	Flux 12-13 gfd; 950 psi operating pressure; 900 ppm TDS product water; 43% recovery.	Reahl, 2004
Port Hueneme	UF	<1	RO not tested	NA	Antrim, 2004
MEW and KISR	Conv.	2.24	RO not tested	NA	Ebrahim et al, 1997
Mekorot Water Company	Conv.	Red Sea: 1.6-2.3 Mediterranean: 2.6-3.8	Hydranautics SWC 4040	Flux 8.2-11.2 gfd, recovery 45-55%. During increased solids in feed water due to storms. Conv. sys would produce poor effluent requiring shut down of RO unit. Iron fouling at high recovery	Gluckstern et al, 2002
	UF	*Red Sea: 0.8-1.2 Mediterranean Sea: 2.1-3.0*	*Hydranautics SWC 4040*	*Flux 8.2-11.2 gfd, recovery 45-55%; During poor feed water period due to storms UF eff quality okay for RO. Iron Fouling at high recovery.*	

Table 4-1: Jar Testing Parameters

Mixing speed (RPM)	*Velocity Gradient (G)* (s^{-1}) *	*Time (min)*
100	97	1
60	51	10
30	22	10
10	4	10
0	0	60

*** According to G-curves from http://www.phippsbird.com/gcurve.html**

Table 4-2: Jar testing Polymer Characteristics

Polymer name	Charge	Molecular Weight
8170	None (non-ionic)	High
7128	Negative (low-charged cationic)	High

Table 4-3: Alternative Coagulant Testing Metal Equivalent Doses

Coagulant	*Molar Dosing (mmol/L)*	*Concentration (mg/L)*
$FeCl_3$	0.0246	4.00
$Fe_2(SO4)_3$	0.0123	4.92
$Al_2(SO4)_3$	0.0123	4.21
PACl	0.0123	2.60

Table 4-4: Bench Scale Testing: Tested Membrane Characteristics

MANUFACTURER	Commercial name	Model
SAEHAN	CSM® Reverse Osmosis Membranes	RE4040-SN
OSMONICS	Desal® Membrane products	AD2540F1969
HYDRANAUTICS	Nitto Denko Corporation	SWC4-4040
TORAY	ROMEMBRA®	TM810
KOCH	Fluid Systems®	TFC HF
DOW	Filmtec	SWHR-LE

Table 4-5: Seawater Fractionation Membranes

Size Fraction	Membrane Type	Membrane Code	Analyses Performed
1.0μm	Glass Fiber Depth Cartridge	-	TOC, UV254
0.1μm	PTFE Depth Cartridge	MFE	TOC, UV254
100kDa	PAN Flat Sheet	MW	TOC, UV254
20kDa	PES Flat Sheet	PW	TOC, UV254

Table 4-6: Conventional Pretreatment Parameters

Run No	Filter Configuration	Water Source	Ferric Dose mg/L	Depths (inches)		Size (mm)	
				Sand	Anthracite	Sand	Anthracite
1-20	72 inch filter	Shelter Island	4	48	24	0.5	1
20-23	90 inch filter	Shelter Island	4	30	60	0.5	1
24,25	90 inch filter	South Bay	7, 13	30	60	0.5	1

Table 4-7: MF Membrane Characteristics

Parameter	Units	Value
Manufacturer		USFilter Memcor
Membrane Element ID Number(s)		119066
Module Commercial Designation		S10V
Number of Elements Tested	#	2
Approximate Size of Element (len x diam)	inch	46.7 x 5.2
Inside Diameter of Fiber	mm	0.5
Outside Diameter of Fiber	mm	0.8
Active Length of Fiber	inch	41.3
Active Membrane Area per Module	sq ft	272.3
Flow Direction	inside out / outside in	outside in
Number of Fibers per Element		9,600
Available Operating Modes	Direct, Crossflow	Direct
Membrane Construction		Hollow fiber
Membrane Material		PVdF
Membrane Surface Characteristics	<slightly> hydrophobic, hydrophillic	Hydrophyllic
Membrane Charge	<slightly> negative / positive	neutral
Nominal Molecular Weight Cutoff	Daltons	approx 500 kD
Nominal Membrane Pore Size	micron	0.1
Design Operating Vacuum	psi	12.3
Design Flux (at Design Pressure)	gfd	52
Maximum Operating Vacuum	psi	12.3
Acceptable Range of Operating pH		2 - 10
Acceptable Range of Operating Temperature	degC	0.5 - 40
Maximum Turbidity	NTU	500
Chlorine / Oxidant Tolerance	ppm-hrs	200,000 at 1,000 ppm max

Table 4-8: Membrane Pretreatment Parameters

Unit	Flow	Operational Parameters	Monitored Parameters
MF Unit	11 gpm	Flux: 30 gfd (initial) Backwash frequency :30 min No Chlorine	Trans Membrane Pressure (TMP), temperature SDI, UV- 254, turbidity, pH

Table 4-9: RO Pilot Specifications

Parameter	Number	Description
Capacity	22 gpm	
Pressure Vessels	4 Vessels	40" by 120" Codeline pressure vessels rated up to 1,000 psi
Elements	12 Elements	Compatible with all manufacturers
High Pressure Pumps	2 pumps	7.5 HP, 10.9 gpm at 950 psi, CAT Model 2537
Low Pressure pumps	2 pumps	2 HP, 11.5 gpm at 35 psi, Sta-Rite Model PDHG-L
Cartridge Filter	2	5 micron, Ryan Herco
Instrumentation	N/A	2 flow meters, 4 pressure sensors, 1 temperature indicator, 1 data logger, 1 control panel
Electrical	N/A	460 Volt, 60 Hz, 3 phase

Table 4-10: Pilot RO Membrane Specifications

Manufacturer		Hydranautics
Element ID		SWC 4
Element Dimensions	dia (in) X length (in)	4 inchx 40 inch
Configuration		Spiral Wound
Material		Polyamide Thin Film Composite
Area (per module)	ft^2	70
Min Salt Rejection*	%	99.7
Nominal Salt Rejection*	%	99.8
Max pressure	psig	1200
Max flow	gpm	16
Max Operating Temp	deg C	45
pH range		10-Mar
Max Feedwater Turbidity	NTU	1
Max feedwater SDI		5
Max Chlorine Tolerance	ppm	<0.1
Max Presure Drop for each element	psi	10

*initial performance data taken after 60 minutes under the following conditions: NaCl solution conc; 32,000ppm, Applied pressure :800psi, 25degC, 10% permeate recovery and pH between 6.5 and 7.5

Table 4-11: Phase II RO Operation Run Information

Run ID	Filter Config	MF Flux (gfd)	Water Source	RO Flux (gfd)	RO Recovery
Phase II - Run 1	72- inch	30	Shelter Island	14	38%
Phase II - Run 2	90- inch	30	Shelter Island	10	26%
Phase II - Run 3	72-inch	30	South Bay	10	26%

Table 4-12: Analytical Methods / Detection Limits for measured Water Quality Parameters

Parameter	Units	Method Number and Type	Detection Limit
Total/Volatile Suspended Solids	mg/L	SM 2540D&E	1.6
Total Hardness	mg/L as CaCO3	EPA 130.1/130.2	0.3
Alkalinity	mg/L as CaCO3	SM 2320 B	1.5
TOC	mg/L	EPA 415.1	0.5
HPC	CFU/mL	SM 9215B	1 CFU/mL

[1] 20th Edition Addendum.

Table 5-1: Raw Water Quality at Three Sites (Ion Data)

	Units	DL	Scripps Pier		Shelter Island		South Bay- Inf	
			7/21/2004	11/4/2004	7/21/2004	11/4/2004	7/21/2004	11/4/2004
TDS	mg/L	10	36,700	34,500	36100	36400	39400	33500
Calcium, Total	mg/L	100	390	390	390	370	430	380
Iron, Total	mg/L	0.2	ND	ND	0.1	ND	0.17	13
Magnesium, Total	mg/L	10	1200	1200	1200	1100	1400	1200
Potassium, Total	mg/L	100	370	380	390	390	450	360
Sodium, Total	mg/L	100	9900	10000	10000	9800	11000	9400
Silica	mg/L	0.5	0.57	ND	1.5	4.8	2.3	5.8
Strontium	mg/L	0.05	7.1	7.2	7.3	6.7	7.9	6.7
Manganese, Total	µg/L	20	ND	ND	ND	ND	54	ND
Bromide	mg/L	1	66	65	65	63	69	61
Fluoride	mg/L	0.05	0.77	0.65	0.78	0.64	0.83	0.66
Boron	mg/L	0.5	4.5	ND	4.5	2.4	5	ND
Orthophosphate-P	mg/L	0.01	0.027	0.026	0.029	0.045	0.078	0.078
Chloride	mg/L	200	19000	1500	19000	14000	20000	17000
Sulfate	mg/L	400	2300	2200	2300	2200	2500	2200

Table 5-2: Raw Water Quality at Three Sites (Other Data)

	Units	DL	Scripps Pier		Shelter Island		South Bay- Inf	
			7/21/2004	11/4/2004	7/21/2004	11/4/2004	7/21/2004	11/4/2004
Temperature	deg C		NA	19	NA	19.4	NA	19.9
pH		0.001	NA	8.1	NA	8	NA	8.1
UV-254 Abs	cm^{-1}	0.001	0.009	0.009	0.015	0.024	0.039	0.07
Turbidity	NTU	0.05	0.68	0.86	0.95	1.14	1.72	56.4
Alkalinity	mg/L as $CaCO_3$	2	115	108	118	108	135	109
Color, True	PtCoCU	3	4	3	4	5	7	10
Color, Apparent	PtCoCU	3	6	8.1	9	8	14	8.1
TOC	mg/L	0.5	5.1	0.6	6	0.8	6.3	1.5
HPC	cfu/ml	1	NA	13	NA	16	NA	180

Table 5-3: Raw Water Quality Comparison Between South Bay Power Plant Influent and Effluent

		Units	DL	South Bay Influent	South Bay Effluent
	TDS	mg/l	10	33600	33500
Cations	Calcium, Total	mg/L	100	380	380
	Iron, Total	mg/L	0.2	13	ND
	Magnesium, Total	mg/L	10	1200	1200
	Potassium, Total	mg/L	100	360	360
	Sodium, Total	mg/L	100	9400	9400
	Silica	mg/L	0.5	5.8	2.5
	Strontium	mg/L	0.05	6.7	6.8
	Manganese, Total	mg/L	0.020	ND	ND
Anions	Bromide	mg/L	1.000	61.000	59.000
	Fluoride	mg/L	0.05	0.66	0.67
	Boron	mg/L	0.5	ND	ND
	Orthophosphate-P	mg/L	0.01	0.078	0.041
	Chloride	mg/L	200	17000	13000
	Sulfate	mg/L	400	2200	2200
Other Water Quality Data	Temp	deg C		19.9	25.3
	pH		0.001	8.1	8
	UV-254 Filt. Abs	cm^{-1}	0.001	0.07	0.056
	Turbidity	NTU	0.05	56.4	8.1
	Alkalinity	mg/L as $CaCO_3$	2	109	109
	Color, True	Pt Co CU	3	10	10
	Color, Apparent	Pt Co CU	3	10	10
	TOC	mg/L	0.5	1.5	1.3
	HPC	cfu/ml	1	180	260

Table 5-4: Results of SDI Analysis

$FeCl_3$ Dose	(mg/L) as Fe	0.0	2.0		4.0		6.0	10.0
7128 Polymer	(mg/L)	-	0.0	0.5	0.0	0.5	-	-
SDI_{15}		5.88	4.86	5.01	3.78	4.23	4.58	4.76

Table 5-5: Raw Seawater Quality for Alternate Coagulant Comparison

Coagulant	pH	Cond. mS/cm	UV254 cm^{-1}	Turb. NTU	Alk. mg/l
Feed	7.92	53.1	0.016	1.720	118
$FeCl_3$	7.75	53.3	0.014	0.301	101
$Fe_2(SO_4)_3$	7.74	53.3	0.014	0.280	107
$Al_2(SO_4)_3$	7.69	53.2	0.015	0.250	106
PACl	7.81	53.3	0.016	0.260	107

Table 5-6: Settled Water Quality for Alternate Coagulant Comparison

Coagulant	UV254 cm^{-1}	DOC mg/l	SUVA l/mg-m
$FeCl_3$	0.014	1.6	0.875
$Fe_2(SO_4)_3$	0.014	1.2	1.17
$Al_2(SO_4)_3$	0.015	1.1	1.36
PACl	0.016	0.99	1.61

Table 5-7: RO Ion Rejection Data

RO Membranes			Saehan	Osmonics	Hydranautics	Toray
Parameter	Unit	Raw seawater	Rejection (%)	Rejection (%)	Rejection (%)	Rejection (%)
TDS	mg/L	36413	97.9%	98.3%	**98.7%**	97.8%
DOC	mg/L	1.32	85.6%	**91.7%**	89.4%	87.9%
Calcium	mg/L	380	99.2%	99.3%	**99.4%**	98.4%
Magnesium	mg/L	1200	99.2%	99.3%	**99.4%**	98.5%
Sodium	mg/L	11000	98.4%	98.6%	**98.9%**	97.5%
Potassium	mg/L	440	97.9%	98.6%	**98.7%**	97.3%
Strontium	mg/L	7.3	99.3%	99.3%	**99.5%**	99.2%
Boron	mg/L	5.0	85.2%	83.6%	**90.2%**	88.0%
Chloride	mg/L	20000	98.3%	98.7%	**98.9%**	97.5%
Sulfate	mg/L	2400	97.8%	99.4%	**99.6%**	98.7%
Nitrate	mg/L	2.0	91.0%	90.0%	**91.5%**	91.0%
Phosphate	mg/L	0.7	97.1%	97.1%	**97.1%**	97.1%
Bromide	mg/L	63.0	96.7%	98.4%	**98.7%**	97.1%
Fluoride	mg/L	6.7	> 98.5%	> 98.5%	**> 98.5%**	> 98.5%

Table 5-8: Seawater Fraction TOC and UV 254 Results

Fraction	TOC mg/L	UV254 cm^{-1}
1.0um	6.6	0.016
0.1um	5.7	0.015
100kDa	6.1	0.014
20kDa	6.2	0.015

Table 6-1: Raw and Settled Seawater Quality for Pretreatment Evaluation Phase : Shelter Island Water

	units	Raw Water		Settled Water		n
		Average	Range	Average	Range	
Temp	deg C	24.5	20.4 - 31.4	23.9	19.6 - 30.4	77
Conductivity	mS/cm	49.3	48.2 - 52.3	48.8	41.4 - 52	68
pH		8.0	7.68 - 8.31	7.9	7.39 - 8.28	73
UV-254	cm-1	0.024	0.014 - 0.035	0.021	0.013 - 0.03	47
Turbidity	NTU	3.48	0.972 - 8.79	3.16	1.24 - 7.44	74

Table 6-2: Raw and Settled Seawater Quality for Pretreatment Evaluation Phase : South Bay Water

	units	Raw Water		Settled Water		n
		Average	Range	Average	Range	
Temp	deg C	22.4	21 - 24.2	22.2	20.3 - 23.8	6
Conductivity	mS/cm	44.3	42.4 - 45.7	44.0	42.5 - 45.7	6
pH		N/A	N/A	N/A	N/A	N/A
UV-254	cm-1	0.046	0.043 - 0.05	0.043	0.041 - 0.048	6
Turbidity	NTU	5.39	2.98 - 10.8	4.11	3.04 - 5.29	6

Table 6-3: Raw Seawater Quality for Pretreatment Evaluation (Lab Data): Shelter Island Water

	units	Average Value	Range	n
Conductivity	mS/cm	50.2	48.3 - 52.6	13
Total Dissolved Solids	mg/L	35069	31100 - 37000	13
Total Hardness	mg/L as $CaCO_3$	6281	6080 - 6420	13
Total Alkalinity	mg/L as $CaCO_3$	112	109 - 114	13
TSS	mg/L	8.32	1.95 - 20.6	13
HPC	cfu/mL	2626	24 - 15000	13
TOC	mg/L	5.8	4.2 - 6.9	8

Table 6-4: Summary of Conventional Filter Runs

Run No	Filter Configuration	Water Source	run time (hrs)		turbidity (NTU)		SDI	
			avg	range	avg	range	avg	range
1-20	72 inch filter	Shelter Island	79	64-118	0.049	0.024-0.079	3.8	2.0-5.3
21-23	90 inch filter	Shelter Island	78	57-105	0.059	0.042-0.07	3.02	2.8-3.2
24,25	90 inch filter	South Bay	101	76-126	0.068	0.065-0.069	3.6	3.4-3.8

Table 6-5: Comparison of selected lab data for MF and conventionally filtered Seawater during Phase I testing (Shelter Island Water)

	units	Conventional Filtrate		MF Filtrate	
		Average Value	Range	Average Value	Range
TSS	mg/L	6.20	1.6 - 15	7.02	2.36 - 12.2
HPC	cfu/mL	403	8 - 1600	23	1 - 98
TOC	mg/L	6.0	3.3 - 8.1	5.7	3.6 - 6.8

Table 6-6: RO run Data During RO operation phase

Run ID	Filter Config	MF Flux (gfd)	Water Source	RO Flux (gfd)	RO Recovery
Phase II - Run 1	72- inch	30	Shelter Island	14	38%
Phase II - Run 2	90- inch	30	Shelter Island	10	26%
Phase II - Run 3	72-inch	30	South Bay	10	26%

Table 6-7: MF RO Feed Water Quality During RO operation phase

	units	Average Value	Range	n
Conductivity	mS/cm	51.6	47.7 - 53.2	9
Total Dissolved Solids	mg/L	35711	33500 - 37400	9
Total Hardness	mg/L as $CaCO_3$	6182	5960 - 6320	9
Total Alkalinity	mg/L as $CaCO_3$	115	111 - 128	9
HPC	cfu/mL	329	135 - 750	5
TOC	mg/L	6.2	4.1 - 9.1	3
Chl-A	mg/m^3	2.0	0 - 5.34	8

Table 6-8: Conventional RO Feed Water Quality During RO operation phase

	units	Average Value	Range	n
Conductivity	mS/cm	52.0	47.8 - 54	9
Total Dissolved Solids	mg/L	35867	33300 - 37900	9
Total Hardness	mg/L as $CaCO_3$	6260	5950 - 6699	9
Total Alkalinity	mg/L as $CaCO_3$	110	103 - 113	9
HPC	cfu/mL	360	190 - 630	4
TOC	mg/L	6.5	5 - 8.8	3
Chl-A	mg/m^3	1.8	0 - 5.34	9

Table 6-9: MF RO permeate Quality During RO operation phase

	units	Average Value	Range	n
Conductivity	mS/cm	288.8	229 - 402	8
Total Dissolved Solids	mg/L	125	70 - 190	8
Total Hardness	mg/L as $CaCO_3$	10	9 - 11	2
Total Alkalinity	mg/L as $CaCO_3$	15	8.49 - 17	8

Table 6-10: Conventional RO permeate Quality During RO operation phase

	units	Average Value	Range	n
Conductivity	mS/cm	331.4	254 - 399	8
Total Dissolved Solids	mg/L	161	120 - 230	8
Total Hardness	mg/L as $CaCO_3$	20	18 - 21	2
Total Alkalinity	mg/L as $CaCO_3$	10	8.36 - 12.6	8

Table 6-11: Water Quality Data from Run 2

	units	DL	RO Feed	Conventional RO	MF RO
Specific Conductance	µmho/cm	2	44300	284	260
TDS	mg/L	10	38000	160	130
Alkalinity	mg/L as CaCO$_3$	2	110	ND	ND
Total Hardness	mg/L as CaCO$_3$	3	6350	16.1	10.7
pH			8.3	7.9	8
Anion Sum	meq/L	0.001	590	2.34	2.34
Cation Sum	meq/L	0.001	572	2.6	2.32
Calcium	mg/L	10	400	1	ND
Magnesium	mg/L	1	1300	3.3	2.6
Manganese	mg/L	0.010	0.01	ND	ND
Potassium	mg/L	10	410	2.5	2.6
Silica	mg/L	0.5	0.95	ND	ND
Sodium	mg/L	100	10,000	51	47
Strontium	mg/L	0.1	7.5	0.018	0.014
Boron	mg/L	20	5.1	0.46	0.46
Chloride	mg/L	200	19000	83	80
Sulfate	mg/L	400	2500	ND	3.9
Bromide	mg/L	1	68	0.33	0.31
Fluoride	mg/L	0.050	0.63	ND	ND
Orthophosphate	mg/L	0.010	0.017	0.018	0.017
Nitrate	mg/L	0.44	ND	ND	ND
Apparent Color	PtCo units	3	3	ND	ND
True Color	PtCo units	3	3	ND	ND

Table 7-1: Capital and O&M cost Component codes

O&M Cost Components	Includes	O&M Cost Codes
Cleaning chemicals	Citric Acid, NaoH, Other proprietary and generic chemicals for pretreatment as well as SWRO process	1
Power	Electrical consumption for pumping as well as instrumentation and auxiliary equipment	2
Labor	Regular operations, maintenance and management	3
Cartridge filter replacement	Cost of Cartridge Filters	4
Replacement parts	Spare parts for mechanical equipment	5
Membrane replacement	Membrane Element Replacement costs	6
Chemical Addition	Including pretreatment chemicals (coagulant, chlorine, bisulfite), RO chemicials (antiscalant, acid) and posttreatment chemicals (lime, NaoH, etc)	7

Capital Cost Components	Includes	Capital Cost Codes
# of Membranes	All RO membrane Elements	A
Foot Print	Size of the plant	B
Vessels	RO vessels	C
Equipment	Including pumps, monitoring equipment, and cleaning equipment	D
Pretreatment	pretreatment train cost	E

Table 7-2: Capital and O&M Cost Factors

Cost Factor	Associated O&M Costs	Associated Capital Costs	IMPACTED BY PRETREATMENT (Y/N?)
Membrane Properties: (a) specific flux, (b) temperature tolerance, (c) salt rejection, (d) fouling resistant treatment, (e) effective membrane area	1,2,3,6	A,B,C	No
Operational Flux	1, 2, 3, 4	A,B,D	Yes
Feed Water Recovery	1,2 ,3	E	Yes
Specific Flux Decline Rate (Fouling Rate)	1,2,3,6	none	Yes
RO Feed Water Quality (salinity, temperature, turbidity, SDI, PC, Biological, Organics)	1,2,3,4,5,6	A,B,C,D, E	Yes
Permeate Water Quality Goals	2,7	A,B,C,D	No.
Membrane Life	2,6	none	Yes.
Chemical Costs (biocide, antiscalant, acid etc)	3,7	D	Yes.
Post Treatment	7	D	No
Membrane Configuration : Number of elements/vessel; single or two pass	none	A,B,C	No
Type of energy recovery equipment	2	D	No
Seawater intake and ocean outfall location	2	D	No
Characteristics of the product conveyance line	2	D	No
Plant Utilization Factor		affect cost / gallon	No
Product Water Blend Ratio	1,2,3,4,5,6,7	A,B,C,D, E	No
Type of Intake	1,2,3,4,6,7	E	No

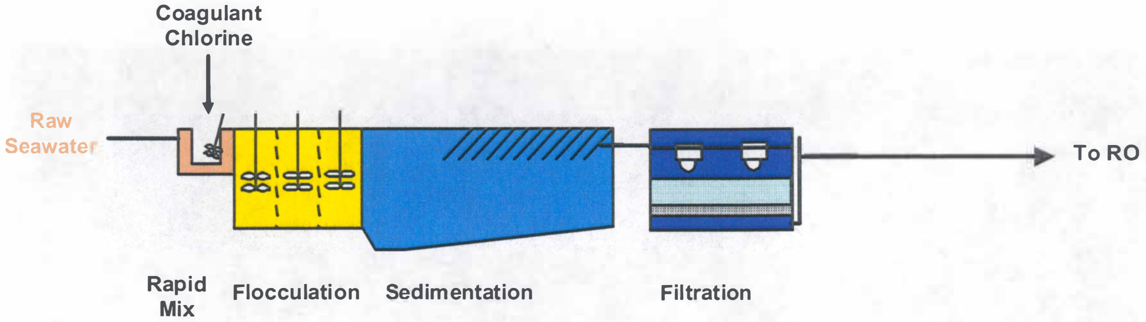


Figure 2-1: Schematic of typical conventional SWRO pretreatment system.

