MWH

PREPARED FOR THE
BAY AREA REGIONAL DESALINATION PROJECT

Pilot Testing at Mallard Slough
Pilot Plant Engineering Report

JUNE
2010

MWH

June 8, 2010

Mr. Hasan Abdullah
East Bay Municipal Utilities District
375 11th Street, MS407
Oakland, CA 94607

Subject: Bay Area Regional Desalination Project
Pilot Plant Engineering Report, Final

Dear Mr. Abdullah,

The MWH Team is pleased to submit the final Pilot Plant Engineering Report for the Bay Area Regional Desalination Project. Per your request, enclosed are two copies of the document. Two copies have been sent under separate cover to Contra Costa Water District, San Francisco Public Utilities Commission, Santa Clara Valley Water District, and Alameda County Flood Control and Water Conservation District Zone 7. Six copies and an additional CD were sent under separate cover to the California Department of Water Resources.

The pilot study was conducted with a nominal 50 gpm pilot plant installed and operated at the existing Mallard Slough Pump Station. Pilot testing verified the technical feasibility of a full-scale desalination facility at that site to meet the water quality targets of the four agencies, despite the complex feedwater quality. Pilot testing also provided a set of design criteria for evaluating a full-scale facility at that site, including capital and operational costs.

The pilot study demonstrates that the desalination treatment facility would experience a wide range of salinities due to the influence of the Sacramento River and San Joaquin River, and due to tidal effects within San Francisco Bay and Suisun Bay. If the plant is located at the East Contra Costa Site, our team recommends for the plant to be designed for a maximum Total Dissolved Solids concentration between 11,500 mg/L and 12,000 mg/L, based on historical dry year conditions. Data also indicate that the plant will be subjected to a normal TDS range between 500 mg/L and 5,500 mg/L during a dry year.

Our analysis indicates that facility design should be based on pretreatment utilizing ultrafiltration membranes followed by a two stage reverse osmosis process. Brackish and seawater membranes within the first and second stages will provide a high level of recovery exceeding 80% during average dry year TDS conditions.

Capital cost for a facility that would utilize 25 mgd of feedwater to produce 19.8 mgd of treated water, including the intake and pipeline for conveyance to existing transmission system, is estimated to be $168.5M, or approximately $8.50 per gpd. This value includes contingencies and planning, permitting, engineering, and administrative costs which would be incurred during the course of the project. With an annual operating cost estimated at $10.45M, the present worth of the facility is determined to be $373M, or approximately $550 per acre-foot of water.

We have thoroughly enjoyed working with the four agencies on this interesting project and look forward to assisting with future phases of work. This report is the final outcome of extensive work which was initiated in 2007. It reflects the combined efforts of MWH as well as all four agencies. We are pleased with the final results, which could not have been possible without the time and effort spent by agency representatives to review the various interim draft reports and to insure conformance with project requirements and goals.

Please contact me if you wish to discuss any of the findings or recommendations contained in this final report.

Sincerely,

Charles Bromley, P.E., BCEE
Project Manager

ACKNOWLEDGEMENTS

We acknowledge and thank the project team for their contributions, dedication and hard work on this successful project, without which this project would not have been possible:

PROJECT CONTACTS FROM THE FOUR AGENCIES

- Hasan Abdullah, PE, of the East Bay Municipal Utility District
- Marie Valmores, PE, of the Contra Costa Water District
- Manisha Kothari and Paula Kehoe of the San Francisco Public Utilities Commission
- Pam John, PE and Ray Wong, PE of the Santa Clara Valley Water District

PROJECT CONTRIBUTORS

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- MWH Research, MWH Laboratories, CRG Maine Laboratories, EBMUD Laboratory, and CCWD Laboratory

EQUIPMENT VENDORS WITH SIGNIFICANT CONTRIBUTIONS

- Entities providing membrane pilot units and support
  - Siemens Water Technologies
  - Layne Christensen Company
- Entities providing membrane modules/elements
  - Norit Americas, Inc.
  - Dow/Filmtec Corp.
- Ferguson Enterprises, local piping and equipment supplier

MWH TEAM

- MWH Americas, Inc.
- GSE Construction Company, Inc.
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- SRT Consultants
- Tenera Environmental
- Malcolm Pirnie, Inc.
- M. Lee Corporation
- Applied Marine Sciences
- Kearns and West, Inc.

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AEL	Adult Equivalent Loss Model
AF	acre-feet
AFY	acre-feet per year
AMTA	American Membrane Technology Association
AWWA	American Water Works Association
BARDP	Bay Area Regional Desalination Project
BW	Backwash
BWRO	Brackish Water Reverse Osmosis
CCPP	Calcium Carbonate Precipitation Potential
CCCSD	Central Contra Costa Sanitary District
CCWD	Contra Costa Water District
CDFG	California Department of Fish and Game
CEB	Chemically-Enhanced Backwash
CEQA	California Environmental Quality Act
CIP	Clean-In-Place
CT	Concentration x Time product
DBP	Disinfection Byproducts
DDSD	Delta Diablo Sanitation District
DPH	California Department of Public Health
deg F	Degrees Fahrenheit
DWR	California Department of Water Resources
EBMUD	East Bay Municipal Utilities District
EPA	United States Environmental Protection Agency
FH	Fecundity Hindcast Model
gfd	gallons per square foot per day
gpd	gallons per day
gpm	gallons per minute
HAA5	Halo-Acetic Acids
kW	kiloWatts
LSI	Langelier Saturation Index
MCC	Motor Control Center
MCL	Maximum Contaminant Limit
MF	Microfiltration
mgd	million gallons per day
mS/cm	milliSiemens per centimeter
MSPS	Mallard Slough Pump Station
MW	Maintenance Wash
NaOCl	Sodium Hypochlorite

