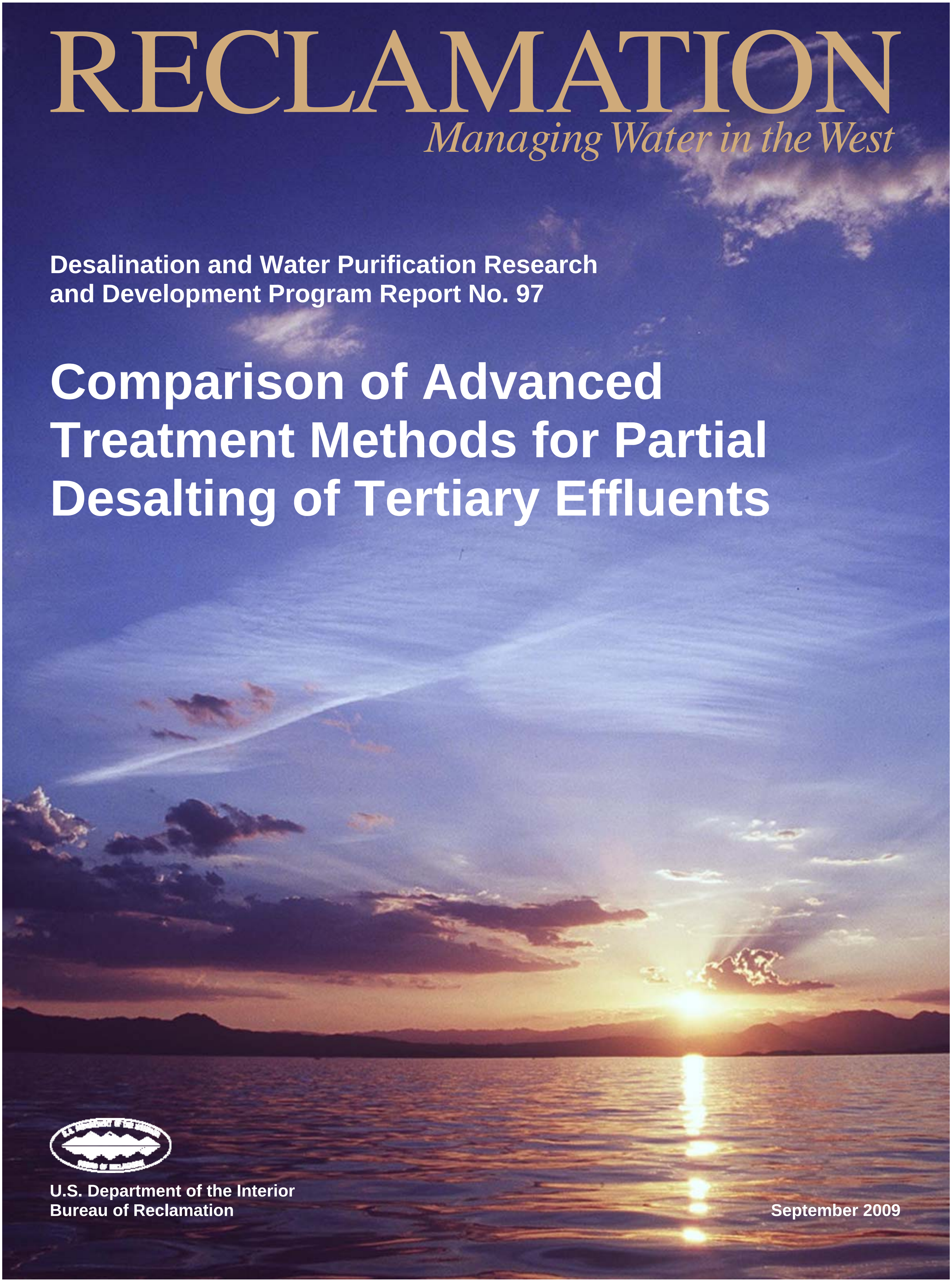

# RECLAMATION

*Managing Water in the West*

Desalination and Water Purification Research and Development Program Report No. 97

# Comparison of Advanced Treatment Methods for Partial Desalting of Tertiary Effluents

U.S. Department of the Interior
Bureau of Reclamation

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This study investigated the feasibility of using two advanced water treatment alternatives—microfiltration (MF) followed by reverse osmosis (RO) and electrodialysis (EDR)—to reduce the salinity of recycled water from the San Jose/Santa Clara Water Pollution Control Plant (SJ/SC WPCP) from a concentration of 750±50 milligrams per liter (mg/L) total dissolved solids (TDS) to either 500 mg/L (38-percent [%] reduction) or 350 mg/L (56% reduction). Pilot scale equipment for the two treatment alternatives was provided by two separate vendors (USFilter and Ionics, Inc.) and operated for approximately 6 months. Lastly, a cost analysis was performed to determine capital and operation and maintenance costs associated with a 50-million-gallons-per-day (mgd) advanced water treatment facility.

Results from this study showed that both advanced water treatment systems (MF/RO and EDR) operated successfully on recycled wastewater effluent from the SJ/SC WPCP and achieved excellent removal of total dissolved solids. This finding is significant as it increases the number of suppliers and feed water sources municipalities can choose to meet reclamation needs using the partial desalination advanced treatment processes. Additionally, based on pilot-scale testing results, it was determined that the high-quality reclaimed water produced by SJ/SC WPCP may be suitable for operation of an EDR system without extensive pretreatment and that a significant cost savings could be realized for an EDR system versus MF/RO. These results, however, were based on short-term pilot testing designed to achieve the goals of this research only (advance treatment comparison for fatal flaws, evaluation of pretreatment alternatives, etc.).

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# Comparison of Advanced Treatment Methods for Partial Desalting of Tertiary Effluents

**Prepared for Reclamation Under Agreement No. 99-FC-81-0189**

*by*


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**Bureau of Reclamation**
**Technical Service Center**
**Water and Environmental Resources Division**
**Water Treatment Engineering Research Group**
**Denver, Colorado**

**September 2009**

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## Disclaimer

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Finally, the authors would like to extend their appreciation to the Bureau of Reclamation for their support in the implementation of this project.

# Acronyms and Abbreviations

| | |
|---|---|
| **Abs** | absorbance |
| **ASME** | American Society of Mechanical Engineers |
| **AWWA** | American Water Works Association |
| **BAT** | best available technology |
| **BOD** | biological oxygen demand |
| **°C** | degrees Celsius |
| **cfu** | colony forming unit |
| **CIP** | clean-in-place |
| **CMF-S** | continuous microfiltration – submerged |
| **dc** | direct current |
| **District** | Santa Clara Valley Water District |
| **DRIP** | Desalination Research and Innovation Partnership |
| **ED** | electrodialysis |
| **EDR** | electrodialysis reversal |
| **$ft^2$** | square feet |
| **FWR** | feed water recovery |
| **GAC** | granular activated carbon |
| **gfd** | gallons per square foot per day |
| **gfd/psi** | gallons per square foot per day per pounds per square inch |
| **gpm** | gallons per minute |
| **Ionics** | Ionics Ultrapure Water Corporation |
| **kgal** | 1,000 gallons |
| **kPa** | kilopascal |
| **MF** | microfiltration |
| **mgd** | million gallons per day |
| **MMF** | multimedia sand filtration |
| **MPN** | most probable number |
| **MWH** | Montgomery Watson Harza |
| **NCWRP** | North City Water Reclamation Plant |
| **NDP** | net driving pressure |
| **NTU** | nephelometric turbidity unit |
| **O&M** | operation and maintenance |

## Acronyms and Abbreviations (continued)

| | |
|---|---|
| **PLC** | programmable logic controller |
| **PP** | polypropylene |
| **psi** | pounds per square inch |
| **psig** | pounds per square inch gauge |
| **ppm** | parts per million |
| **PVDF** | polyvinylidene fluoride |
| **QA** | quality assurance |
| **QC** | quality control |
| **RO** | reverse osmosis |
| **SBWR** | South Bay Water Recycling |
| **SDI** | silt density index |
| **SJ/SC WPCP** | San Jose/Santa Clara Water Pollution Control Plant |
| **TDS** | total dissolved solids |
| **TMP** | trans-membrane pressure |
| **TOC** | total organic carbon |
| **TPS** | transmission pumping station |
| **TSS** | total suspended solids |
| **UF** | ultrafiltration |
| **USEPA** | United States Environmental Protection Agency |
| °F | degrees Fahrenheit |
| **$/kgal** | dollars per 1,000 gallons |
| **%** | percent |
| **µm** | micrometer |
| µS | micro Siemens |

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# 1. Executive Summary

South Bay Water Recycling (SBWR) is a program of the San Jose/Santa Clara Water Pollution Control Plant (SJ/SC WPCP) that supplies recycled water for irrigation and industrial use in the Silicon Valley area of Northern California. Beginning full-scale operation in 1998, SBWR supplies over 10 million gallons per day (mgd) of recycled water to more than 400 customers. The concentration of total dissolved solids (TDS) ranges between 770 and 820 milligrams per liter (mg/L) such that the water is suitable for all current uses. However, as recycled water is used more extensively for evaporative cooling, the TDS likely increases to the point that the water may no longer be suitable for irrigating some salt-sensitive plants and the pretreatment cost required for industrial use increases.

This study investigated the feasibility of two treatment alternatives: 1) microfiltration (MF) followed by reverse osmosis (RO) and 2) electrodialysis reversal (EDR) with various pretreatments to reduce the salinity of recycled water from the SJ/SC WPCP. This study investigates the feasibility of reducing the recycled water salinity from a concentration of 750±50 mg/L TDS to either 500 mg/L (38-percent [%]reduction) or 350 mg/L (56% reduction). Pilot scale equipment for the two treatment alternatives was provided by two separate vendors and operated for approximately 6 months.

MF/RO pilot testing included two phases of operation to evaluate RO performance at an applied flux of 15 gallons per square foot per day (gfd) and feed water recovery of 50–65%. During Phase I, polypropylene (PP) MF membranes were used for pretreatment of tertiary effluents before being fed to the RO pilot. The PP membranes, operating at an 18-gfd flux, were effective at reducing the silt density index (SDI) from 8 to less than 1 during operation (greater than 1,000 hours) such that the RO membranes produced water with a TDS below 10 mg/L (98% salt rejection). MF/RO testing using the PP membranes for pretreatment was terminated due to MF membrane damage by free chlorine in the feed.

In Phase II of MF/RO testing, the PP membranes were replaced with new chlorine-tolerant polyvinylidene fluoride (PVDF) membranes. The PVDF membranes operated at a flux of 30 gfd for 500 hours and maintained the SDI below 1.0 with no observed operational failures. A 97% salt rejection was observed with the RO pilot during Phase II. Results from Phase I and II confirmed that MF/RO is a suitable advanced treatment technology for the reclaimed water produced by the SJ/SC WPCP. Typically, reclaimed wastewater applications have used PP membranes. These test results represent one of the first pilot trials in which PVDF membranes were utilized for the pretreatment of RO feed water in a reclaimed wastewater application.

Pilot testing of EDR was conducted to ensure the system could produce water that met the desired treatment goal and to assess the impact of three different alternative pretreatment methods: granular activated carbon (GAC) followed by multimedia filtration (MMF), MF, and cartridge filtration. Pilot testing results demonstrated that the EDR was capable of meeting the effluent water quality goals for all pretreatment methods tested. Because of the high quality effluent produced by the SJ/SC WPCP (i.e., low suspended solids, turbidity less than (<) 1.0 nephelometric turbidity unit, low plant chemical residuals), the EDR performed well without pretreatment (cartridge filtration only). However, it is important to note that these conclusions were based on short-term testing results (<500 hours per pretreatment condition) and primarily focused on comparing the three pretreatment technologies for EDR membranes. Long-term testing would be required to investigate fully the O&M impacts of operating EDR membranes without significant pretreatment.

The cost of producing 50 mgd of desalinated recycled water with either 350 or 500 mg/L TDS from the San Jose/Santa Clara Water Pollution Control Plant using MF/RO and EDR treatment is shown in the table below. Treatment costs, expressed per thousand gallons ($/kgal), include capital and operating expenses calculated for production of a blended product water with a final salinity of either 350 or 500 mg/L.

| Type of Treatment | Cost of Treatment[1] (Capital and O&M) $/kgal | |
|---|---|---|
| | **350 mg/L** | **500 mg/L** |
| MF/RO | $0.86 | $0.51 |
| MF/EDR | $0.85 | $0.55 |
| EDR[2] | $0.57 | $0.32 |

[1] O&M = operation and maintenance; $/kgal = dollars per kilogallon.

[2] EDR with cartridge filtration only.

Based on the results of this investigation, additional pilot testing is recommended to confirm long-term performance of EDR without pretreatment. We also recommend further investigation to increase the productivity of MF/RO under optimized conditions for the relatively high-quality reclaimed water produced by the SJ/SC WPCP. Such an evaluation should be designed to provide data on the following parameters:

- Operational experience including maintenance requirements
- Evaluation of membrane performance during plant “upsets”
- Full-scale design criteria including a refined cost analysis
- Operator training

The data obtained in this investigation will support the sponsoring local agencies in future decisions on advanced water treatment. The data generated in this study are useful for selecting the most appropriate technology for improving the quality of recycled water in this area. Since the study indicates that electrodialysis is the lower cost alternative, the agencies could facilitate the design of full-scale facilities by validating the pretreatment requirements and cost estimates reported here. If, however, it is determined that nonionic contaminants (e.g., pharmaceutically active and endocrine disrupting compounds) also need to be removed, RO may be required. This study will allow the agencies to estimate the cost required to use reverse osmosis to both remove emerging compounds and reduce the TDS of the recycled water. As a result of this study, the agencies are better able to evaluate advanced water treatment designs and are better prepared to operate and maintain future facilities. The City of San Jose is particularly grateful to the Bureau of Reclamation for their early and continued support of this work as well as to the other project co-sponsors.

# 2. Introduction

## 2.1 Background

South Bay Water Recycling (SBWR) is a program of the San Jose/Santa Clara Water Pollution Control Plant (SJ/SC WPCP) that supplies recycled water for landscape irrigation and industrial use to three cities in the Silicon Valley area of northern California. This study was designed to evaluate the feasibility of improving recycled water quality through using two advanced treatment technologies, microfiltration (MF) followed by reverse osmosis (RO) and electrodialysis reversal (EDR).

The SBWR system was constructed primarily to reduce effluent discharges into the south end of San Francisco Bay. In 1989, the San Francisco Regional Water Quality Control Board limited the plant's discharge to 120 million gallons per day (mgd) when they determined that plant effluent converted salt marsh to fresh marsh and reduced the habitat of two endangered species; the salt marsh harvest mouse and the California clapper rail. In response to this order, the cities of San Jose and Santa Clara (joint owners of the plant that also serves six other cities and three sanitary districts) prepared the South Bay Action Plan. The plan consisted of three components: 1) water conservation, 2) marsh mitigation and 3) water reuse. The water reuse component was accomplished through constructing a 60-mile recycled water distribution system, including four pump stations and a reservoir, with a capacity to distribute peak recycled water flows of up to 50 mgd. The project was partially funded by a construction grant from the Bureau of Reclamation (Reclamation) through its Title XVI program (Public Law 102-575).

SBWR began full-scale operation in 1998 and now supplies over 10 mgd of recycled water during the summer months to more than 400 customers for landscape irrigation and industrial use. The total dissolved solids (TDS) concentration ranges between 770 and 820 milligrams per liter (mg/L) which is suitable for all current uses. However, the TDS likely may increase to the point that recycled water no longer may be suitable for irrigating some salt-sensitive plants. The pretreatment cost for industrial use may increase because the water is used mainly for evaporative cooling.

## 2.2 Objectives of the Study

In 1999, Reclamation awarded the City of San Jose a research cooperative agreement (Reclamation agreement number 99-FC-81-0189) to investigate the feasibility of using advanced water treatment to reduce the salinity of recycled water for industrial use. Montgomery Watson Harza (MWH) was selected as the principal investigator for the study, and additional funding was obtained from the Santa Clara Valley Water District (in conjunction with the Metropolitan Water

Districts of Southern California) and the WateReuse Foundation. USFilter and Ionics Ultrapure Water Corporation (Ionics) provided pilot water treatment equipment that was used in this investigation.

The purpose of this pilot study was to compare nonthermal demineralization processes for partial desalination of nonpotable recycled water and determine if operation issues, such as excessive membrane fouling causing process interruption/failure, existed with the water quality of tertiary treated wastewater produced by the SJ/SC WPCP. MF/RO and EDR were evaluated to reduce effluent salinity from a concentration of approximately 750±50 mg/L TDS to either 500 mg/L (38-percent [%] reduction) or 350 mg/L (56% reduction).

## 2.3 Advanced Water Treatment Processes

RO and EDR, two nonthermal demineralization processes, were chosen because they are recognized by the water treatment community as viable alternatives for demineralization of tertiary treated wastewater to produce water that could be used for nonpotable applications, including industrial and landscape irrigation uses. With further treatment, partially desalinated water also could be adapted to specialized applications (like ultra-pure water for manufacturing electronic products). The different treatment processes were analyzed to determine the most cost-effective demineralization technology.

### 2.3.1 Reverse Osmosis

RO is a pressure driven membrane separation process where dissolved solutes are separated from the solution by forcing the water through a semi-permeable membrane under a pressure greater than the osmotic pressure of the solution. The most common type of RO membrane module used is the spiral-wound configuration. As shown in figure 2-1, two sheets of the membrane are placed back to back, separated by a spacing fabric that acts as a permeate channel. Three sides of the sheet are glued together to form the envelope or leaf. The open end of the leaf is attached to the central permeate tube. A feed stream spacer is placed between a pair of membrane leafs to allow the feed water to flow across the membrane surface. Finally, the leaves and feed spacers are spirally rolled into a cylindrical shape and sealed to create a tightly wound element.

Individual RO membrane elements are housed in cylindrical pressure vessels. As shown in figure 2-2, feed and concentrate flow through the feed-side channels in a straight path parallel to the direction of the permeate collection tube. Water penetrates the membrane and is collected in the center permeate tube. The remaining water passes the element and exits through the concentrate port of the pressure vessel. Typically, several elements are housed in series in a pressure vessel in which the concentrate from one element serves as the feed to the next in series.

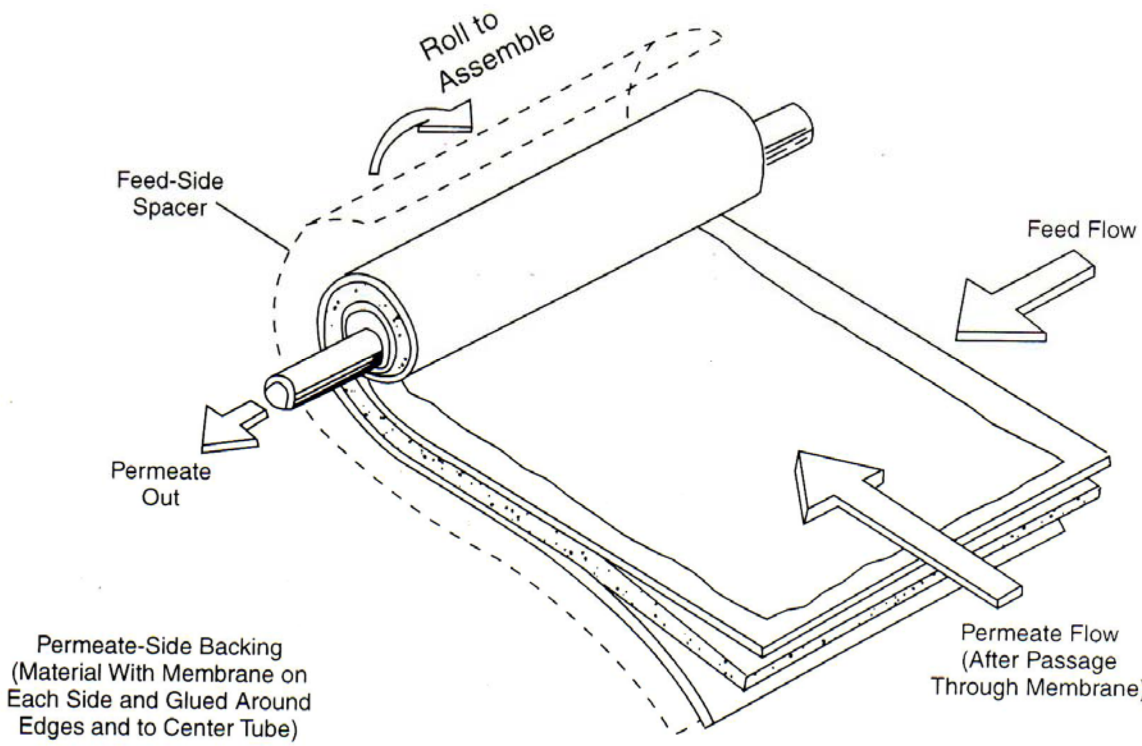


**Figure 2-1. Reverse osmosis spiral wound module (American Water Works Association [AWWA] 1999).**

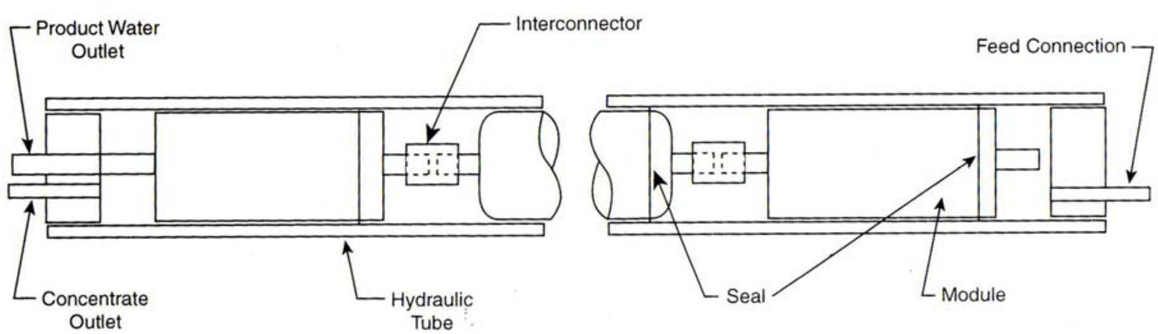


**Figure 2-2. Reverse osmosis pressure vessel assembly (AWWA 1999).**

RO has been selected as a best available technology (BAT) by the United States Environmental Protection Agency (USEPA) for removing inorganic contaminants, such as sulfate and nitrate, which can comprise a large percentage of the TDS present in a water or wastewater. It has been tested extensively for treatment of reclaimed water and several full-scale facilities have been constructed. A list of recent literature references discussing RO full-scale installations, operational experience, and applications are provided in appendix A.