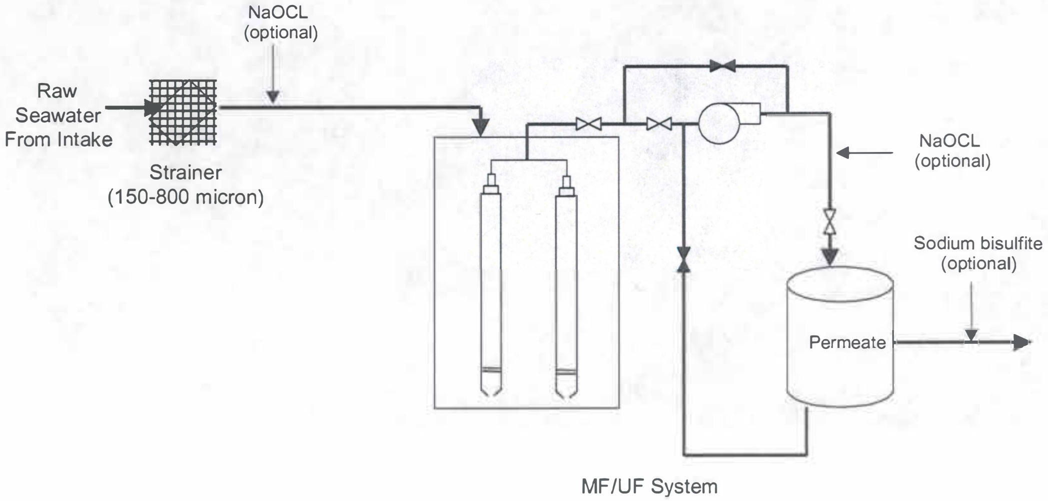


Figure 2-2: Schematic of typical membrane SWRO pretreatment system.