NEPA	National Environmental Protection Act
NF	Nanofiltration
NF3	NF System (Train) No, 3
NMFS	National Marine Fisheries Service
NPDES	National Pollutant Discharge Elimination System
NTU	Nephelometric Turbidity Units
O&M	Operations and Maintenance
OD/ID	Outer Diameter/Inner Diameter
ORP	Oxidation Reduction Potential
PDT	Pressure Decay Test
PE	Proportional Entrainments estimate
PES	Polyethersulfone
PG&E	Pacific Gas & Electric
PLC	Programmable Logic Controller
PPS	Pilot Plant Study
ppd	Pounds per day
ppm	Parts per million
psi	pounds per square inch
QA/QC	Quality Assurance/Quality Control
RO	Reverse Osmosis
RO1	RO System (Train) No. 1
RO2	RO System (Train) No. 2
RWQCB	Regional Water Quality Control Board
SCVWD	Santa Clara Valley Water District
SDI	Silt Density Index
SFPUC	San Francisco Public Utilities Commission
SMCL	Secondary Maximum Contaminant Limit
SWRO	Seawater Reverse Osmosis
SWTR	Surface Water Treatment Rule
TCF	Temperature Correction Factor
TDS	Total Dissolved Solids
TM	Technical Memorandum
TMP	Trans-Membrane Pressure
TOC	Total Organic Carbon
TSS	Total Suspended Solids
TTHM	Total Trihalomethanes
UF	Ultrafiltration
USFWS	United States Fish and Wildlife Service
UV	Ultraviolet radiation
VFD	Variable frequency drive
μS/cm	microSiemens per centimeter

1.0 EXECUTIVE SUMMARY

The Bay Area's four largest water agencies, the Contra Costa Water District (CCWD), the East Bay Municipal Utility District (EBMUD), the San Francisco Public Utilities Commission (SFPUC), and the Santa Clara Valley Water District (SCVWD) are jointly evaluating the development of a shared regional desalination facility to improve water supply reliability of the Bay Area and benefit approximately 5.4 million residents and businesses served by the four agencies. The goal of the Bay Area Regional Desalination Project (BARDP) is to develop one or more desalination plants operating on brackish water or on seawater to provide supplemental water supply during droughts, emergencies such as earthquakes and levee failures and maintenance-related outages.

The four agencies have worked together to better utilize and leverage existing infrastructure and assets owned by the four water agencies to receive and distribute desalination product water. By pooling resources together and leveraging existing infrastructure and assets, a regional treatment plant would:

- Minimize potential adverse environmental impacts associated with the construction of separate desalination plants in close proximity to one another and construction of new facilities;
- Provide substantial cost savings through economies of scale, such as pooling resources and sharing of project administration, as compared to individual projects conducted separately by the agencies; and
- Promote a strong regional cooperation concept by joint ownership, operation, and management of a regional desalination facility that will serve the needs of four major water providers in northern California.

Since 2003 the four agencies have worked together to conduct a Pre-Feasibility Study for the BARDP. The Pre-Feasibility Study findings concluded that a regional desalination facility in the Bay Area is feasible. The environmental, permitting, institutional, and public outreach aspects of such facilities need to be systematically addressed and the viability of the BARDP will depend on the commitment of each agency's stakeholders, including board members, management and staff.

Next, the agencies continued work on a Feasibility Study which was completed in June 2007. The Feasibility Study developed a process for evaluating regional desalination facilities. It evaluated institutional options such as Joint Powers Authority and other institutional mechanisms, developed a process and criteria to evaluate optimal desalination sites, and began the public stakeholder outreach for the BARDP. The Feasibility Study concluded that there are at least three Bay Area locations that are suitable for siting such a regional desalination facility (see Figure 1-1).

The Feasibility Study recommended conducting a pilot test at CCWD's Mallard Slough Pump Station site located in the eastern part of Contra Costa County. The goals of the pilot test are to collect data on technical feasibility (pretreatment options, membrane performance, design parameters) and to assess the potential environmental impacts (brine disposal, marine life) of an East Contra Costa desalination facility. Additionally, the East Contra Costa site was selected for piloting to fill in the data gap that currently exists regarding desalination piloting in an estuarine environment; other agencies have recently conducted pilot tests in the San Francisco Bay (Marin Municipal Water District) and the Pacific Ocean (Santa Cruz).

The pilot test was started in October 2008 and continued through April 2009. Approximately 50 gallons per minute (gpm) were drawn from CCWD's Mallard Slough intake. Performance data were collected for treatment by two types of ultra-filtration pre-treatment membranes, two types of Reverse Osmosis (RO) membranes, and one Nanofiltration (NF) membrane. This report presents the pilot test findings and provides recommendations for future steps.

With the State of California facing water supply challenges, the State Department of Water Resources (DWR) provided grant funds to the four agencies to conduct a Feasibility Study through Proposition 50 — the Water Security, Clean Drinking Water, Coastal and Beach Protection Act. Later, following successful completion of the Feasibility Study, another State Proposition 50 grant was awarded to conduct the Pilot Study. The combined State grants paid for 50% of the projects' costs. In addition, the BARDP is authorized to receive $4 million in federal grants under the Water Resources Development Act of 2007, Section 5158 (88). Obtaining these grants underscores the importance of this BARDP effort.

1.1 Pilot Study Objectives and Background

The primary objectives of the Pilot Plant Study (PPS) were to:

- Establish an organizational structure to implement the pilot study;
- Maximize the efficiency of operating and maintaining a regional desalination facility including sludge/solids disposal evaluation and water quality testing;
- Identify potential environmental impacts and evaluate methods to mitigate these potential impacts;
- Identify the preferred pre-treatment for this site;
- Identify the preferred RO system configuration for this site; and
- Develop an information sharing platform to share test data, methodologies and project information with other interested parties in the State.

The Mallard Slough Pump Station (MSPS) located in eastern Contra Costa was chosen as the site of the pilot study (Figure 1-1). The MSPS site had several benefits: ready access to potential source water in the Suisun Bay; available power and related utilities; and ease of operations and site use as the site is owned by CCWD. Existing fish screens that were already in use at the pump station intake were used for drawing water from the slough, eliminating the need for a dedicated pilot intake while minimizing biological or other potential environmental impacts.