### 2.3.2 Electrodialysis Reversal

In the EDR process, charged ions are removed from the solution by applying an electrical potential across a stream of water. This causes the ions to move towards the opposite charged electrode (figure 2-3). Ion selective membranes separate the stream from the electrode, allowing only positive or negatively charged ions to pass through. These membranes are arranged alternately, with an anion selective membrane followed by a cation selective membrane. A spacer sheet is then placed between these two membranes forming channels in the EDR cell. As the electrodes are charged and feed water flows along the product water spacer at right angles to the electrodes, the anions (like chloride and carbonate) in the water are attracted and diverted through the anion selective membrane towards the positive electrode. This dilutes the salt content of the water in the product water channel. The anions pass through the anion selective membranes but cannot pass through the cation selective membrane and, hence, the anions are concentrated in the brine channel. Similarly, cations, like calcium and sodium under the influence of the negatively charged electrode, pass through the cation selective membrane and are trapped in the brine channel on the other side. This results in concentrated and dilute solutions being created in the spaces between the alternating membranes.

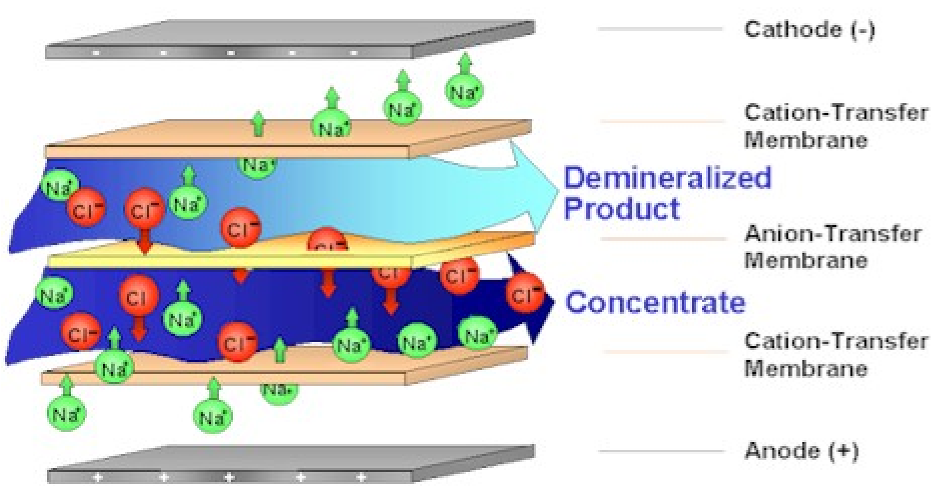


**Figure 2-3. Schematic to illustrate the electrodialysis (ED) process (Ionics, Inc.).**

These spaces, bound by two membranes (one cationic and one anionic) are called cells. The cell pair consists of two cells, one from which the ions migrated (dilute cell for product water) and the other in which the ions concentrate (the concentrate cell for the brine). The basic EDR unit consists of several hundred cell pairs, bound together with electrodes on the outside, and is referred to as a membrane stack. Feed water passes through the feed paths in parallel providing a continuous flow of desalted water and concentrate from the stack.

Currently, the City of San Diego's North City Water Reclamation Plant (NCWRP) operates a full-scale demineralization facility utilizing EDR technology to reduce the salinity of reclaimed water. The water quality after tertiary treatment is similar to that of the SJ/SC WPCP and minimal pretreatment (only cartridge filtration) is being used. During the first year of full-scale operation, extensive EDR membrane fouling occurred; and frequent clean in place (CIP) cleanings (even EDR membrane replacements) were required to maintain membrane integrity.

The NCWRP and Ionics determined that the EDR failures were associated with excess amounts of alum that was fed to the secondary and tertiary processes to help reduce the total suspended solids (TSS). The excess coagulant ended up getting into the membrane stack, causing them to become severely damaged. Additionally, it was later discovered that a nearby agency was routinely dumping clarified sludge down the sewer directly adjacent to the wastewater plant. As a result, this dumping caused a plant upset condition; and excessive chemical feeds were required to treat the wastewater. Consequently, when this plant upset condition occurred, the EDR membranes were damaged shortly afterwards and needed to be replaced.

Today, the NCWRP uses ferric chloride, instead of alum. Operation of the EDR also has been modified such that if an "upset" condition is experienced (i.e., nephelometric turbidity unit (NTU) greater than (>) 2, excess chemical feed, etc), the EDR is temporarily bypassed for that "upset." After implementing these process changes, the EDR operation has been successful for 3.5 years, and a normal (less aggressive and less costly) maintenance schedule has been followed (Chou 2004, Reahl 2004). In addition to this recent experience with EDR, a list of literature references discussing EDR full-scale installations, operational experience, and applications is provided in appendix A.

### 2.3.3 Membrane Pretreatment

An important aspect of the advanced treatment of recycled water is the selection of the pretreatment process. Membrane filtration has been found to be the ideal pretreatment for selected RO processes in studies conducted by MWH at San Diego (MWH 1997, Desalination Research and Innovation Partnership [DRIP] 2002). Pretreatment of the recycled water using MF or ultrafiltration (UF) helps in particle removal and provides a higher quality of feed water to the advanced treatment process as compared to conventional pretreatment processes.

Additionally, since reclaimed water quality can be highly variable, membrane pretreatment processes provide the additional benefit that the product water quality from the membranes is not dependent on the feed water quality.

# 3. Conclusions and Recommendations

This study has demonstrated that both RO and EDR systems can treat successfully tertiary effluent containing high TDS salts. It was further determined that the EDR system can treat the high quality recycled water produced by the SJ/SC WPCP with only cartridge filtration pretreatment. EDR with cartridge filtration is more cost effective than MF/RO for producing a comparable volume and quality of partially desalted water. This conclusion is based on the successful operation of the EDR unit for about 3 weeks. However, during this time, fouling of the EDR membranes was observed indicating a tendency toward fouling over a longer duration of operation.

## 3.1 Operational Performance

The system flux, feed water recovery (FWR), and product water quality were used to evaluate the operational performance of each desalination process.

### 3.1.1 Reverse Osmosis

- The RO system operated for more than 2,000 hours at an applied flux of 15 gallons per square foot per day (gfd) with a FWR of 50 to 65% using 12 RO membrane elements (DOW BW30-4040) configured in a three-vessel, 2:1 array.

- The RO membranes achieved excellent salt rejection (> 97%).

- Membrane pretreatment (for the RO) using MF membranes reduced influent turbidity from 1 to 0.1 NTU.

- Two types of MF membranes were evaluated: polypropylene (PP) and polyvinylidene fluoride (PVDF). Both types of membranes were capable of meeting the desired product water goals. However, PVDF membranes performed better than the PP membranes due to their chlorine-tolerant characteristic.

- MF fluxes up to 30 gfd were used without excessive membrane fouling

- The silt density index (SDI) was consistently reduced from values as high as 20 to below 1 by the MF pretreatment throughout the course of testing.

### 3.1.2 Electrodialysis Reversal

- The Ionics EDR pilot system operated for more than 2,500 hours as a single stack configuration with up to two electrical stages.

- The EDR system achieved salt rejections ranging from 26 to 57%
- Stable operation was achieved during EDR operation without significant EDR membrane fouling or operational errors.
- Three different pretreatment scenarios were evaluated for the EDR pilot, including granular activated carbon (GAC) followed by multimedia sand filtration (MMF), membrane MF, and cartridge filtration only.
- EDR can be operated using a minimal amount of pretreatment (cartridge filtration only) without excessive fouling, operational failure, or a decrease in product water quality.

## 3.2 Costing Analysis

- Cost estimates (dollars per 1,000 gallons [$/kgal]) for 50-mgd MF/RO advanced treatment were estimated for a blended product of 350 and 500 mg/L (see table 3-1).
- Because GAC for chlorine reduction is considerably more expensive than chemical addition for chlorine removal (i.e., sodium bisulfite), the cost for GAC pretreatment to EDR was not considered.

**Table 3-1. Cost estimates for the different treatment alternatives**

| Type of Treatment | Cost of Treatment (Including Capital and Operation and Maintenance[O&M]) ($/kgal) | |
|---|---|---|
| | 350 mg/L | 500 mg/L |
| MF/RO | $0.86 | $0.51 |
| MF/EDR | $0.85 | $0.55 |
| EDR[1] | $0.57 | $0.32 |

[1] EDR with cartridge filtration only.

## 3.3 Recommended Future Work

Based on the pilot testing of RO/MF and EDR as advanced water treatment technologies for tertiary treated effluent, we recommended long-term performance testing of EDR membranes with and without pretreatment and MF/RO under optimized conditions. Longer term testing will help to provide the following information:

- Long-term operation data
- O&M requirement
- Full-scale design criteria
- Evaluation of membrane performance during planned and unplanned plant "upsets"
- Refined cost analysis
- Operator training

# 4. Materials and Methods

This section contains information concerning the materials and methods used in this study. Details concerning the operation of the pilot equipment are included in "Appendix B, San Jose Operator Experience During Pilot Study."

## 4.1 Testing Site

The test site was located at the SJ/SC WPCP transmission pumping station (TPS) at 700 Los Esteros Road in San Jose, California.

### 4.1.1 Site Background Information

The SJ/SC WPCP is one of the largest advanced wastewater treatment facilities in California. This facility treats and cleans the wastewater of over 1.5 million people that live and work in the 300-square-mile area encompassing San Jose, Santa Clara, Milpitas, Campbell, Cupertino, Los Gatos, Saratoga, and Monte Sereno.

The SJ/SC WPCP has the capacity to treat 167 million gallons of wastewater per day. It is located in the Alviso neighborhood of north San Jose, at the southernmost tip of the San Francisco Bay. The SJ/SC WPCP treatment train includes the following processes:

1. Pretreatment (screening, sedimentation, and grit removal)
2. Primary settling and scum removal
3. Flow equalization
4. Secondary biological nutrient removal consisting of a four-chamber aerobic/anoxic suspended growth activated sludge treatment with partial denitrification
5. Gravity filtration with anthracite coal and sand
6. Chlorine disinfection followed by sulfur dioxide dechlorination (prior to discharge) or gaseous chlorine rechlorination (prior to reuse)

The plant is designed to remove more than 98% of biochemical oxygen demand (BOD) and more than 99% of the TSS. The current advanced wastewater treatment plant has the capacity to treat up to 167 mgd. The flow diagram is presented in figure 4-1.

Most of SJ/SC WPCP final treated water is discharged through Artesian Slough and into south San Francisco Bay. During the summer months, about 10% (10–12 mgd) is recycled through South Bay Water Recycling pipelines for landscaping, agricultural irrigation, and industrial needs around the South Bay.

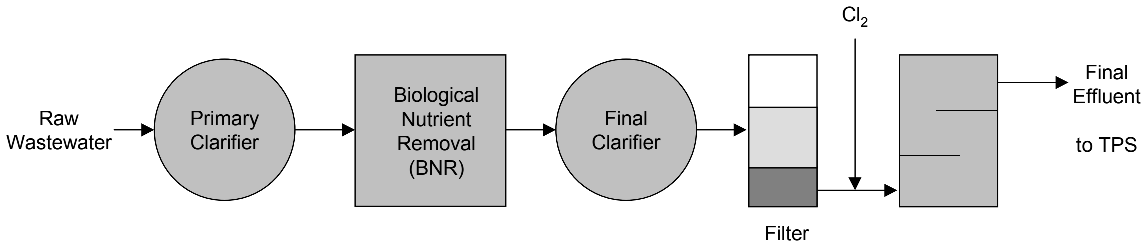


**Figure 4-1. SJ/SC WPCP advanced wastewater treatment flow diagram.**

## 4.2 Feed Water Quality

Tertiary treated wastewater is characterized by relatively high levels of TDS, hardness, and alkalinity; moderate levels of organic material; and relatively low turbidity. This type of water was used as the feed water for pilot testing. All offsite water quality analyses were performed at the City of San Jose's Environmental Services Department laboratory. Table 4-1 presents the typical feed water quality at the pilot site and the analytical method used for all laboratory analyses performed.

**Table 4-1. SJ/SC WPCP average pilot influent water quality[1]**

| Parameter | Concentration | Units | Method |
|---|---|---|---|
| $Cl^-$ | 188 | mg/L | EPA 300 |
| $NO_3$-N | 7.1 | mg/L | EPA 300 |
| $SO_4$ | 96 | mg/L | EPA 300 |
| $Br^-$ | < 1.0 | mg/L | EPA 300 |
| $NO_2$-N | <0.05 | mg/L | EPA 354.1 |
| | | | |
| Al | 0.06 | mg/L | EPA 200.7 |
| Ba | 0.02 | mg/L | EPA 200.7 |
| B | 0.51 | mg/L | EPA 200.7 |
| Ca | 59.1 | mg/L | EPA 200.7 |
| Cr (Total) | < 0.002 | mg/L | EPA 200.7 |
| Fe | 0.07 | mg/L | EPA 200.7 |
| Mg | 31.7 | mg/L | EPA 200.7 |
| $SiO_2$ | 24.0 | mg/L | EPA 200.7 |
| Na | 156 | mg/L | EPA 200.7 |
| Sr | 0.387 | mg/L | EPA 200.7 |

Table 4-1. SJ/SC WPCP average pilot influent water quality[1] (continued)

| Parameter | Concentration | Units | Method |
|---|---|---|---|
| $NH_3$-N | <0.1 | mg/L | SM 4500-(NH3)H |
| Conductivity | 1,250 | umhos/cm | SM 2510B |
| pH | 7.3 | SU | SM 4500H+ |
| TOC | 9 | mg/L | SM 5310B |
| TKN | 0.4 | mg/L | SM 4500 N(org)-C |
| TSS | < 2 | mg/L | SM 2540D |
| Turbidity | 0.7 | NTU | SM 2130B |
| Hardness, total ($CaCO_3$) | 250 | mg/L | SM 2340C |
| Alkalinity, total ($CaCO_3$) | 190 | mg/L | SM 2320B |
| TDS | 750 | mg/L | SM 2540C |
| $UV_{254}$ | 0.109 | 1/cm or $cm^{-1}$ | SM 5910B |
| Fecal coliforms | <1 | MPN; cfu/mL | SM 9221A/9222A |
| Total coliforms | 1 | MPN; cfu/mL | SM 9221D/9222D |
| Heterotrophic plate count | 300 | cfu/mL | SM 9215 |

[1] $Cl^-$ = chloride; $NO_3$-N = nitrate nitrogen; $SO_4$ = sulfate; $Br^-$ = bromide; $NO_2$-N = nitrite nitrogen; Al = aluminum; Ba = barium; B = boron; Ca = calcium; Cr = chromium; Fe = iron; Mg = magnesium; $SiO_2$ = silicon dioxide; Na = sodium; Sr = strontium; $NH_3$-N = ammonia nitrogen; TOC = total organic carbon; TKN = total Kjeldahl nitrogen; $CaCO_3$ = calcium carbonate; cfu/mL = colony forming unit per milliliter; $cm^{-1}$ = a reciprocal centimeter (or wavenumber) used as an energy unit; MPN = most possible number.

### 4.2.1 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.

### 4.2.2 Quality Assurance/Quality Control (QA/QC)

The following QA/QC procedures were followed during pilot testing.

#### *4.2.2.1 Pilot Plants Auxiliary Units*

The pilot plant auxiliary equipment such as electronic pressure sensors, flow meters, volt and amperage meters, and safety switches were not calibrated onsite during the pilot testing startup period as outlined in the *Advance Water Treatment Pilot Study Work Plan*. Selected equipment calibrations occurred during the testing period.

#### *4.2.2.2 Online Monitoring Devices*

The readings from online pH meters, conductivity meters, and thermocouples were verified by comparison to grab samples collected, submitted and analyzed by the City of San Jose Environmental Services Department laboratory.

#### *4.2.2.3 Data Analyses*

Data collected onsite was regularly merged with data obtained from offsite laboratory analyses to form a comprehensive database for analysis, retrieval, reporting, and graphics. A modular database program was developed for this project to include all produced data. All data was checked and verified by the project engineer before and after entry into the database program.

## 4.3 General Pretreatment

To protect the membranes in the advanced treatment processes investigated in this study, effluent from the SJ/SC WPCP was pretreated prior to the RO and EDR units. Different types of chemicals were added to remove free chlorine that can degrade the membranes. Also, various types of filters (including MF) were used to remove suspended solids that can foul the membranes, increasing maintenance costs and reducing runtimes.

### 4.3.1 Dechlorination

Tertiary treated water from the SJ/SC WPCP is disinfected with chlorine before being diverted to the SBWR transmission pumping station. Free chlorine concentrations observed in the effluent, which feeds the pilot plant typically average 1–2 mg/L, with peaks of 4–8 mg/L. However, on occasion, the concentration of free chlorine can reach 25 mg/L. Such spikes occur most often when the recycled water demand drops rapidly after a period of high use and are thought to result from delays in the automatic reduction of chlorination rates. GAC, addition of chloramines, and addition of sodium bisulfite were investigated to reduce the free chlorine concentration in the feed water.

#### *4.3.1.1 Granular Activated Carbon*

Two Ionics TurboFlo GAC contactors were used for removing free chlorine in the feed water by reduction (figure 4-2). The vessels were 100 pounds per square inch (psi) American Society of Mechanical Engineers code-stamped carbon steel with a capacity of 42 cubic feet. The GAC contactors were plumbed in series to ensure that, given the variability of free chlorine present in the feed water, complete dechlorination would be maintained. GAC contactors were used exclusively for the EDR pilot system.


**Figure 4-2. Ionics TurboFlo GAC contactors.**

#### *4.3.1.2 Chloramination (Ammonia)*

Addition of ammonia to the feed converts harmful free chlorine concentrations to chloramines. The ammonia feed pump was regulated and adjusted daily to ensure complete conversion of free chlorine to chloramines. Chloramination was used exclusively on the MF pretreatment equipment while the PP (free-chlorine sensitive) hollow fiber membranes were in use.

#### *4.3.1.3 Sodium Bisulfite*

Dechlorination also was achieved using sodium bisulfite (figure 4-3). A dedicated pump was used to maintain a 4- to 5-mg/L dose of sodium bisulfite to chlorinated water entering the demineralization equipment. Sodium bisulfite was used as a pretreatment for both the RO and EDR pilot systems.

### 4.3.2 Particulate Removal

For many membrane treatment processes, pretreatment is required to reduce suspended solids and colloidal matter concentrations for the prevention of membrane fouling. Suspended or undissolved matter in the feed water may deposit on the surface of the membrane as the water passes along or through the membrane. A buildup of these deposits may eventually reduce the flow of water through the membrane and cause the applied pressure to increase. For both RO and EDR systems, membrane fouling may, in part, contribute to a decrease in the salt rejection of the system, causing deterioration in the product water quality.


**Figure 4-3. Sodium bisulfite dechlorination feed pump and storage tank.**

#### *4.3.2.1 Conventional Pretreatment*

The use of prefilters as a pretreatment is common among all membrane systems to help prevent membrane fouling and minimize mechanical damage that may be caused by particulate matter. Prefiltration typically is accomplished using cartridge filters (AWWA 1999). Cartridge filters (5–15 micrometers [μm]) were used as pretreatment for both RO and EDR processes.

Additionally, MMF was pilot tested as a pretreatment to the EDR system. The pilot multimedia prefilter contained 10 cubic feet of filter media consisting of various sizes of distinctly layered sand. MMF can accommodate a flow of up to 100 gallons per minute (gpm) with a maximum pressure loss of 26 psi.

A major feature of current RO plants is the use of conventional pretreatment (Wilf 2001). Conventional pretreatment has several disadvantages including:

- Lack of an absolute barrier to suspended particles and colloidal matter that can severely limit RO performance.
- Fluctuation in feed water quality to RO.
- Need for frequent backwashing of the filters used.
- Biological growth in filters leading to RO membrane biofouling.
- Chlorine used for biofouling control in filters can reach RO membrane and cause damage

These operational issues lead to higher costs because operation of RO systems at conservative operational parameters is necessary.

#### *4.3.2.2 Membrane Pretreatment*

Membrane pretreatment using MF or UF provides several advantages over conventional pretreatment. MF or UF can provide an absolute barrier to microorganisms, suspended particles, and colloids, leading to stable and high quality RO feed water. Consequently, the RO system can be operated at more aggressive conditions resulting in savings in both operational and overall costs.

**USFilter Memcor Pilot Equipment**

The USFilter Memcor CMF-S 16S10T pilot system tested included the following components (figure 4-4):

- Feed pump
- Up to 16 membrane modules
- Air compressor
- Data logger


**Figure 4-4. Memcor CMF-S 16S10T pilot unit.**

The pilot system is equipped with a centrifugal pump and is run in direct, or dead-end, filtration mode; that is, all feed passes through the membrane while filtering. An inlet feed valve, responding to level sensors in the tank, controls the water level above the modules, so that they remain completely submerged. During filtration, water is drawn through the fiber walls (outside to inside) under suction. The microfiltered water then is directed to the filtrate tank or to service. All solids and particulate matter are removed from the feed water and accumulate on the outside of the fiber walls. A timer initiates regular backwashes after 30 minutes. The backwash uses air to scour the fibers, while a small amount of filtrate is pushed backwards through the fibers (inside to outside) to remove the fouling layer. The backwash duration is approximately 2.5 minutes. Backwash water is drained, and the tank is refilled prior to restarting filtration.