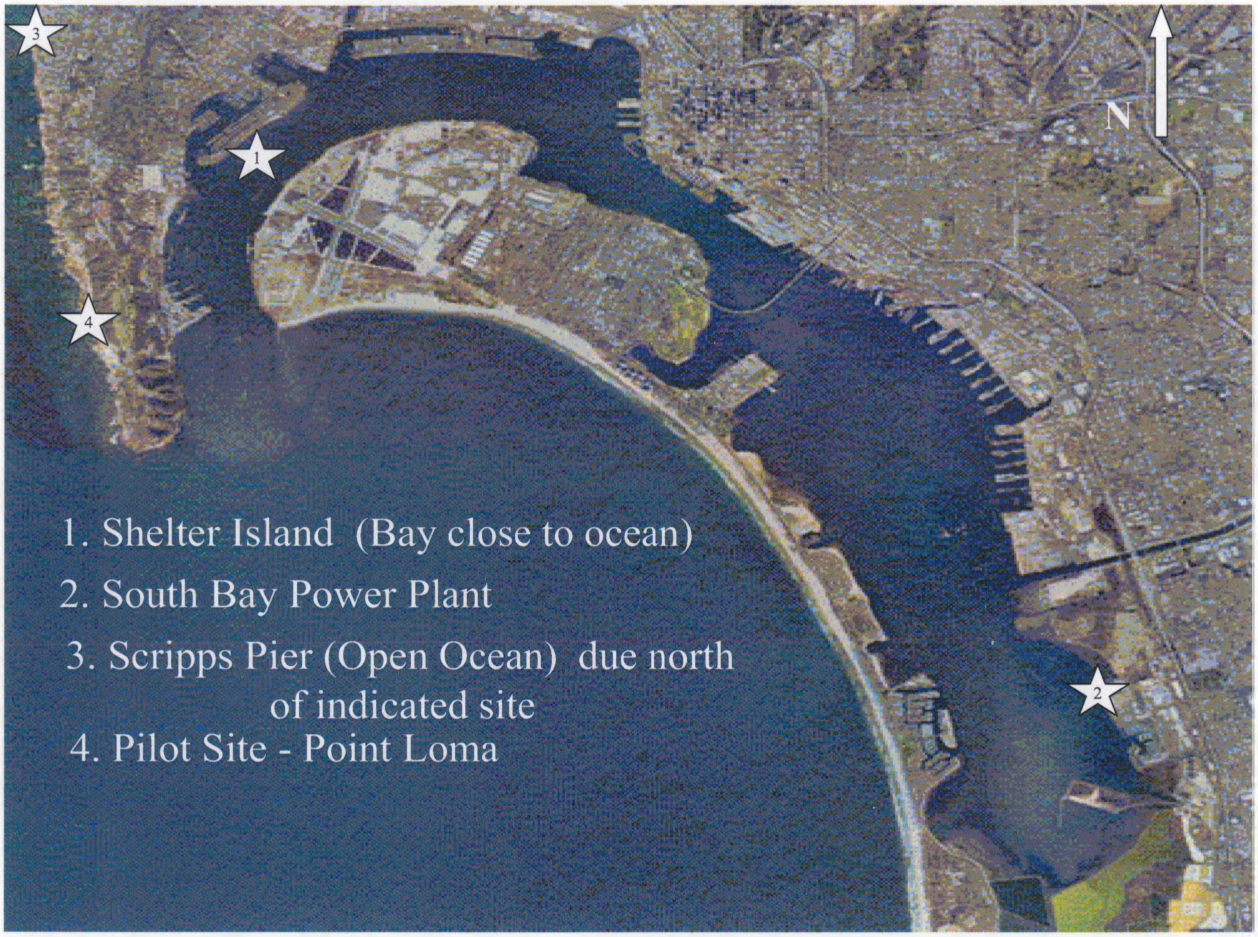


Figure 4-1: San Diego Bay and locations of sampling and pilot sites

Figure 4-2: Shelter Island Boat Ramp Sample Collection site

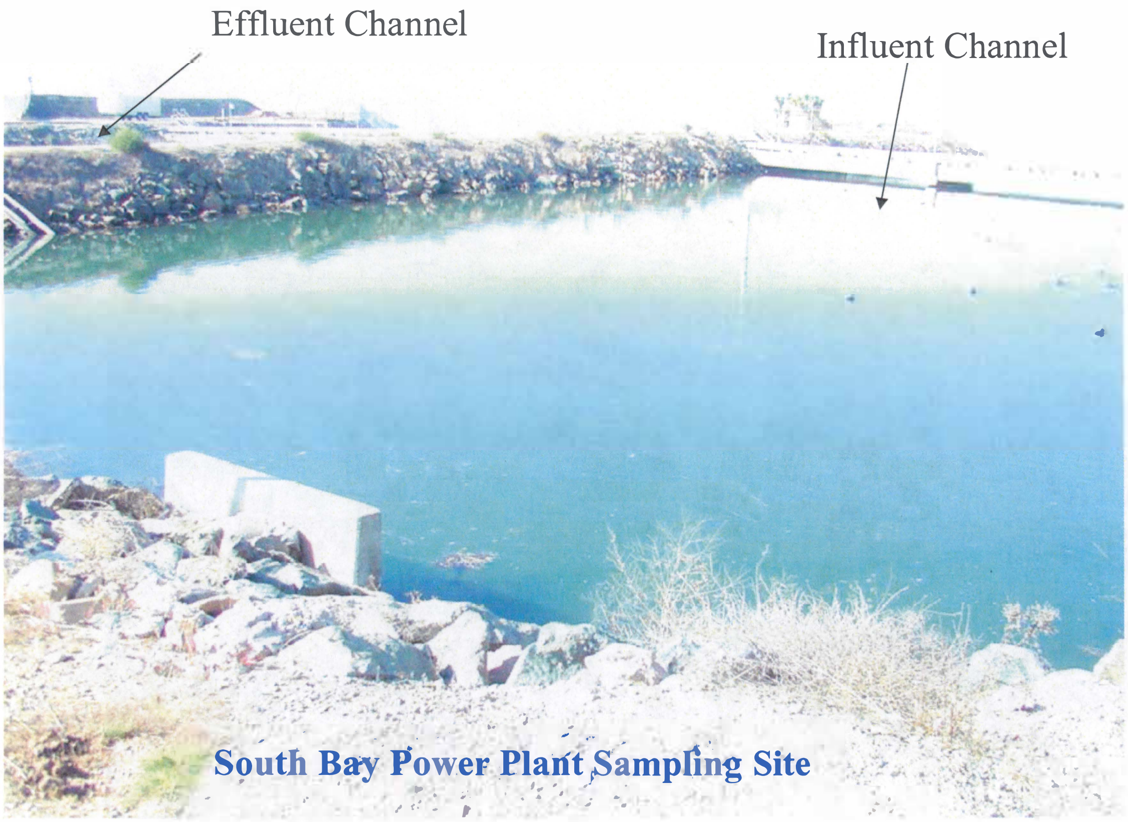


Figure 4-3: South Bay Power Plant Sample Collection site

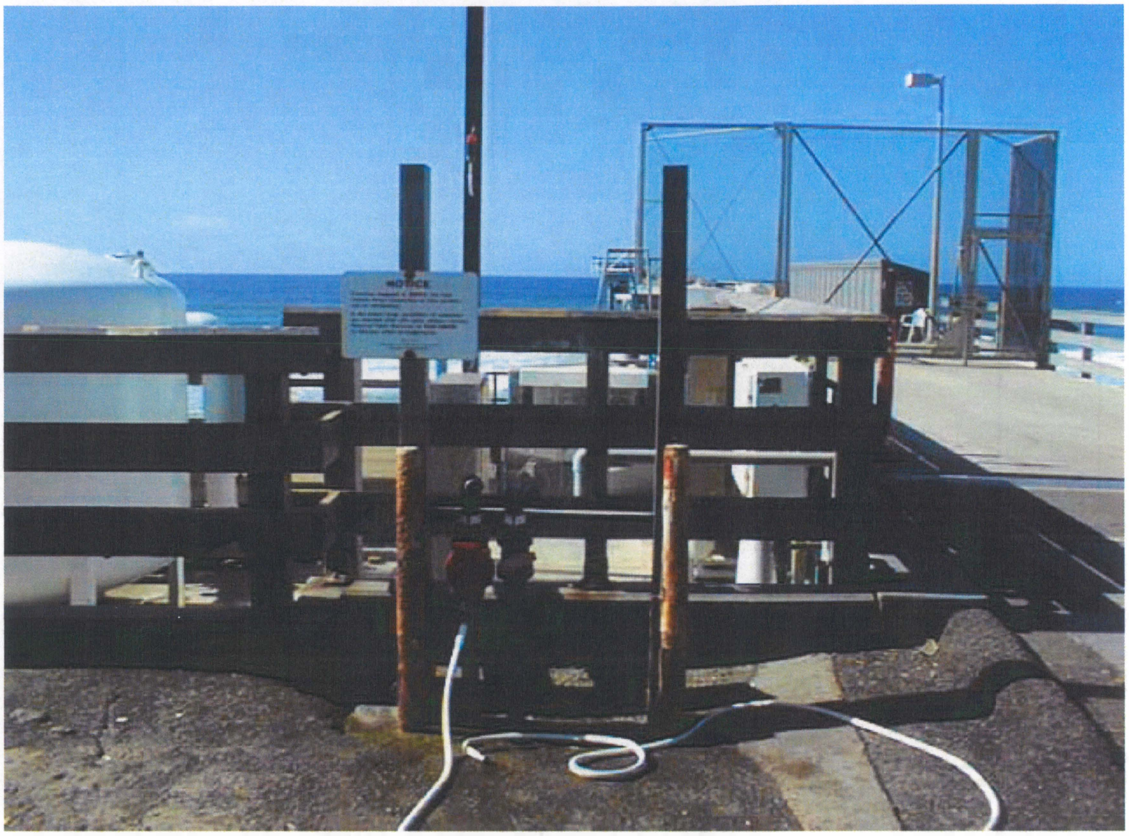


Figure 4-4: Scripps Pier Sample Collection site

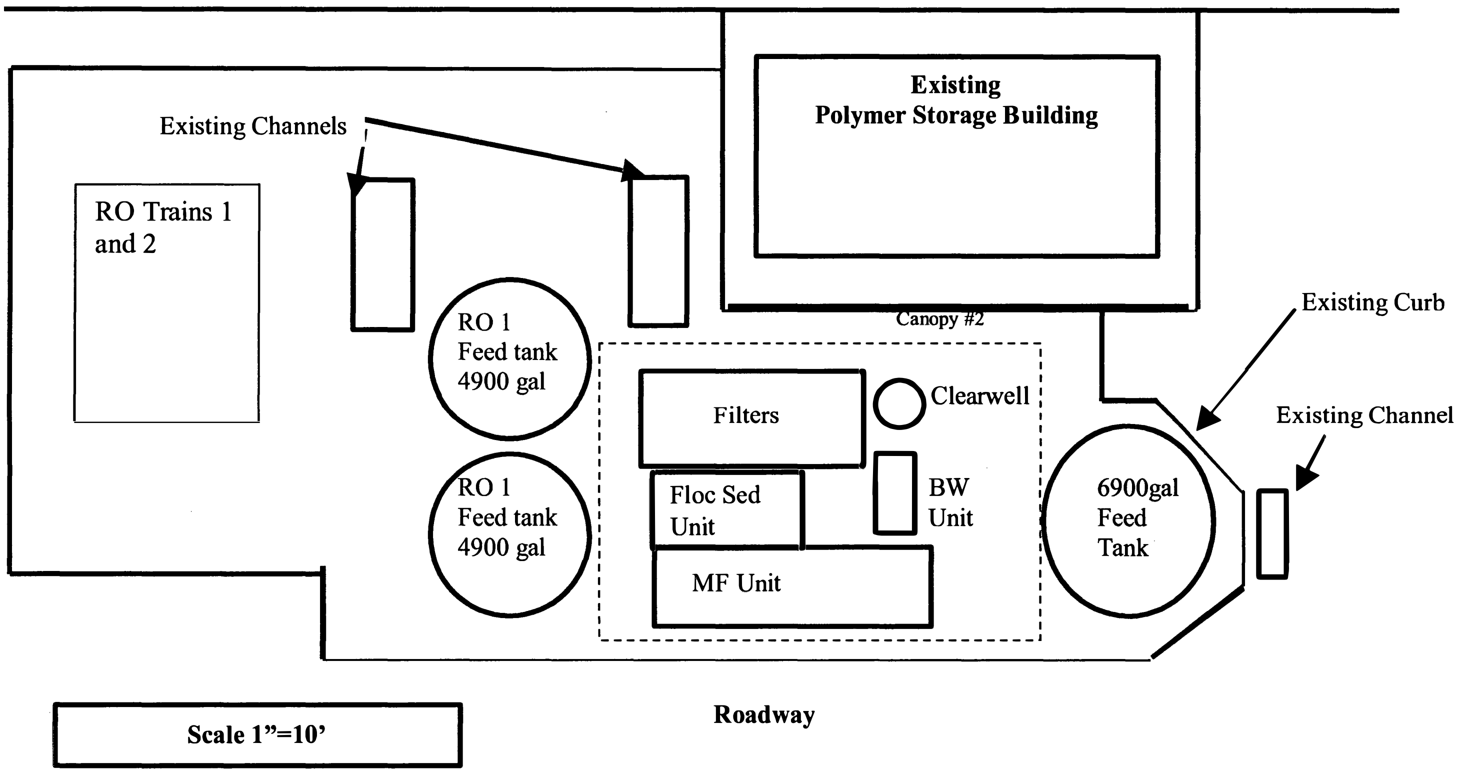


Figure 4-5: Pilot Site Layout

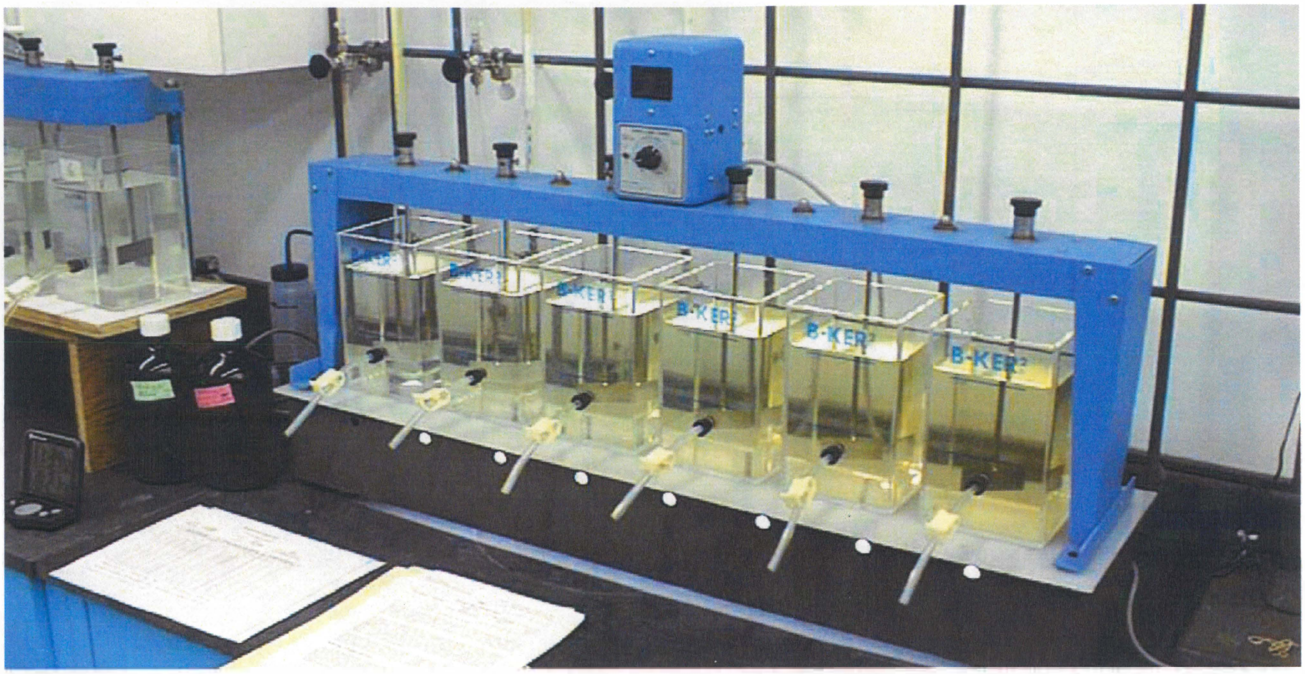


Figure 4-6: Jar Testing Apparatus

Figure 4-7: Osmonics Flat Sheet Sepa® Membrane Cell System

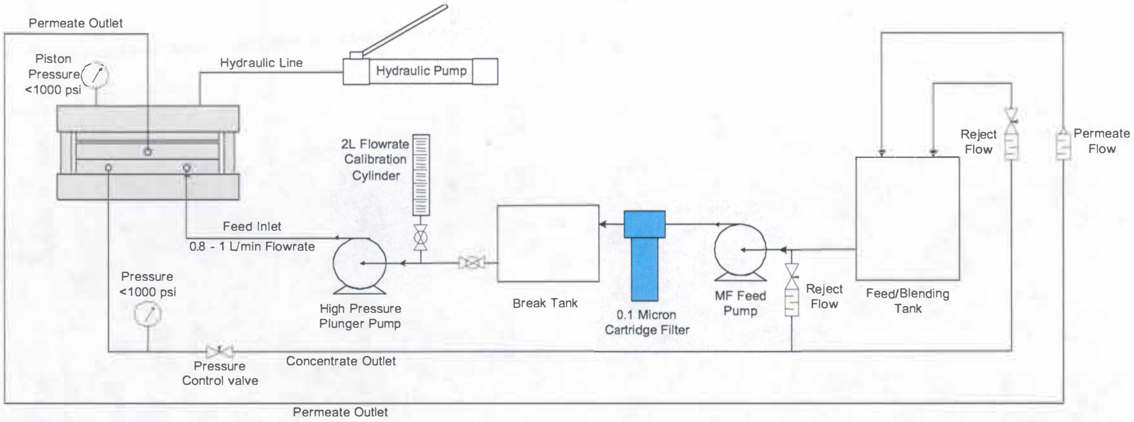


Figure 4-8: BAIReMT Experimental Setup

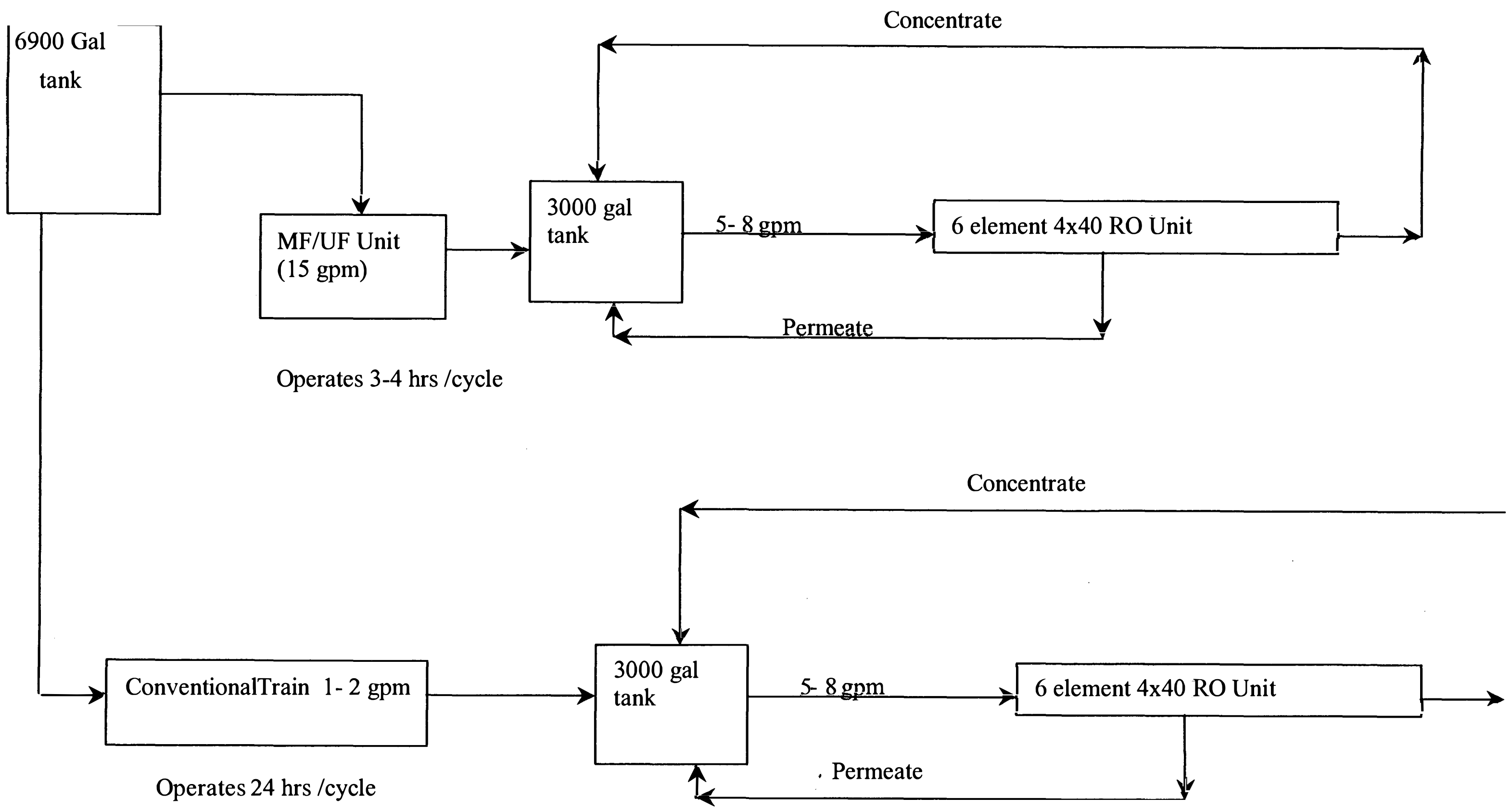
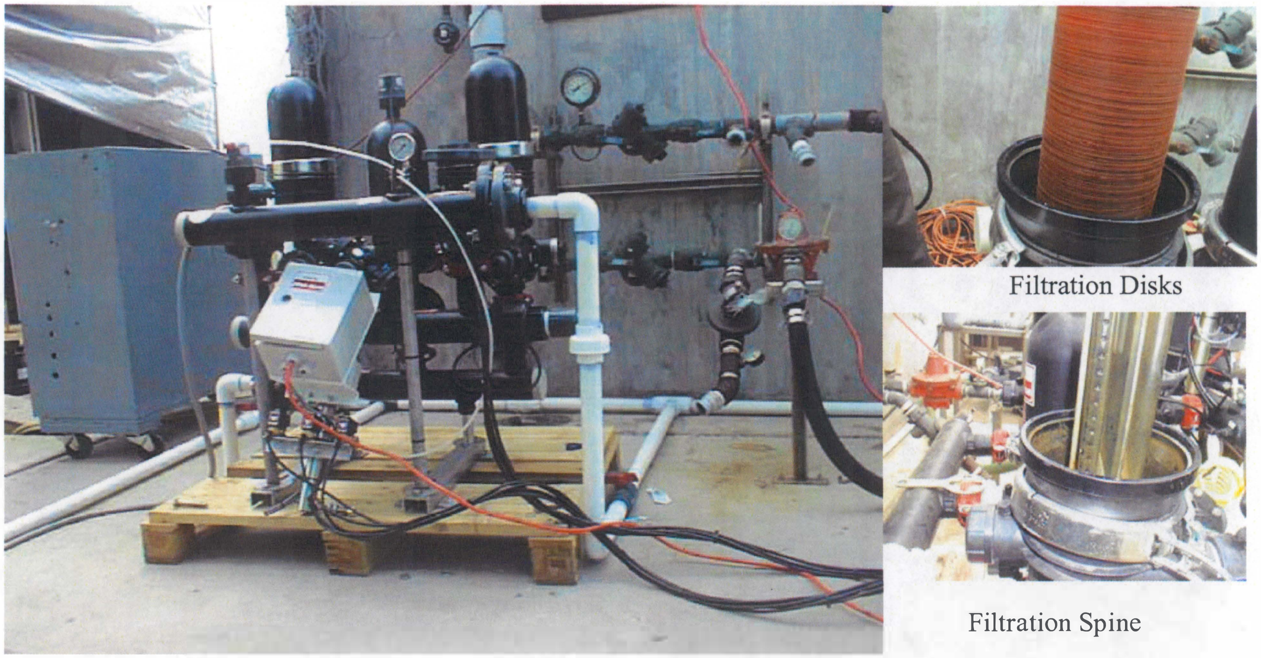


Figure 4-9: Pilot Testing Treatment Train Schematic

Figure 4-10: Coagulation/Flocculation/Sedimentation Unit Setup

Figure 4-11: Dual Media Filtration Unit Setup



Filtration Disks

Filtration Spine

Figure 4-12: Spin Klin Disk Filtration System

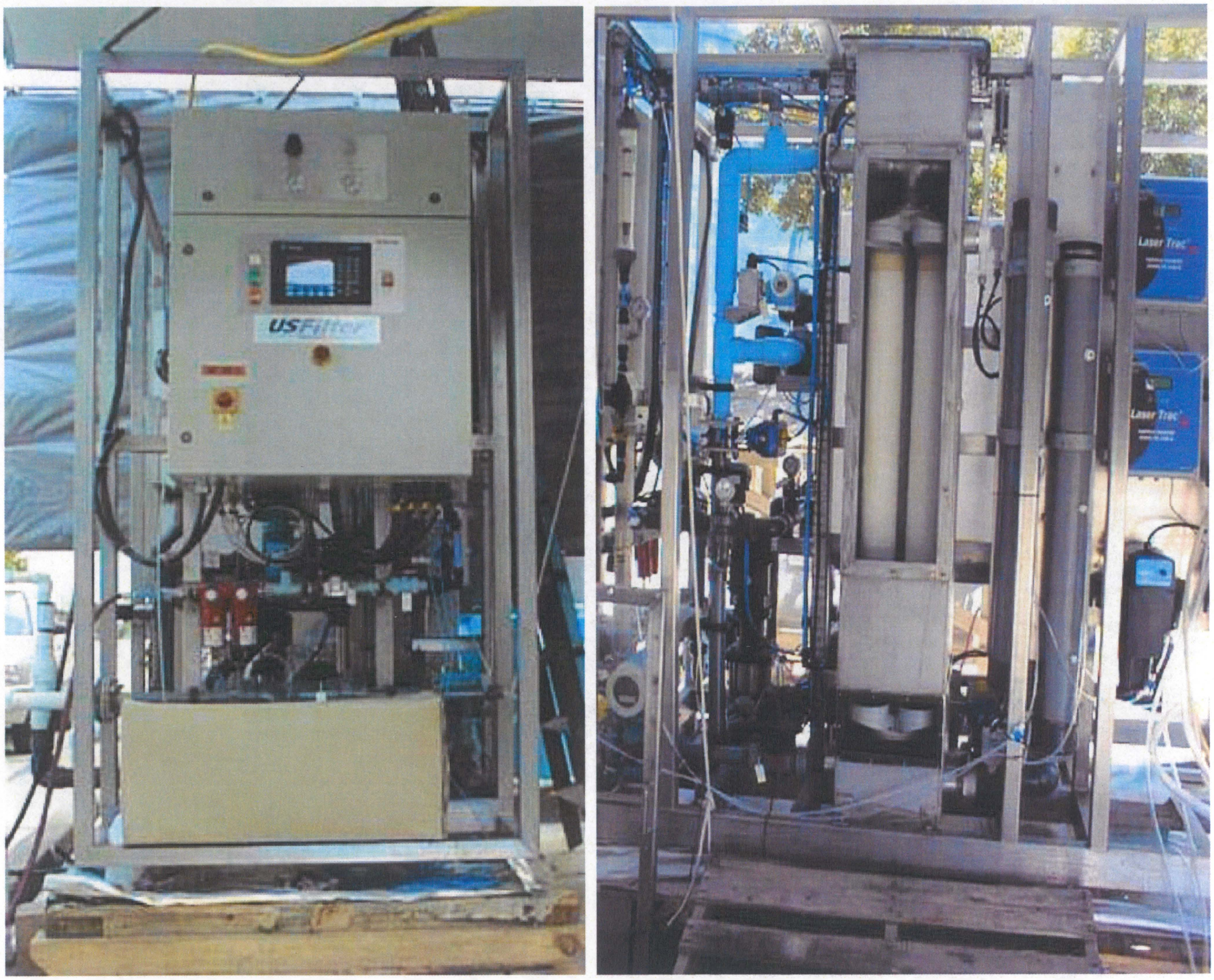


Figure 4-13: MF System (US Filter)