Figure 1-1: Pilot Plant Location

The MSPS serves to furnish untreated water to CCWD during periods of low salinity and so consequently does not operate for much of the year. It is located in a remote area in unincorporated Pittsburg and is a fenced site with controlled access via several locked gates. The site is shown in Figure 1-2. For security purposes the piloting equipment was installed inside of the temporary containers and trailers within the area designated on the left side of the Figure 1-2.

Water in Suisun Bay is a blend of fresh water from the Sacramento and San Joaquin Rivers and naturally occurring seawater entering tidally through the San Francisco Bay. As evidenced by existing water quality monitoring stations, and confirmed during the pilot study, source water is subject to significant tidal influence, resulting in wide variations of total dissolved solids (TDS) observed on a daily, monthly and seasonal basis.

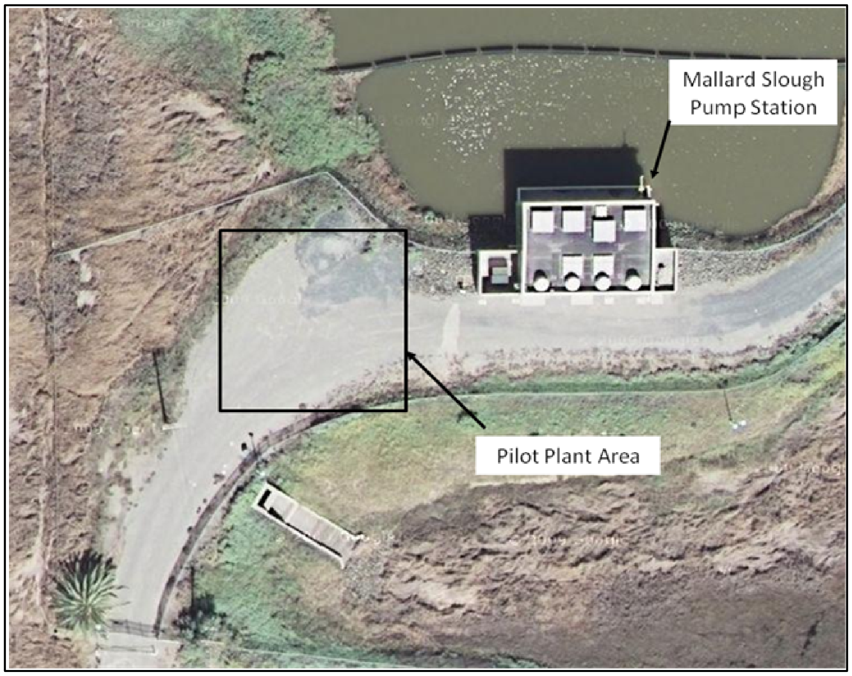


Figure 1-2: Pilot Plant at Mallard Slough Pump Station

1.2 Pilot Description and Activities

The overall pilot process was established to mimic a full scale treatment plant. To meet the pilot study objectives stated above, work was planned and conducted as shown in Table 1-1.

Table 1-1: Summary of Project Objectives

Objective	Findings
1. Establish an organizational structure to implement pilot study	A memorandum of agreement between the four partner agencies was approved on May 22, 2007 to implement the pilot study.

Objective	Findings
2. Minimize potential adverse environmental effects to aquatic organisms.	An evaluation of potential screens, including Aquatic Filter Barriers such as Marine Life Exclusion System (MLES) by Gunderboom, Inc., was conducted. Based on the results of the evaluation, with DWR's concurrence, existing screens at the MSPS site was used for piloting. Using existing screens helped minimize potential adverse impacts associated with new intake construction.
3. Conduct biological sampling and analysis to determine the existence or absence of biological species at the pilot study intake location	Biological sampling and associated evaluations were conducted to gain an understanding of aquatic species populations. Sensitive fish (Longfin and Delta Smelt) were found to be present during February and March 2009.
4. Verify level of screening required upstream of the pretreatment systems.	The 100 micron screen was effective in protecting the pretreatment membranes from particulate fouling. The automatic cleaning feature will be an important step in minimizing maintenance for the desalination plant operations and maintenance staff.
5. Identify preferred pretreatment method.	The pilot study confirmed that both pressurized and submerged membranes are able to meet filtrate water quality goals. The pressurized system was able to operate at high sustained flux and could be acceptably cleaned. The submerged system experienced fouling; however, acceptable operation was achieved once the fouled membranes were replaced.
6. Evaluate disposal options for solids generated by the desalination facility.	Solids produced in the pilot are suitable for thickening and dewatering using commonly available and commercially proven processes. Final disposal into a landfill or potential reuse is feasible.
7. Test technologies and methods to maximize the efficiencies.	Desalination can be effectively achieved using a two stage membrane RO process. Other potential innovative options are explored in this report.
8. Identify and test concentrate toxicity levels and related impacts to receiving waters.	Toxicity of the desalination membrane concentrate was tested utilizing bioassays with serial dilutions and three selected aquatic organisms. No significant toxicity effects were found.

Objective	Findings
9. Test product water quality.	Treated water quality was tested throughout this project for a wide variety of parameters. All of the systems tested were able to meet agency water quality goals during period of low feed water salinity. Nanofiltration membranes, however, were challenged during higher salinity periods and would require multiple stages or passes if considered in the full scale.
10. Conduct water compatibility tests to evaluate whether the permeate is compatible with EBMUD aqueduct or CCWD multi-purpose pipeline water	Water compatibility tests indicate the treated water can be safely stabilized and matched to existing waters in EBMUD aqueduct or CCWD multi-purpose pipeline.
11. Conduct public outreach in support of the project.	The following public outreach events were conducted during the course of this project. • Open House, San Francisco, SFPUC headquarters, December 16, 2008 • Presentation to Richmond-Pinole Lions Club, San Pablo, February 18, 2009 • Presentation to Walnut Creek Lions Club, Walnut Creek, April 8, 2009 • Presentation to WateReuse Chapter meeting, San Francisco, December 4, 2009 • Open House, EBMUD headquarters, Oakland, December 9, 2009 • Presentation to SIR-51 meeting, Los Altos Hills, January 6, 2009 Additional efforts and activities are planned by the four member agencies.