**Figure 4-5. CMF-S hollow fiber membrane modules.**

The hollow fiber MF membrane modules can be arranged in four groups of four modules each (figure 4-5). PP and PVDF submerged modules, each with a nominal pore size of 0.1 micron, were evaluated during pilot testing. The membrane element specifications are presented in table 4-2.

The unit was equipped with a programmable logic controller (PLC) from Allen-Bradley Inc., pressure transmitters, flow meter, chlorine meter, conductivity meter, and temperature measurement. The pressure transmitters monitored the transmembrane pressure (TMP). Online instruments were connected to the PLC as well as a MEMLOG™ data logger. Control functions and data display were accessed via an operator interface mounted on the front of the control panel.

**Chemical Consumption**

Ammonia was added to the feed water at a dose of approximately 4 to 5 mg/L to eliminate free chlorine from harming the PP membranes. When PVDF membranes were used; no ammonia was necessary to reduce free chlorine. No chemicals were used during routine backwash. CIP using chemicals was performed periodically to remove foulants. CIPs were performed with citric acid (10 pounds) as needed during the testing period.

**Waste Production**

Backwash waste contains naturally occurring particulates and organics at significantly higher concentrations (20 to 30 times) than the raw water. Approximately 220 to 270 gallons of backwash waste were generated with each backwash. Cleaning chemical wastes consisted of citric acid solutions at pH 2 to 2.5. Approximately 140 to 170 gallons of waste were generated per cleaning.

**Table 4-2. CMF-S membrane module specifications[1]**

| | Units | Value | Value |
|---|---|---|---|
| Manufacturer | | US Filter | US Filter |
| Membrane model and ID number | | 119066 (for CMF-S) | 119018 (for CMF-S) |
| Membrane commercial designation | | S10V | S10T |
| Approximate size of element (length x diameter) | ft (m) | 1.186 x 0.131 (3.892 x 0.433) | 1.186 x 0.131 (3.892 x 0.433) |
| Active membrane area | $ft^2$ ($m^2$) | 272 (25.3) | 335 (31.09) |
| Number of fibers per module | | 9,600 | 14,500 |
| Number of modules (operational) | | 9 in 16S10T pilot unit | 11 in 16S10T pilot unit |
| Inside diameter of fiber | mm | 0.5 | 0.39 |
| Outside diameter of fiber | mm | 0.8 | 0.65 |
| Approximate length of fiber | m | 1.1 m exposed length | 1.1 meter exposed length |
| Flow direction | | Outside-in | Outside-in |
| Nominal membrane pore size | micro | 0.1 µm | 0.1 µm |
| Absolute membrane pore size | micron | 0.2 µm | 0.2 µm |
| Membrane material/construction | | Polypropylene | Polypropylene |
| Membrane surface characteristics | | Hydrophobic | Hydrophobic |
| Membrane charge | | Neutral | Neutral |
| Maximum transmembrane pressure | kPa (psig) | 120 (17.4) | 85 (12.3) |
| Acceptable range of operating pH values | | 2–10 | 2–14 |
| Acceptable range of operating temperatures | °F (°C) | 32–104 (0–40) | 34–104 (0–40) |
| Chlorine/oxidant tolerance | ppm | 200 | <0.05 |

[1] ft = foot; m = meter; $ft^2$ = square foot; $m^2$ = square meter; mm = millimeter; kPa = kilopascal; psig = pounds per square inch gauge; °F = degree Fahrenheit; °C = degrees Celsius; ppm = parts per million.

### 4.3.3 Phase I – MF Pretreatment Using PP Membranes

A process flow schematic of the Phase I pilot operation is provided in figure 4-6. Tertiary treated wastewater from the SJ/SC WPCP was dosed with 4 to 5 mg/L ammonia to form chloramines before entering the USFilter CMF-S. The CMF-S operated at a total flow rate of 45 gpm using 11 PP hollow fiber membrane modules (approximately 4 gpm per module). At this flow rate, the operating flux was 22 gfd. A portion of the MF permeate was stored in a separate backwash tank and used for regular backwashing of the membranes every 30 minutes. The remaining MF permeate was stored in a large break tank downstream, which served as the feed to the RO system.

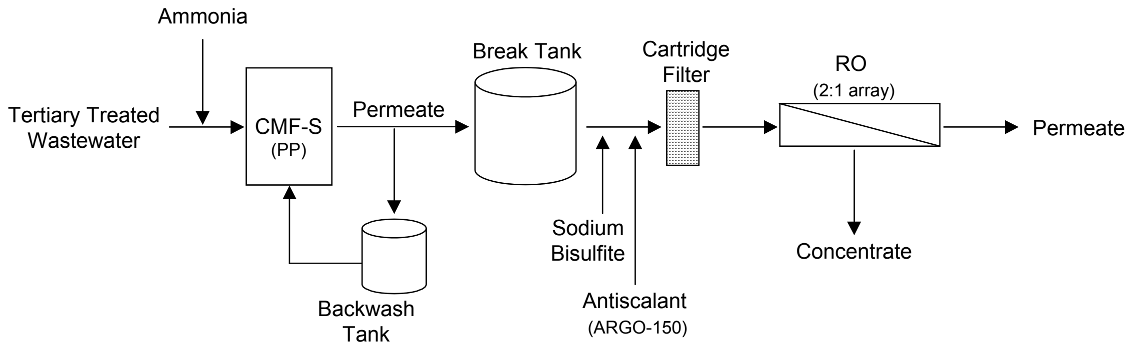


**Figure 4-6. RO pilot testing schematic (PP membrane pretreatment).**

### 4.3.4 Phase II – MF Pretreatment Using PVDF Membranes

A process flow schematic of the Phase II pilot operation is provided in figure 4-7. Tertiary treated wastewater from the SJ/SC WPCP, with an average free chlorine residual of 1 mg/L, was fed directly to the USFilter CMF-S system. Dechlorination, however, was not necessary since PVDF hollow fiber membrane modules were used.

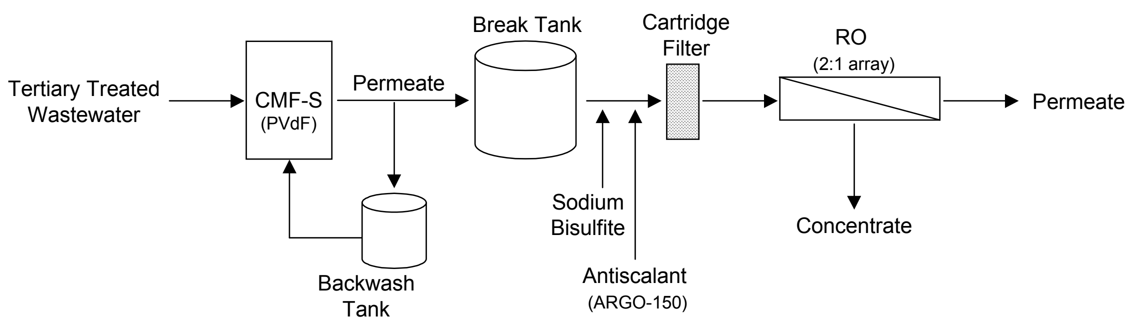


**Figure 4-7. RO pilot testing schematic (PVDF membrane pretreatment).**

## 4.4 Reverse Osmosis Equipment

A modified USFilter "H" series RO, model number ROSLH 3180 (figure 4-8), was used. Three vessels configured in a 2:1 array with a minimum FWR of 50% were used. It originally was proposed to test two different types of RO membranes. However, during commissioning of the pilot unit, the equipment vendor recommended evaluating two different pretreatment membranes instead of two different RO membranes due to the presence of variable free chlorine concentrations in the feed water. The RO membranes elements evaluated were Dow Filmtec brackish water membranes (part number BW30-4040). Each vessel housed four RO membrane elements. The RO membrane element specifications are presented in table 4-3.


**Figure 4-8. RO pilot system.**

**Table 4-3. RO membrane element specifications[1]**

| | Units | Value |
|---|---|---|
| Manufacturer | | Dow Filmtec |
| Membrane model and ID number | | 80783 |
| Membrane commercial designation | | BW30-4040 |
| Approximate size of element | length x diameter – in (mm) | 40 x 3.9 (1,016x99) |
| Active membrane area | $ft^2$ ($m^2$) | 82 (7.6) |
| Number of modules | | 12 in ROSLH 3180 pilot unit |
| Applied pressure | psig (bar) | 225 (15.5) |
| Permeate flow rate | gpd ($m^3$/d) | 2,400 (9.1) |
| Stabilized salt rejection | % | 99.5 |
| Membrane type | | Polyamide thin-film composite |
| Maximum operating temperature | °F (°C) | 113 (45) |
| Maximum operating pressure | psi (bar) | 600 (41) |
| Maximum feed flow rate | gpd ($m^3$/d) | 16 (3.6) |
| Maximum pressure drop | psig (bar) | 15 (1.0) |
| pH range, continuous operation | | 2 to 11 |
| pH range, short-term cleaning | | 1 to 12 |
| Maximum feed SDI | | SDI 5 |
| Free chlorine tolerance | ppm | <0.1 |

[1] $m^3$/h = cubic meters per hour.

### 4.4.1 Pretreatment

Several RO pretreatment options were available and used for pilot testing including:

- MF using chlorine-sensitive PP membranes
- MF using chlorine-tolerant PVDF membranes
- Sodium bisulfite
- Antiscalant (Argo 150; 1 to 2 mg/L)
- 5- to 15-µm cartridge filtration

### 4.4.2 Waste Production

The RO was operated at FWR between 50 and 65%. The maximum feed water flow was 20 gpm resulting in 5 to 10 gpm of concentrate generated by the RO process.

## 4.5 Electrodialysis Reversal Equipment

The EDR pilot system was tested using various pretreatment technologies to determine which is most suitable.

### 4.5.1 Phase I – Baseline Testing of EDR

A general process flow schematic of the EDR pilot operation is provided in figure 4-9. Tertiary treated wastewater from the SJ/SC WPCP was dechlorinated by two GAC contactors in series. Suspended solids removal was achieved using a multimedia sand filter and a 5- to 10-µm cartridge filter. To prevent biological fouling on the EDR membranes, a small amount of the chlorinated feed water was bypassed to maintain a chlorine residual of 0 to 0.5 mg/L.

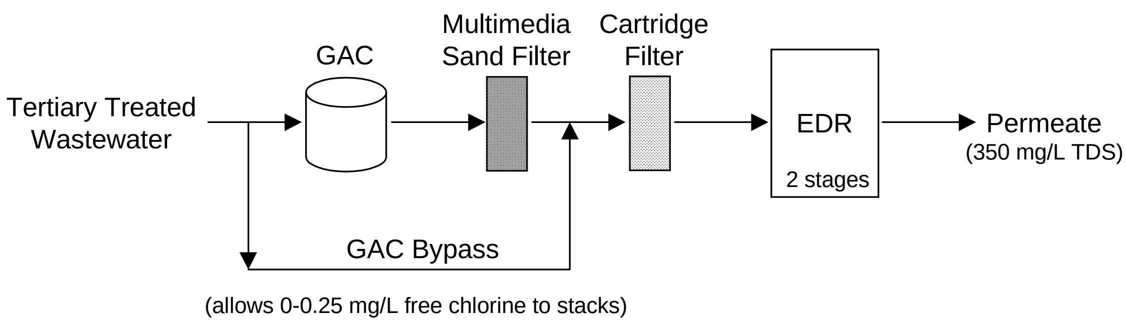


**Figure 4-9. EDR pilot testing schematic (Phase I – baseline).**

### 4.5.2 Phase II – Evaluation of Different Pretreatment Alternatives for EDR

Tertiary treated wastewater from the SJ/SC WPCP was fed to the EDR pilot plant after being pretreated with either (a) GAC/MMF, (b) MF, or (c) cartridge filtration only (see figure 4-10). For the GAC/MMF pretreatment configuration, a small amount of the chlorinated feed water was allowed to enter the stacks to maintain a free chlorine residual of 0.5 to 1 mg/L to minimize biological fouling on the EDR membranes. In the remaining configurations, no dechlorination of the feed water was performed. Final suspended solids removal was achieved using a 5- to 10-µm cartridge filter.

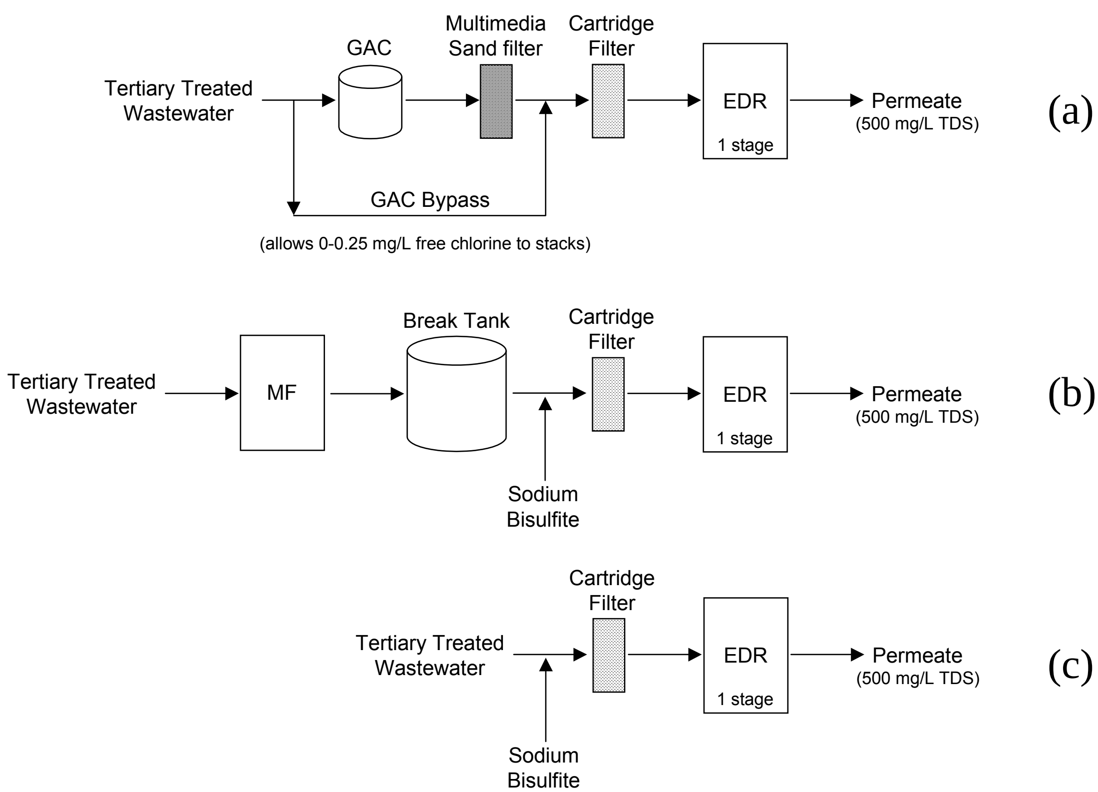


**Figure 4-10. EDR pilot testing schematic (pretreatment alternatives).**

The trial equipment consisted of an Aquamite V with a bipolar membrane stack. The capacity of the Aquamite V was 15,000 to 35,000 gpd. The maximum feed flow for this unit was 60,000 gpd. The Aquamite V supported an electric power supply of 480/460/380/220 volts, 50/60 Hertz, three-phase and was supplied by direct current (dc) at three phases, full wave with silicon diode rectifiers.

As shown in figure 4-11, the EDR pilot system was installed in Ionics' mobile pilot plant trailer. The trailer housed the EDR unit, control panel, multimedia filter, and cartridge filter. Located outside of the trailer were the two GAC filters and the sodium bisulfite feed pump and tank.

**Figure 4-11. Electrodialysis reversal pilot plant.**

The EDR operated at 22 to 27 gpm to continually produce demineralized water without constant chemical addition during normal operation. Current was supplied at 2 to 4 amps depending on the specific water quality goals to be achieved. Membrane fouling and scaling was controlled by using electrical polarity reversal every 15 minutes.

Typically, EDR is configured using multiple stages to provide the maximum membrane surface area and retention time to remove a specified fraction of salt from the demineralized stream. Two types of staging are used: hydraulic and electrical. For this study, the Aquamite V pilot unit operated as a single stack with two electrical stages that could be controlled independently to achieve a desired water quality. Electrical staging was accomplished by inserting additional electrode pairs into the membrane stack to provide maximum salt removal rates while avoiding polarization and hydraulic pressure limitations.

#### *4.5.2.1 Pretreatment*

Several pretreatment options were available for pilot testing prior to EDR treatment including:

- GAC
- Sodium bisulfite
- MMF
- MF
- 5- to 10-µm cartridge filtration

Although GAC and sodium bisulfite were used for dechlorination, it was determined that membrane biofouling due to algal growth could be controlled by maintaining a small amount of residual free chlorine in the feed to the EDR stack.

This was achieved by bypassing a portion of the chlorinated feed water and allowing it to enter the dechlorinated feed stream to the EDR stack. During this study, the average free and total chlorine concentrations in the pilot feed water were 1–2 mg/L and 4–5 mg/L, respectively. Bypassing approximately 25% of the flow to the EDR allowed up to 0.5 mg/L free chlorine to be maintained in the EDR stack.

# 5. Results and Discussion

## 5.1 Reverse Osmosis with Membrane Pretreatment

MF pretreated water was continuously fed to the RO pilot system at a flow rate of approximately 20 gpm. Sodium bisulfite and antiscalant were added to the MF pretreated water to control RO membrane fouling and protect the membrane elements from chemical damage due to free chlorine or chloramines. Two types of membranes were evaluated for use in the CMF-S: PP hollow fiber membranes and PVDF membranes.

### 5.1.1 Phase I – MF Pretreatment using PP Membranes

The silt density index, a measure of the suspended solids concentration in water, was measured both on the feed and permeate of the CMF-S to characterize the fouling potential of the RO feed water. The average SDI in the tertiary treated wastewater was 7.0 (figure 5-1). It is important to note that certain feed samples caused the 0.45-µm filter used for SDI measurement to become plugged within 15 minutes. In those instances, a modified SDI was calculated by determining the time required for the filter to become completed plugged. After membrane pretreatment, the SDI of the feed water to the RO averaged 1.0. The maximum SDI allowed to the RO membranes, as recommended by the manufacturer, is SDI 5.

The CMF-S was operated continuously for 1,100 hours at a 20-degree-Celsius (°C) temperature-corrected flux of approximately 19 gfd (figure 5-2). This flux was conservative and recommended by the equipment vendor. The transmembrane pressure was maintained between 1 to 1.5 psi during the first 450 hours of operation and then increased to 3 psi during the remainder of operation (figure 5-3). The increase in TMP was due to the fouling of the PP membranes over time. The resulting drop in the membrane permeability or specific flux can be seen figure 5-2 for the last 650 hours of operation. Despite the minimal fouling trend observed during this test period, the CMF-S was able to remove continually suspended solids in the raw feed water, making it suitable to be subsequently treated using RO membranes. Additionally, it generally is expected that performing a CIP chemical cleaning at the end of this operational period could recover the membrane permeability. Unfortunately, membrane performance after a CIP cleaning was not evaluated due to the ammonia feed system failure that damaged the PP membrane by exposing them to free chlorine.

The RO system operated continuously for 1100 hours using MF pretreated water. A summary of the RO operation and performance data is presented in table 5-1.