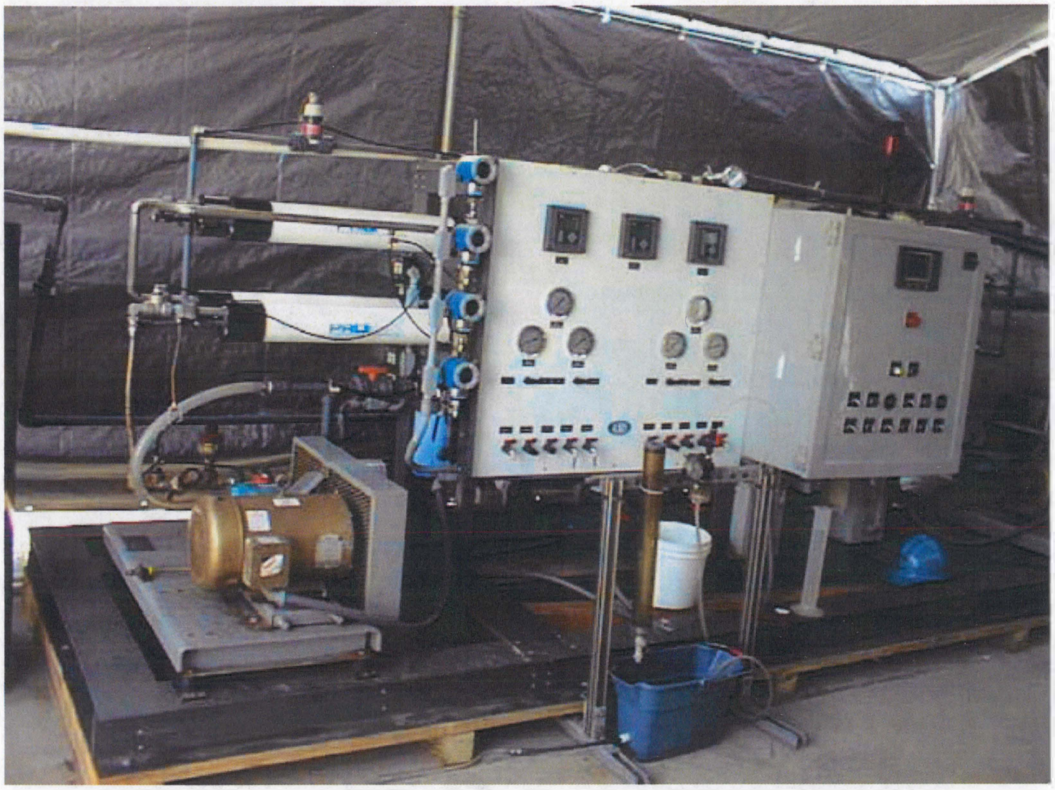


Figure 4-14: RO Pilot

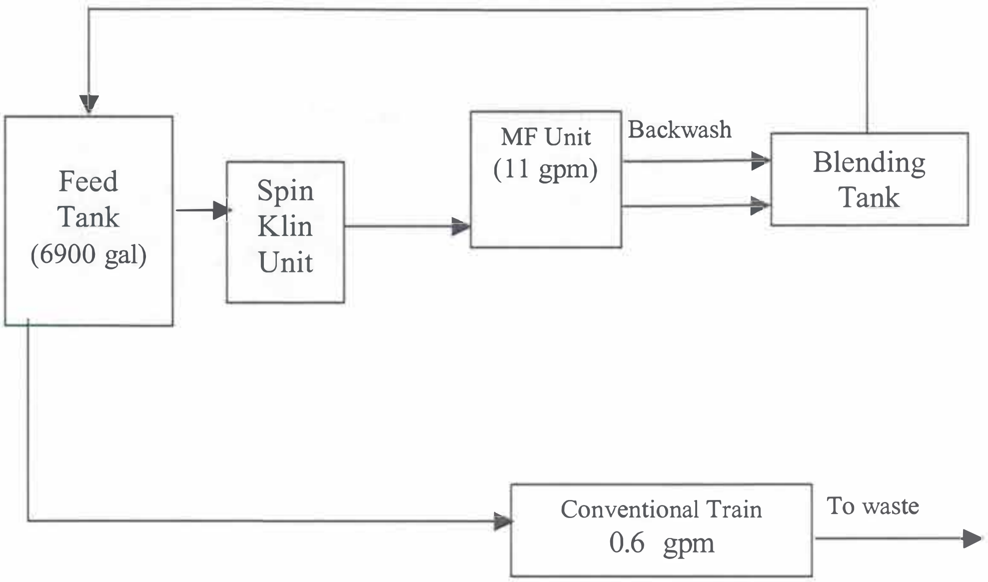


Figure 4-15: Conventional Pretreatment Phase: Pilot Setup

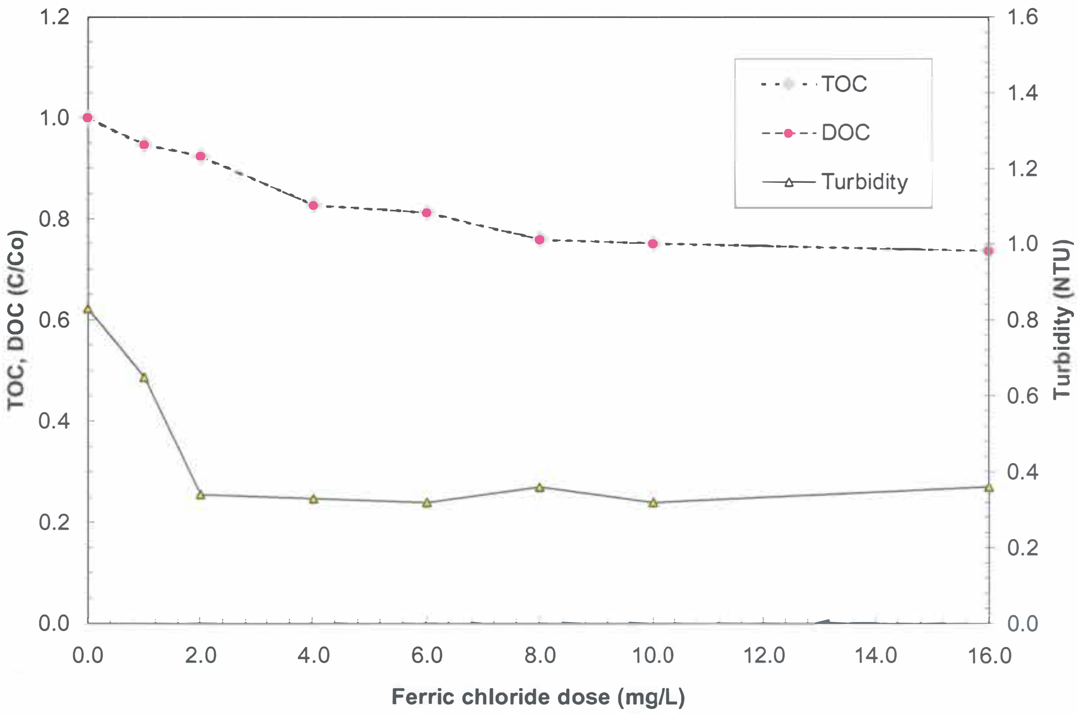


Figure 5-1: Optimization of Ferric Chloride dose

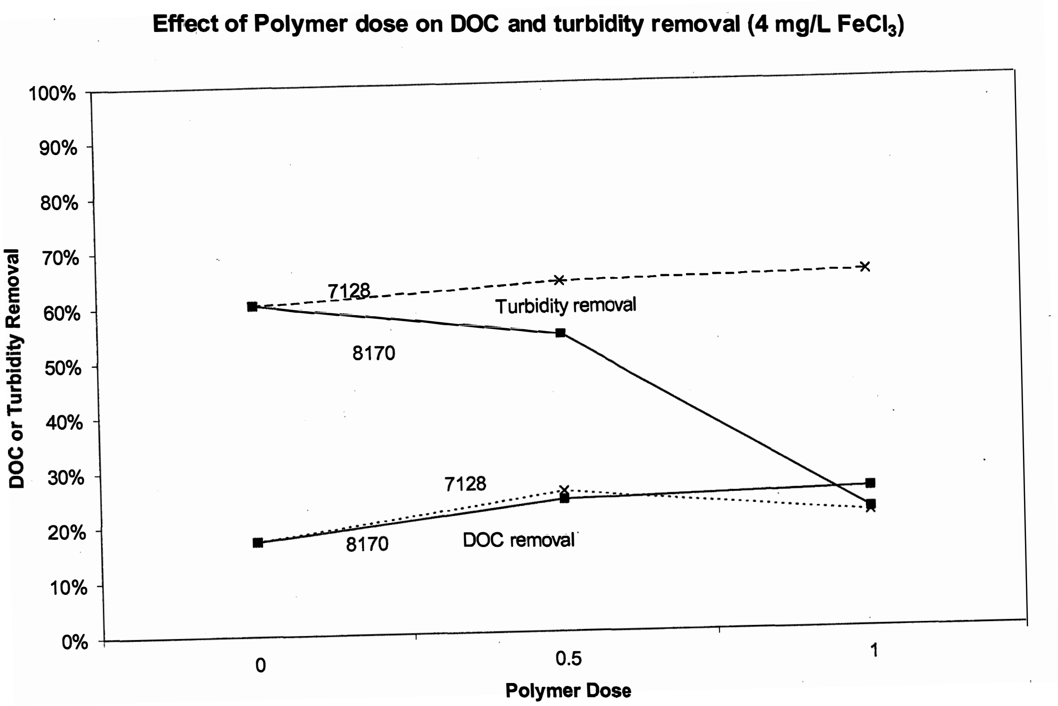


Figure 5-2: Effect of Polymer addition

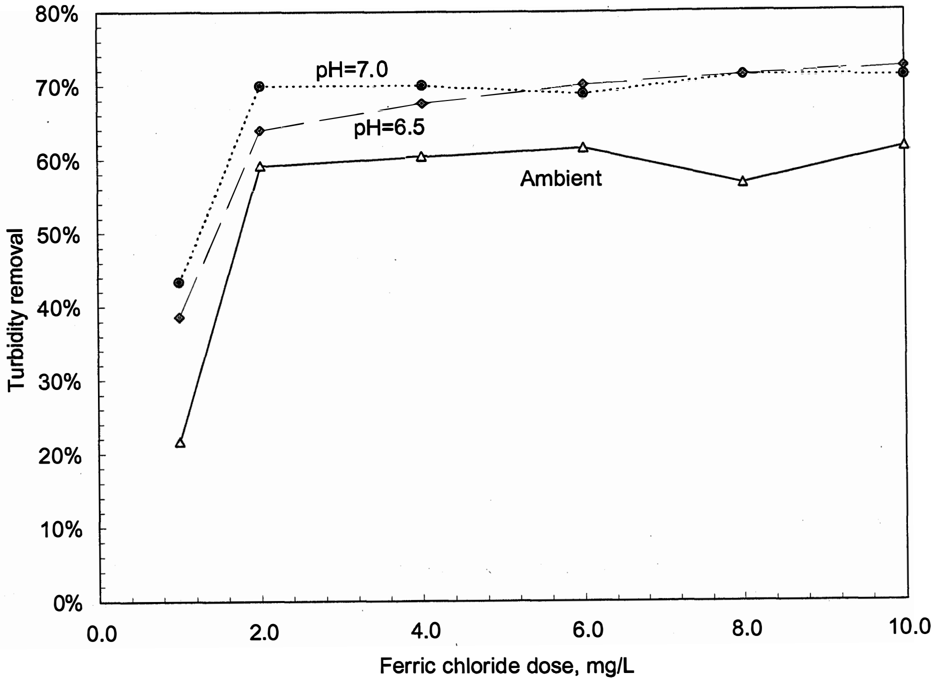


Figure 5-3: Results of pH adjustment- Turbidity removal

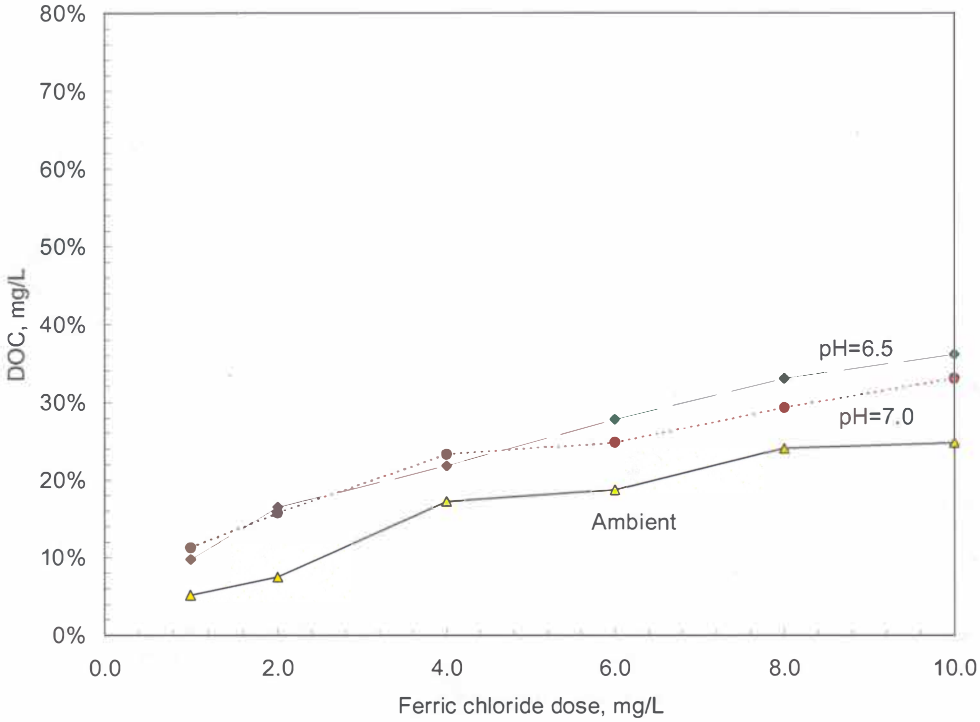


Figure 5-4: Results of pH adjustment- DOC removal

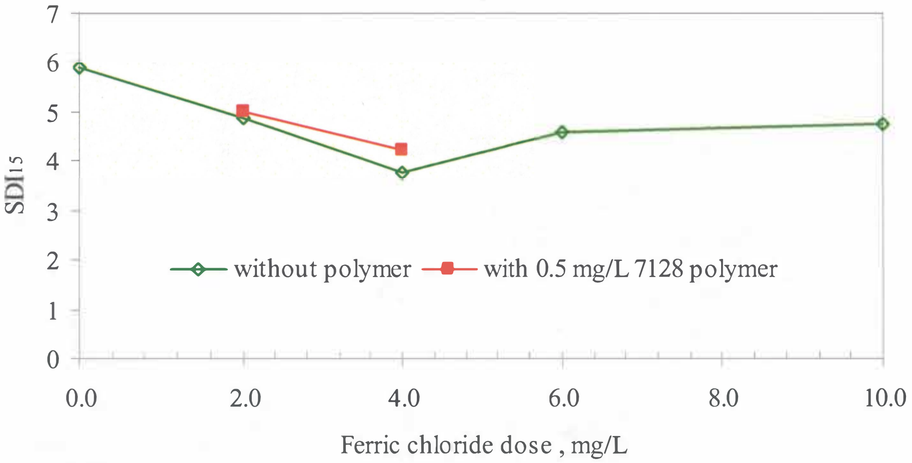


Figure 5-5: Results: Effect of ferric dose on SDI

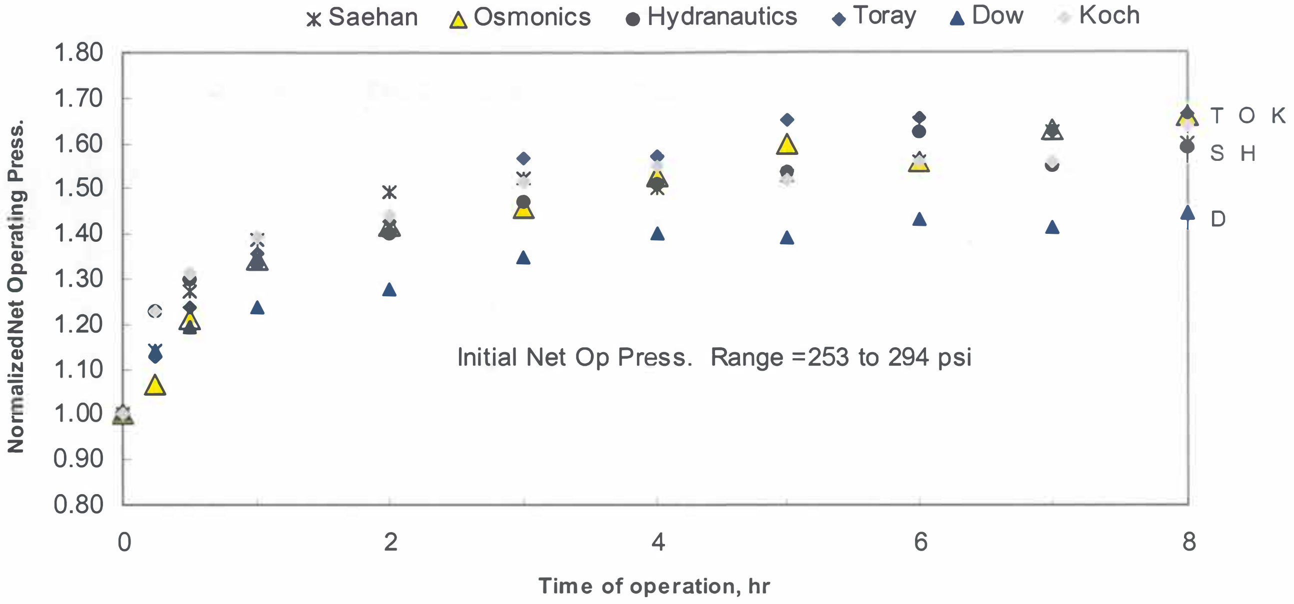


Figure 5-6: Results of RO membrane testing- Net Operating Pressure

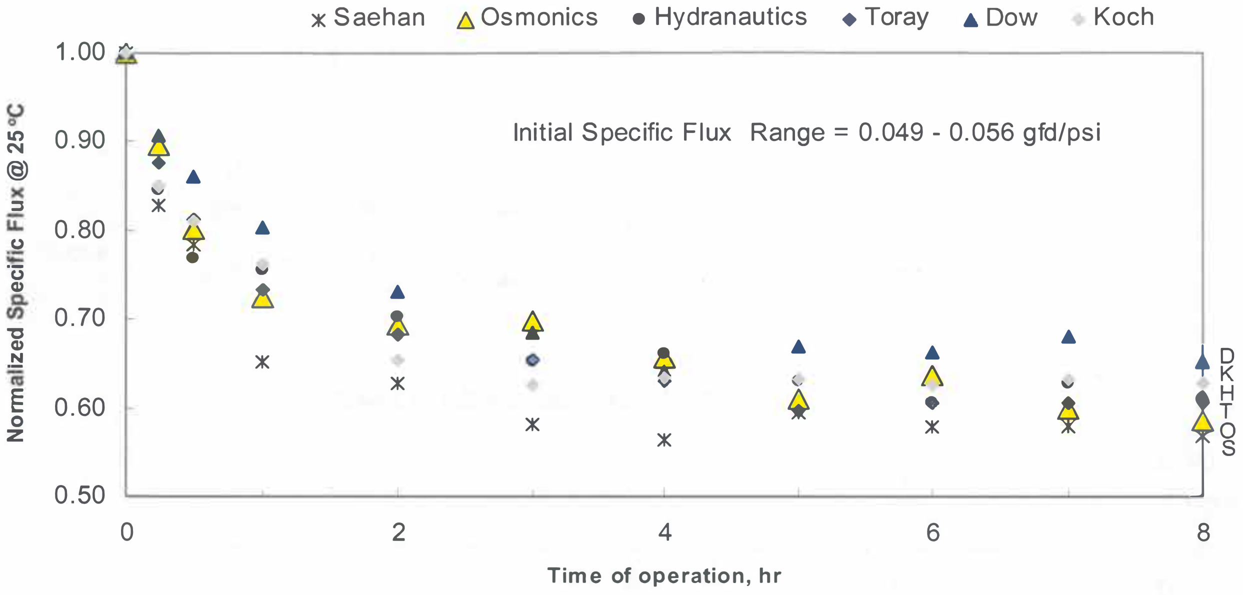


Figure 5-7: Results of RO membrane testing- Normalized Specific Flux Decline at 25 deg C

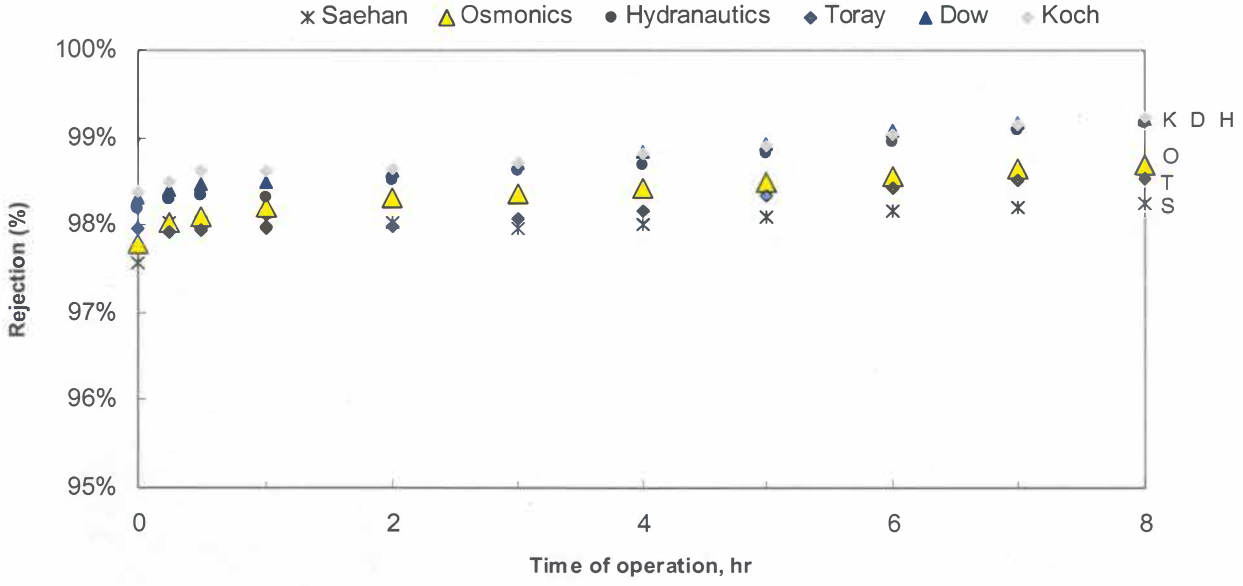


Figure 5-8: Results of RO membrane testing- Rejection

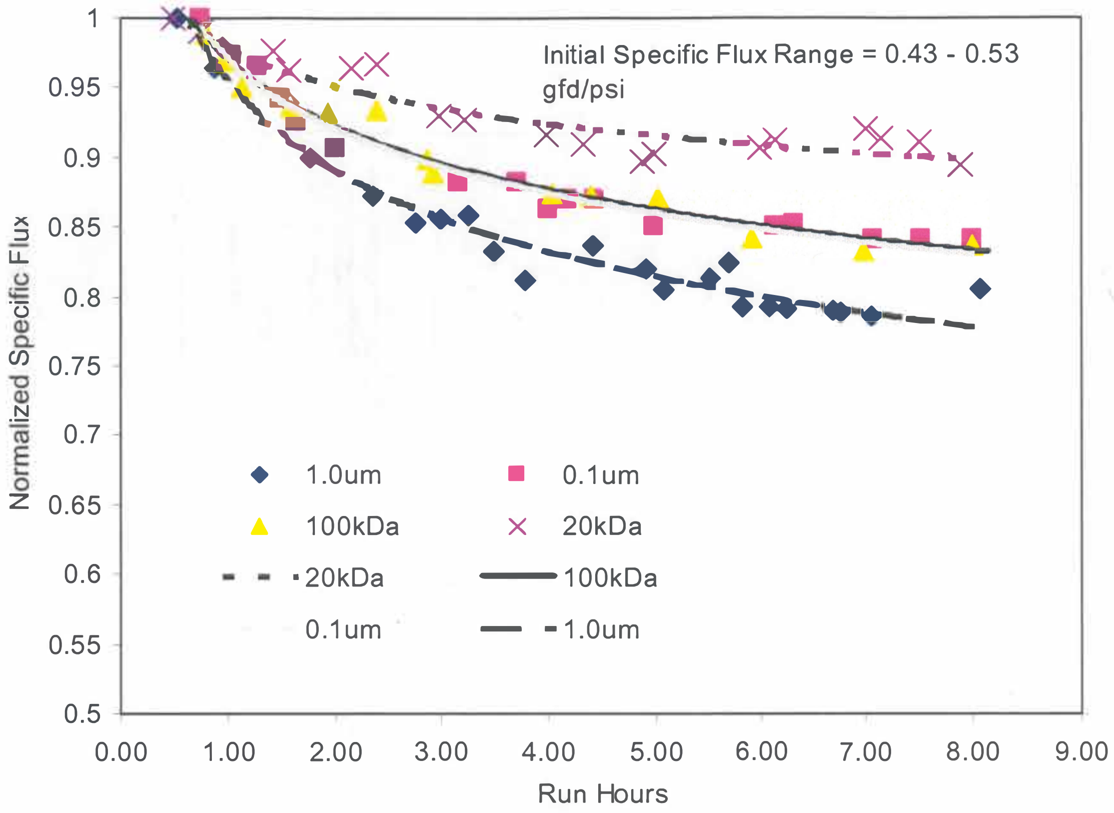


Figure 5-9: Fractionation RO Results: Average Normalized Eight Hour Flux Decline

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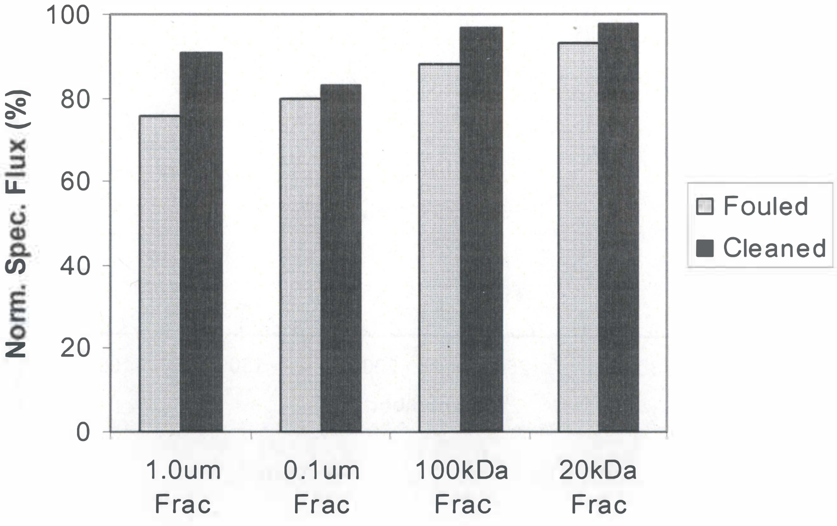


Figure 5-10: Fractionation RO Results: Cleaning Results

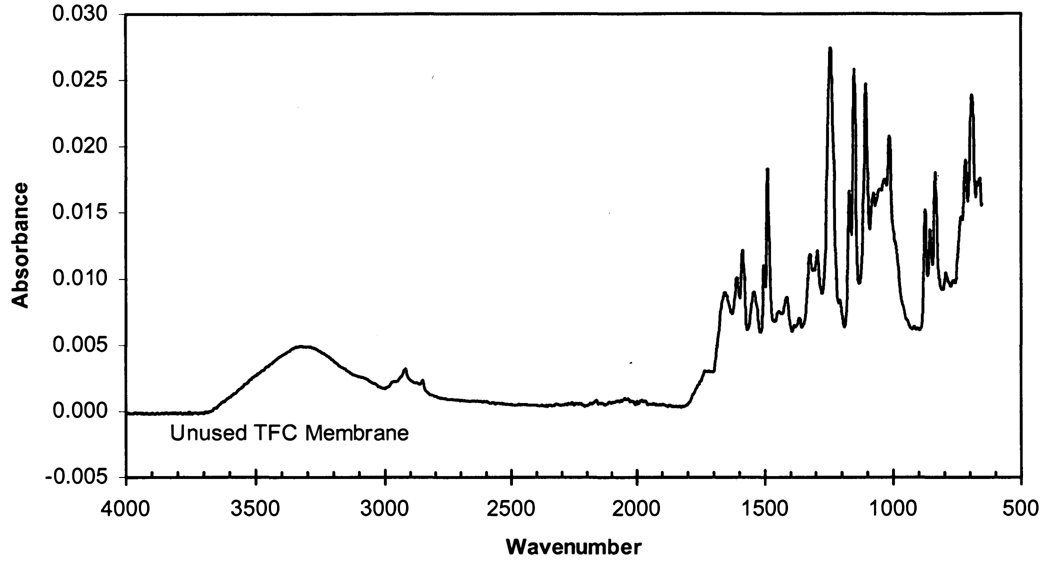


Figure 5-11: ATR/FTIR spectrum of an unused thin-film composite membrane

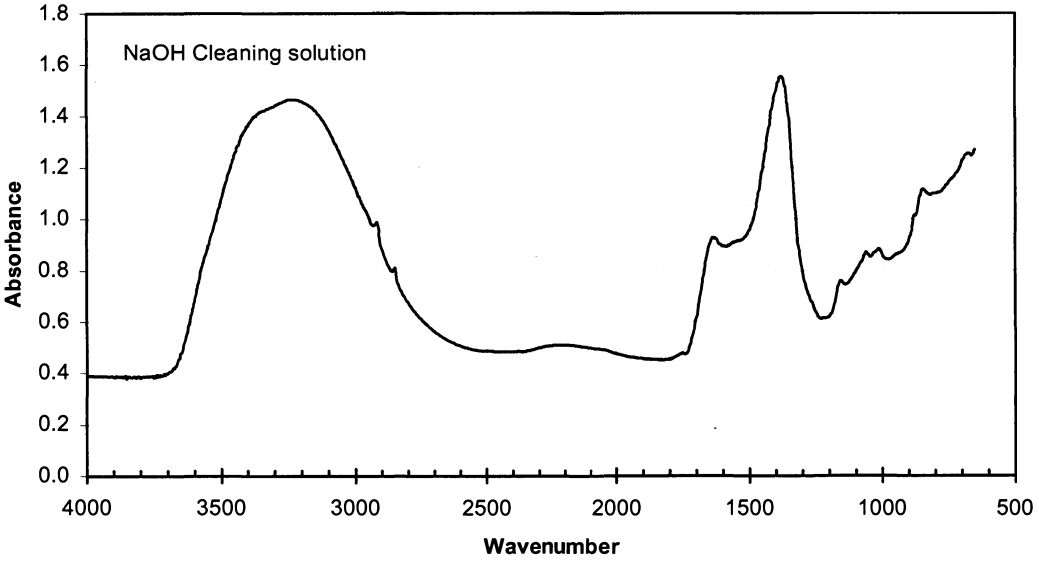


Figure 5-12: ATR/FTIR spectrum of the NaOH cleaning solution

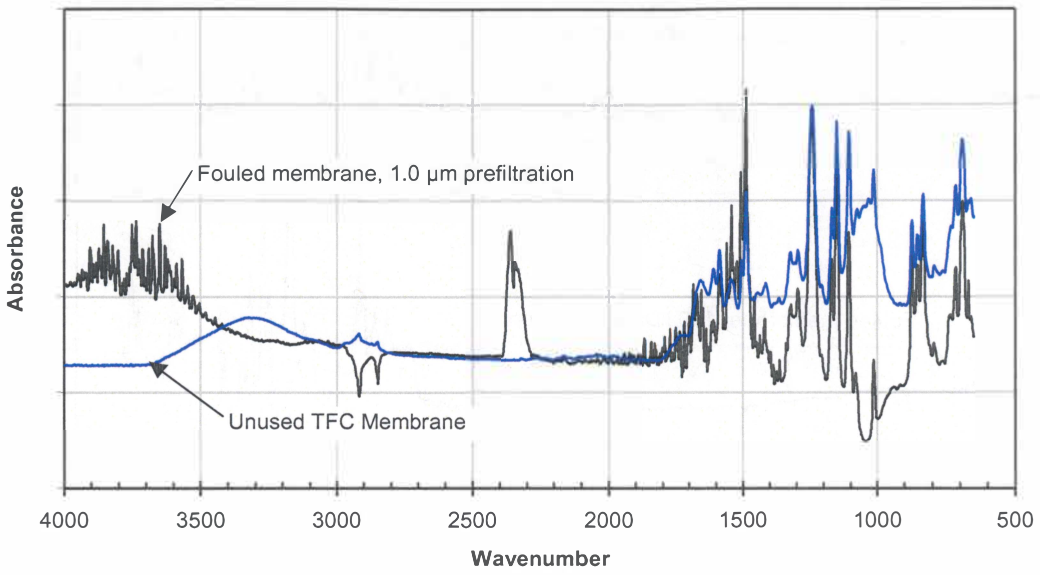


Figure 5-13: Comparative ATR/FTIR spectrum of an RO membrane fouled after 1.0 µm membrane prefiltration and unused thin-film composite membrane

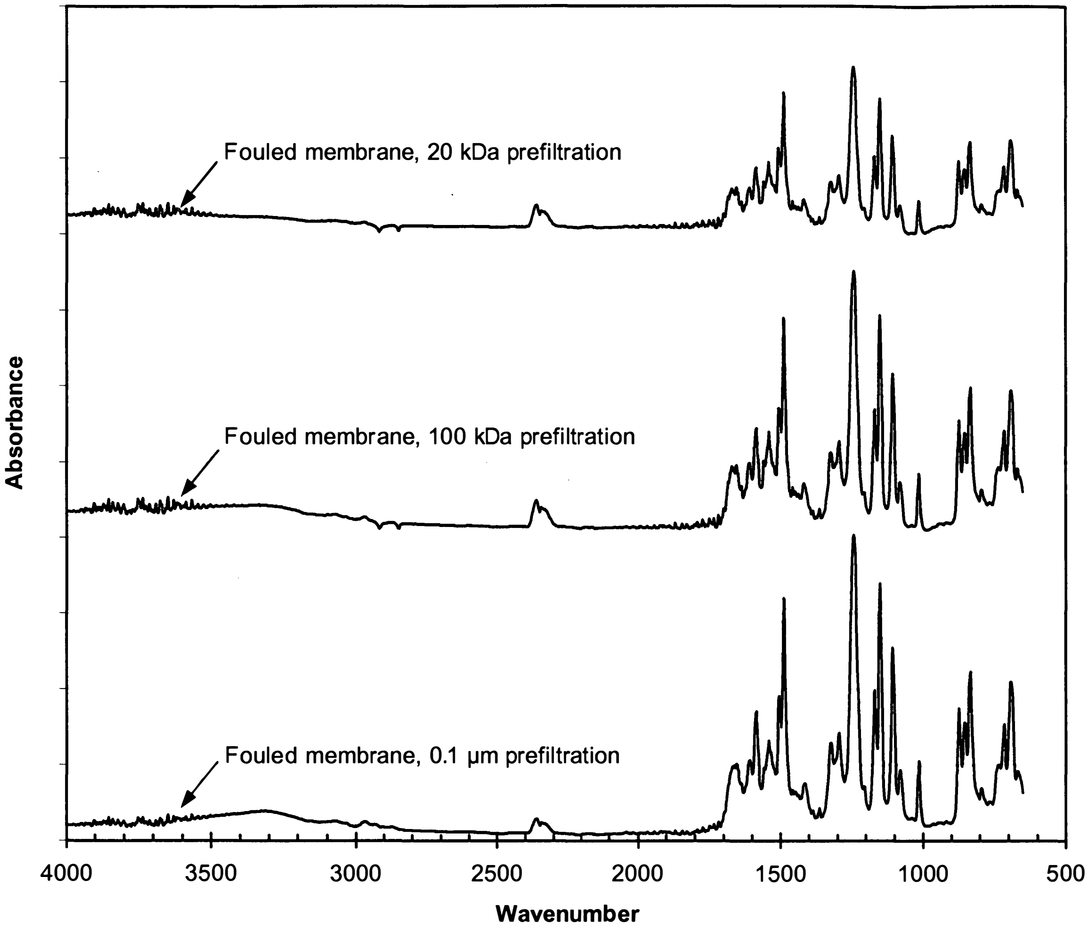


Figure 5-14: ATR/FTIR spectra of RO membranes fouled after 0.1 µm, 100 kDa, and 20 kDa prefiltration.