All pilot objectives were met for this project. Further discussions of key objectives are provided in the following paragraphs and may be found within this report.

Initial studies, workshops, and technical evaluations were initiated in July 2007 and completed between August 2007 and September 2008 to evaluate water quality, obtain a permit to discharge pilot wastes to the sanitary sewer, develop pilot equipment specifications and arrangements, procure equipment, erect the pilot equipment, interconnect with available utilities, and complete startup operations.

Based on this work, pilot systems were designed as illustrated in Figure 1-3. Feed water was pumped through a 100 micron self-cleaning screen, followed by two parallel ultrafiltration (UF) units. These systems featured Norit Americas, Inc. pressurized membranes and

Siemens/Memcor submerged membranes. Equipment was fabricated and leased from Layne Christensen Company and Siemens Water Technologies Corporation.

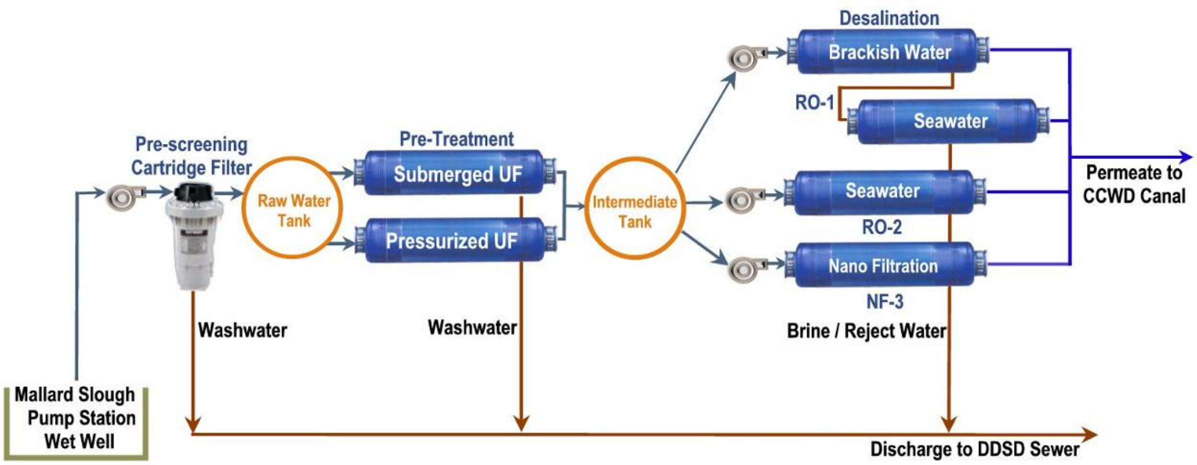


Figure 1-3: Pilot Plant Schematic

Combined filtrate produced by these UF units was pumped into three parallel RO membrane desalination systems. The configurations of the RO trains are as follows:

- RO Train No. 1: Two-Stage RO desalination, brackish water membrane treatment followed by seawater membrane treatment. Concentrate from brackish water membranes is fed into the seawater membranes to increase overall permeate recovery.
- RO Train No. 2: Single stage RO desalination using seawater membranes.
- NF Train No. 3: Single stage desalination using NF membranes.

Permeate collected from these parallel desalination processes was stabilized and pumped into the CCWD untreated water supply pipeline. Concentrate and other pilot plant wastes were collected and pumped to the nearby Delta Diablo Sanitation District (DDSD) sewer pipeline.

The overall process was established to mimic a full scale treatment plant as it might be designed on behalf of the four agencies. Chemicals were utilized for disinfection, coagulation, dechlorination, filtrate conditioning, membrane cleaning, and permeate stabilization. Specific chemicals included sodium hypochlorite, aqueous ammonia, ferric chloride, sodium bisulfite, citric acid, antiscalant, and caustic soda.

All pilot test work including water intake and discharge of wastes were conducted under current permits, therefore no new permits were required.

Field operations, laboratory work, and special studies were conducted while the pilot plant was operational from October 2008 through April 2009. Three separate runs of approximately six weeks duration each were conducted and were designed in response to actual water quality conditions encountered at the site:

- Run No. 1: Operated during period from November 2008 to December 2008 to establish baseline system operating parameters.

- Run No. 2: Operated during high salinity period from January 2009 to February 2009 to establish higher flux and mid-challenge system operating parameters.

- Run No. 3: Operated during low salinity period from March 2009 to April 2009 to establish highest manufacturer recommended flux and challenge system operating parameters.

Data were collected for each pilot run by observing and recording various piloting parameters, such as pressure, temperature, pH, chemical tank levels, and flow, at regular intervals and at various points in the treatment processes. Computerized equipment was used to record and archive data, while manual entry forms were completed during daily pilot plant inspections and experiments. Routine field analyses were performed for:

- Physical parameters of the test water at various points in the process,
- Silt Density Index (SDI), pressure decay tests, etc., and
- Verification and calibration of various instruments and field devices.

1.3 Key Pilot Study Findings

The pilot study has led to a greater understanding of water quality and treatment issues for this potential water source.

1.3.1 Water Quality Parameters

Feed water quality observed during the pilot study exhibited significant seasonal and tidal fluctuations in salinity, with high salinity in the dry season and low salinity in the wet season with the freshwater influence. Parameters are summarized in Table 1-2.

The concentration of total dissolved solids (TDS) during the pilot testing period ranged as high as 12,000 mg/L in the dry season and below 1,000 mg/L in the wet season. Conversely, turbidity was low in the dry season (5-15 Nephelometric Turbidity Units [NTU]) and high in the wet season due to runoff (20-30 NTU with spikes greater than 40 NTU). Because of the shallow source water, water temperature patterns mirrored those of the ambient air temperature, ranging from approximately 46 deg F to 70 deg F.