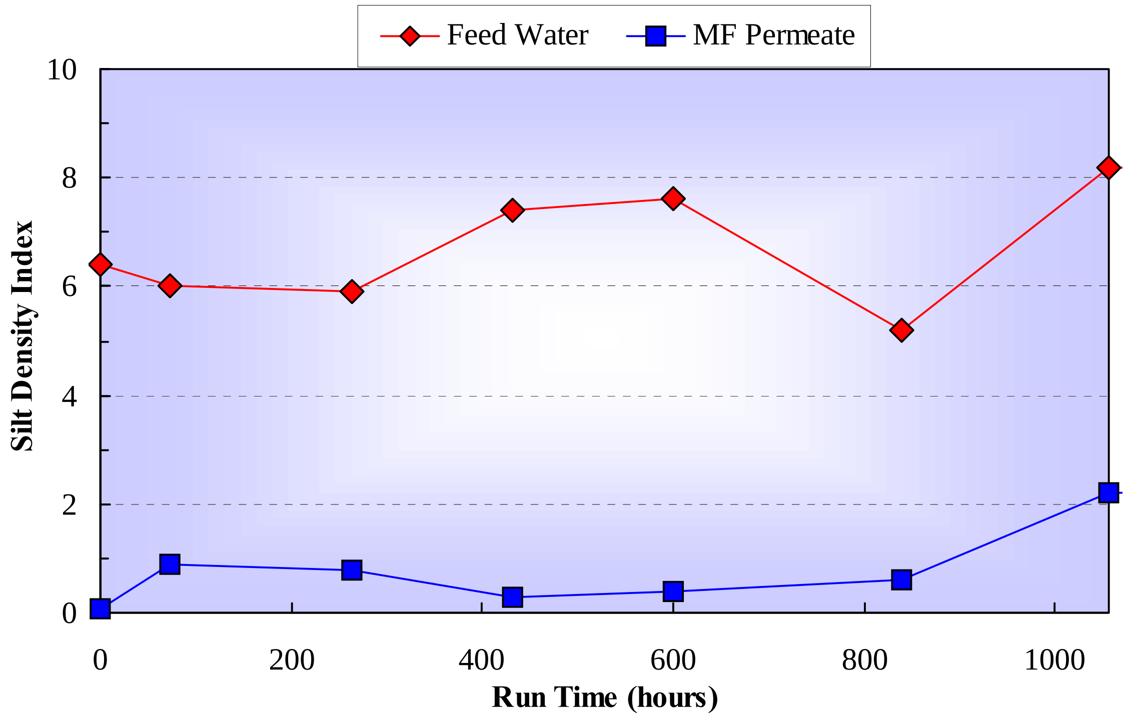


**Figure 5-1. SDI measurements for Phase I using PP membranes.**

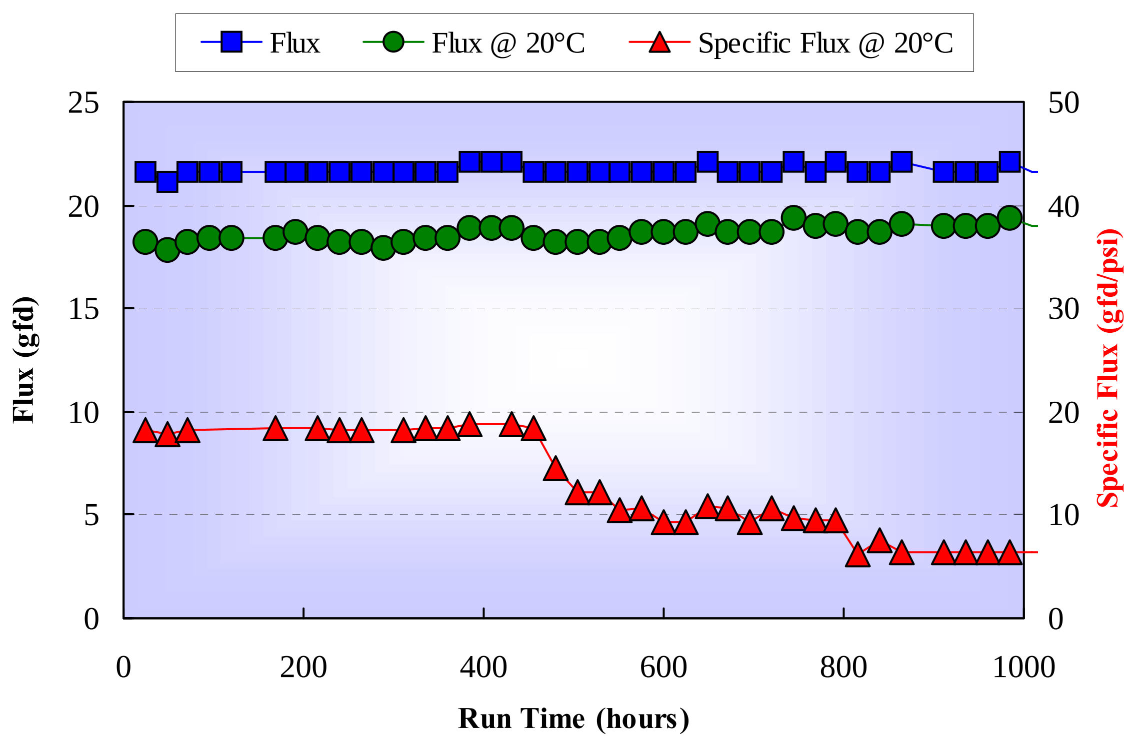


**Figure 5-2. CMF-S temperature-corrected operating flux and specific flux for Phase 1 using PP membranes.**

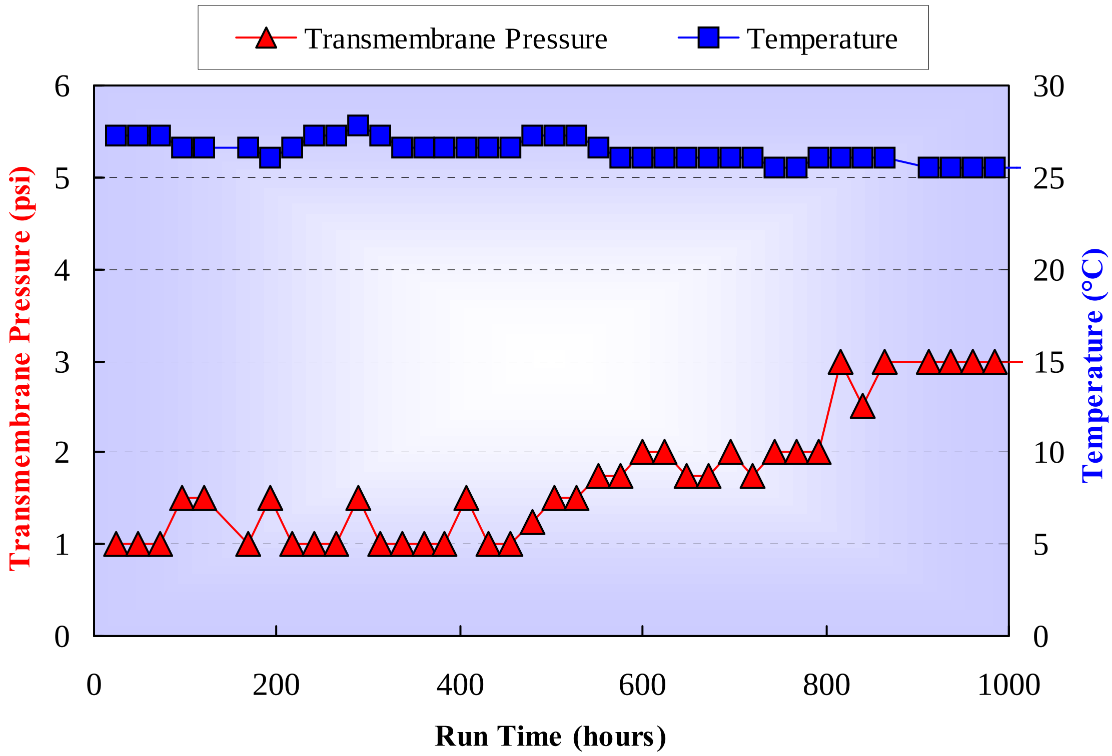


**Figure 5-3. CMF-S transmembrane pressure and temperature for Phase I using PP membranes.**

**Table 5-1. Summary of RO operation and performance for Phase I using PP membranes**

| Parameter | Range | Average |
|---|---|---|
| Feed water flow rate (gpm) | 19.3–23 | 20.7 |
| Product flow rate (gpm) | 10–11.5 | 10.6 |
| Feed water recovery (%) | 48–55 | 51 |
| Flux at 25 °C (gfd) | 14.5–17 | 15 |
| Specific flux at 25 °C (gfd/psi) | 0.07–0.09 | 0.08 |
| Feed water temperature (ºC) | 25–27 | 26.5 |
| Feed water pressure (psi) | 215–245 | 230 |
| Feed TDS (mg/L) | 660–795 | 720 |
| Product TDS (mg/L) | 6–10 | 7 |
| Salt rejection (%) | 98.5–99 | 99 |
| Feed water SDI | 0.1–2 | 0.9 |

RO feed water (MF permeate) was dosed with 4 to 5 mg/L sodium bisulfite and 1 to 2 mg/L antiscalant (ARGO-150). Before entering the first stage of the RO, a 5-µm cartridge prefilter was used to remove fine particulate matter that might enter the system as a result of a failure in the pretreatment system.

Variations in performance due to temperature fluctuations were negligible for the testing period; the RO influent water temperature averaged 26 °C. As shown in figure 5-4, the RO operated at a temperature corrected flux of 15 gfd. Fed at a flow rate of 20 gpm, the RO operated at a FWR of 50% (figure 5-5) producing 10 gpm of permeate while the remaining 10 gpm was disposed of as concentrate. The calculated specific flux based on the net driving pressure (NDP) was 0.07–0.08 gallons per square foot per day per pounds per square inch (gfd/psi). The RO was operated with a net driving pressure of approximately 200 psi (figure 5-6). Despite some slight variation in the operating pressure, the NDP was maintained throughout the testing period without significant increase or loss.

Up to 99% TDS rejection (1% salt passage) was achieved under these operating conditions, as shown in figure 5-7. The feed water TDS averaged 720 mg/L, and the permeate TDS averaged 7 mg/L (figure 5-8). Although the overall RO permeate quality was not affected, it is interesting to note that the influent TDS increased and decreased on a weekly basis. This may have been due to regular operation and/or the demand experienced by the SJ/SC WPCP. Table 5-2 summarizes the amount of salt rejected for specific ions by the RO.

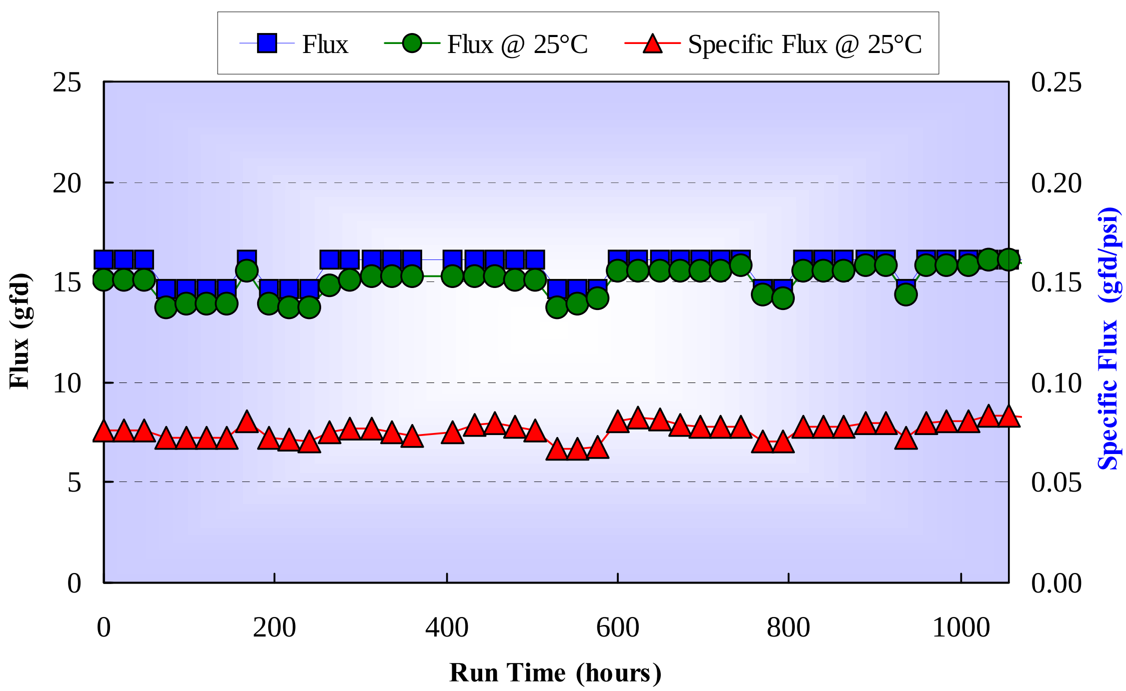


**Figure 5-4. RO temperature-corrected operating flux and specific flux for Phase I using PP membranes.**

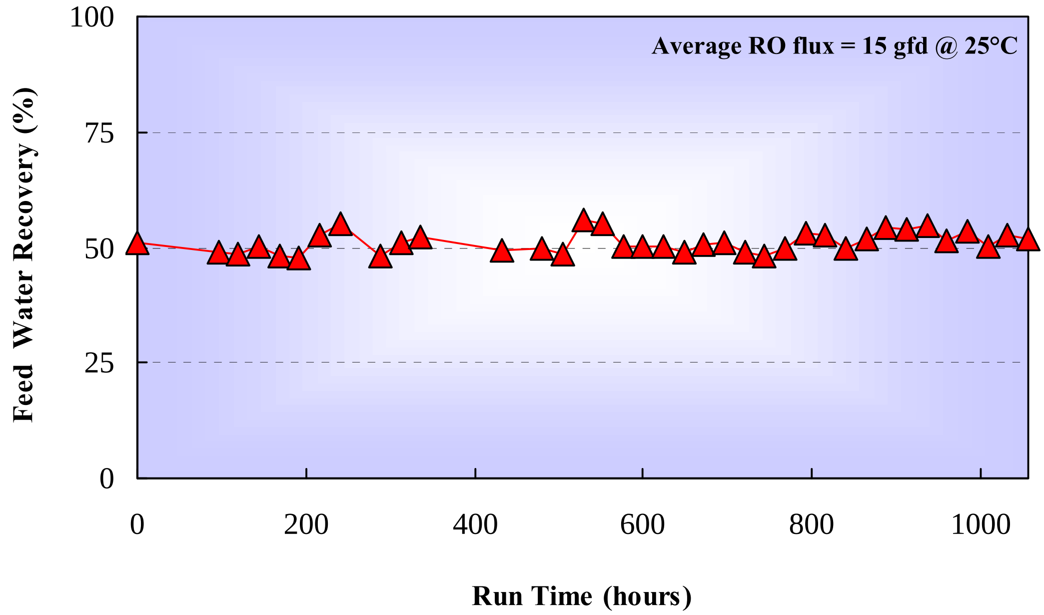


**Figure 5-5. RO feed water recovery for Phase I using PP membranes.**

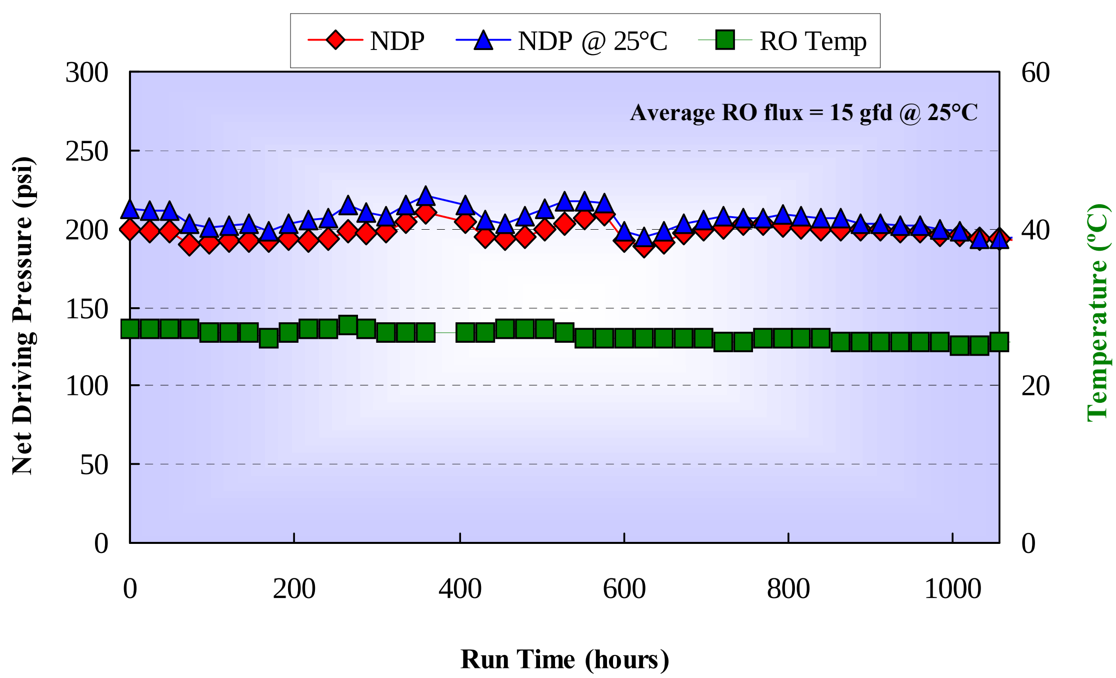


**Figure 5-6. RO net driving pressure for Phase I using PP membranes.**

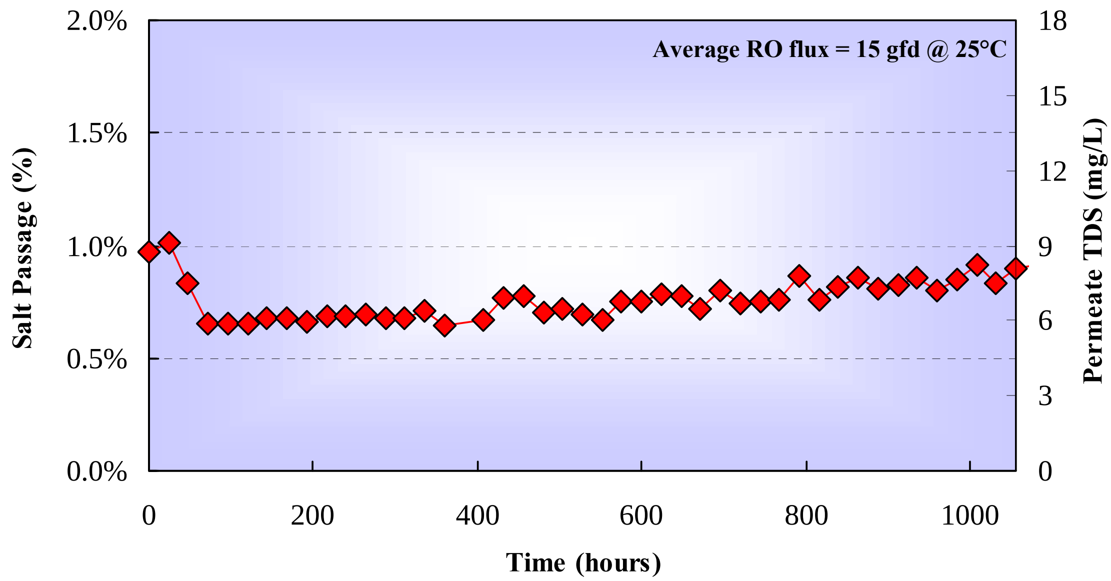


**Figure 5-7. RO salt passage for Phase I using PP membranes.**

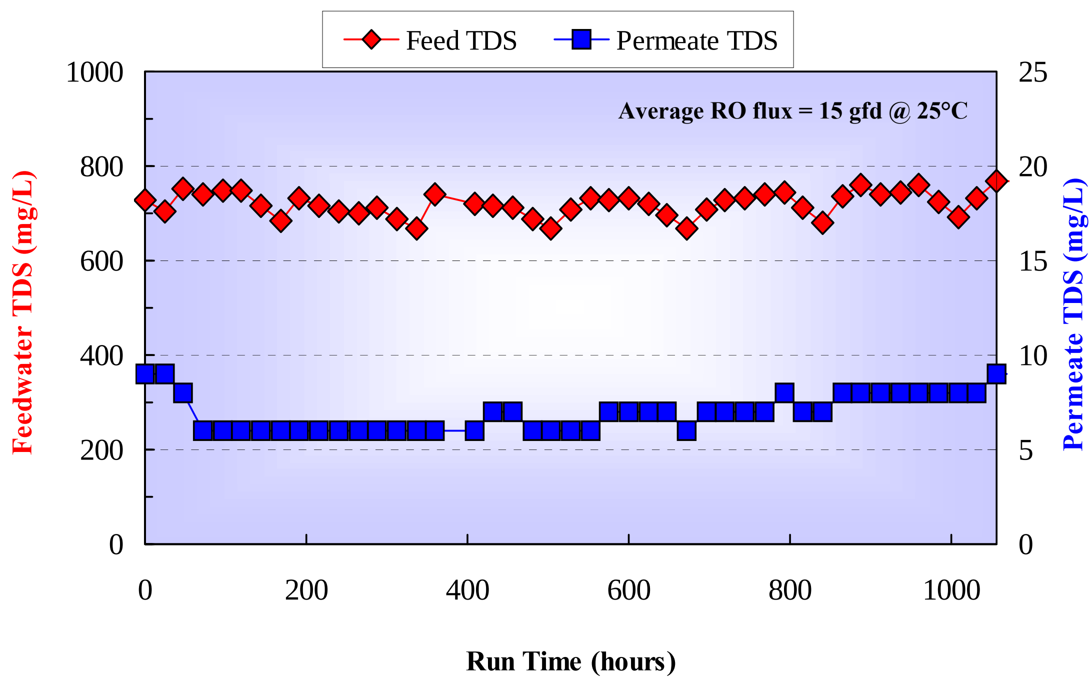


**Figure 5-8. RO TDS levels for Phase I using PP membranes.**

**Table 5-2. Average RO salt rejection**

| Ion | Feed (mg/L) | Product (mg/L) |
|---|---|---|
| Chloride | 200 | 1.4 |
| Nitrate as N | 8 | 0.2 |
| Sulfate | 100 | <4 |
| Calcium | 54 | 0.3 |
| Magnesium | 31 | 0.1 |
| Silica | 25 | 0.4 |
| Sodium | 150 | 3.5 |
| Conductivity (µS)[1] | 1,230 | 12 |
| TDS (mg/L) | 720 | 7 |

[1] µS = microsiemens.

As previously discussed, the ammonia feed pump for the CMF-S feed water failed after 1,100 hours of continuous operation. As a result, free chlorine in the pilot plant feed water came into contact with the chlorine-sensitive PP membranes and caused the CMF-S hollow fibers to break. Concurrently, the RO performance also decreased; and it was quickly discovered that the O-ring seals used in the RO vessels had become worn causing poor water quality and damaged RO membrane elements.

This event concluded Phase I testing and encouraged the project team to explore a more robust option to protect RO membranes from harmful damage. After discussion with USFilter, it was recommended that chlorine-tolerant PVDF membranes be used for this application to ensure the performance of the RO process. Additionally, replacement RO membrane elements were provided and used for additional testing.

### 5.1.2 Phase II – MF Pretreatment Using PVDF Membranes

The objective of Phase II testing was to operate the RO membranes using MF pretreatment with chlorine-tolerant PVDF membranes. As in Phase I, the CMF-S feed and permeate SDI were monitored weekly to characterize the fouling potential of the RO feed water. As shown in figure 5-9, the modified SDI in the tertiary treated wastewater averaged 13.0. After membrane pretreatment, the SDI averaged 0.3.