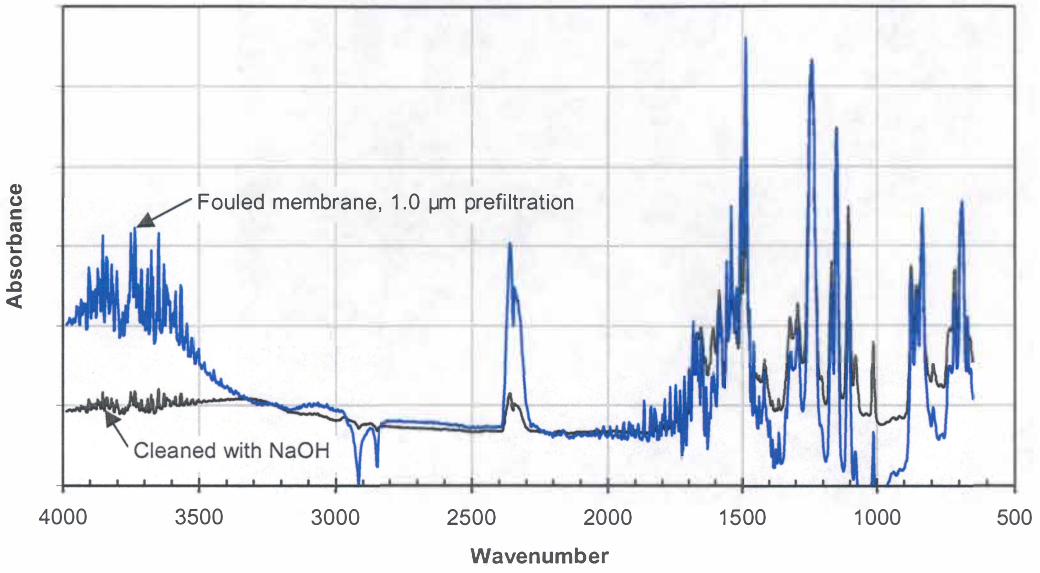


Figure 5-15: Comparative ATR/FTIR spectra of fouled and cleaned RO membranes fouled after 1 µm prefiltration.

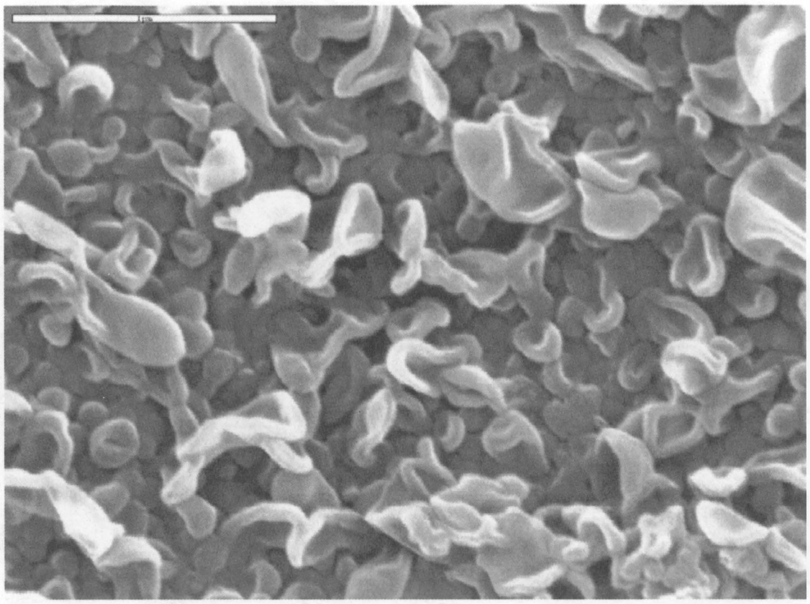
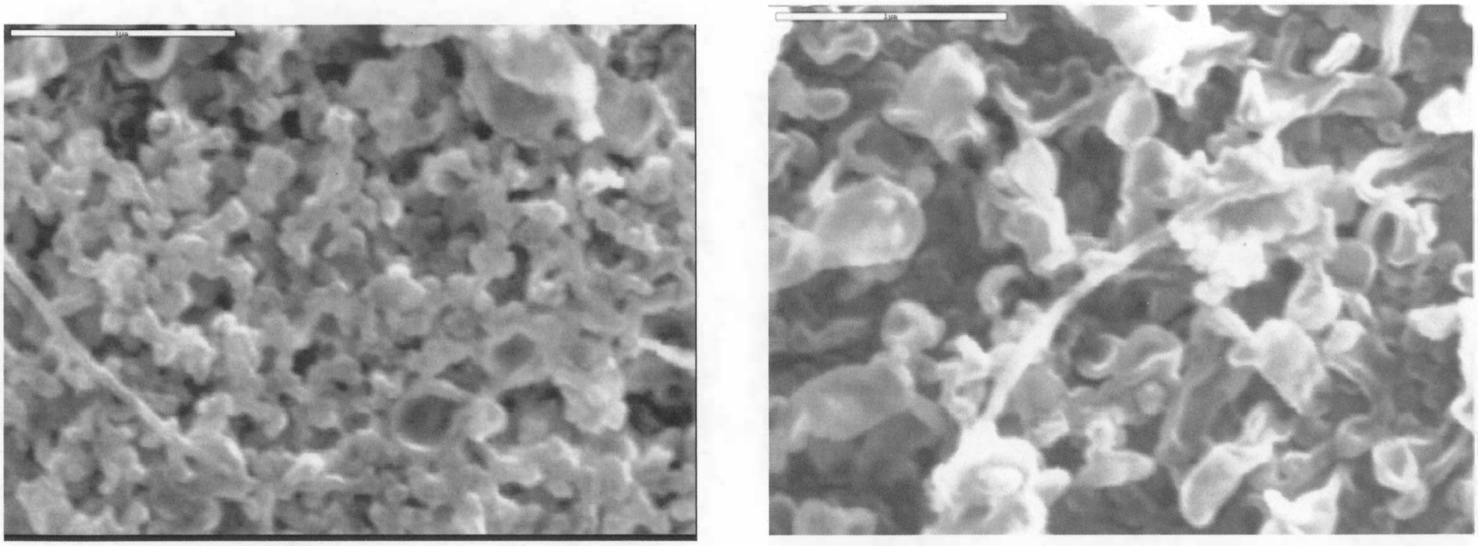


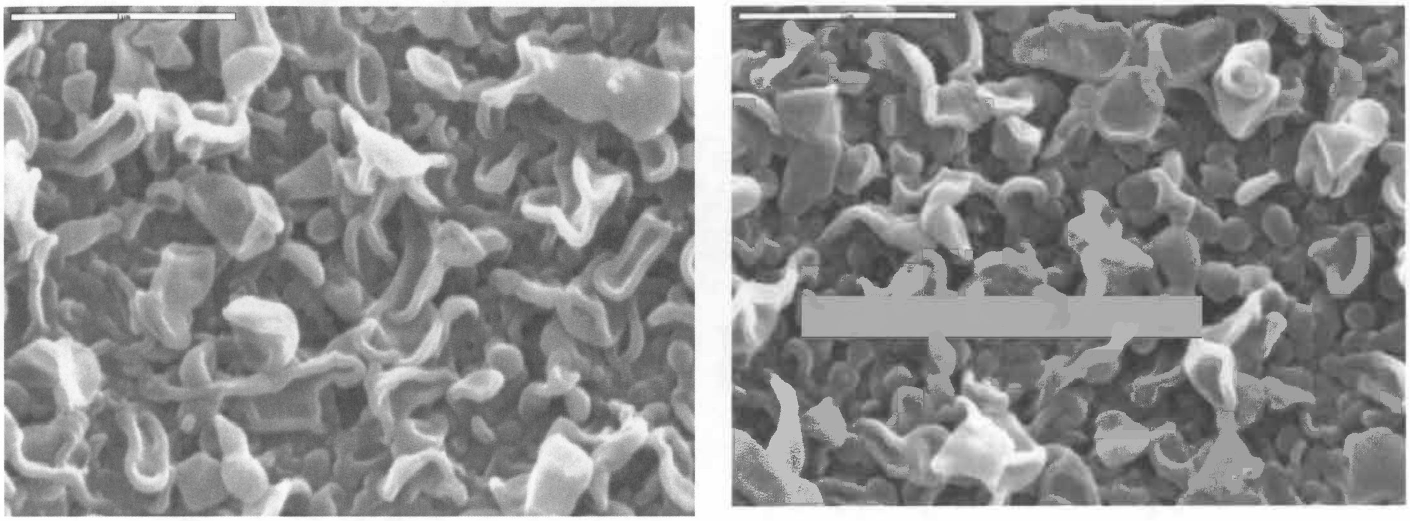
Figure 5-16: SEM Image of clean RO membrane (bar represents 1 micron)



a) Fouled

b) Cleaned

Figure 5-17: SEM Image of RO membranes fouled by 1 μm fraction and cleaned (bar represents 1 micron)



a) Fouled

b) Cleaned

Figure 5-18: SEM Image of RO membranes fouled by 0.1 μm fraction and cleaned (bar represents 1 micron)