Table 1-2: Pilot Feed Water Quality Parameters

	Dry Season[1]	Wet Season[1]
Total Dissolved Solids (TDS)	Up to 12,000 mg/L	<1,000 mg/L
Boron	1 mg/L	0.2 mg/L
Sodium	1944 mg/L	198 mg/L
Chloride	3259 mg/L	311 mg/L
Turbidity	5 to 15 NTU	20 to 30 NTU with spikes >40 NTU
Total Organic Carbon (TOC)	1.8 mg/L	2.4 mg/L
Iron	406 µg/L	1040 µg/L
Aluminum	249 µg/L	493 µg/L

Note 1: Dry Season feedwater quality spanned November 2008 through February 2009 (Runs 1 and 2). Wet Season spans March and April of 2009 (Run 3). The seasons are further discussed in Section 4.1.2.

During the pilot study, conductivity was measured continuously via online instruments. Conductivity as well as TDS were measured by offsite labs on a weekly bases. Based on these empirical lab results, TDS and conductivity are correlated, but the TDS:Conductivity ratio varies between waters, as shown:

Pilot Plant Stream	TDS:Conductivity Mean Ratio (with Standard Deviation)
Brackish Water (pre-RO)	0.61 (SD = 0.05)
NF/RO Concentrate	0.71 (SD = 0.05)
NF/RO Permeate	0.49 (SD = 0.06)

These ratios are used throughout the report to convert conductivity values (µS/cm) into TDS values (mg/L).

1.3.2 Pretreatment

The pilot data collected confirmed that both the pressurized and submerged UF membrane systems produced a suitable feedwater for the RO systems, meeting the filtrate water quality goals of SDI < 3 and turbidity < 0.15 NTU. However, the pilot study was not of sufficient duration to fully optimize all operating parameters.

The specific flux that the pressurized UF system was able to sustain was significantly higher in magnitude than that of the submerged UF system, indicating that more water could be produced per membrane area with a lower transmembrane pressure (TMP). The trade-off for the higher specific flux is that the pressurized UF membranes required the use of small amounts of ferric chloride as a coagulant, adding iron to the backwash waste, and also featured a lower water recovery over the pilot study. More feedwater was consequently drawn into the system to produce the same amount of filtrate. Additionally, using coagulant in the pretreatment system will adversely affect solids handling by increasing the necessary size and cost of the facilities.

The submerged membranes experienced irreversible fouling that was not recoverable by CIPs, even after modifying the CIP procedures for more aggressive cleaning. Replacement of the submerged membranes was required during the pilot study. The pressurized membranes, on the other hand, experienced consistent permeability and no significant fouling regardless of the flux. Because the long term operability of the submerged membranes is not clear, and since the pressurized membranes performed technically superior over the submerged type during the pilot study, the pressurized UF membranes were selected for the scale-up evaluation.

1.3.3 Desalination

Each desalination system had distinct operational advantages and disadvantages which are broadly summarized in Table 1-3. Running the systems at pilot scale over the diverse feedwater conditions in the testing period provided specific performance data that can be used to project full-scale capital and operational expenditures.

The RO Train No. 1 system was a 2:1 array (twice the number of vessels in Stage 1 compared to Stage 2) with brackish water membranes in the first stage and seawater membranes in the second stage. Brackish water membranes were most suitable for the anticipated feedwater quality. A second stage, which received concentrate produced by the first stage, was used to increase recovery. Seawater membranes were selected for Stage 2 because of their higher rejection characteristics since the salt concentration in the Stage 2 feed was higher than the Stage 1 feed.

The RO Train No. 2 system was a single vessel of seawater membranes. Seawater elements were tested because they were potentially suitable when the feedwater salinity was at its highest, but it was also desirable to understand how the seawater elements would perform given low salinity feedwater in the wet season.

The NF Train No. 3 system was a single vessel of NF membranes. NF elements were tested because they were potentially suitable when the feedwater salinity was at its lowest, but it was also desirable to understand how the NF elements would perform given high salinity feedwater in the dry season.

Table 1-3: Desalination System Performance Comparison

Description	Goal	RO Train No. 1 Two Stage Brackish and Seawater Membranes	RO Train No. 2 Single Stage Seawater Membranes	NF Train No. 3 Single Stage Nanofiltration Membranes
Recovery	High	70-82%	50-62%	50-63%
Specific Flux, gfd/psi	High	Approx. 0.1	0.07-0.075	0.19-0.26
Permeate TDS[1], mg/L	< 500	23-120	9.5-27	44-220
Permeate Boron, mg/L	< 0.5	0.14-0.48	0.08-0.2	0.2-0.69
Permeate Chloride, mg/L	< 100	19-67	6.5-11	56-130
Permeate Sodium, mg/L[2]		12-43	2.8-7.8	13-82
Permeate Turbidity, NTU[2]		<0.05	<0.05	<0.05
Permeate TOC, mg/L[2]		<0.1 – 0.5	<0.1 – 0.5	<0.1 – 0.5
Permeate Iron, mg/L[2]		< 0.011	< 0.011	< 0.011
Permeate Aluminum, mg/L[2]		< 0.021	< 0.021	< 0.021

Note 1: All water quality values are from dry season conditions (Runs 1 and 2).
Note 2: No specific goal established.

1.3.4 Desalination Alternatives Based on Pilot Findings

RO Train No. 1 is a very suitable system for a full-scale plant because it achieves a high recovery, while providing very good water quality and meeting the four agencies' goals for TDS, chloride and other parameters. RO Train No. 2 features the best permeate water quality but is also subject to high feed pressure requirements. NF Train No. 3, although having the lowest energy requirements, does not meet the project's goal for chloride or boron in the dry season. The chloride goal in particular is exceeded by up to 30% based on pilot evidence. Adding a second NF stage may provide advantages for improved recovery with lower total operating pressure during certain monthly periods, although degradation of permeate quality must be considered.