The CMF-S operated at a flow rate of 45 gpm for 500 hours with an applied flux of 30 gfd (specific flux = 8 to 10 gfd/psi) (figure 5-10). While the flux of the PVDF membranes was significantly higher than the PP membranes, the specific flux was approximately the same because the PVDF membranes were operated at a higher pressure than the PP membranes. As shown in figure 5-11, the transmembrane pressure was stable throughout all conditions and no significant

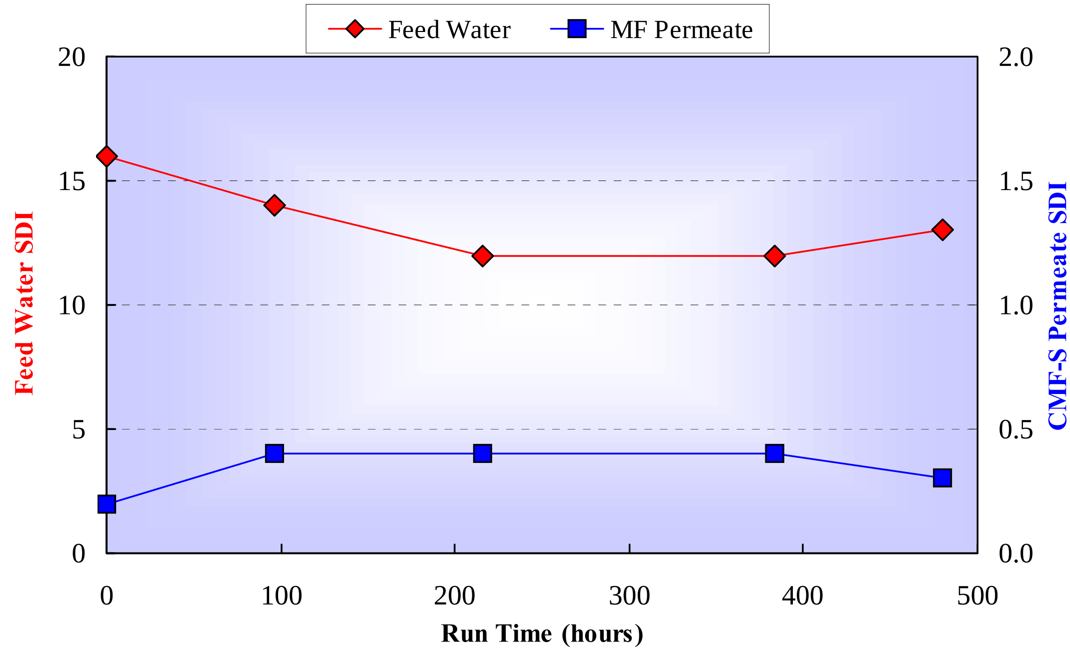


**Figure 5-9. SDI measurements for Phase II using PVDF membranes.**

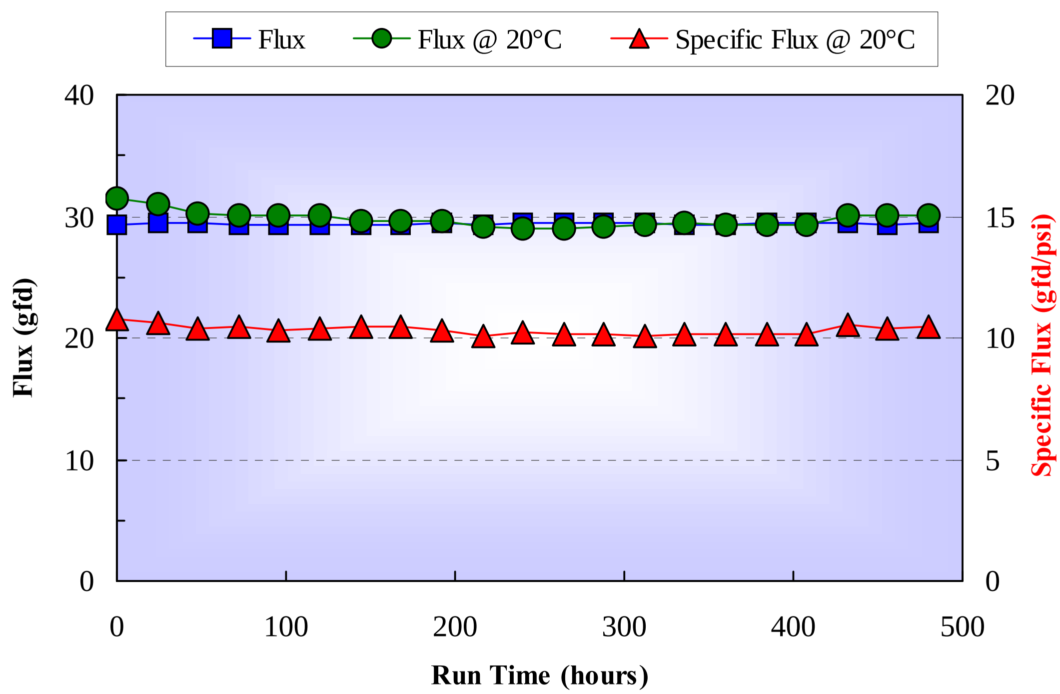


**Figure 5-10. CMF-S temperature-corrected operating flux and specific flux for Phase II using PVDF membranes.**

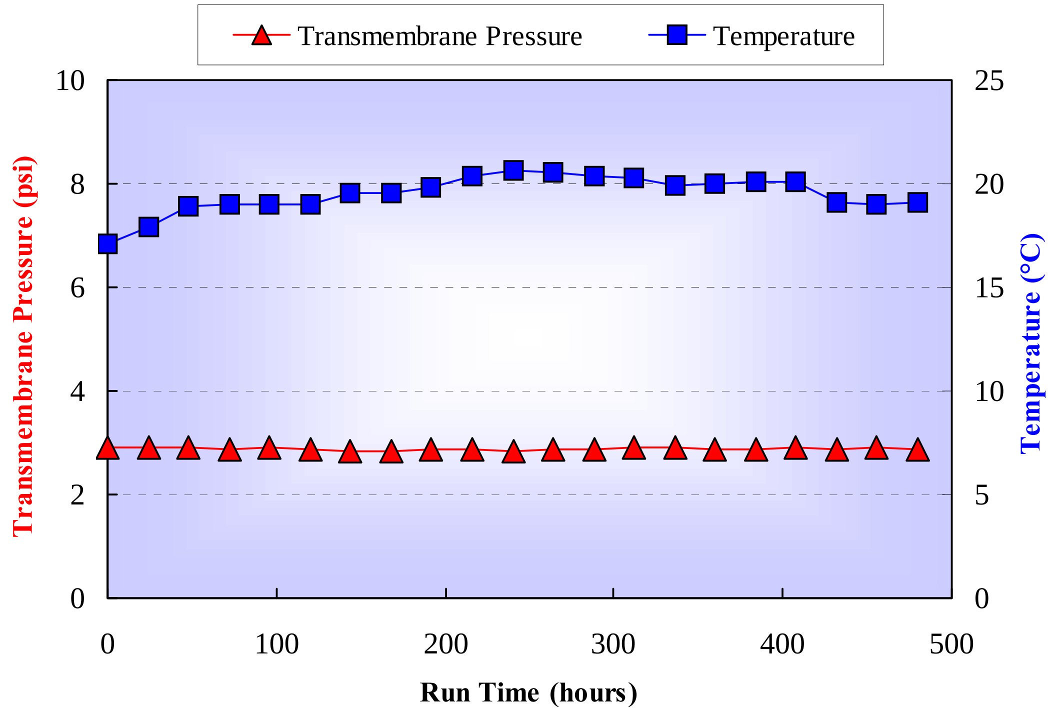


**Figure 5-11. CMF-S TMP and temperature for Phase II using PVDF membranes.**

membrane fouling was observed. It is important to note that no fatal flaws in the membrane pretreatment scheme were observed when operated with PVDF membranes on reclaimed wastewater from the SJ/SC WPCP. Overall, the PVDF membrane performance was slightly better than the PP membrane performance. An additional benefit was that the membranes were protected from free chlorine and any temporary spikes that may occur.

A summary of the RO operation, performance data, and salt rejection of specific ions is presented in table 5-3. The RO system was operated continuously for 500 hours at an applied flux of 15 gfd at 25 °C and an average FWR of 65% (figures 5-12 and 5-13). CMF-S pretreated reclaimed wastewater was dosed with 4–5 mg/L sodium bisulfite and 1–2 mg/L antiscalant (ARGO-150) and fed to the RO at 14 gpm. At a FWR of 65%, 9 gpm of RO permeate was produced.

The NDP (corrected to 25 ºC) of less than 90 psi (figure 5-14) was measured during operation with new membranes, compared to 200 psi previously observed during Phase I. The lower NDP required during Phase II most likely was because new RO membranes were used to replace the fouled RO membranes. Discussions with USFilter revealed that the original RO elements used during Phase I testing were refurbished membranes that previously had been used.

**Table 5-3. Summary of RO operation and performance for Phase II using PVDF membranes**

| Parameter | Range | Average |
|---|---|---|
| Feed water flow rate (gpm) | 13–15 | 14 |
| Product flow rate (gpm) | 8–9.5 | 9 |
| Feed water recovery (%) | 60–69 | 65 |
| Operation flux (gfd) | 14.5–15 | 15 |
| Specific flux (gfd/psi) | 0.14–0.17 | 0.15 |
| Feed water temperature (ºC) | 21–22 | 22 |
| Feed water pressure (psi) | 100–120 | 102 |
| Feed TDS (mg/L) | 640–760 | 720 |
| Product TDS (mg/L) | 16–30 | 20 |
| Salt rejection (%) | 96–98 | 98 |
| Feed water SDI | 0.2–0.4 | 0.3 |

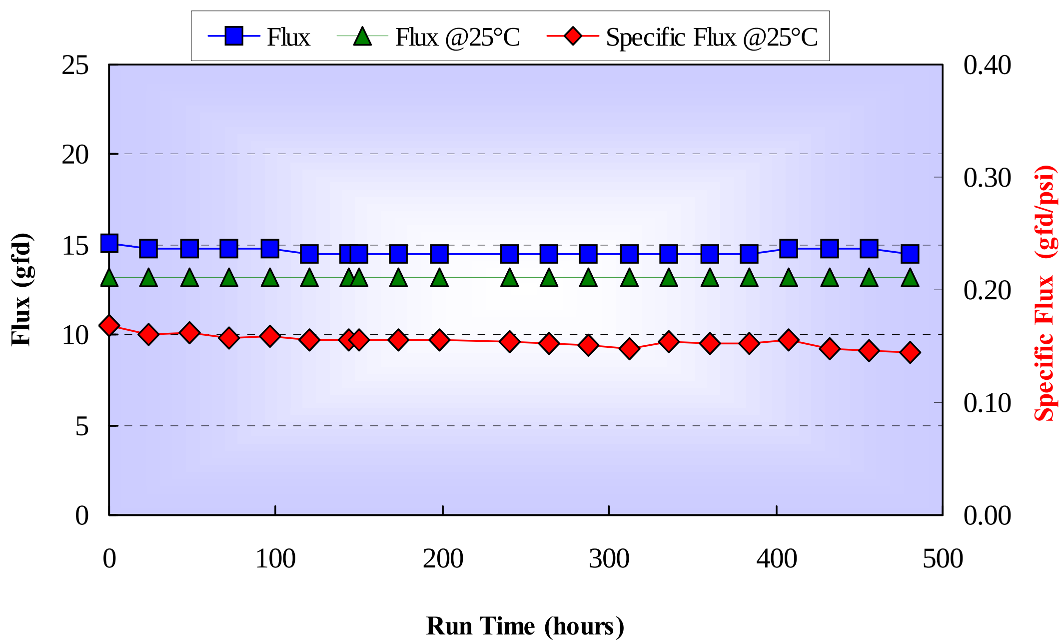


**Figure 5-12. RO temperature corrected operating flux and specific flux for Phase II using PVDF membranes.**

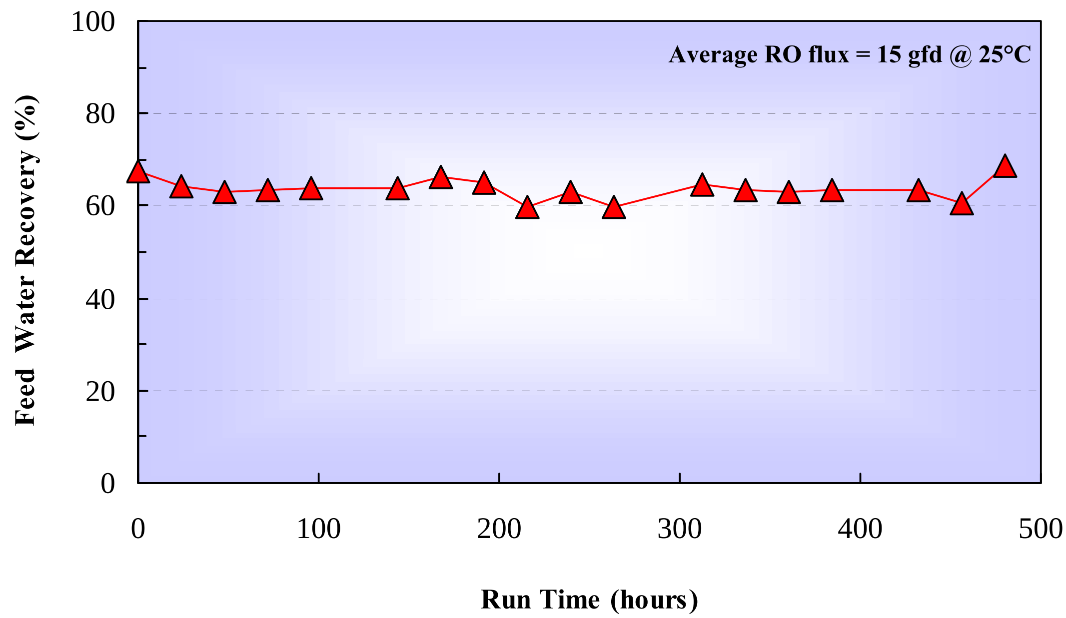


**Figure 5-13. RO feed water recovery for Phase II using PVDF membranes.**

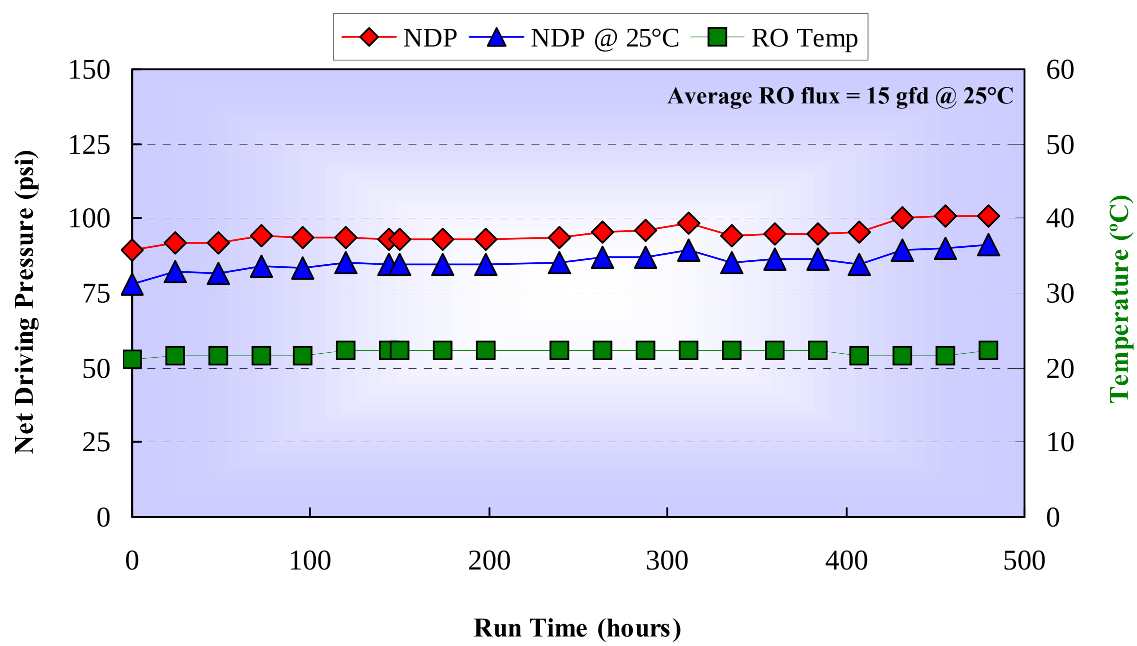


**Figure 5-14. RO net driving pressure for Phase II using PVDF membranes.**

A slight increase in the NDP was observed throughout the test period from an initial 80 to 90 psi. The pressure increase may have been due to slow fouling over time. Additional long-term pilot testing would be required at this flux to determine the point at which the potential fouling of the membranes would result in diminished product water quality.

A TDS rejection of 98% (2% salt passage) was achieved under these operating conditions, as shown in figure 5-15. The feed water TDS averaged 720 mg/L, and the permeate TDS averaged 20 mg/L (figure 5-16). Table 5-4 summarizes the amount of salt rejected for specific ions by the RO.

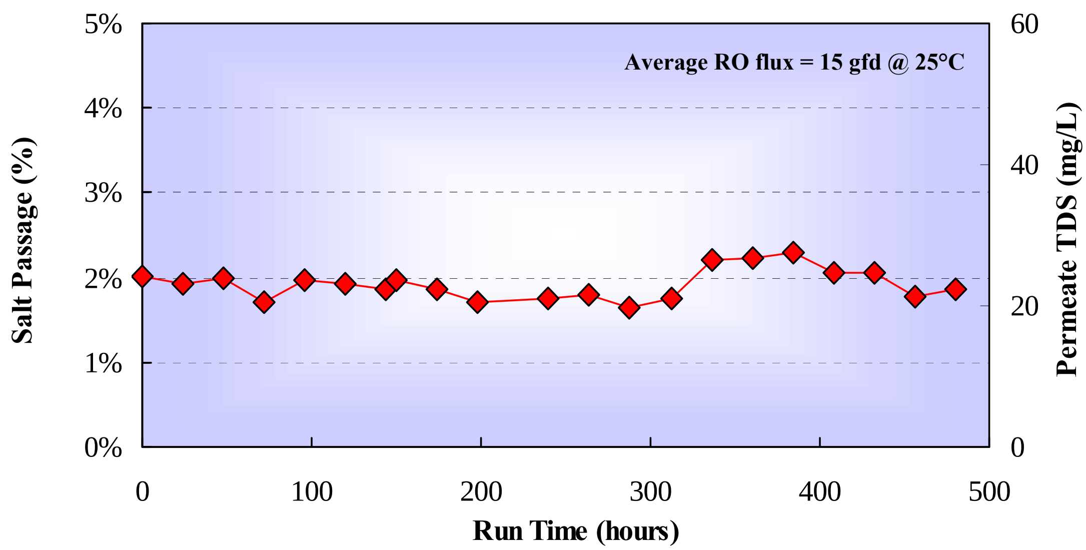


**Figure 5-15. RO salt passage and permeate TDS for Phase II using PVDF membranes.**

**Table 5-4. RO salt rejections – Phase II**

| Ion | Feed (mg/L) | Product (mg/L) |
|---|---|---|
| Chloride | 191 | 13 |
| Nitrate as N | 9.2 | 1.2 |
| Sulfate | 119 | <4 |
| Calcium | 52 | 0.6 |
| Magnesium | 31 | 0.4 |
| Silica | 25 | 3.5 |
| Sodium | 155 | 14.5 |
| Conductivity (µS) | 1,250 | 33 |
| TDS (mg/L) | 720 | 20 |

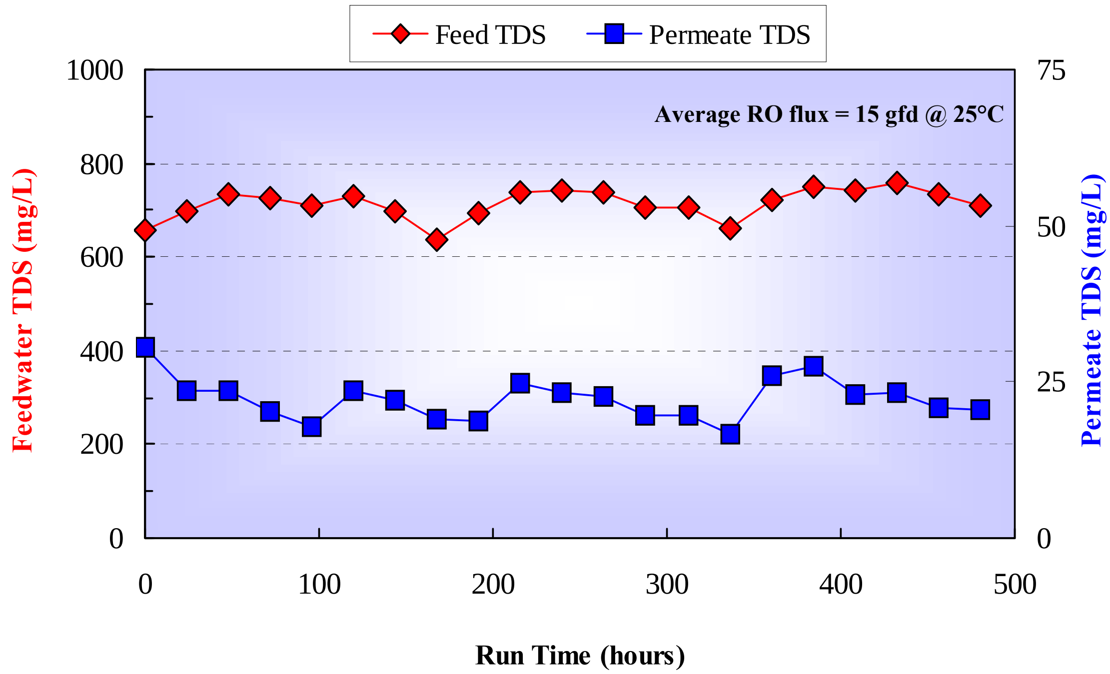


**Figure 5-16. RO total dissolved solids levels for Phase II using PVDF membranes.**

## 5.2 Electrodialysis Reversal

The first phase of testing was to demonstrate the ability of the EDR pilot to reduce the influent TDS to the 350-mg/L target treatment goal (baseline performance). Baseline performance was established using the manufacturer's recommended pretreatment scheme; this included GAC/MMF. Once baseline performance had been established, the EDR was adjusted to produce water with 500-mg/L TDS to establish performance and identify any potential cost savings due to anticipated lower energy consumption. Additionally, during this second phase of operation, the three different types of pretreatment options (GAC/MMF, MF, cartridge filtration only) were evaluated.