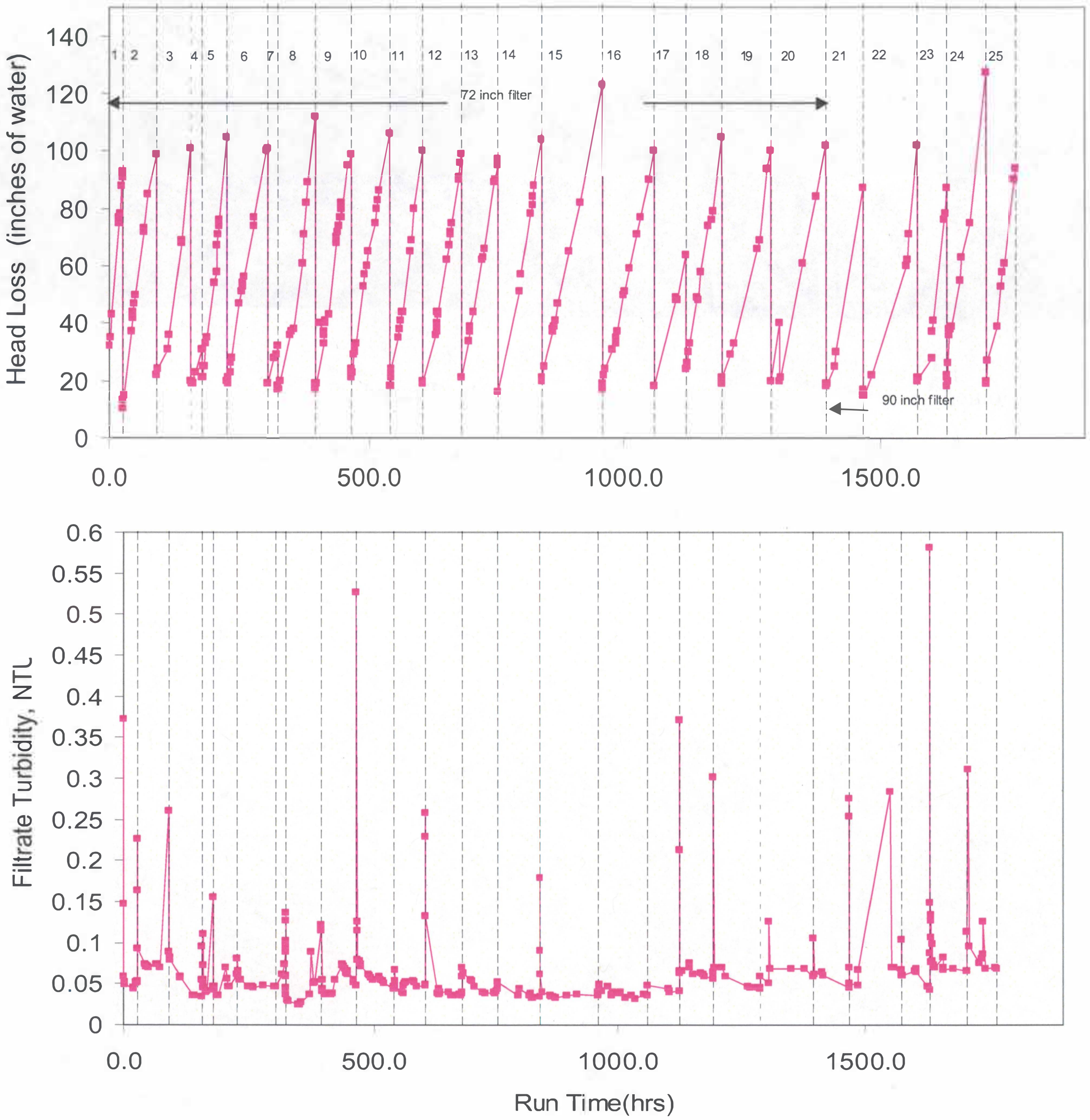


Figure 6-1: Filter Run Info: Head Loss and Turbidity

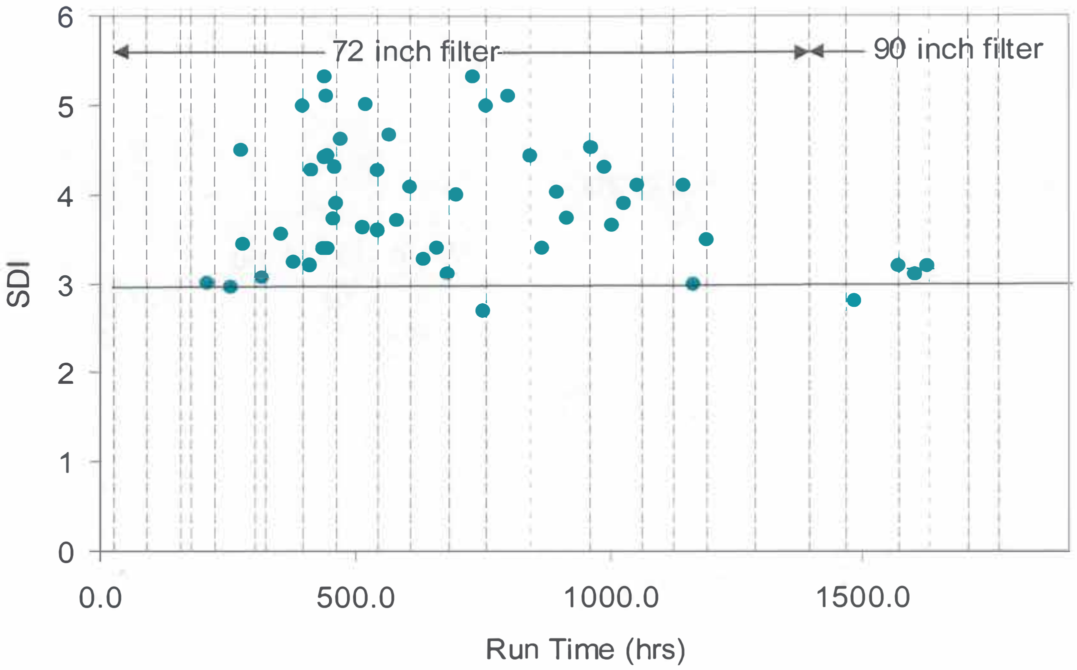


Figure 6-2: Effect of Filter configuration on filtrate SDI

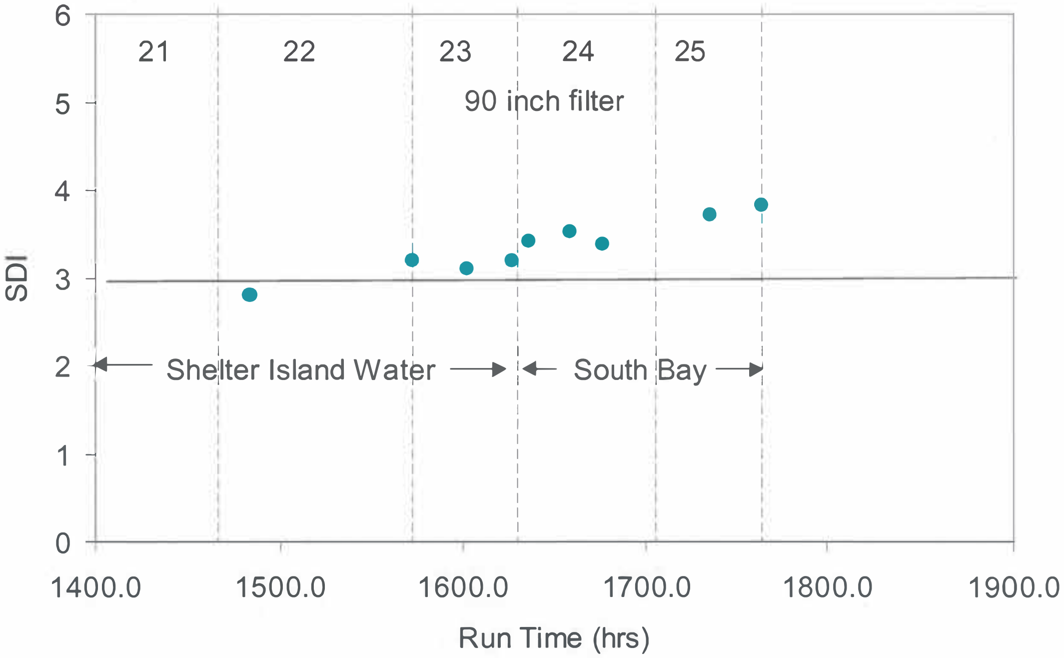


Figure 6-3: Effect of Source Water on filtrate SDI

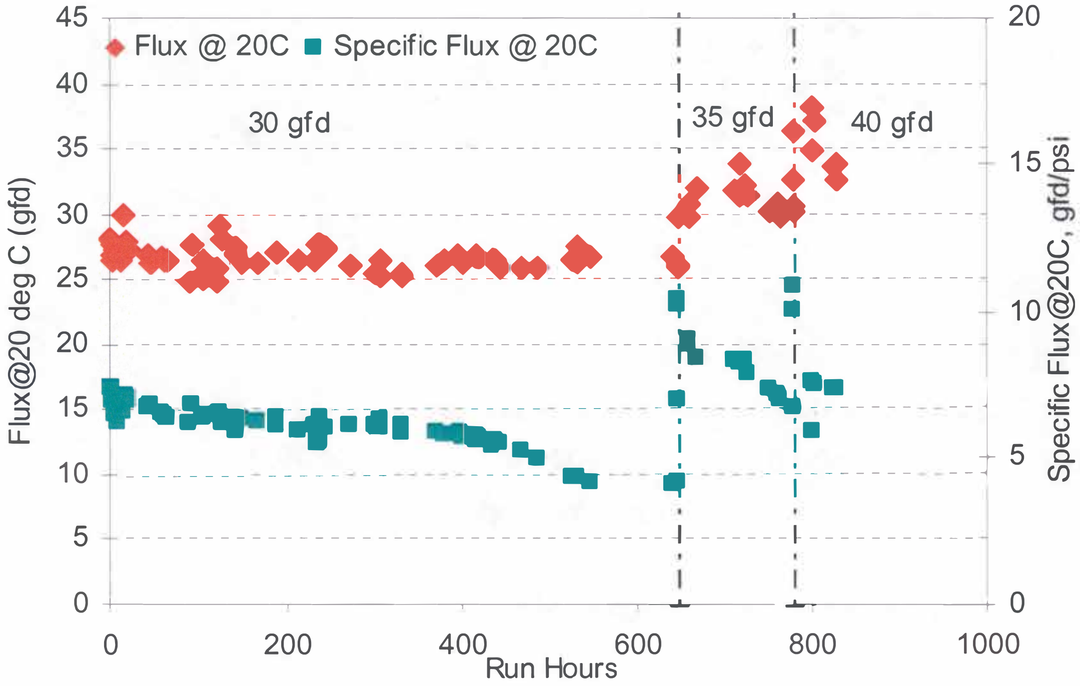


Figure 6-4: Membrane Performance of the US Filter MF System: Operational

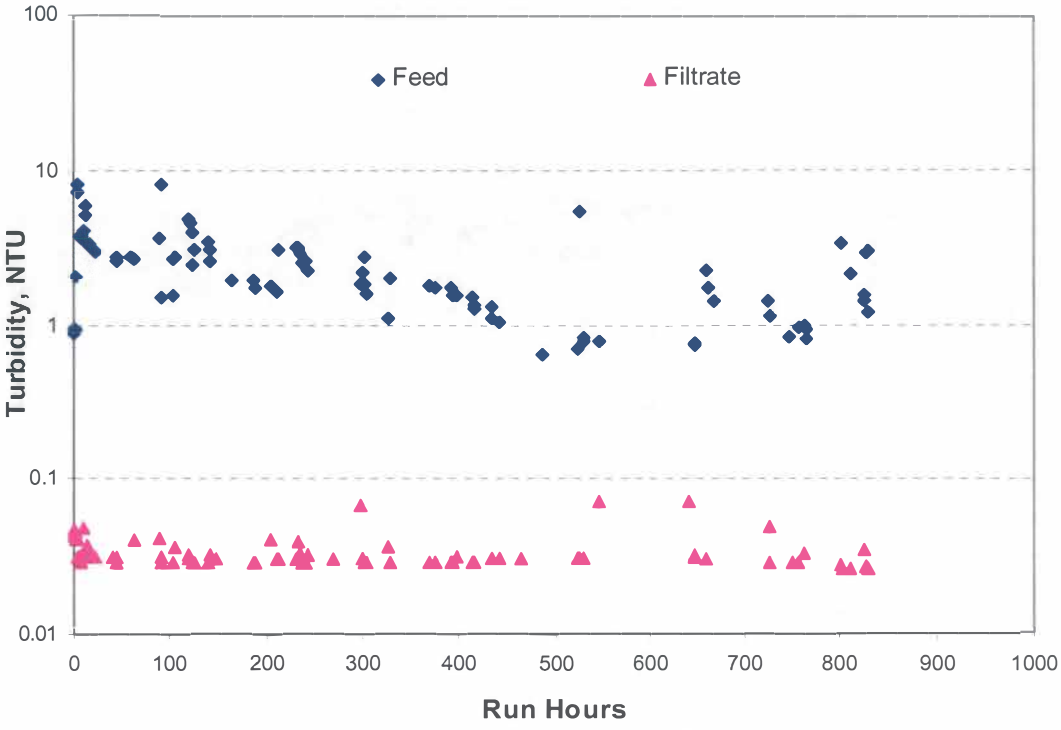


Figure 6-5: Membrane Performance of US Filter MF System: Turbidity

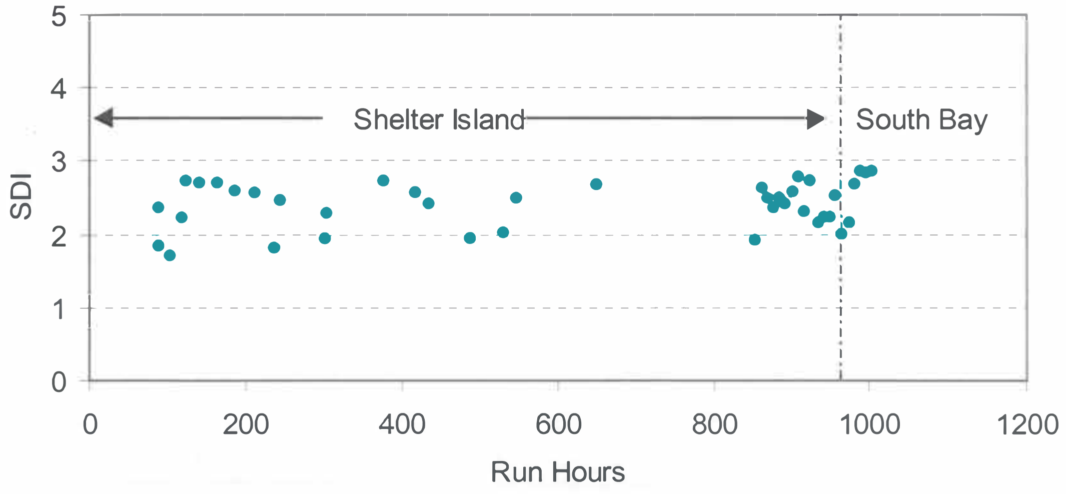


Figure 6-6: Membrane Performance of US Filter MF System: SDI

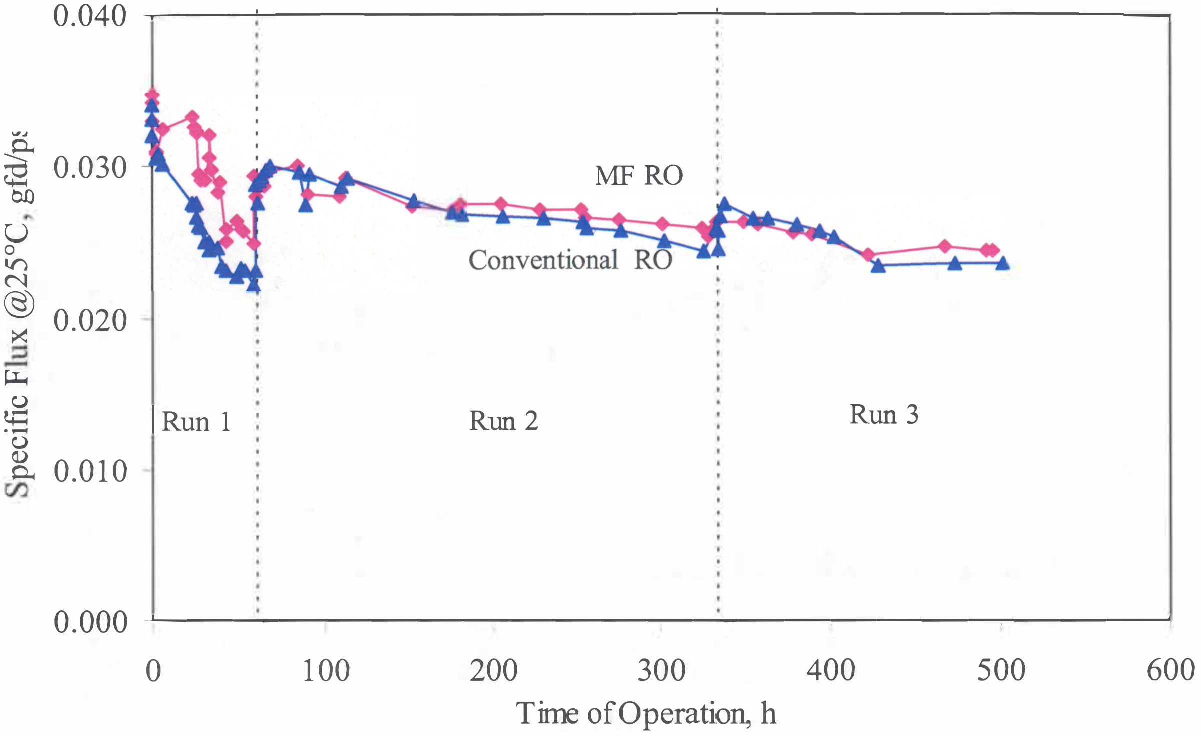


Figure 6-7: Overall Flux Decline for the two RO trains

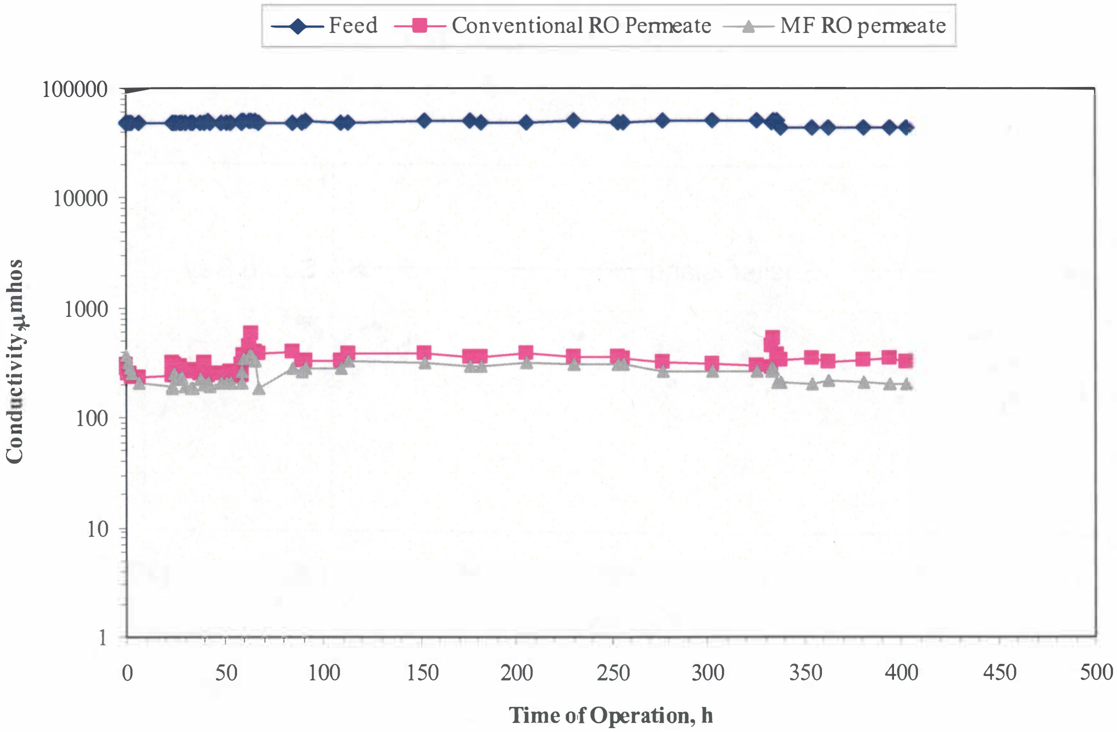


Figure 6-8: Overall Rejection values for the two RO trains

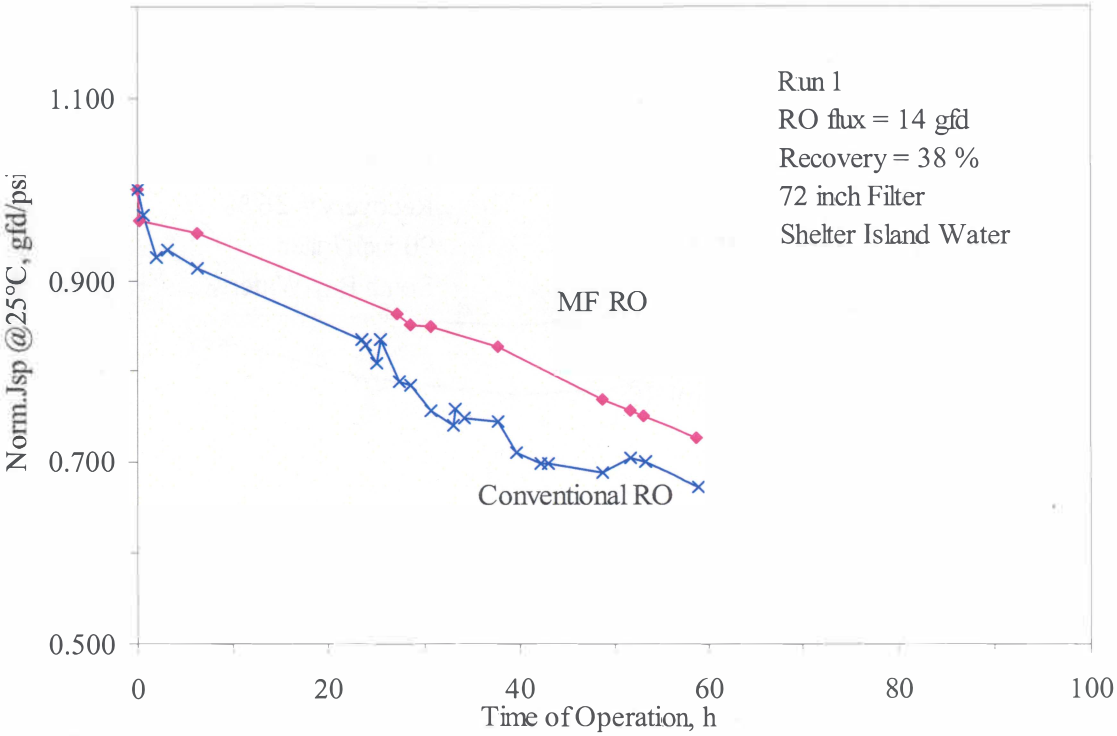


Figure 6-9: Normalized Specific Flux Curves: Run 1

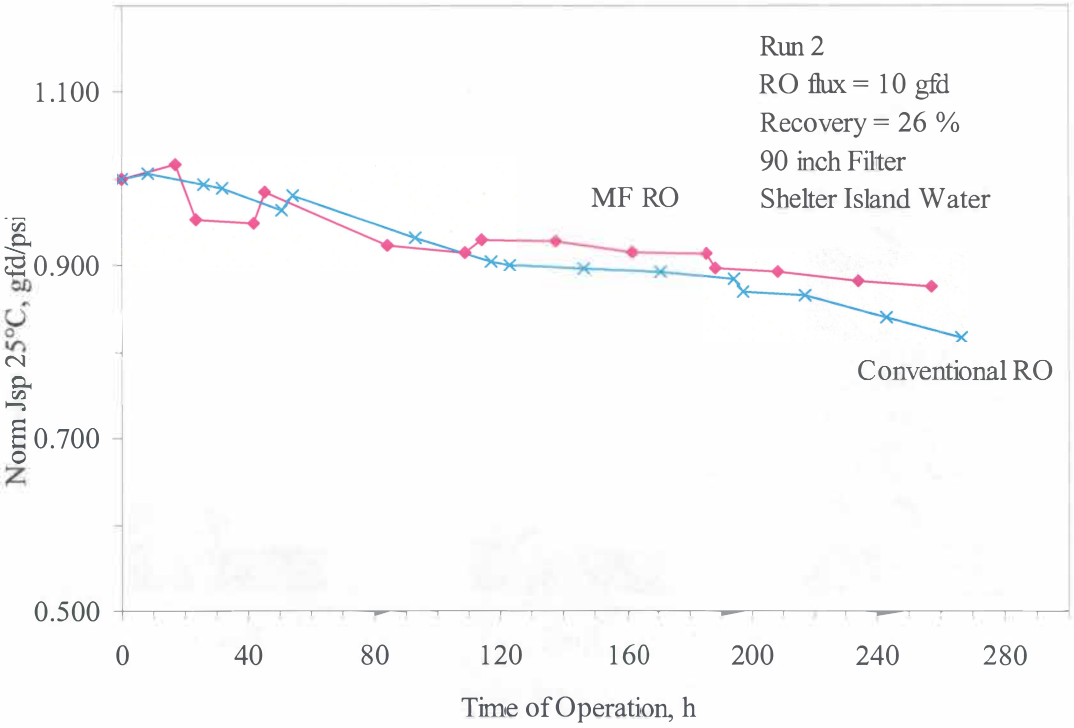


Figure 6-10: Normalized Specific Flux Curve: Run 2

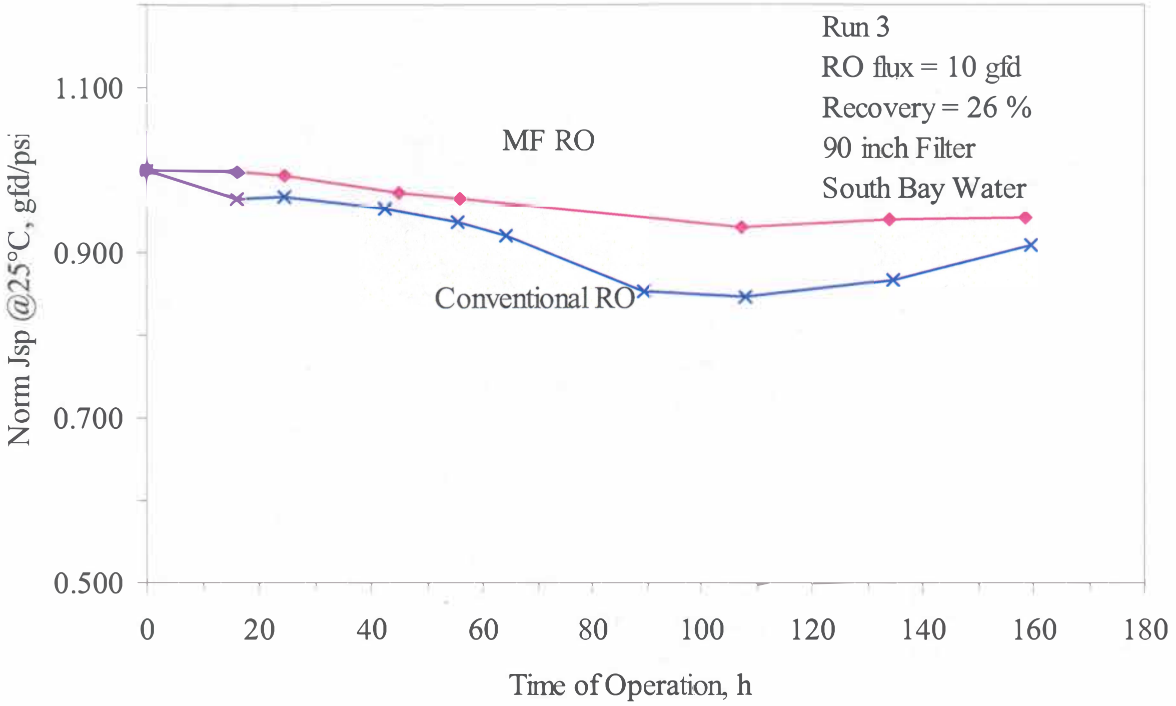


Figure 6-11: Normalized Specific Flux Curve: Run 3

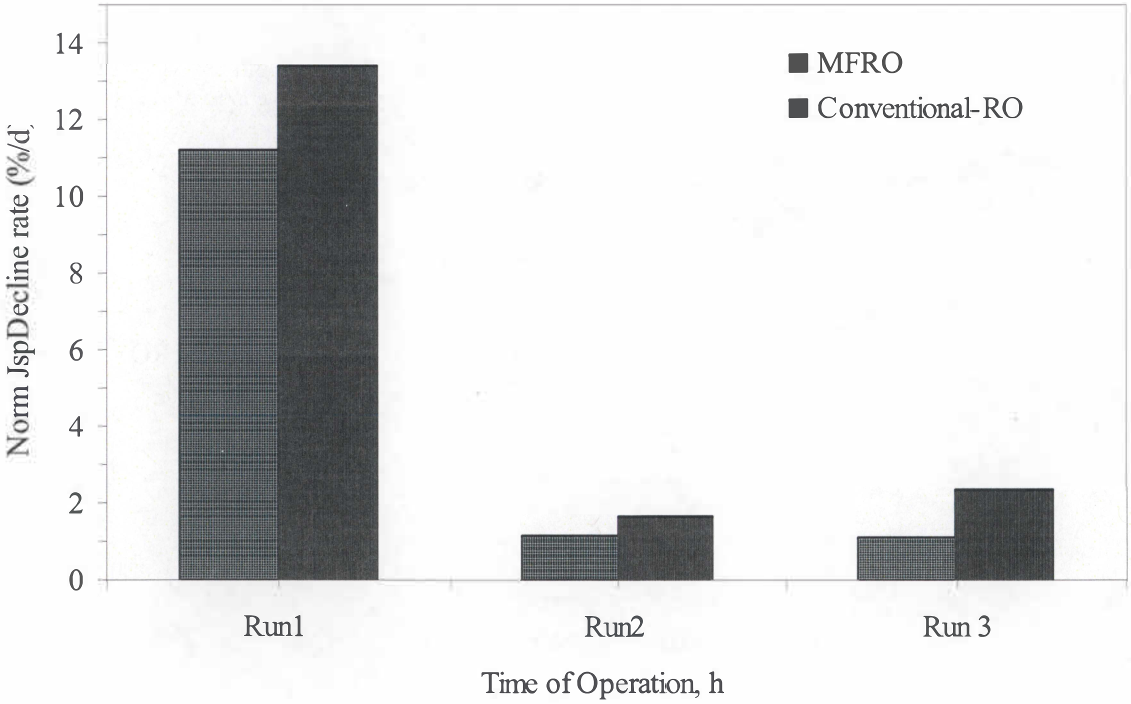


Figure 6-12: Comparative Normalized Specific Flux Decline rate

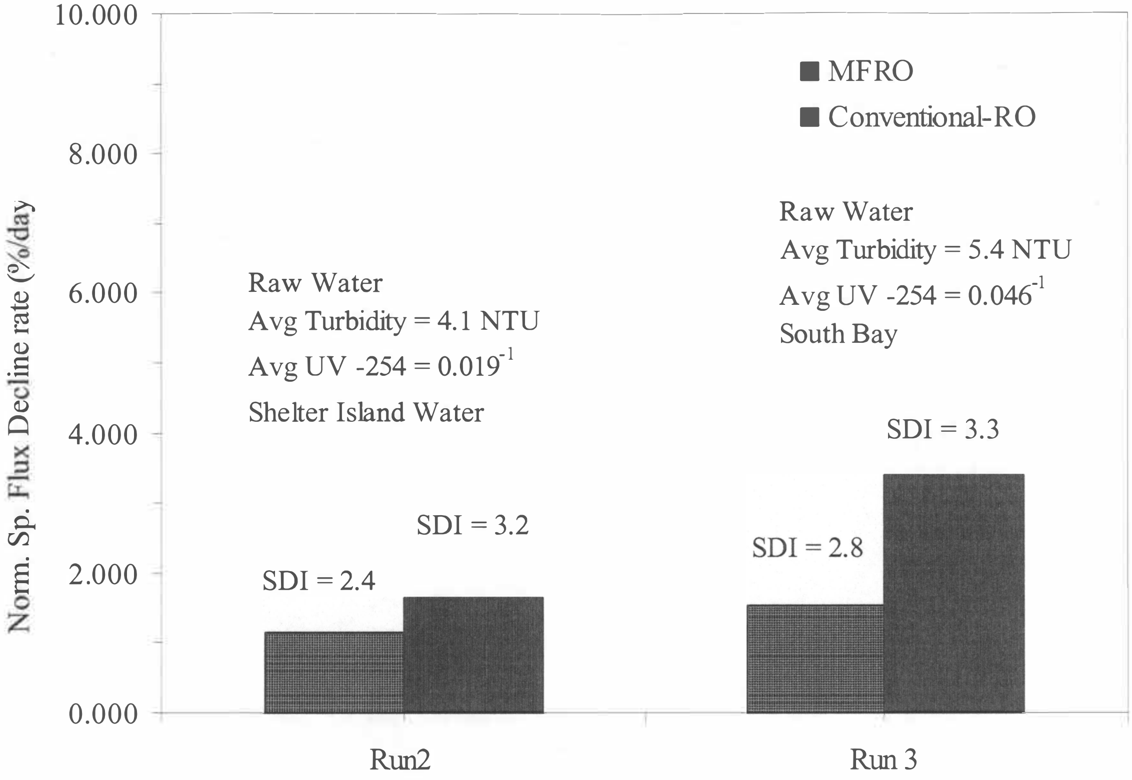


Figure 7-1: Specific Flux Decline rates for RO Runs 2 and 3.