Combining RO Train No. 2 and NF Train No. 3 into a single integrated system may nevertheless have advantages to the four agencies. Operating an NF system during low TDS periods would presumably provide substantial operating cost savings due to low feed pressures. However, in order to achieve sufficient recovery, the NF concentrate would be sent to the seawater membranes, even during low TDS periods, increasing the power requirements of the overall system. During high TDS periods, dividing the influent flow between the NF and seawater membrane could achieve water quality goals through blending of permeate produced by the two systems. Consequently, two alternatives, both using pressurized pretreatment membranes, were evaluated further and subjected to a present worth analysis, as described below.

Alternative No. 1: RO Train No. 1, as piloted

Alternative No. 1 evaluates a two stage system designed to match the pilot RO Train No. 1. In a two stage system, concentrate from Stage 1 is sent to a second set of membranes (Stage 2) to increase the overall recovery. Brackish water membranes were determined to be most suitable for the Stage 1 feedwater quality. Seawater membranes were determined the most suitable for Stage 2 because of their higher rejection characteristics in response to the higher salt concentration in the Stage 2 feedwater.

Because desalination is an energy intensive process, a number of techniques have been developed to reduce the energy consumption. For Alternative No. 1, a pressure exchanger will transfer the residual pressure in the Stage 2 concentrate to the Stage 1 feed, reducing the demand on the high pressure feed pump. In addition, an interstage boost pump with a variable frequency drive (VFD) will supply additional pressure to Stage 2 further reducing the demand on the high pressure feed pumps while improving the life of the membranes by balancing the flux between stages.

Alternative No. 1 Summary

- Brackish water RO membranes in Stage 1 and seawater RO membranes in Stage 2.
- 2:1 array (twice the number of membranes in Stage 1 compared to Stage 2).
- Approximately 70% total system recovery during high TDS conditions.
- Pressure exchanger for energy recovery.
- Interstage boosting with VFD.
- Pressure exchanger to transfer residual pressure to Stage 2.

Alternative No. 2: A hybrid plant that has independent trains for RO Train No. 2 and NF Train No. 3.

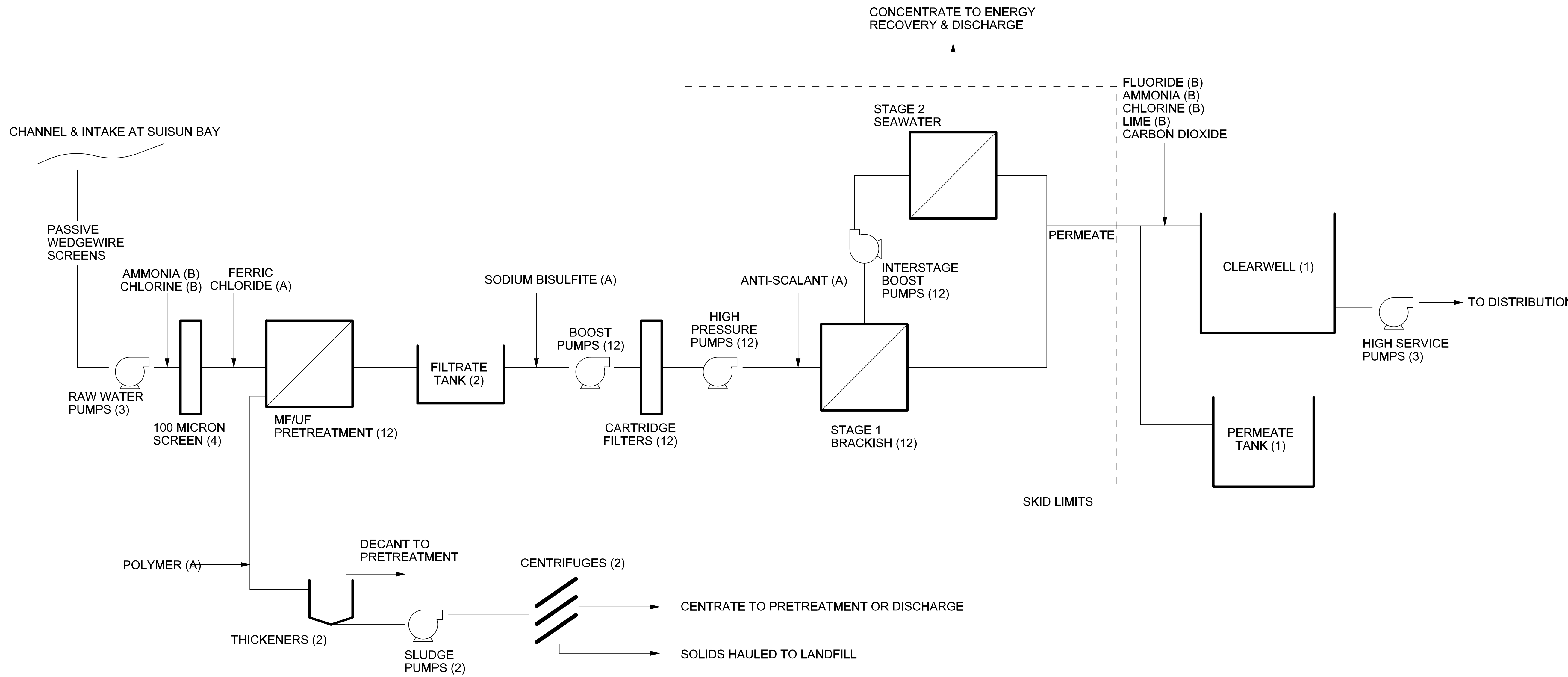
Alternative No. 2 is a combination of the piloted RO Train 2 and NF Train 3. Feedwater is split between the two types of membranes to meet treated water goals. During the dry, high salinity, season the seawater RO train will handle approximately 70% of the feed flow due to its superior salt rejection characteristics. During low TDS conditions the NF train would treat nearly 100% of the feed flow due to its lower feedwater pressure requirement. In order to achieve the highest recoveries possible, all NF concentrate will be sent to the seawater membranes.