### 5.2.1 Phase I – Baseline Operation (350-mg/L TDS Treatment Goal)

In Phase I, the EDR pilot operated continuously for 1,050 hours at a feed flow rate of 28 gpm (figure 5-17). Table 5-5 summarizes the EDR operation and performance in this test period. Approximately 28 gpm of feed water was provided to produce 24 gpm of demineralized water, resulting in an 85% feed water recovery. A consistent effluent water quality was maintained throughout the test period (figure 5-18). Table 5-6 summarizes the amount of salt rejected for specific ions by the EDR.

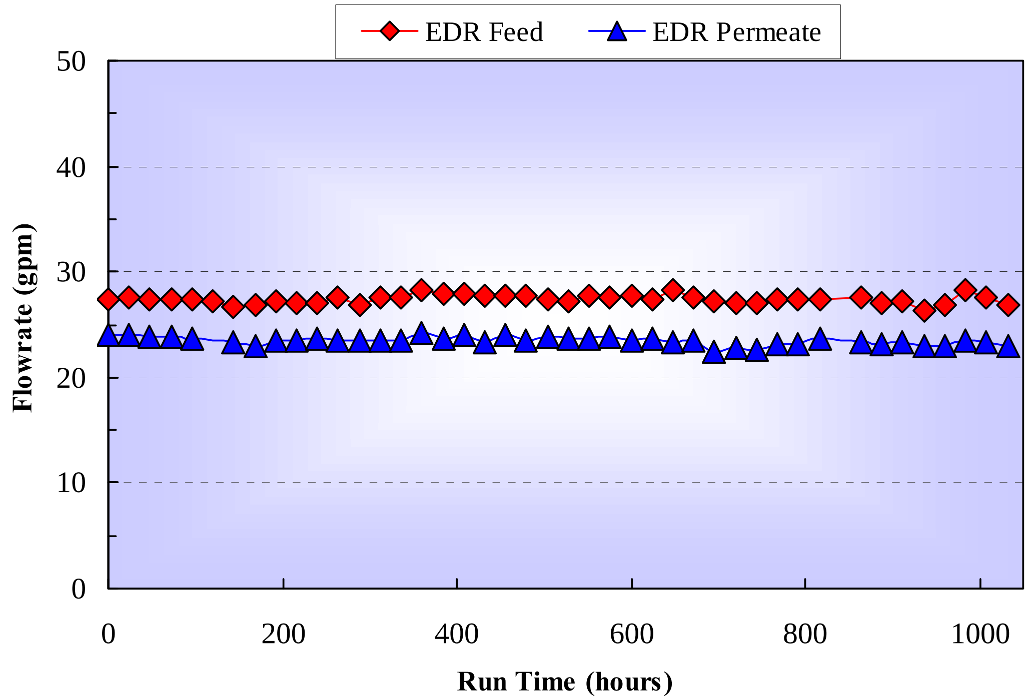


**Figure 5-17. EDR pilot flowrates – Phase I.**

**Table 5-5. Summary of baseline EDR operation and performance**

| Parameter | Range | Average |
|---|---|---|
| Feed water flow rate (gpm) | 26–28 | 27.5 |
| Product flow rate (gpm) | 22.5–24 | 23.5 |
| Feed water recovery (%) | 83–88 | 85.5 |
| Feed water temperature (ºC) | 25–27 | 26.5 |
| Feed TDS (mg/L) | 660–750 | 720 |
| Product TDS (mg/L) | 330–390 | 360 |
| Stage 1 voltage (V) | 53–54 | 53.3 |
| Stage 1 current (amps) | 3.3–4.3 | 3.8 |
| Stage 2 voltage (V) | 49–51 | 50 |
| Stage 2 current (amps) | 3–3.7 | 3.3 |
| Stack inlet pressure (psi) | 24–34 | 28 |
| Stack inlet DP (inches $H_2O$) | 74–100 | 87.5 |
| Stack outlet pressure (psi) | 5–8 | 6.5 |
| Stack outlet DP (inches $H_2O$) | 28–52 | 40 |
| Salt rejection (%) | 53–57 | 53 |

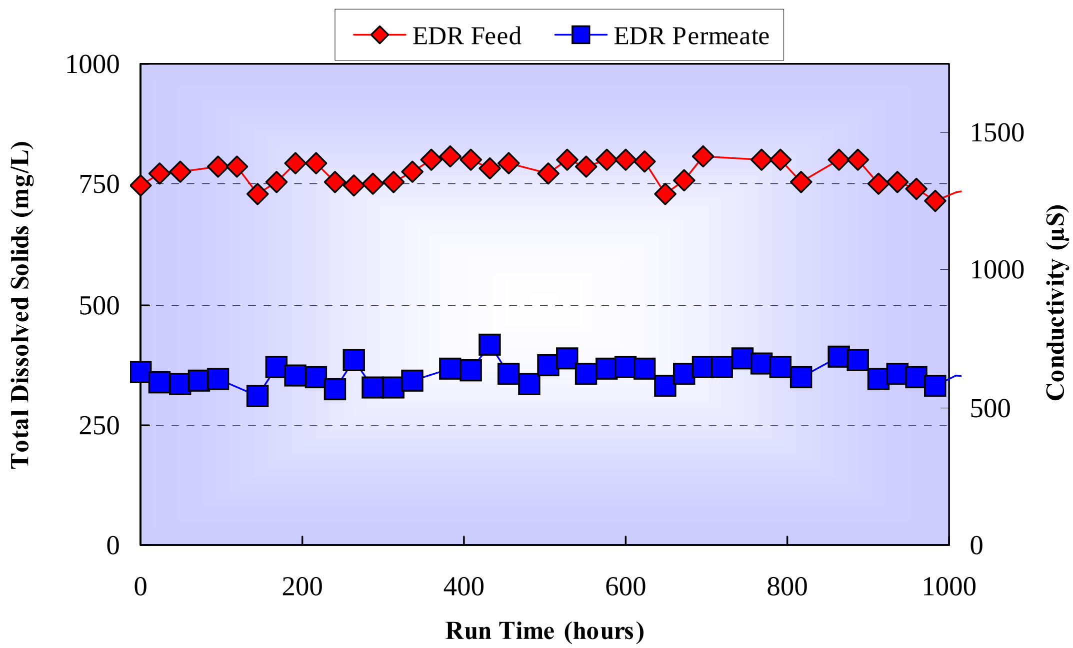


**Figure 5-18. EDR pilot TDS levels – Phase I.**

**Table 5-6. Summary of baseline EDR water quality**

| Ion | Feed (mg/L) | Product (mg/L) |
|---|---|---|
| Chloride | 200 | 88 |
| Nitrate as N | 8 | 3.9 |
| Sulfate | 100 | 36 |
| Calcium | 54 | 12 |
| Magnesium | 31 | 8 |
| Silica | 25 | 23 |
| Sodium | 150 | 106 |
| Conductivity (µS) | 1230 | 650 |
| TDS (mg/L) | 750 | 350 |

The pressure drop through the membrane stack is dependent upon the spacer type, flow per stage, and number of pairs in each stage and also may increase as a result of membrane fouling. As specified by the equipment vendor, the Aquamite V pilot should operate such that the stack inlet pressure does not exceed 50 psi. As shown in figure 5-19, the stack inlet pressure was maintained below the 50-psi limit. However, during the testing period, the inlet pressure increased from 25 to 35 psi. This may be indicative of fouling of the membrane stack and could be related to the slight increase in both product TDS and electrical resistance that was observed after 400 hours of run time. The stack outlet pressure was measured to be approximately 6 psi throughout the testing period.

In order to achieve TDS removal to 350 mg/L, approximately 4 amps/53 volts and 2.8 amps/48 volts were applied to the first and second electrical stages, respectively. The voltage and amperage were monitored throughout the testing period and ensured that a constant electric potential was supplied to the EDR. Membrane fouling may cause a decrease in the applied current (at a constant voltage) and, as a result, will cause the electrical resistance to increase. As seen in figure 5-20, a slight increase in the electrical resistance was observed after 400 hours, which may have indicated that slight membrane fouling was occurring.

Despite the slight increase in both stack pressure and resistance, the EDR was able to consistently meet the water quality goal during the pilot test program. The observed fouling rate was minimal and is characteristic of normal operation.

### 5.2.2 Phase II – Pretreatment Alternatives (500-mg/L TDS Treatment Goal)

In Phase II, the EDR pilot operated continuously for nearly 1,300 hours, during which the three pretreatment configurations were evaluated. As shown in figure 5-21, the EDR feed flow rate varied between 20 and 27, depending on which pretreatment option was being evaluated at the time. Table 5-7 summarizes the operation and performance of the EDR in this test period. A lower flow rate, 20 gpm, was fed to the EDR during the MF pretreatment condition, because the MF permeate was divided between the EDR and RO. Table 5-8 summarizes specific water qualities and the amount of salt rejected for specific ions by the EDR.

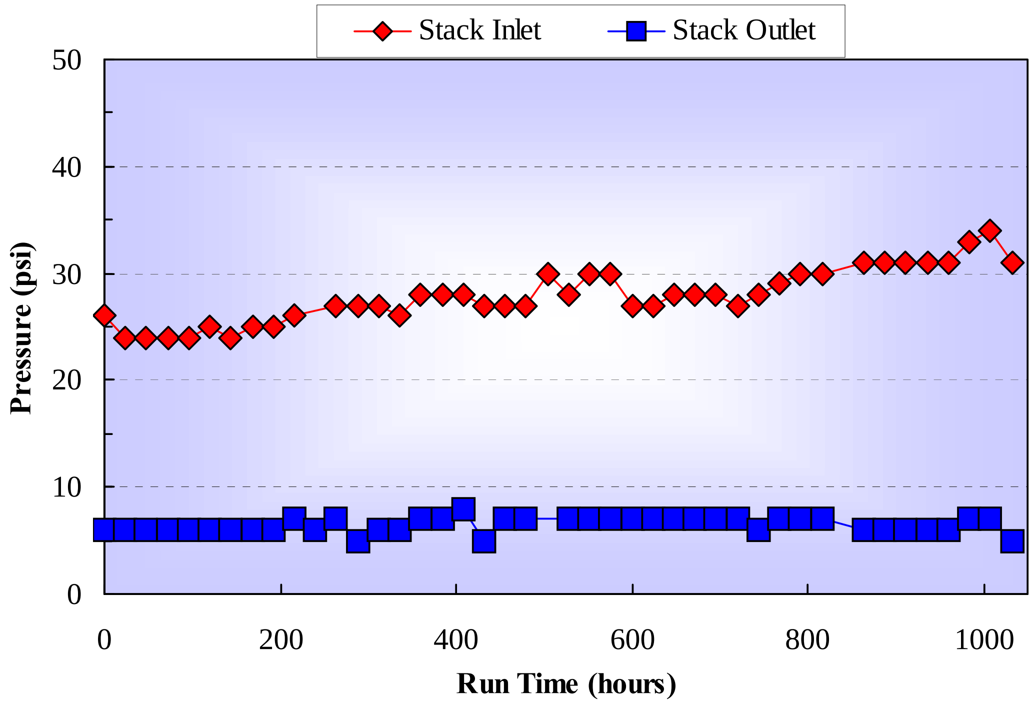


**Figure 5-19. EDR pilot stack pressures – Phase I.**

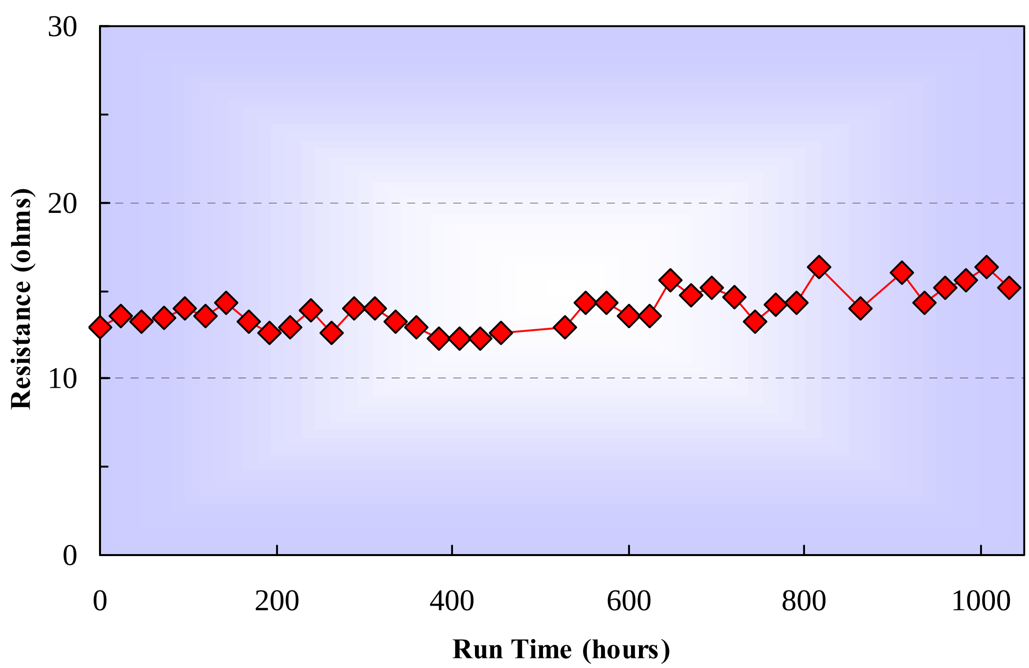


**Figure 5-20. EDR pilot electrical resistance – Phase I.**

**Table 5-7. Summary of EDR operation and performance – Phase II**

| Parameter | GAC/MMF | | MF (PVDF) | | Cartridge Filtration | |
|---|---|---|---|---|---|---|
| | Range | Avg | Range | Avg | Range | Avg |
| Feed water flow rate (gpm) | 25–27.5 | 27 | 20–21 | 20 | 25–29 | 25 |
| Product flow rate (gpm) | 21–23 | 22 | 16–17 | 16.5 | 22–24 | 22 |
| Feed water recovery (%) | 78–92 | 82 | 75–86 | 81 | 86–89 | 88 |
| Feed water temperature (ºC) | 20–20 | 20 | 19.5–20 | 20 | 21–22 | 22 |
| Feed TDS (mg/L) | 675–790 | 750 | 707–780 | 740 | 670–770 | 760 |
| Product TDS (mg/L) | 500–590 | 550 | 475–550 | 550 | 450–550 | 480 |
| Stage 1 voltage (V) | 57–67 | 61 | 66–68 | 67 | 56–67 | 57 |
| Stage 1 current (amps) | 3.3–4.1 | 3.8 | 3–4 | 3.5 | 2.6–3 | 2.8 |
| Stack inlet pressure (psi) | 32–38 | 34 | 24–28 | 24 | 43–50 | 50 |
| Stack inlet DP (inches $H_2O$) | 78–98 | 82 | 60–84 | 61 | 10–26 | 12 |
| Stack outlet pressure (psi) | 5–7 | 5.4 | 5–7 | 6 | 7–9 | 7 |
| Stack outlet DP (inches $H_2O$) | 32–80 | 60 | 50–60 | 60 | 10–17 | 10 |
| Salt rejection (%) | 24–30 | 27 | 26–34 | 26 | 23–36 | 36 |

**Table 5-8. Summary of EDR water quality – Phase II**

| Parameter | GAC/MMF | | MF (PVDF) | | Cartridge Filtration | |
|---|---|---|---|---|---|---|
| | Feed (mg/L) | Product (mg/L) | Feed (mg/L) | Product (mg/L) | Feed (mg/L) | Product (mg/L) |
| Chloride | 188 | 130 | 195 | 130 | 195 | 121 |
| Nitrate as N | 8.6 | 6 | 9 | 5.7 | 8.4 | 5.4 |
| Sulfate | 98 | 57 | 100 | 55 | 100 | 63 |
| Calcium | 52 | 30 | 54 | 28 | 57 | 28 |
| Magnesium | 32 | 19 | 32 | 17 | 33 | 17 |
| Silica | 25 | 25 | 25 | 25 | 25 | 25 |
| Sodium | 140 | 120 | 150 | 120 | 158 | 123 |
| Turbidity | 1.0 | 0.8 | 0.1 | 0.1 | 1.0 | 0.6 |
| TOC | 8 | 7 | 7 | 6 | 9 | 8 |
| Conductivity (µS) | 1250 | 900 | 1200 | 830 | 1250 | 830 |
| TDS (mg/L) | 730 | 520 | 700 | 480 | 730 | 500 |

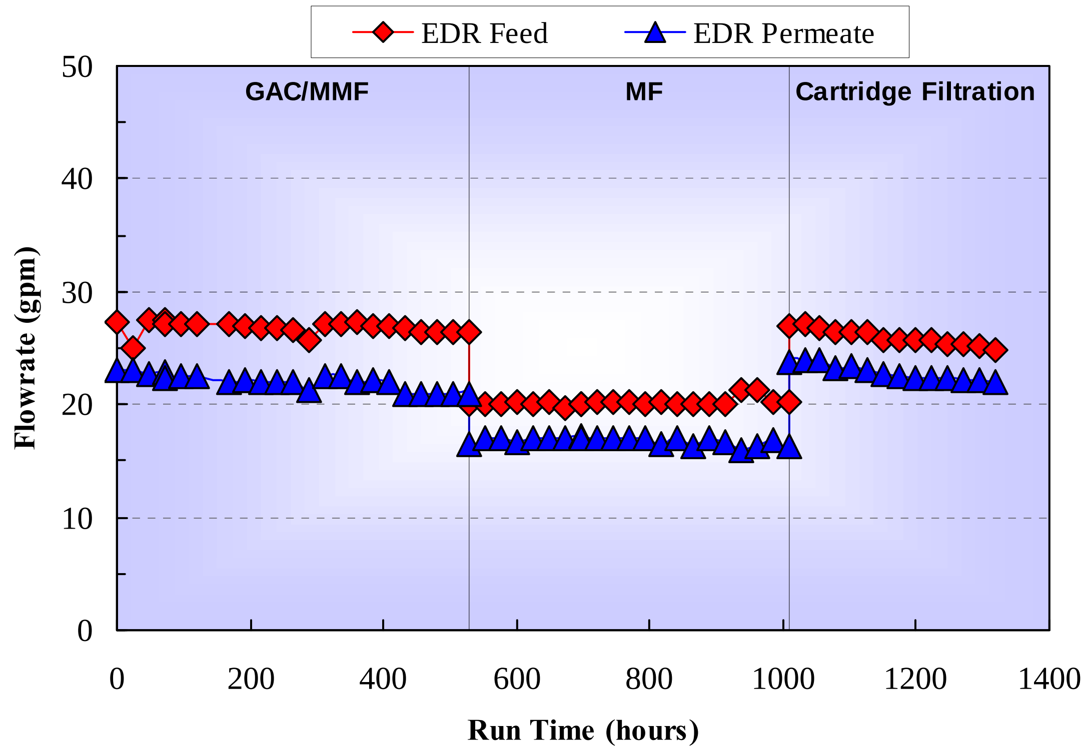


**Figure 5-21. EDR pilot flowrates – Phase II.**

During all three pretreatment configurations, the EDR operated consistently at a feed water recovery of approximately 85%. The pilot unit utilized a single stack with one electrical stage to reduce the influent TDS to 500 mg/L, as shown in figure 5-22. This consistent effluent water quality was maintained for the entire Phase II test period, and no operation failures occurred.

The pressure drop through the membrane stack was monitored to indicate if any membrane fouling was occurring. As shown in figure 5-23, the stack outlet pressure was maintained below 10 psi for all conditions. During GAC/MMF pretreatment, the initial stack pressure represented a continuation of operation from Phase I, and a similar fouling rate was observed as the stack inlet pressure increased from 32 to 38 psi. The EDR operated for approximately 500 hours using GAC/MMF as a pretreatment step.