APPENDIX B
Membrane Cleaning Protocols

US FILTER CMF-S MEMRANE CLEANING PROTOCOL

Maintenance Wash (MW)

The main process for recovering membrane permeability is the backwash however, regular use of a maintenance wash can extend the operation time between full Clean-In-Place (CIP) procedures. The Maintenance Wash sequence is similar to the CIP sequence, only shorter in duration.

The spent chlorine solution is discharged along with the rinse water. The Maintenance Wash sequence involves the following steps:

1. Step 1 – Backwash. The cell is backwashed to remove excess solids and maximize chemical efficiency. The cell is drained to waste; there is no refill with feed water at the end of this first step.
2. Step 2 – Chemical Solution Fill. The cell is filled with MF filtrate through extensive liquid backwashing. The cell is filled until the water level in the cell reaches the backwash level (i.e. the top of the modules only to minimize chemical waste).
3. Step 3 – Recirculation and Chemical Dosing. The filtrate pump starts and recirculates the water to the cell in a closed loop. The cell is fully isolated from the rest of the system. The dosing of chlorine or acid starts at the beginning of the recirculation.
4. Step 4 – Soak/Aeration and Recirculation Alternate. The modules are left to soak for a preset time (adjustable through the HMI). The soak and recirculation phases alternate automatically. Aeration is only used in certain cases, and then typically only with an acid solution.
5. Step 5 – Maintenance Wash Solution Draindown. The maintenance wash solution is drained away from the cell to the neutralization tank.
6. Step 6 – Rinse Backwash. The cell is refilled with feed water and backwashed to remove residual chemical. All rinse water is directed to the waste holding tank. On completion of the rinse steps, the cell returns to Standby or Filtration.

The complete sequence takes 20 minutes for completion. The number of soak and recirculation sequences is adjustable. Shorter or longer Maintenance Wash sequences can be enabled if deemed necessary.

Clean in Place (CIP)

The Clean-In-Place sequence is used to maintain long-term membrane performance. The CIP is initiated based either on transmembrane pressure (TMP), resistance, volume filtered (throughput) or time. In practice, the control system will allow the operator to initiate the CIP on time (by default) or on a resistance or permeability level.

A CIP typically involves two chemicals. PVDF modules are cleaned with an acid and a chlorine solution. The acid may be citric, hydrochloric, phosphoric, or sulfuric; a pH 2.0 solution is typically used. The chlorine solution is typically about 400 ppm.

The CIP sequence involves similar steps as the maintenance wash.

1. Step 1 – Backwash. The cell is backwashed to remove excess solids and maximize chemical efficiency. The cell is drained to waste; there is no refill with feed water at the end of this first step.
2. Step 2 – Chemical Solution Fill. The cell is filled with MF filtrate through extensive liquid backwashing. The cell is filled until the water level in the cell reaches the backwash level (i.e. the top of the modules only to minimize chemical waste).
3. Step 3 – Recirculation and Chemical Dosing. The filtrate pump starts and recirculates the water to the cell in a closed loop. The cell is fully isolated from the rest of the system. The dosing of chlorine or acid starts at the beginning of the recirculation.
4. Step 4 – Soak/Aeration and Recirculation Alternate. The modules are left to soak for a preset time (adjustable through the HMI). The soak and recirculation phases alternate automatically. Aeration is only used in certain cases, and then typically only with an acid solution.
5. Step 5 – CIP Solution Draindown. The cleaning solution is drained away from the cell to the neutralization tank.
6. Step 6 – Rinse Backwash. The cell is refilled with feed water and backwashed to remove residual chemical. All rinse water is directed to the waste holding tank. On completion of the rinse steps, the cell returns to Standby or Filtration.

On the pilot unit the chemicals used for CIP are manually added to the membrane tank. The CIP sequence is two hours per chemical. There's a 30-minute filtrate recirculation step followed by a 90-minute period of soaking. During an acid clean, aeration may be utilized intermittently during the 90-minute soak period. This is typically performed on applications that have had direct dosing or treat waters with high suspended solids.

([illegible])

HYDRANAUTICS POLYAMIDE TFC MEMRANE CLEANING PROTOCOLS

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Technical Service Bulletin

July 2003 TSB107.10

Foulants and Cleaning Procedures for composite polyamide RO Membrane Elements (ESPA, ESNA, CPA, LFC, and SWC)

This bulletin provides general information about the usual foulants affecting the performance of Hydranautics' Composite Polyamide Reverse Osmosis (RO) membrane elements and the removal of these foulants. The information in this bulletin applies to 4-inch, 6-inch, 8-inch, and 8.5-inch diameter RO membrane elements.

Note: The Composite Polyamide type of RO membrane elements may not be exposed to chlorinated water under any circumstances. Any such exposure will cause irreparable damage to the membrane. Absolute care must be taken following any disinfection of piping or equipment or the preparation of cleaning or storage solutions to ensure that no trace of chlorine is present in the feedwater to the RO membrane elements. If there is any doubt about the presence of chlorine, perform chemical testing to make sure. Neutralize any chlorine residual with a sodium bisulfite solution, and ensure adequate mixing and contact time to accomplish complete dechlorination. Dosing rate is 1.8 to 3.0 ppm sodium bisulfite per 1.0 ppm of free chlorine.

Note: It is recommended that all RO membrane cleaning operations should be closely coordinated with Hydranautics during the RO membrane element warranty period. Hydranautics field service personnel are available to be on site for cleaning assistance, should the need arise. Please contact Hydranautics for current charges for this service.

Note: The use of cationic surfactant should be avoided in cleaning solutions, since irreversible fouling of the membrane elements may occur.

If additional information is needed, please contact the Technical Services Department at:

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RO Membrane Fouling and Cleaning

During normal operation over a period of time, RO membrane elements are subject to fouling by suspended or sparingly soluble materials that may be present in the feedwater. Common examples of foulants are:

- Calcium carbonate scale
- Sulfate scale of calcium, barium or strontium
- Metal oxides (iron, manganese, copper, nickel, aluminum, etc.)
- Polymerized silica scale
- Inorganic colloidal deposits
- Mixed inorganic/organic colloidal deposits
- NOM organic material (Natural Organic Matter)
- Man-made organic material (e.g. antiscalant/dispersants, cationic polyelectrolytes)
- Biological (bacterial bioslime, algae, mold, or fungi)

The nature and rapidity of fouling depends on a number of factors, such as the quality of the feedwater and the system recovery rate. Typically, fouling is progressive, and if not controlled early, will impair the RO membrane element performance in a relatively short time. Cleaning is recommended when the RO shows evidence of fouling, just prior to a long-term shutdown, or as a matter of scheduled routine maintenance. Fouling characteristics that signal the need to clean are:

- A 10-15% decrease in normalized permeate flow.
- A 10-15% decrease in normalized permeate quality.
- A 10-15% increase in normalized pressure drop, as measured between the feed and concentrate headers.
- In the event you do not normalize your operating data, the above values still apply if you do not have major changes in critical operating parameters. The operating parameters that have to stay constant are permeate flow, permeate back-pressure, recovery, temperature, and feed TDS. If these operating parameters fluctuate, then it is highly recommended that you normalize the data to determine if fouling is occurring or if the RO is actually operating normally based on the change in a critical operating parameter. Hydranautics offers a free normalization software program called ROData, which can be downloaded from our web site at www.membranes.com.

Monitoring overall plant performance on a regular basis is an essential step in recognizing when membrane elements are becoming fouled. Performance is affected progressively and in varying degrees, depending on the nature of the foulants. Table 1 “RO Troubleshooting Matrix” provides a summary of the expected effects that common foulants have on performance.

RO cleaning frequency due to fouling will vary by site. A rough rule of thumb as to an acceptable cleaning frequency is once every 3 to 12 months. If you have to clean more than once a month, you should be able to justify further capital expenditures for improved RO pretreatment or a re-design of the RO operation. If the cleaning frequency is every one to three months, you may want to focus on improving the operation of your existing equipment but further capital expenditure may be harder to justify.

It is important to clean the membranes when they are only lightly fouled, not heavily fouled. Heavy fouling can impair the effectiveness of the cleaning chemical by impeding the penetration of the chemical deep into the foulant and in the flushing of the foulant out of the elements. If normalized membrane performance drops 30 to 50%, it may be impossible to fully restore the performance back to baseline conditions.

One RO design feature that is commonly over-looked in reducing RO cleaning frequency is the use of RO permeate water for flushing foulants from the system. Soaking the RO elements during standby with permeate can help dissolve scale and loosen precipitates, reducing the frequency of chemical cleaning.

What you clean for can vary site by site depending on the foulant. Complicating the situation frequently is that more than one foulant can be present, which explains why cleanings frequently require a low pH and high pH cleaning regiment.

Table 1: RO Troubleshooting Matrix

(Pressure Drop is defined as the Feed pressure minus the Concentrate pressure)

Possible Cause	Possible Location	Pressure Drop	Feed Pressure	Salt Passage
Metal Oxide Fouling (e.g. Fe,Mn,Cu,Ni,Zn)	1st stage lead elements	Rapid increase	Rapid increase	Rapid increase
Colloidal Fouling (organic and/or inorganic complexes)	1st stage lead elements	Gradual increase	Gradual increase	Slight increase
Mineral Scaling (e.g. Ca, Mg, Ba, Sr)	Last stage tail elements	Moderate Increase	Slight increase	Marked increase
Polymerized Silica	Last stage tail elements	Normal to increased	Increased	Normal to increased
Biological Fouling	Any stage, usually lead elements	Marked increase	Marked increase	Normal to increased
Organic Fouling (dissolved NOM)	All stages	Gradual increase	Increased	Decreased
Antiscalant Fouling	2nd stage most severe	Normal to increased	Increased	Normal to increased
Oxidant damage (e.g Cl_2, ozone, $KMnO_4$)	1st stage most severe	Normal to decreased	Decreased	Increased
Hydrolysis damage (out of range pH)	All stages	Normal to decreased	Decreased	Increased
Abrasion damage (carbon fines, etc)	1st stage most severe	Normal to decreased	Decreased	Increased
O-ring leaks (at interconnectors or adapters)	Random (typically at feed adapter)	Normal to decreased	Normal to decreased	Increased
Glue line leaks (due to permeate back-pressure in service or standby)	1st stage most severe	Normal to decreased	Normal to decreased	Increased
Glue line leaks (due to closed permeate valve while cleaning or flushing)	Tail element of a stage	Increased (based on prior fouling & high delta P)	Increased (based on prior fouling & and high delta P)	Increased

Discussion on Foulants

Calcium Carbonate Scale: Calcium carbonate is a mineral scale and may be deposited from almost any feedwater if there is a failure in the antiscalant/dispersant addition system or in the acid injection pH control system that results in a high feedwater pH. An early detection of the resulting calcium carbonate scaling is absolutely essential to prevent the damage that crystals can cause on the active membrane layers. Calcium carbonate scale that has been detected early can be removed by lowering the feedwater pH to between 3.0 and 5.0 for one or two hours. Longer resident accumulations of calcium carbonate scale can be removed by a low pH cleaning with a citric acid solution.

Calcium, Barium & Strontium Sulfate Scale: Sulfate scale is a much "harder" mineral scale than calcium carbonate and is harder to remove. Sulfate scale may be deposited if there is a failure in the antiscalant/dispersant feed system or if there is an over feed of sulfuric acid in pH adjustment. Early detection of the resulting sulfate scaling is absolutely essential to prevent the damage that crystals can cause on the active membrane layers. Barium and strontium sulfate scales are particularly difficult to remove as they are insoluble in almost all cleaning solutions, so special care should be taken to prevent their formation.

Calcium Phosphate Scale: This scale is particularly common in municipal waste waters and polluted water supplies which may contain high levels of phosphate. This scale can generally be removed with acidic pH cleaners. At this time, phosphate scaling calculations are not performed by the Hydranautics RO Design software. As a rule of thumb, contact Hydranautics technical department if phosphate levels in the feed are 5 ppm or higher.

Metal Oxide/Hydroxide Foulants: Typical metal oxide and metal hydroxide foulants are iron, zinc, manganese, copper, aluminum, etc. They can be the result of corrosion products from unlined pipes and tanks, or result from the oxidation of the soluble metal ion with air, chlorine, ozone, potassium permanganate, or they can be the result of a pretreatment filter system upset that utilizes iron or aluminum-based coagulant aids.

Polymerized Silica Coating: A silica gel coating resulting from the super-saturation and polymerization of soluble silica can be very difficult to remove. It should be noted that this type of silica fouling is different from silica-based colloidal foulants, which may be associated with either metal hydroxides or organic matter. Silica scale can be very difficult to remove by traditional chemical cleaning methods. Contact Hydranautics technical department if the traditional methods are unsuccessful. There does exist harsher cleaning chemicals, like ammonium biflouride, that have been used successfully at some sites but are considered rather hazardous to handle and can damage equipment.

Colloidal Foulants: Colloids are inorganic or mixed inorganic/organic based particles that are suspended in water and will not settle out due to gravity. Colloidal matter typically contains one or more of the following major components: iron, aluminum, silica, sulfur, or organic matter.

Dissolved NOM Organic Foulants: The sources of dissolved NOM (Natural Organic Matter) foulants are typically derived from the decomposition of vegetative material into surface waters or shallow wells. The chemistry of organic foulants is very complex, with the major organic components being either humic acid or fulvic acid. Dissolved NOMs can quickly foul RO membranes by being absorbed onto the membrane surface. Once absorption has occurred, then a slower fouling process of gel or cake formation starts. It should be noted that the mechanism of fouling with dissolved NOM should not be confused with the mechanism of fouling created by NOM organic material that is bound up with colloidal particles.

Microbiological Deposits: Organic-based deposits resulting from bacterial slimes, fungi, molds, etc. can be difficult to remove, particularly if the feed path is plugged. Plugging of the feed path makes it difficult to introduce and distribute the cleaning solutions. To inhibit additional growth, it is important to clean and sanitize not only the RO system, but also the pretreatment, piping, dead-legs, etc. The membranes, once chemically cleaned, will require the use of a Hydranautics approved biocide and an extended exposure requirement to be effective. For further information on biocides, refer to Hydranautics Technical Service Bulletin TSB-110 “Biocides for Disinfection and Storage of Hydranautics Membrane Elements”.

Selection and Use of Cleaning Chemicals

There are a number of factors involved in the selection of a suitable cleaning chemical (or chemicals) and proper cleaning protocol. The first time you have to perform a cleaning, it is recommended to contact the manufacturer of the equipment, the RO element manufacturer, or a RO specialty chemical and service supplier. Once the suspected foulant(s) are identified, one or more cleaning chemicals will be recommended. These cleaning chemical(s) can be generic or can be private-labeled proprietary chemicals. Typically, the generic chemicals can be of technical grades and are available from local chemical supply companies. The proprietary RO cleaning chemicals can be more expensive, but may be easier to use and you cannot rule out the advantage of the intellectual knowledge supplied by these companies. Some independent RO service companies can determine the proper chemicals and cleaning protocol for your situation by testing at their facility a fouled element pulled from your system.

It is not unusual to use a number of different cleaning chemicals in a specific sequence to achieve the optimum cleaning. Typically, a low pH cleaning is used first to remove foulants like mineral scale, followed by a high pH cleaning to remove organic material. There are times that a high pH cleaning is used first to remove foulants like oil or biological matter, followed by a low pH cleaning. Some cleaning solutions have detergents added to aid in the removal of heavy biological and organic debris, while others have a chelating agent like EDTA added to aid in the removal of colloidal material, organic and biological material, and sulfate scale. An important thing to remember is that the improper selection of a cleaning chemical, or the sequence of chemical introduction, can make the foulant worse.

General Precautions in Cleaning Chemical Selection and Usage

- If you are using a proprietary chemical, make sure the chemical has been qualified for use with your Hydranautics membrane by the chemical supplier. The chemical supplier’s instructions should not be in conflict with Hydranautics recommended cleaning parameters and limits listed in this Technical Service Bulletin.
- If you are using generic chemicals, make sure the chemical has been qualified for use with your Hydranautics membrane in this Technical Service Bulletin.
- Use the least harshest cleaning regiment to get the job done. This includes the cleaning parameters of pH, temperature, and contact time. This will optimize the useful life of the membrane.
- Clean at the recommended target temperatures to optimize cleaning efficiency and membrane life.
- Use the minimal amount of chemical contact time to optimize membrane life.
- Be prudent in the adjustment of pH at the low and high pH range to extend the useful life of the membrane. A “gentle” pH range is 4 to 10, while the harshest is 2 to 12.

- Typically, the most effective clean-up sequence is low pH followed by high pH solutions. One known exception is that oil-fouled membranes should not use a low pH clean-up first as the oil will congeal.
- Cleaning and flushing flows should be in the same direction as the normal feed flow to avoid potential telescoping and element damage.
- When cleaning a multi-stage RO, the most effective cleaning is one stage at a time so cleaning flow velocities can be optimized and foulants from upstream stages don't have to pass through down-stream stages.
- Flushing out detergents with higher pH permeate can reduce foaming problems.
- Verify that proper disposal requirements for the cleaning solution are followed.
- If your system has been fouled biologically, you may want to consider the extra step of introducing a sanitizing biocide chemical after a successful cleaning. Biocides can be introduced immediately after cleaning, periodically (e.g. once a week), or continuously during service. You must be sure that the biocide is compatible with the membrane, does not create any health risks, is effective in controlling biological activity, and is not cost prohibitive.
- For safety reasons, make sure all hoses and piping can handle the temperatures, pressures and pH's encountered during a cleaning.
- For safety reasons, always add chemicals slowly to an agitated batch of make-up water.
- For safety reason, always wear safety glasses and protective gear when working with chemicals.
- For safety reasons, don't mix acids with caustics. Thoroughly rinse the 1st cleaning solution from the RO system before introducing the next solution.

Selecting a Cleaning Solution

Table 2 lists the recommended generic chemical solutions for cleaning an RO membrane element based on the foulant to be removed.

Important: It is recommended that the MSDS of the cleaning chemicals be procured from the chemical supplier and that all safety precautions be utilized in the handling and storage of all chemicals.

Table 2: Hydranautics Recommended Chemical Cleaning Solutions

Foulant	**Gentle Cleaning Solution**	**Harsher Cleaning Solution**
Calcium carbonate scale	1	4
Calcium, barium or strontium sulfate scale	2	4
Metal oxides/hydroxides (Fe, Mn, Zn, Cu, Al)	1	5
Inorganic colloidal foulants	1	4
Mixed Inorganic/organic colloidal foulants	2	6
Polymerized silica coating	None	7
Biological matter	2	3 or 6
NOM organic matter (naturally occurring)	2	3 or 6

Table 3 "Hydranautics Recipes for Cleaning Solutions" offers instructions on the volumes of bulk chemical to be added to 100 U.S. gallons (379 liters) of make-up water. Prepare the solutions by proportioning the amount of chemicals to the amount of make-up water to be used. Make-up water quality should be of RO permeate or deionized (DI) quality, and be free of chlorine and hardness. Before forwarding the cleaning solution to the membranes, it is important to thoroughly mix it, adjust the pH according to the target pH, and stabilize the temperature at the target temperature. Unless otherwise instructed, the cleaning design parameters are based on a chemical recirculation flow period of one hour and an optional chemical soak period of one hour.

Table 4 "Hydranautics Maximum pH and Temperature Limits for Cleaning" highlights the maximum pH and temperature limits for specific membranes, after which irreparable membrane damage can occur. A suggested minimum temperature limit is 70 F (21 C), but cleaning effectiveness and the solubility of the cleaning chemical is significantly improved at higher temperatures.

Description of Cleaning Solutions

Note: The notation (w) denotes that the diluted chemical solution strength is based on the actual weight of the 100% pure chemical or active ingredient.

Solution 1: This is a low pH cleaning solution (target pH of 4.0) of 2.0% (w) citric acid ($C_6H_8O_7$). It is useful in removing inorganic scale (e.g. calcium carbonate, calcium sulfate, barium sulfate, strontium sulfate) and metal oxides/hydroxides (e.g. iron, manganese, nickel, copper, zinc), and inorganic-based colloidal material. Note: Citric acid has chelating properties that function better when an upward pH adjustment is performed using ammonium hydroxide. Sodium hydroxide should not be used for pH adjustment. Citric acid is available as a powder.

Solution 2: This is a high pH cleaning solution (target pH of 10.0) of 2.0% (w) of STPP (sodium tripolyphosphate) ($Na_5P_3O_{10}$) and 0.8% (w) of Na-EDTA (sodium salt of ethylaminediaminetetraacetic acid). It is specifically recommended for removing calcium sulfate scale and light to moderate levels of organic foulants of natural origin. STPP functions as an inorganic-based chelating agent and detergent. Na-EDTA is an organic-based chelating cleaning agent that aids in the sequestering and removal of divalent and trivalent cations and metal ions. STPP and Na-EDTA are available as powders.

Solution 3: This is a high pH cleaning solution (target pH of 10.0) of 2.0% % (w) of STPP (sodium tripolyphosphate) ($Na_5P_3O_{10}$) and 0.25% (w) Na-DDBS ($C_6H_5(CH_2)_{12}$-SO_3Na) (sodium salt of dodecylbenzene sulfonate). It is specifically recommended for removing heavier levels of organic foulants of natural origin. STPP functions as an inorganic-based chelating agent and detergent. Na-DDBS functions as an anionic detergent.

Solution 4: This is a low pH cleaning solution (target pH of 2.5) of 0.5% (w) of HCL (hydrochloric) acid. It is useful in removing inorganic scale (e.g. calcium carbonate, calcium sulfate, barium sulfate, strontium sulfate and metal oxides/hydroxides (e.g. iron, manganese, nickel, copper, zinc) and inorganic-based colloidal material. This cleaning solution is considered to be harsher than Solution 1. HCL acid, a strong mineral acid, is also known as muriatic acid. HCL acid is available in a number of concentrations: (18 ^{0} Baume = 27.9%), (20 ^{0} Baume = 31.4%), (22 ^{0} Baume = 36.0%).

Solution 5: This is a high pH cleaning solution (target pH of 11.5) of 1.0% (w) of $Na_2S_2O_4$ (sodium hydrosulfite). It is useful in the removal of metal oxides and hydroxides, and to a lesser extent calcium sulfate, barium sulfate and strontium sulfate. Sodium hydrosulfite is strong reducing agent and is also known as sodium dithionite. Sodium hydrosulfite is available as a powder.

Solution 6: This is a high pH cleaning solution (target pH of 11.5) of 0.1% (w) of NaOH (sodium hydroxide) and 0.03% (w) of SDS (sodium dodecylsulfate). It is useful in the removal of organic foulants of natural origin, colloidal foulants of mixed organic/inorganic origin, and biological material (fungi, mold, slimes and biofilm). SDS is a detergent that is an anionic surfactant that will cause some foaming. This is considered to be a harsh cleaning regiment.