In order to reduce energy demands in Alternative 2, a pressure exchanger would be used on the seawater membranes, transferring pressure from the seawater concentrate to the seawater feed.

Alternative No. 2 Summary

- Two single-stage membrane systems operating in parallel: NF membranes in one train, and seawater membranes in the second train.
- NF concentrate partially recovered by blending with feedwater being pumped to the seawater RO train.
- Approximately 58% total system recovery during high TDS conditions.
- Pressure exchanger to recovery energy in the seawater train.

Process flow diagrams for both alternatives are furnished in Figure 1-4 and Figure 1-5.



NOTES

1. NUMBER OF UNITS SHOWN IN ( )
2. MEMBRANE CLEANING SYSTEMS NOT SHOWN
3. (A) INDICATES CHEMICALS ARE STORED IN CHEMICAL BUILDING A.
4. (B) INDICATES CHEMICALS ARE STORED IN CHEMICAL BUILDING B.

REV	DATE	BY	DESCRIPTION

SCALE: NONE

WARNING
0 ½ 1
IF THIS BAR DOES NOT MEASURE 1" THEN DRAWING IS NOT TO SCALE

DESIGNED C. BROMLEY
DRAWN B. BEHNKE
CHECKED G. BEHNKE

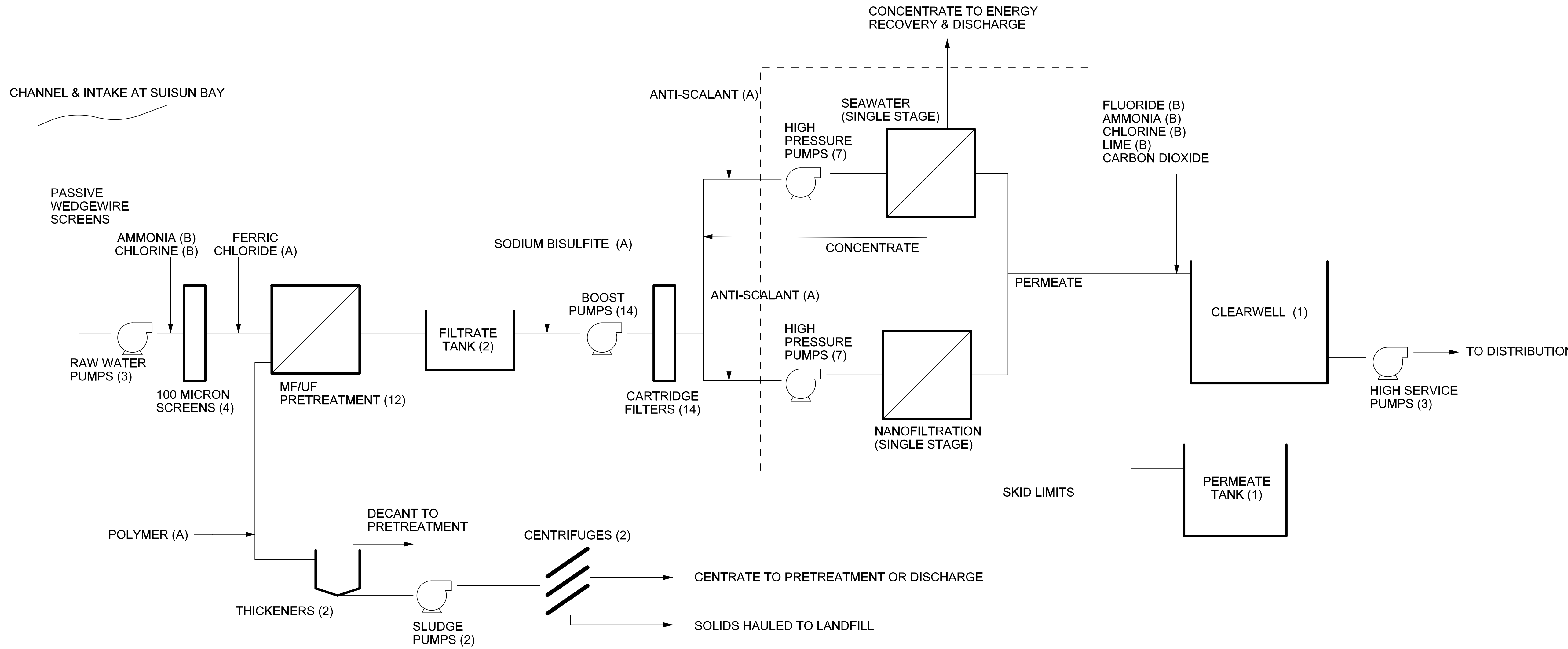
MWH AMERICAS, INC.
2121 N. CALIFORNIA BLVD
WALNUT CREEK, CALIF. 94596

MWH

BAY AREA REGIONAL DESALINATION PROJECT

FIGURE 1-4
ALTERNATE NO. 1 SCHEMATIC DIAGRAM
2-STAGE BRACKISH-SEAWATER MEMBRANES

SHEET



NOTES

1. NUMBER OF UNITS SHOWN IN ( )
2. MEMBRANE CLEANING SYSTEMS NOT SHOWN
3. (A) INDICATES CHEMICALS ARE STORED IN CHEMICAL BUILDING A.
4. (B) INDICATES CHEMICALS ARE STORED IN CHEMICAL BUILDING B.

REV	DATE	BY	DESCRIPTION	SCALE	WARNING					SHEET
				NONE	0 ½ 1 IF THIS BAR DOES NOT MEASURE 1" THEN DRAWING IS NOT TO SCALE	DESIGNED C. BROMLEY DRAWN B. BEHNKE CHECKED G. BEHNKE	MWH AMERICAS, INC. 2121 N. CALIFORNIA BLVD WALNUT CREEK, CALIF. 94596	MWH	BAY AREA REGIONAL DESALINATION PROJECT	FIGURE 1-5 ALTERNATE NO. 2 SCHEMATIC DIAGRAM PARALLEL SEAWATER AND NF MEMBRANES

1.4 Special Studies

Three special studies were conducted during the course of the piloting effort to assist in resolving key issues which the design of a full-scale desalination plant would face.