Next, MF pretreatment was used, and the EDR was operated for an additional 500 hours during which the stack inlet pressure was observed to be lower, at approximately 25 psi. The lower stack inlet pressure was expected to be due to the lower feed flow rate applied to the EDR during this period of testing (20 gpm versus 27 gpm, previously). Additionally, the stable inlet pressure observed during operation with MF pretreated wastewater most likely was due to the superior water quality of the EDR feed, with respect to suspended solids. As

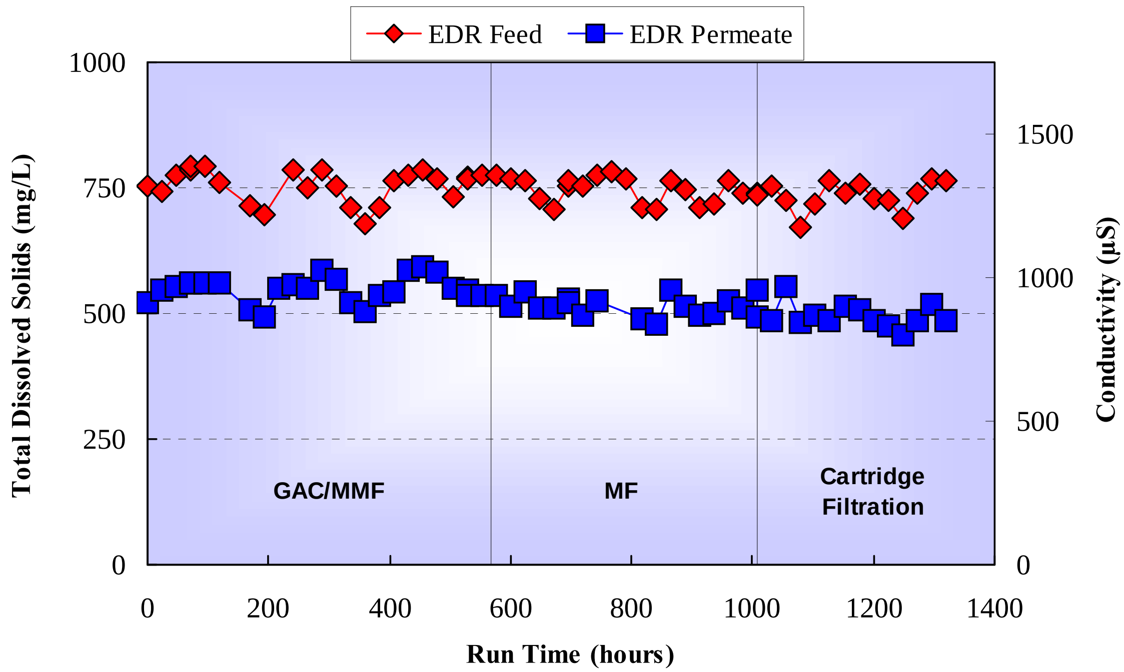


**Figure 5-22. EDR pilot TDS levels – Phase II.**

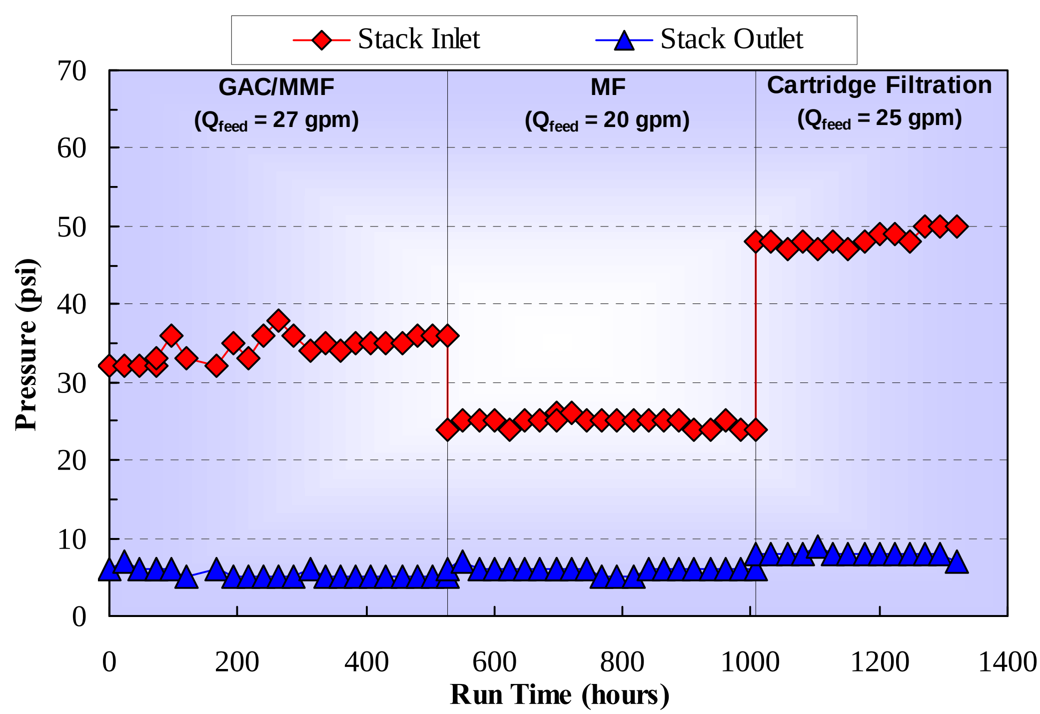


**Figure 5-23. EDR pilot stack pressures – Phase II.**

shown in table 5-8, the feed turbidity of the GAC/MMF pretreated water was about 1 NTU, and the MF pretreated water was consistently measured below 0.1 NTU.

After the MF pretreatment test phase was complete, the EDR operated directly with reclaimed wastewater from the SJ/SC WPCP with only cartridge filtration. The purpose of this phase was to operate the EDR with a minimal amount of pretreatment (cartridge filtration only) to determine if the reclaimed water quality was sufficiently clean to avoid harmful fouling of the EDR membranes. Unfortunately, only 300 hours of continuous operation were achieved. An unexpected power surge in the electrical supply seriously damaged the system and prematurely terminated this evaluation. Based on the data collected, however, the stack inlet pressure began at 43 psi (most likely due to the higher flow rates used similar to the GAC/MMF operation) and approached the 50-psi limit by the end of operation. Typically, 50 psi is the recommended upper limit of operation, as recommended by the equipment manufacturer.

In order to achieve TDS removal to 500 mg/L, the applied voltage was set to achieve the desired water quality goal. For the different flow rates, different voltages and resulting currents were required during each pretreatment condition to achieve the desired 500-mg/L effluent TDS level (table 5-7). This variability makes it difficult to directly compare the three based on these parameters. The electrical resistance, however, provides a reasonable basis for comparing the three pretreatment conditions, in addition to providing insight about potential EDR membrane fouling.

Figure 5-24 shows that, for the entire test period (all pretreatment conditions), a constant increase in the electrical resistance occurred. Similar to the increase in the inlet stack pressures, a slight fouling trend was observed; however, neither pretreatment condition can be attributed to an accelerated fouling rate over the other. It has been assumed that the observed fouling rate was minimal and was characteristic of normal operation. During extended operation, recovery of performance would be achieved through regular maintenance and CIP cleans.

These important findings indicated that, when fed the high quality effluent from the SJ/SC WPCP, the EDR can be operated using a minimal amount of pretreatment without excessive fouling, operational failure, or a decrease in product water quality. Additionally, removing extraneous pretreatment equipment could provide a significant cost saving in operation without sacrificing the performance of the EDR system.

It is important to note that further study is required to fully understand the impact of irregular and/or intermittent full-scale plant changes and their effect on the reclaimed water quality. Temporary changes made during full-scale operation of the SJ/SC WPCP to deal with unexpected events (i.e., high flow/demand, plant upsets, increased free chlorine, extra or alternative coagulant addition, etc.) potentially could create a harmful condition for the EDR membranes.

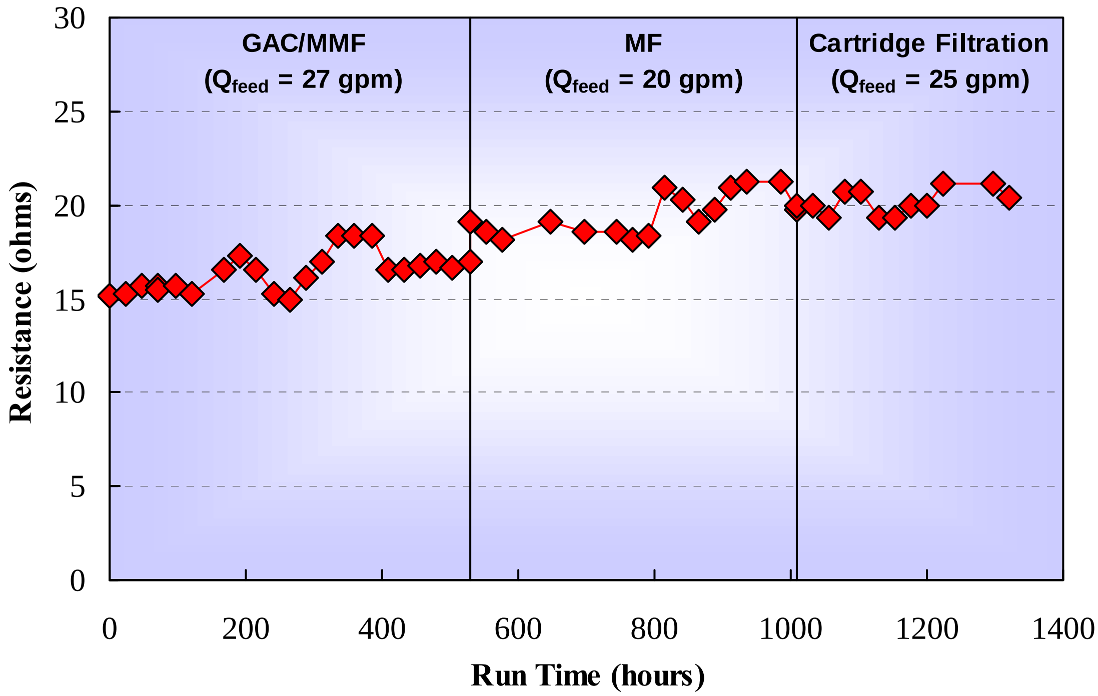


**Figure 5-24. EDR pilot electrical resistance – Phase II.**

# 6. Process Comparison

Both MF/RO and EDR were found to be capable of reducing effluent salinity of the tertiary treated wastewater produced by the SJ/SC WPCP. This section compares the performance of the processes with respect to the following:

- **Operational issues** including maintenance requirements and process failures experienced during pilot testing

- **Water quality** produced from each demineralization process

- **Projected costs** for construction and operation of full-scale facilities required to produce 50 mgd of recycled water with a final TDS concentration of either 350 or 500 mg/L

## 6.1 Operation

The following operational issues were experienced during startup and operation of the pilot equipment. A detailed memo of operational experiences noted by plant operators of the SJ/SC WPCP is included in Appendix B.

### 6.1.1 USFilter MF/RO Operation

#### *6.1.1.1 CMF-S Membrane Damage*

During the initial phase of testing, PP hollow-fiber membrane modules were chosen for the RO feed pretreatment. These PP membranes are known to be sensitive to free chlorine concentrations and require dechlorination strategies to prevent damage. An ammonia feed system was installed to generate chloramines in the tertiary treated wastewater feed water and protect the PP membranes. Midway through the MF/RO pilot testing, the ammonia feed system failed causing the PP membrane fibers to break and become damaged. As a result, the CMF-S was not able effectively to reduce the SDI below 5 which caused RO membrane fouling and reduced the salt rejection efficiency. Although the first phase of MF/RO pilot testing was terminated prematurely, sufficient data was collected, as presented in this report to characterize MF/RO operation under baseline conditions. Upon further consideration, the project team and USFilter decided 1) to replace the PP membranes with PVDF membranes which are more tolerant of free chlorine and (2) to clean and replace the RO membrane elements as necessary and test their integrity to ensure continued performance.

### 6.1.2 Ionics EDR Operation

#### *6.1.2.1 Algal Fouling*

As noted in section 4, initially sodium bisulfite was used to remove chlorine from the EDR system. Chemical addition subsequently was replaced by GAC filters, and two GAC contactors were connected in series to remove chlorine from EDR feed water. The GAC contactors proved to be effective at removing free chlorine; in fact, there was evidence that complete dechlorination occurred in the first GAC contactor allowing algae to grow in the second GAC contactor, as indicated by inspection through the clear viewing ports. This eventually led to biofouling of the EDR membranes, producing a dramatic increase in the inlet stack pressure. Corrective measures were taken by flushing the EDR stack with the undechlorinated EDR influent to remove algae, restoring the inlet operating pressure to 25 psi. To prevent further biofouling from occurring, a small stream of chlorinated feed water was allowed to bypass the GAC contactors to maintain a chlorine residual to the EDR stack of 0.5–1 mg/L.

## 6.2 Water Quality

Pilot testing demonstrated that both EDR and MF/RO could reduce the TDS of recycled water to 350 mg/L and 10 mg/L, respectively. Based on the pilot EDR configuration, the treatment goal could be achieved without additional blending of the product water with untreated recycled water. The RO product water, however, would need to be blended to achieve the appropriate treatment goal. It is often common for full-scale demineralization plants (both EDR and MF/RO) to blend some source water with the permeate of the plant. An advantage is gained by blending because plant sizes are reduced, resulting in lower capital and operating costs (AWWA 1999).

Table 6-1 compares the water quality of EDR permeate to MF/RO permeate blended with source water. For this example, the MF/RO permeate was normalized to the average TDS of the EDR permeate. This table shows that similar water quality is achieved for both processes. Slightly higher concentrations of sulfate, calcium, and magnesium would be present in a MF/RO blended product quality. However, MF/RO blended water had lower silica and sodium concentrations than the EDR product water.

**Table 6-1. EDR versus estimated blended MF/RO pilot product water quality[1]**

| Parameter | Unit | Feed | RO Blended Product[1] | EDR Product |
|---|---|---|---|---|
| $Cl^-$ | mg/L | 188 | 87.0 | 88 |
| $NO_3$-N | mg/L | 7.1 | 3.3 | 3.4 |
| $SO_4$ | mg/L | 96 | 44.7 | 33 |
| $Br^-$ | mg/L | < 1.0 | <1.0 | <1.0 |
| $NO_2$-N | mg/L | <0.05 | <0.05 | <0.05 |
| | | | | |
| Al | mg/L | 0.06 | 0.1 | <0.005 |
| Ba | mg/L | 0.020 | 0.01 | 0.007 |
| B | mg/L | 0.510 | 0.3 | 0.481 |
| Ca | mg/L | 59.1 | 27.2 | 10.8 |
| Cr (Total) | mg/L | < 0.002 | <0.002 | <0.002 |
| Fe | mg/L | 0.07 | 0.1 | 0.05 |
| Mg | mg/L | 31.7 | 14.6 | 7.11 |
| $SiO_2$ | mg/L | 24.0 | 11.2 | 23.5 |
| Na | mg/L | 156 | 73.3 | 101 |
| Sr | mg/L | 0.387 | 0.2 | 0.078 |
| | | | | |
| $NH_3$-N | mg/L | <0.1 | <0.1 | <0.1 |
| Conductivity | Umhos/cm | 1,250 | 591 | 614 |
| pH | SU | 7.3 | 6.5 | 6.9 |
| TOC | mg/L | 9 | 5.2 | 6 |
| TKN | mgL | 0.4 | <0.3 | <0.3 |
| TSS | mg/L | <2 | <2 | <2 |
| Turbidity | NTU | 0.7 | 0.4 | 0.5 |
| Hardness, total ($CaCO_3$) | mg/L | 250 | 115 | 55 |
| Alkalinity, total ($CaCO_3$) | mg/L | 190 | 90 | 120 |
| TDS | mg/L | 750 | 350 | 350 |
| $UV_{254}$ | 1/cm or $cm^{-1}$ | 0.109 | 0 | 0.068 |

[1] Based on pilot operating RO flux=15 gfd; FWR=65%.

## 6.3 Treatment Costs

A cost analysis was performed to estimate capital and operational costs associated with a full-scale EDR and MF/RO system with a production capacity of 50 mgd. Both Ionics and USFilter were contacted to provide costs associated with their respective systems. Design criteria used to estimate advanced treatment costs were based on information collected during pilot testing and manufacturer recommendations. The cost estimate encompassed the following criteria:

- Treatment capacity of 50 mgd
- Construction and labor costs
- Product water (including blending water) to achieve effluent with 350 and 500 mg/L of TDS
- O&M costs including consumables, power and parts based on pilot performance
- Membrane system capital costs (obtained from Ionics and USFilter).
- Ancillary (pre)treatment costs

Construction-related costs include engineering design, site work, legal, and administrative work involved in the construction of a new brine treatment facility. These values are calculated using a range of construction related costs (as a percentage of the capital cost), based on experience with water treatment plant construction (table 6-2).

**Table 6-2. Range of construction related costs**

| Construction Related Cost | Average Range* (%) |
|---|---|
| Civil site work | 1–10 |
| Instrumentation | 3–15 |
| Electrical site work | 7–12 |
| Piping | 5–12 |
| Construction contingency | 10–35 |
| Contractor overhead and profit, bonds, and insurance | 10–20 |
| Engineering, legal and administrative | 10–30 |

[1] Average ranges based on experience with surface water treatment plant design.

Construction contingencies are applied to the cost estimate to account for items not specifically included in a project scope but found to be necessary. The level of contingency selected should reflect the level of detail provided during predesign. A low contingency budget reflects a high degree of confidence in the predesign, and a high contingency budget reflects a low level of confidence in the predesign. A low degree of confidence may be due to limited availability of detailed costs or an experimental treatment technology. The recommended contingency levels for the varying types of cost estimates are listed in table 6-3. A 20% contingency was applied to the engineering analysis in this study.

**Table 6-3. Recommended contingency for corresponding level of estimate**

| Type of Cost Estimate | Level of Accuracy (%) | Recommended Contingency (%) |
|---|---|---|
| Order-of-magnitude | +50 to -30 | 20 to 30 |
| Conceptual | +40 to -20 | 20 to 15 |
| Preliminary design | +30 to -15 | 15 to 10 |
| Definitive | +15 to -5 | 10 to 5 |

### 6.3.1 Conceptual Level Full-Scale Treatment Costs

A preliminary conceptual cost analysis was performed using capital and O&M costs provided by equipment manufacturers and a MWH cost model of MF/RO treatment facilities (DRIP 2004). These full-scale EDR and MF/RO treatment costs were developed for a 50-mgd system including blending of product water with untreated tertiary effluent to achieve a finished water TDS of both 350 and 500 mg/L.

Based on results of pilot-scale testing, three treatment scenarios were evaluated:

- RO using MF pretreated water
- EDR using MF pretreated water
- EDR without pretreatment (cartridge filtration only)

#### *6.3.1.1 Design Criteria*

As GAC for chlorine reduction is considerably more expensive than chemical (i.e., sodium bisulfite reduction), a separate cost for GAC pretreatment was not considered. The following assumptions (table 6-4) were used in the cost analysis of MF/RO full-scale plants designed to produced 50-mgd of recycled water. Additionally, the blending ratios used to achieve 350- and 500-mg/L TDS recycled water are outlined in figure 6-1.

**Table 6-4. MF/RO design conditions used for conceptual cost analysis**

| Parameter | 350-mg/L TDS Goal | 500-mg/L TDS Goal |
|---|---|---|
| RO Flux | 12 gfd | 12 gfd |
| RO FWR | 85% | 85% |
| MF Flux | 20 gfd | 20 gfd |
| MF FWR | 90% | 90% |
| | | |
| MF feed flow | 43 mgd | 21 mgd |
| RO permeate flow | 35 mgd | 17 mgd |
| Blending flow rate | 15 mgd | 33 mgd |
| | | |
| RO product TDS | 20 mg/L | 20 mg/L |
| Blending water TDS | 750 mg/L | 750 mg/L |

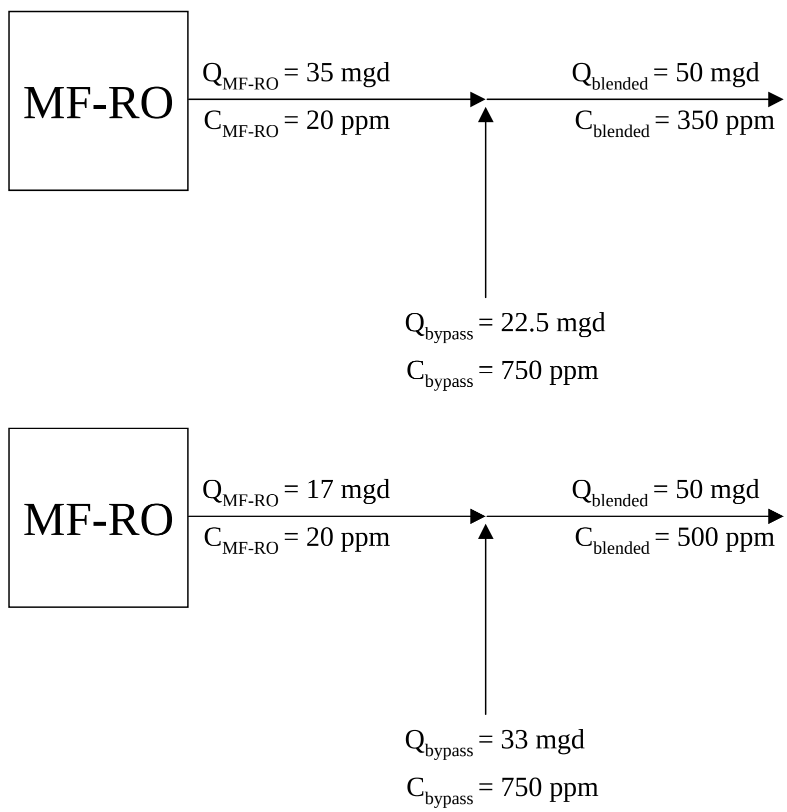


**Figure 6-1. Schematic of MF/RO 50-MGD blending at 350- and 500-mg/L treatment goals.**

The full-scale EDR facility was based on using a three-stage design with a total 85% feed water recovery. Based on pilot-scale testing results, it was determined that the high quality reclaimed water produced by SJ/SC WPCP might be suitable for operation of an EDR system without any pretreatment. It is important to note that additional pilot testing of the EDR without pretreatment is recommended to confirm the long-term performance of the EDR membranes, since operation during this test program was focused on evaluating multiple pretreatment options (with short-term test periods) to determine if any immediate fatal flaws would be encountered. Additionally, EDR testing at the "Cartridge Filtration" condition was prematurely terminated due to a pilot plant power failure. The following assumptions (table 6-5) were used in the cost analysis of MF/RO full-scale plants designed to produce 50 mgd of recycled water. Additionally, the blending ratios used to achieve 350- and 500-mg/L TDS recycled water are outlined in table 6-5.

In order to provide perspective regarding the estimated cost of full-scale EDR treatment, additional costs were developed to include MF pretreatment. A cost estimate of MF/EDR would provide the high range for the treatment costs as compared to the EDR without pretreatment. Similar to the MF/RO cost estimate, the EDR costs were estimated for the production of 50-mgd blended product for both 350- and 500-mg/L TDS treatment goals by utilizing appropriately sized systems at 34- and 19-mgd production rates, respectively (figure 6-2).

**Table 6-5. EDR design criteria used for conceptual cost analysis**

| Parameter | 350-mg/L TDS Goal | 500-mg/L TDS Goal |
|---|---|---|
| EDR operation | Three-stage | Three-stage |
| EDR FWR | 85% | 85% |
| EDR feed flow | 40 mgd | 22.3 mgd |
| EDR product flow | 34 mgd | 19 mgd |
| Blending flow rate | 19.6 mgd | 31.2 mgd |
| RO product TDS | 82 mg/L | 82 mg/L |
| Blending water TDS | 750 mg/L | 750 mg/L |

#### *6.3.1.2 Estimated Treatment Costs*

As shown in table 6-6, the conceptual cost of treating 750-mg/L TDS water to produce 50 mgd of 350-mg/L TDS was $0.57 per 1,000 gallons (kgal) and $0.86 per kgal for EDR and MF/RO, respectively. A significant cost savings was estimated for the EDR without required pretreatment. Once MF pretreatment was factored into the EDR costing, the cost of treatment was $0.85 per kgal and was similar to the cost of MF/RO ($0.86 per kgal).