Solution 7: This is a high pH cleaning solution (target pH of 11.5) of 0.1% (w) of NaOH (sodium hydroxide). It is useful in the removal of polymerized silica. This is considered to be a harsh cleaning regiment.

> **Important: It is recommended that the MSDS of the cleaning chemicals be procured from the chemical supplier and that all safety precautions be utilized in the handling and storage of all chemicals.**

Table 3: Hydranautics Recipes for Cleaning Solutions

The quantities listed below are to be added to 100 U.S.gallons (379 liters) of dilution water.

Cleaning Solution	Bulk Ingredients	Quantity	Target pH Adjustment	Target Temp.
1	**Citric acid** (as 100% powder)	17.0 pounds (7.7 kg)	Adjust to pH 4.0 with ammonium hydroxide.	104 F (40 C)
2	**STPP** (sodium tripolyphosphate) (as 100% powder) **Na-EDTA** (Versene 220 or equal) (as 100% powder)	17.0 pounds (7.7 kg) 7.0 pounds (3.18 kg)	Adjust to pH 10.0 with sulfuric or hydrochloric acid.	104 F (40 C)
3	**STPP** (sodium tripolyphosphate) (as 100% powder) **Na-DDBS** Na-dodecylbenzene sulfonate	17 pounds (7.7 kg) 2.13 pounds (0.97 kg)	Adjust down to pH 10.0 with sulfuric or hydrochloric acid.	104 F (40 C)
4	**HCl acid** (hydrochloric acid (as 22^0 Baume or 36% HCL)	0.47 gallons (1.78 liters)	Slowly adjust pH down to 2.5 with HCL acid. Adjust pH up with sodium hydroxide.	95 F (35 C)
5	**Sodium hydrosulfite** (as 100% powder)	8.5 pounds (3.86 kg)	No pH adjustment is required.	95 F (35 C)
6	**NaOH** (sodium hydroxide) (as 100% powder) (or as 50% liquid) **SDS** (sodium dodecylsulfate)	0.83 pounds (0.38 kg) 0.13 gallons (0.49 liters) 0.25 pounds (0.11 kg)	Slowly adjust pH up to 11.5 with sodium hydroxide. Adjust pH down to 11.5 by adding HCL acid.	86 F (30 C)
7	**NaOH** (sodium hydroxide) (as 100% powder) (or as 50% liquid)	0.83 pounds (0.38 kg) 0.13 gallons (0.49 liters)	Slowly adjust pH up to 11.5 with sodium hydroxide. Adjust pH down to 11.5 by adding HCL acid.	86 F (30 C)

Table 4: Hydranautics pH and Temperature Limits for Cleaning

(See Table 3 for target pH and temperatures)

Membrane	45 C (113 F)	35 C (95 F)	30 C (86 F)
CPA	2-10	2-11.5	2-12
ESPA	2-10	2-11.5	2-12
LFC	2-10	2-11.5	2-12
SWC	2-10	2-11	2-12
ESNA	3-10	2-11.5	2-12

Note: The above cleaning parameters denote the maximum temperature limits for a corresponding range of pH. Cleaning operations performed at the extremes may result in a more effective cleaning, but can shorten the useful life of the membrane due to hydrolysis. To optimize the useful life of a membrane, it is recommended to use the least harshest cleaning solutions and minimize the contact time whenever possible.

Table 5: Cleaning and Flushing Flow Rates per RO Pressure Tube

(Pressures are not to exceed 60 psi (4 bar) at inlet to tubes.)

Element Diameter	GPM	LPM
4-inches	6 to 10	23 to 38
6-inches	12 to 20	46 to 76
8-inches	24 to 40	91 to 151
8.5-inches	27 to 45	102 to 170

Table 6: Cleaning Solution Volume Requirement per RO Element

(these volumes do not include volumes required for piping, filters, etc)

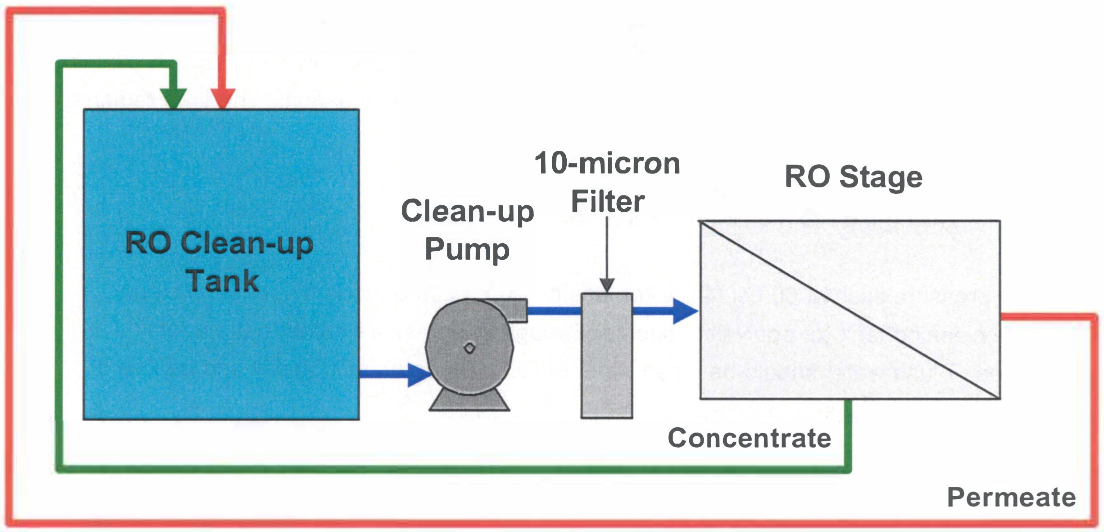
(these volumes do not include initial 20% of volume dumped to drain)

Element Size	Normal	Heavy	Normal	Heavy
	Fouling	Fouling	Fouling	Fouling
	(Gallons)	(Gallons)	(Liters)	(Liters)
4 x 40 inches	2.5	5	9.5	19
6 x 40 inches	5	10	19	38
8 x 40 inches	9	18	34	68
8.5 x 40 inches	10	20	38	76

RO Cleaning Skid

The successful cleaning of an RO on-site requires a well designed RO cleaning skid. Normally this skid is not hard piped to the RO skid and uses temporary hosing for connections. It is recommended to clean a multi-stage RO one stage at a time to optimize cross-flow cleaning velocity. The source water for chemical solution make-up and rinsing should be clean RO permeate or DI water and be free of hardness, transition metals (e.g. iron), and chlorine. Components must be corrosion proof. Major cleaning system components are:

RO Cleanup Skid



- RO Cleaning Tank: This tank needs to be sized properly to accommodate the displacement of water in the hose, piping, and RO elements. The table below denotes the amount of chemical solution that needs to be made for a single RO element. The tank should be designed to allow 100 % drainage, easy access for chemical introduction and mixing, a recirculation line from the RO Cleaning Pump, proper venting, overflow, and a return line located near the bottom to minimize foam formation when using a surfactant.
- RO Cleaning Pump: This pump needs to be sized to develop the proper cross-flow velocity to scrub the membrane clean. The maximum recommended pressure is 60 psi (4 bar) at the inlet to the pressure vessels to minimize the production of permeate during cleaning and reduce the convective redeposition of foulant back on to the membrane surface. The table below denotes the flow rate ranges for each pressure tube.
- RO Cleaning Cartridge Filter: Normally 5 to 10-micron and is designed to remove foulants that have been displaced from the cleaning process.
- RO Tank Heater or Cooler: The maximum design temperature for cleaning is 113^{0} F (45^{0} C). It should be noted that heat is generated and imparted by the RO Cleaning Pump during recirculation.
- RO Tank Mixer: This is recommended to get optimal mixing of chemical, though some designers rely solely on the slow introduction of chemical while maintaining a recirculation through the RO Cleaning Pump back to the tank.
- Instrumentation: Cleaning system instrumentation should be included to monitor flow, temperature, pressure, and tank level.
- Sample Points: Sample valves should be located to allow pH and TDS measurements off the RO Cleaning Pump discharge and the concentrate side recirculation return line.

- Permeate Return Line: A small amount of the cleaning solution can permeate through the membranes and so a permeate-side return line back to the RO Cleaning Tank is required.

  Important: The permeate line and any permeate valves must always be open to atmospheric pressure during the cleaning and flushing steps or damage to RO elements can occur. If the permeate line is closed, the permeate pressure can build up and become higher than the feed-side pressure of the tail elements. This can result in excessive permeate back-pressure which can damage the membrane glue lines in the tail elements.

RO Membrane Element Cleaning and Flushing Procedures

The RO membrane elements can be cleaned in place in the pressure tubes by recirculating the cleaning solution across the high-pressure side of the membrane at low pressure and relatively high flow. A cleaning unit is needed to do this. RO cleaning procedures may vary dependent on the situation. The time required to clean a stage can take from 4 to 8 hours.

A general procedure for cleaning the RO membrane elements is as follows:

1. Perform a low pressure flush at 60 psi (4 bar) or less of the pressure tubes by pumping clean water from the cleaning tank (or equivalent source) through the pressure tubes to drain for several minutes. Flush water should be clean water of RO permeate or DI quality and be free of hardness, transition metals, and chlorine.

2. Mix a fresh batch of the selected cleaning solution in the cleaning tank. The dilution water should be clean water of RO permeate or DI quality and be free of hardness, transition metals, and chlorine. The temperature and pH should be adjusted to their target levels.

3. Circulate the cleaning solution through the pressure tubes for approximately one hour or the desired period of time. At the start, send the displaced water to drain so you don't dilute the cleaning chemical and then divert up to 20% of the most highly fouled cleaning solution to drain before returning the cleaning solution back to the RO Cleaning Tank. For the first 5 minutes, slowly throttle the flow rate to 1/3 of the maximum design flow rate. This is to minimize the potential plugging of the feed path with a large amount of dislodged foulant.. For the second 5 minutes, increase the flow rate to 2/3 of the maximum design flow rate, and then increase the flow rate to the maximum design flow rate. If required, readjust the pH back to the target when it changes more than 0.5 pH units.

4. An optional soak and recirculation sequence can be used, if required. The soak time can be from 1 to 8 hours depending on the manufacturer's recommendations. Caution should be used to maintain the proper temperature and pH. Also note that this does increase the chemical exposure time of the membrane.

5. Upon completion of the chemical cleaning steps, a low pressure Cleaning Rinse with clean water (RO permeate or DI quality and free of hardness, transition metals, and chlorine) is

required to remove all traces of chemical from the Cleaning Skid and the RO Skid. Drain and flush the cleaning tank; then completely refill the Cleaning Tank with clean water for the Cleaning Rinse. Rinse the pressure tubes by pumping all of the rinse water from the Cleaning Tank through the pressure tubes to drain. A second cleaning can be started at this point, if required.

6. Once the RO system is fully rinsed of cleaning chemical with clean water from the Cleaning Tank, a Final Low Pressure Clean-up Flush can be performed using pretreated feed water. The permeate line should remain open to drain. Feed pressure should be less than 60 psi (4 bar). This final flush continues until the flush water flows clean and is free of any foam or residues of cleaning agents. This usually takes 15 to 60 minutes. The operator can sample the flush water going to the drain for detergent removal and lack of foaming by using a clear flask and shaking it. A conductivity meter can be used to test for removal of cleaning chemicals, such that the flush water to drain is within 10-20% of the feed water conductivity. A pH meter can also be used to compare the flush water to drain to the feed pH.

7. Once all the stages of a train are cleaned, and the chemicals flushed out, the RO can be restarted and placed into a Service Rinse. The RO permeate should be diverted to drain until it meets the quality requirements of the process (e.g. conductivity, pH, etc.). It is not unusual for it to take from a few hours to a few days for the RO permeate quality to stabilize, especially after high pH cleanings.

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KING LEE TECHNOLOGIES MEMRANE CLEANING PROTOCOL (FOR DIAMITE BFT)

Diamite™ Series

Liquid Membrane Cleaners

The Diamite Series is a line of liquid membrane cleaners designed to remove a variety of organic and inorganic foulants. All Diamite cleaners are highly concentrated and easy to use.

Product	Ideal For	Membrane	Mixing Ratio	pH
Diamite ACA	Microbiological Matter, Silt, Organics, Particulates, Colloids, Acid Insolubles.	Cellulose Acetate	1 gallon to 40 gallons of water.	Med.
Diamite LpH and ZpH	Fe, CaCO3, Metal Oxides, Inorganic Salts, Acid Solubles.	All Types	1 gallon to 40 gallons of water.	Low
Diamite AFT and BFT	Silt, Organics, Particulates, Colloids, Microbiological Matter, Acid Insolubles.	Thin Film Composite	1 gallon to 40 gallons of water.	High

Diamite ACA

Diamite ACA was designed to optimize cellulose acetate membrane performance by effectively removing microbiological foulants, organics, silt, and particulates from the membrane surface. It is buffered to not cause hydrolysis of cellulose acetate.

Diamite LpH and ZpH

Diamite LpH and ZpH are ideal for the removal of acid soluble scale including iron, calcium carbonate and metal oxides. They are compatible with thin film composite and cellulose acetate membranes. The mild acidic liquids are convenient and safe to use.

Diamite AFT and BFT

Diamite AFT and BFT were designed to aggressively remove silt, organics, particulates, colloids and microbiological foulants from thin film composite membranes. The unique formulations include a highly effective sanitizing agent that eliminates the need for hydrogen peroxide, formaldehyde, and other membrane disinfectants as a post or pretreatment to membrane cleaning.

Cleaning Procedures

1. Prepare system for cleaning and fill cleaning tank with good quality water.
2. Use 1 gallon of Diamite Cleaner for every 40 gallons of cleaning solution mixed.
3. Re-circulate the cleaning solution for a minimum of 1 hour. Heavily fouled membranes may require a static soak.
4. For best results:
   Temperature – elevated, but should not exceed 110°F.
   Pressure – minimal (not to exceed 60 psig).
   Flow rate per unit – 4 inch diameter membrane 9 gpm
   8 inch diameter membrane 35 gpm
5. Rinse system with good quality water for a minimum of 30 minutes at low pressure.
6. Discard system product water for a minimum of 15 minutes after system start-up.

***MSDS available upon request.**

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APPENDIX C
QA/QC Memorandum

MWH
MONTGOMERY WATSON HARZA

To: **Samer Adham, Ph.D.**
From: **Manish Kumar**
Subject: **Evaluation Of Conventional And Membrane Pretreatment For Seawater Reverse Osmosis**
Date: 12/03/04
Reference:

Pilot testing for the Bureau of Reclamation project entitled, *Evaluation Of Conventional And Membrane Pretreatment For Seawater Reverse Osmosis*, was begun in March of 2004 at the Point Loma Waste Water Treatment Plant (PLWWTP) in San Diego, California. To ensure the accuracy and integrity of the data collected, a number of quality assurance and quality control procedures were followed throughout the experiment. This Technical Memorandum (TM) summarizes these procedures for the on-site instrument verification and water quality analysis performed by the project team, including:

- On-line Turbidimeters
- On-line Conductivity Meter
- On-line Dissolved Oxygen (DO) Meters
- Membrane System Thermometers
- Membrane System Pressure Gauges
- Membrane System Rotameters
- Membrane System Run Hour Clock
- Chemical Feed Pumping Rate
- Portable DO/Temperature Meter
- Desktop pH Meter
- Desktop Turbidimeter
- Desktop Ultraviolet (UV) Spectrophotometer
- Desktop Silt Density Index (SDI) Analyzer

The sampling protocol for off-site water quality analysis is also described herein. All off-site water quality analysis were analyzed at one of the following locations: onsite, Point Loma laboratory (PL Lab), MWH Laboratories and Calscience Environmental Laboratories (CEL Lab). All labs have the State of California Department of Health Services (DHS) Environmental Laboratory Accredited Programs (ELAP), and follow the associated QA/QC requirements.

Lastly, this TM provides the QA/QC procedures followed to ensure accurate data management and data analyses of all water quality and operational data collected during this study.

ON-LINE TURBIDIMETERS

Two types of on line turbidimeters systems were used during testing to acquire filtrate turbidities. Permeate turbidities of the US Filter MF system were measured using Hach 1720D turbidimeters while the Conventional filter turbidity was measured using Hach 1720C turbidimeters. Both the 1720 C& D and are designed to accurately measure low range turbidity. Turbidity values were manually collected from each pretreatment train on a daily basis. The following procedures were followed to ensure the integrity and accuracy of this data:

- A primary calibration of the on-line turbidimeters was performed at the beginning of the test period and as needed during testing.
- On-line turbidities were compared to desktop turbidities to verify accurate calibration.
- The manufacturer's specified acceptable discharge flow range for the Hach 1720 C &D is 250 to 750 mL/min. On-line turbidimeter flows were verified weekly (when systems were in operation) with a graduated cylinder and stopwatch, and adjusted as necessary.
- The turbidmeters were periodically cleaned using a 50 ppm free chlorine solution to remove build of ferric hydroxide precipitate and/or algae.

ON-LINE CONDUCTIVITY METERS

On-line conductivity of the RO influent and effluent was monitored using on-line conductivity meters (GLI C53 Contacting Conductivity Analyzer, Hach Co These meters were calibrated at the beginning, and end of the test period using standard solutions; daily comparisons are performed between the on-line conductivity readings and on-site lab results. The first meter was used to measure the feed water to the RO system and was calibrated using a conductivity standard of 2764 µmhos @ 25 °C. The remaining conductivity meters were used for RO permeate and were calibrated using a 23 µmhos @ 25 °C standard.

MEMBRANE SYSTEM TEMPERATURE PROBES

At the beginning of the study, all temperature probes and thermometers were verified at a normal operating temperature (25-30°C) using an NIST thermometer. Monthly verification of system temperature probes and thermometers was performed. The temperature probe on the MF system was not functional for phase II of the testing so the NIST thermometer was directly used for emasuring temperature. The thermometers used to measure the RO influent water were also verified and within 5% error.

MEMBRANE SYSTEM PRESSURE GAUGES

Pressure gauges supplied with the RO systems tested were verified against redundant electronic gages on the RO system . For the MF system the system pressure was verified against recently purchased grade 3A certified pressure and vacuum gauges. The certified pressure and vacuum gauges were manufactured by Ashcroft and have an accuracy of 0.25% over their range (0-60 psi pressure). Where possible, system gauges were removed and tested over the expected range of operating pressures against the verification gauge, using a portable hand pump over the range of normal operating pressures. The pressure gauges for the RO skids were also within 5% error when tested at lower pressures using the Ashcroft gauge.

MEMBRANE SYSTEM ROTAMETERS/MAGMETERS

Membrane system liquid flow rates were verified volumetrically by bucket tests using calibrated containers or graduated cylinders and a stopwatch. The measured flow rate was compared with flows indicated on the rotameters. Measured and indicated flow rates agreed to within 5% for both the MF systems, the conventional systems and the two RO trains. The combined flow rates, concentrate and permeate, of the RO skid were checked volumetrically and were both within 5% error.

MF system air flow rotameters were factory calibrated prior to the study. [*Please note: there exists no practical method of volumetrically verifying the air flow rates during the pilot study.*]

MEMBRANE SYSTEM RUN HOUR CLOCK

All system run hour clocks used during this study are periodically checked for accuracy using a stop watch.

CHEMICAL FEED PUMPING RATE

The peristaltic pumps used for ferric injection to the conventional pilot were continually checked for accuracy. Upon start-up, the pumps were checked on a daily basis; this frequency was decreased to once per week after pumping consistency was demonstrated. The accuracy is verified using a graduated cylinder and stopwatch.

PORTABLE DISSOLVED OXYGEN/TEMPERATURE METER

A hand-held YSI Model 55 dissolved oxygen meter was used to measure DO in the aerobic tank of the MBR systems. The DO meter was factory calibrated prior to the study, and was re-calibrated before every use according to manufacturer's directions. Periodic comparisons between the hand-held meter, and the PL Lab DO sensor were also performed to ensure continued accuracy. The meter membrane and electrolyte solution are replaced as needed.

DESKTOP pH METER

AN Oakton® Acorn pH 6 pH meter was used throughout the study to determine pH of the raw pretreated and RO treated seawater. The meter was calibrated weekly (when used)using a 3 point calibration with buffers 4, 7, and 10.

DESKTOP TURBIDIMETER

A Hach 2100N desktop turbidimeter was used to perform onsite turbidity analyses of feed and permeate samples. Readings were recorded in non-ratio operating mode. The following quality assurance and quality control procedures were followed to ensure the integrity and accuracy of onsite laboratory turbidity data:

- Weekly primary calibration of turbidimeter according to manufacturer's specification.
- Daily secondary standard calibration verification. Two secondary standards (approx. 0.05 NTU, and 19.1 NTU) were recorded after primary calibration and on the remaining working days until the next primary calibration.

DESKTOP UV SPECTROPHOTOMETER

Samples collected for TOC analysis were analyzed for UV-254 absorbency using a Hach DR/4000 UV spectrophoteter. This instrument was returned to the factory for calibration prior to the study; the instrument was "zeroed" prior to each measurement.

WATER QUALITY SAMPLING PROTOCOL

All sample lines were properly sterilized (for microbial samples) and flushed for a minimum of one minute prior to sampling. Sample containers were obtained from the labs performing the analyses and all preservation chemicals were added to the bottles by the lab prior to sampling, when required. Filtering or any other required preparatory steps was also performed by the respective lab performing the analysis. A courier from the MWWD or CEL Labs transported all samples that will be analyzed off site. Standard shipping and packing procedures were followed, including isolating samples and storage of samples in a cooler packed with plastic bubble wrap to prevent breaking of glass sample bottles. Ice packs were added to the coolers containing samples requiring storage at 4 degrees C. The samples were delivered and analyzed within the allotted holding time for each measured parameter.

A chain of custody was filled out on-site by the person performing the sampling and given to the courier when the samples were picked up for delivery. Upon receipt, a representative from the lab signed the Chain of Custody and the samples will be released to their custody. A copy of the signed Chain of Custody was then be sent back to the sampler and was be kept on file at the pilot site.

DATA MANAGEMENT/ANALYSES

All water quality data collected on-site was merged with data obtained from offsite laboratories throughout the study. Operational data was recorded on raw data sheets and routinely inputted into a database. The water quality and operational databases were combined to create a comprehensive database, which was used for data analysis, retrieval, reporting and graphics. All data inputted to the database was checked and verified by the onsite engineer. Lastly, data files were periodically sent to TAC members during the study for analysis.