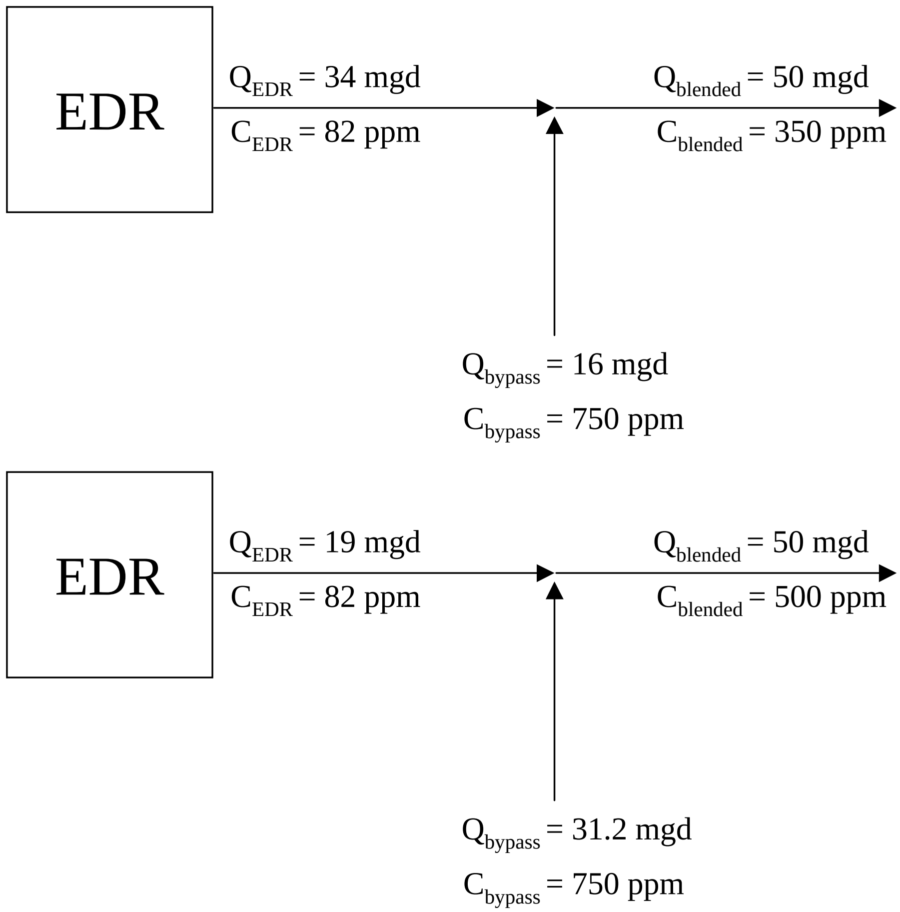


**Figure 6-2. Schematic of EDR 50-MGD blending at 350- and 500-mg/L treatment goals.**

**Table 6-6. Conceptual life cycle costs for desalting treatment options**

| | | Total Capital Cost ($ million) | Annual O&M Cost ($ million) | Total Annualized Treatment Cost[1,2] ($/kgal) |
|---|---|---|---|---|
| **350 mg/L TDS** | | | | |
| MF/RO | 35 mgd | $102 | $6.8 | $0.86 |
| MF/EDR | 34 mgd | $73.6 | $9.0 | $0.85 |
| EDR | 34 mgd | $33.4 | $7.4 | $0.57 |
| **500 mg/L TDS** | | | | |
| MF/RO | 17 mgd | $62.2 | $3.8 | $0.51 |
| MF/EDR | 19 mgd | $52.9 | $5.5 | $0.55 |
| EDR | 19 mgd | $20.2 | $4.2 | $0.32 |

[1] Amortization for 20 years at 5% interest.
[2] Annualized cost normalized to 50-mgd blended product with 350-mg/L TDS.

At the 500-mg/L TDS treatment goal, a significant cost savings for all treatment scenarios would be expected because of the reduction in the size of the treatment plant required. Treatment costs for MF/RO and MF/EDR systems were estimated to be $0.51 per kgal and 0.55 per kgal, respectively. This represents approximately a 40% reduction of the overall treatment costs. Similarly, a full-scale EDR plant without pretreatment is estimated to only cost $0.32 per kgal. A breakdown of the estimated capital and O&M costs is presented in table 6-7.

## 6.4 Process Advantages

Based on the conceptual cost comparison of the two pilot tested demineralization processes, it was shown that EDR may have a significant cost advantage over MF/RO, if no pretreatment beyond cartridge filtration is required. Factoring in a MF pretreatment step to the EDR causes both advanced treatment technologies to be cost competitive. Additional process advantages of both EDR and MF/RO treatment processes may be realized in light of overall treatment costs. The following table 6-8 summarizes the advantages and disadvantages of both technologies, as compared to one another.

**Table 6-7. MF/RO versus EDR full-scale cost breakdown**

| MF/RO Percent Breakdown | | EDR Percent Breakdown | |
|---|---|---|---|
| **Capital** | **%** | **Capital** | **%** |
| MF cost | 21 % | | |
| RO cost | 26 % | EDR cost | 60 % |
| | 17 % | Building | 15 % |
| Site work | 23 % | Site work | 20 % |
| Miscellaneous | 13 % | Miscellaneous | 5 % |
| **O&M** | **%** | **O&M** | % |
| RO power | 13 % | EDR power | 39 % |
| Other power | 9 % | | |
| Chemicals | 6 % | Chemicals | 2 % |
| Membrane replacement | 21 % | Membrane replacement | 18 % |
| Labor/maintenance | 35 % | Labor/maintenance | 25 % |
| Cartridge filter | 1 % | Cartridge filter | 8 % |
| Miscellaneous | 15 % | Miscellaneous | 8 % |

**Table 6-8. Advantages and disadvantages of demineralization processes**

| EDR | RO |
|---|---|
| **Advantages** | |
| ▪ Minimal pretreatment may be required (cartridge filtration is recommended)<br>▪ Operates at a low pressure<br>▪ Process is much quieter because high pressure pumps are not required<br>▪ Antiscalant is not required<br>▪ Membrane life expectancy is longer because foulants continuously are removed during the reversal process<br>▪ Requires less maintenance than RO due to reversal process | ▪ RO membranes provide a barrier to microorganisms and many anthropogenic organic contaminants (for the treated portion of the water produced)<br>▪ More demonstrated experience for wastewater demineralization<br>▪ RO membranes can remove more than 90% of TDS<br>▪ Source water blending will reduce size of systems<br>▪ Flexibility to provide higher quality water, if desired. |
| **Disadvantages** | |
| ▪ Limited to 50% salt rejection for a single membrane stack (stage)<br>▪ Requires larger footprint to produce similar quantity and quality of water if multiple staging is used<br>▪ Electrical safety requirements<br>▪ Less experience for wastewater demineralization in the United States<br>▪ Not as effective at removing microorganisms and many anthropogenic organic contaminants | ▪ Requires high pressure to achieve high salt rejection<br>▪ Requires pretreatment processes to minimize scaling and fouling.<br>▪ Requires chemical addition for MF and RO fouling control<br>▪ More routine maintenance may be required to maintain performance |

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# Appendix A
# Survey of Recent Literature

## A. Reverse Osmosis (RO), Application for Wastewater Treatment and Repurification

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# Appendix B
# SJ/SC WPCP Operator Experience During Pilot Study

One of the more valuable outcomes of the study was obtaining the practical experience in the operation of the pilot equipment. The pilot operations team included three senior operators in rotation and a project engineer. Each had the opportunity to learn the operation of both advanced treatment systems. The experience of the staff and key findings about the operation of the treatment processes are summarized below.

## Installation and Startup

Experience with both advanced treatment pilot units suggests that a high degree of planning is needed to ensure successful installation and startup of the equipment. Prior to equipment delivery, it was necessary to size electrical supplies and install wiring for telemetry and power to the skid-mounted units (or trailer-mounted, in the case of the electrodialysis equipment). In addition, adequate lighting and safety measures including conduit and cord covers and secondary containment for hazardous materials.

Another major effort was to install instrumentation and Supervisory Control and Data Acquisition (SCADA) equipment to monitor the pilot treatment units and ensure that they continuously produced the desired adjustable water quality objective (figure 1). Online monitoring data was verified with discrete sampling and laboratory testing which showed electronic data occasionally drifted out of calibration. There were also occasional lapses (power outages, etc.) when the SCADA system was unable to capture all the online monitoring data that required more operator intervention.

Of significant benefit was the ability of the SCADA system throughout the course of the study to allow remote access of data using the network. From the network, one could download the data in Excel allowing for automated macros that evaluated large volumes of data to assure system performance was achieved. Although time and resources did not permit, it is conceivable that a system could be designed to automatically populate an Excel spreadsheet at specific time intervals to support Statistical Process Control (SPC).

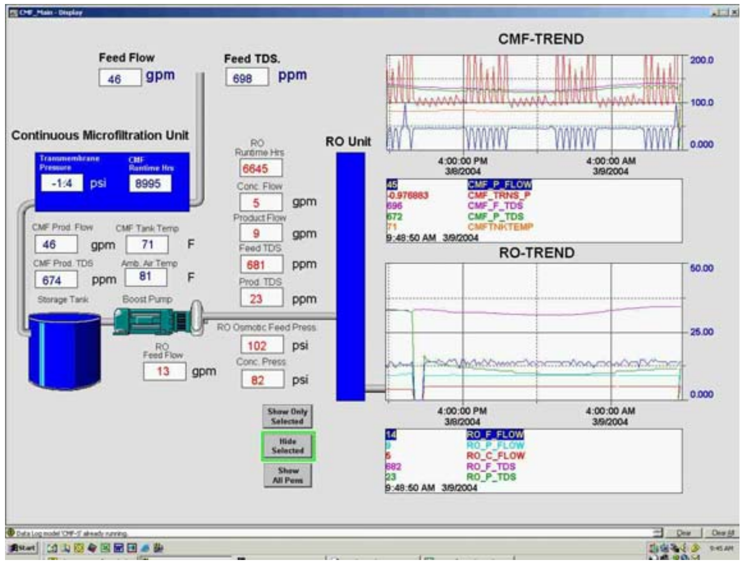


**Figure 1. SCADA personal computer (PC) interface showing the continuous microfiltration (CMF)/reverse osmosis (RO) feedback display**

## MF/RO Installation and Startup

After installation it was found that direct sunlight on the microfiltration facilities encouraged the growth of algae in the system. Green algae grew in the gauges making them hard to read and malfunction, and they eventually had to be replaced. To prevent further growth of algae, a tarp was installed over the microfiltration (MF) unit to protect it from direct sunlight. The tarp was raised only for sampling and inspection, which significantly reduced the problem. In addition, the backwash tank eventually had to be painted opaque as the backwash water was also polluted with algae growth. Algae in the system can increase Trans Membrane Pressure (TMP) ultimately reducing filtration efficiency.

## Electrodialysis Reversal (EDR) Installation and Startup

Prior to placing the EDR facilities into service, the equipment was specifically reviewed for Underwriter Laboratory (UL) compliance. As part of ensuring a safe workplace at the treatment plant, all equipment must have UL certification. Unlike the MF/RO, the EDR was outfitted with older electronic components that were responsible for providing the electromotive force driving filtration. Of primary concern was the voltage supply and rectifier of the Aquamite V providing an electric power supply rated at 480/460/380/220 Volts, 50/60 hertz (Hz), three-phase conditioned through a rectifier supplying the three-phased direct current (dc). Rigorous testing was applied to both of these components by UL, requiring the replacement of several parts and the installation of additional grounds in the trailer prior to operation.

Other issues needing to be corrected prior to operation included the repositioning of the control panel to allow unrestricted opening of the panel cover to a 90-degree door swing. However, only one person at a time would be allowed in the trailer, as the open door would not provide unrestricted exit of personnel from within the trailer in the event of an emergency.

Once UL compliant, the equipment was run in series with two canisters of granulated activated carbon (GAC) to reduce chlorine initially at a concentration of about 4-8 milligrams per liter (mg/L). After several weeks under GAC prefiltration, the influent chlorine concentration was reduced to allow algae to grow from within the transparent port windows in the GAC canisters, thereby polluting the EDR membrane. This algae buildup eventually required in-place cleaning of the EDR membrane with elevated chlorine. This new chlorine feed prefiltration manifold will be reviewed in a later section of the findings.

## Operation and Maintenance

### Operation and Maintenance—MF/RO

The MF/RO equipment functioned with antigravity vacuum filtration through a 0.2-micrometer (µm) polypropylene filter and later through polyvinyldifluoride or PVDF membrane filter. The polypropylene filter required chlorine pretreatment to neutralize free chlorine into its nondestructive chloramine form. Chloramination pretreatment required the dosing through a pump system of a one-to-one solution of ammonia to neutralize the free chlorine forming chloramine compounds that flow nondestructively through the polypropylene membrane. Because of the required chloramination, a reliable pumping system had to be maintained and monitored by hand to assure full conversion of the free chlorine to chloramines. Eventually, not having a backup ammonia pump had its toll, and the chlorammination pretreatment eventually failed causing catastrophic failure to the free chlorine sensitive polypropylene filter. As a result, the fibers began to break as shown in figure 2.

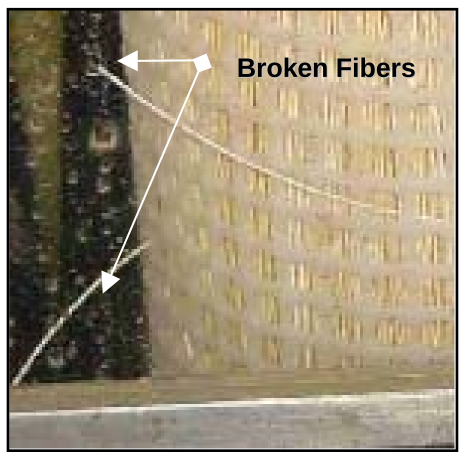


**Figure 2. Photo shows broken polypropylene microtubes (fibers) caused by free chlorine oxidation.**

Eventually, the polypropylene filter was replaced with PVDF (polyvinyl-difluoride), a free chlorine tolerant membrane not requiring ammination pretreatment. Other pretreatment safeguards included dosing RO influent with sodium bisulfite followed by antiscalent and cartridge filtration to assure the complete removal of chlorine, scale-forming compounds and particulates prior to passing through the RO. Of primary concern was to assure pre-RO water quality maintained an Silt Density Index (SDI) below 3, and the turbidity was kept below 0.5 nephelometric turbidity unit (NTU) to avoid particulate degradation of the RO elements, reducing filtration integrity.

Other difficulties included the failure of a level switch that would stick in the “tank empty” position causing the system to shut down. The purpose of the level switch was to assure the submersion and hydration of the microfilter membrane, as it is highly sensitive to dehydration. Throughout the study, computer safeguards were set into place to protect the system from dehydration or filtration failure (high TMP). The failed level switch must have caused in excess of 10 filtration shutdowns.

### *Chloramination Failure and Membrane Replacement*

Recycled water is normally disinfected by chloramination, a process whereby a small concentration of ammonia is added to the chlorinated effluent. The chlorine and ammonia combine to form chloramines, a stable form of chlorine that provides disinfection for an extended period of time. However, on occasion, fluctuations in flow can result in abnormally high chlorine concentrations at which time insufficient amount of ammonia may be added such that chlorine becomes available in a “free” or uncombined state. Unlike chloramines, free chlorine can have a highly oxidative effect on membranes.

These variations in chlorine concentration result from changes in the demand for recycled water. California Department of Health Services (DHS) regulations require recycled water to receive disinfection (defined as the product of detention time and chlorine concentration) of at least 450 mg/L-minutes on a continuous basis. On average, the system delivers about 10–15 million gallons per day (mgd), with a corresponding detention time between 200 and 300 minutes and a residual chlorine concentration between 2 and 5 mg/L. However, although a 4-million gallon reservoir provides minimal storage, the South Bay Water Recycling (SBWR) system essentially operates “on demand” such that increases in customer usage directly correspond to increased flow from the treatment plant. Increased flows result in shorter detention times and correspondingly require higher doses of chlorine, so that when peak demand reaches 25 or even 30 mgd, chlorine concentrations can go as high as 20 mg/L, outstripping the ability of the ammonia pumps to ensure complete chloramine formation in the recycled water, allowing free chlorine to pass into the system.

The microfilter membranes were composed of polypropylene and required complete hydration in an oxidant free solution to protect the modules from chemical degradation. Since recycled water occasionally had free-chlorine concentrations as high as 10 mg/L, the pilot system included an ammonia pretreatment pump to convert any remaining free chlorine into nondestructive chloramines. Unfortunately, the pilot plant ammonia pump failed within a few months of operation, and the untreated free chlorine degraded the polypropylene fibers of the microfilter membranes to the extent that the entire MF/RO system had to be shut down. As an alternative to designing and building a more reliable ammonia pumping system, the vendor instead chose to replace the polypropylene membranes with microfiltration modules made of polyvinylidene fluoride (PVDF), a fiber which is resistant to free chlorine.

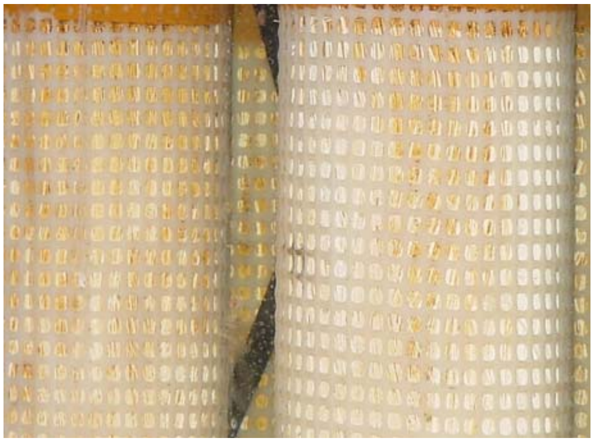

**Figure 3. Photo shows build up of suspected iron oxide deposits during the first days of operation.**

### *Failure and Replacement of RO O-Rings*

As a performance indicator, flux, feed water recovery (FWR) and TDS removal are the more critical parameters to monitor when assuring optimal RO performance. RO integrity became suspect any time the TDS was above 10 mg/L and the flux exceeded 19 gallons per square feet (gfd) with the feed water recovery exceeding 65 percent (%) for the USFilter H Series 3180. Early in the study, RO total dissolved solid (TDS) removal was achieving close to 99% reduction but at a feed water recovery of only 50%, so not much strain was put onto the system early in the study.

As the study progressed, the TDS would increase from under 10 mg/L to over 20 mg/L, which became indicative of integrity failure of the RO membrane system. When RO integrity is suspect, downtime is so undesirable, that replacement of both the membrane elements and the o-rings are a routine practice to assure optimal pressure integrity after any RO maintenance. When time permits, and the data are necessary, equipment can be deployed into the RO vessel to determine at what RO stage the integrity breach has occurred. This level of detailed analysis was not necessary for this study.

## Operation and Maintenance—EDR

Different TDS meters were deployed to measure the TDS concentration of the feed water, concentrate, and product waters. By design, the EDR control panel incorporated a TDS readout display for the meters integrated into the

EDR system.  These meters, were not always in agreements with the additional online meters deployed at various points during water production and were very difficult to calibrate.  Because of this, the laboratory data became a valuable part of the study when assisting decisions for system optimization.

By design, the EDR required little maintenance as the system employed a reversal, which shifted the polarity of the membrane reducing the accumulated buildup on any one electrode.  Throughout the study, one event of algal buildup on the EDR membrane requiring in-place cleaning of the membrane.

### *Regular Operation and Maintenance*

Free-chlorine was also of concern to the anion and cation stack transfer membranes of the EDR.  Initially, two GAC filters were installed in series to remove both free-chlorine and any particulates.  However, the TMP increased after a few months due to algal growth on the membrane stack.

Cleaning the membranes stacks is an onerous process (the stack weighs over 1,500 pounds and should be opened only when all other fouling-prevention remedies fail), so the feed water manifold was redesigned to temporarily bypass the GAC filters and send chlorinated feed water directly into the EDR stacks for several hours.  This cured the fouling problem, and the new manifold blended chlorinated feed water with filtered water to maintain a continuous chlorine concentration of around 0.25 mg/L, which was incorporated as part of normal operations.

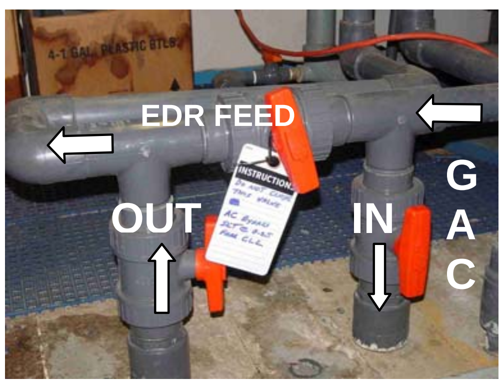


**Figure 4.  Photo shows EDR system with GAC bypass, trimming raw chlorinated water to reduce algal growth in the stack.**

# OBSERVED OPPORTUNITIES FOR FULL-SCALE DESIGN

The objective of water supply and treatment operations is to maintain cost-efficient systems with minimal downtime. Also of importance is the necessity to flexibly meet expected ranges of influent water quality and fluctuating volumes while maintaining the production of consistent water quality and quantity. In addition to quality and quantity (capacity), other design considerations such as safety, redundancy, and automation should also be considered as part of the design. While these ideas are not new, every new treatment system should keep these principles in mind when evaluating designs for optimal system performance. Having the ability to test a pilot prior to the design of a full-scale system allows the testing of equipment and treatment systems with actual system conditions.

In order to maintain consistent feed water quality and stabilize fluctuations in demand, the need for a reservoir prior to advanced treatment becomes more apparent. The construction of a reservoir would also assure sufficient contact time and minimize chemical supply costs currently expended based on swings in volume demand. The reservoir will also dampen spikes in water quality routinely experienced on an inline system.