1.4.1 Finished Water Compatibility Study

Finished water compatibility studies were conducted to verify the capability of the desalinated product water to be rendered compatible with water being conveyed in the EBMUD Mokelumne Aqueduct and the CCWD Multipurpose Pipeline. Post-treatment is necessary to enable compatibility between water produced by the proposed desalination plant, water in the transmission systems, and the integrity of the transmission systems. Several post-treatment options are available, such as lime in combination with carbon dioxide, which will result in suitable alkalinity, pH and appropriate corrosion indices. The study determined that treated water when mixed with existing supplies could be rendered compatible with the either transmission system with lime and carbon dioxide.

1.4.2 Biological Sampling

Biological sampling activities were initiated in November 2008 and continued into October 2009, well beyond the end of the pilot study, so that an extensive array of data could be obtained from varying seasonal periods. Samples of water from the slough and from the pilot intake were obtained to gain an understanding of species population and variability throughout these seasons and to identify the impact of the intake screens on aquatic life found in this Mallard Slough. Biological samples were collected during November and December of 2008 and February, March, July, and October of 2009.

The species composition of larval fishes collected during entrainment and source water sampling was consistent with published life history information for species found in Suisun Bay. The estimated small annual loss of adult prickly sculpin and bluegill/redear sunfishes is unlikely to affect adult populations. Entrainment of longfin/delta smelts occurred during the sensitive fish period of January through June when these larvae are normally present in the vicinity of Mallard Slough. Entrainment of these listed species at a full-scale desalination facility would require Endangered Species Act consultation with USFWS and CDFG for delta smelt and CDFG for longfin smelt.

1.4.3 Concentrate Toxicity Study

Concentrate toxicity studies were conducted using RO concentrate produced by the pilot plant to evaluate its toxicity to a variety of selected test organisms locally representing fish, crustaceans, and plant life. Two tests were conducted with RO concentrate samples collected during November 2008 (high salinity period) and February 2009 (low salinity period). Bioassay results showed no significant effects on survival or growth of any of the species tested. Assuming that concentrate samples tested in this study are representative of those produced by an operational desalination plant at Mallard Slough, there would be no expected toxic effects of the concentrate on biota were the concentrate to be discharged into the Delta.

1.5 Full Scale Facility

For the purposes of this project, it is assumed that a full-scale desalination plant would take advantage of the CCWD water right at Mallard Slough.

The finished water production of a full scale facility will vary depending on maximum and average TDS conditions. To define potential TDS conditions, the last three years of Water Year Hydrologic Classification Index data compiled by DWR were analyzed. For the San Joaquin and Sacramento Rivers, these years have been classified as either dry or critically dry year types. Conductivity data obtained from a DWR monitoring station in Suisun Bay, along with source water data obtained during the pilot study, are used to identify anticipated feed water conditions for the proposed desalination facility.

Recovery and treated water production values for the two alternatives under consideration for this analysis are shown in Table 1-4. Both Alternatives were developed based on the piloted pressurized UF pretreatment system. Recovery is defined as the percentage of feed water which is recovered as treated water and available for distribution to customers. Alternative No. 1 generally offers higher recovery because the first stage brackish membranes perform quite well for the anticipated water quality and because a high percentage of the first stage concentrate is recovered by the second stage. The NF and seawater membranes which in Alternative No. 2 have somewhat lower feedwater recovery and only a portion of the NF concentrate can be recovered. This has a direct bearing on the unit cost of water.

Table 1-4: Estimated Recovery and Treated Water Production

Parameter	Salinity Condition	Alternative No. 1	Alternative No. 2
Recovery:			
Total System[2]	Max. TDS Conditions[1]	68%	57%
Total System[2]	Avg. TDS Conditions[1]	79%	77%
RO & NF only	Max. TDS Conditions[1]	70%	58%
RO & NF only	Avg. TDS Conditions[1]	81%	79%
Initial Capacity:			
Feed Water		25 mgd	25 mgd
Treated Water (permeate)	Max. TDS Conditions[1]	17.0 mgd	14.1 mgd
Treated Water (permeate)	Avg. TDS Conditions[1]	19.8 mgd	19.2 mgd

Note 1. Maximum TDS during historical dry years ranges between 11,500 mg/L to 12,000 mg/L, with occasional peaks reaching as high as 15,000 mg/L. Average TDS during historical dry years ranges between 500 mg/L and 5,500 mg/L.

Note 2. Total system includes 88% average MF/UF recovery.

For the initial 25 mgd feed condition, Alternative No. 1 is able to produce 19.8 mgd total treated water at average conditions, but this declines to 17.0 mgd during the highest TDS period. For Alternative No. 2, the production rate drops to 14.1 mgd.

A potential layout for Alternative No. 1 is provided in Figure 1-6. Alternative No. 2 would be configured in a very similar arrangement, requiring a slightly larger RO Building to accommodate additional skidded desalination equipment. A total of approximately 7 acres is necessary for the arrangement illustrated in Figure 1-6. The acreage does not include a buffer zone around the desalination facility, which may be desirable in the final design. Due to groundwater and subsurface conditions in the vicinity near Mallard Slough, all tanks, clearwells, and other water-holding structures have been placed above-ground.

Treatment processes included for each alternative are:

Intake and Source Water Pumping Station which will draw source water and feed the self cleaning screens and MF/UF system at the proper flow rate and pressure.

MF/UF System which will remove large particles, suspended materials, algae and large molecules from the source water and create a filtrate suitable for RO desalination. The MF/UF system includes filtrate storage tanks.

RO System which will remove dissolved salts and small organic molecules from the filtrate. The RO system includes cartridge filters, energy recovery, permeate storage and chemical cleaning systems.

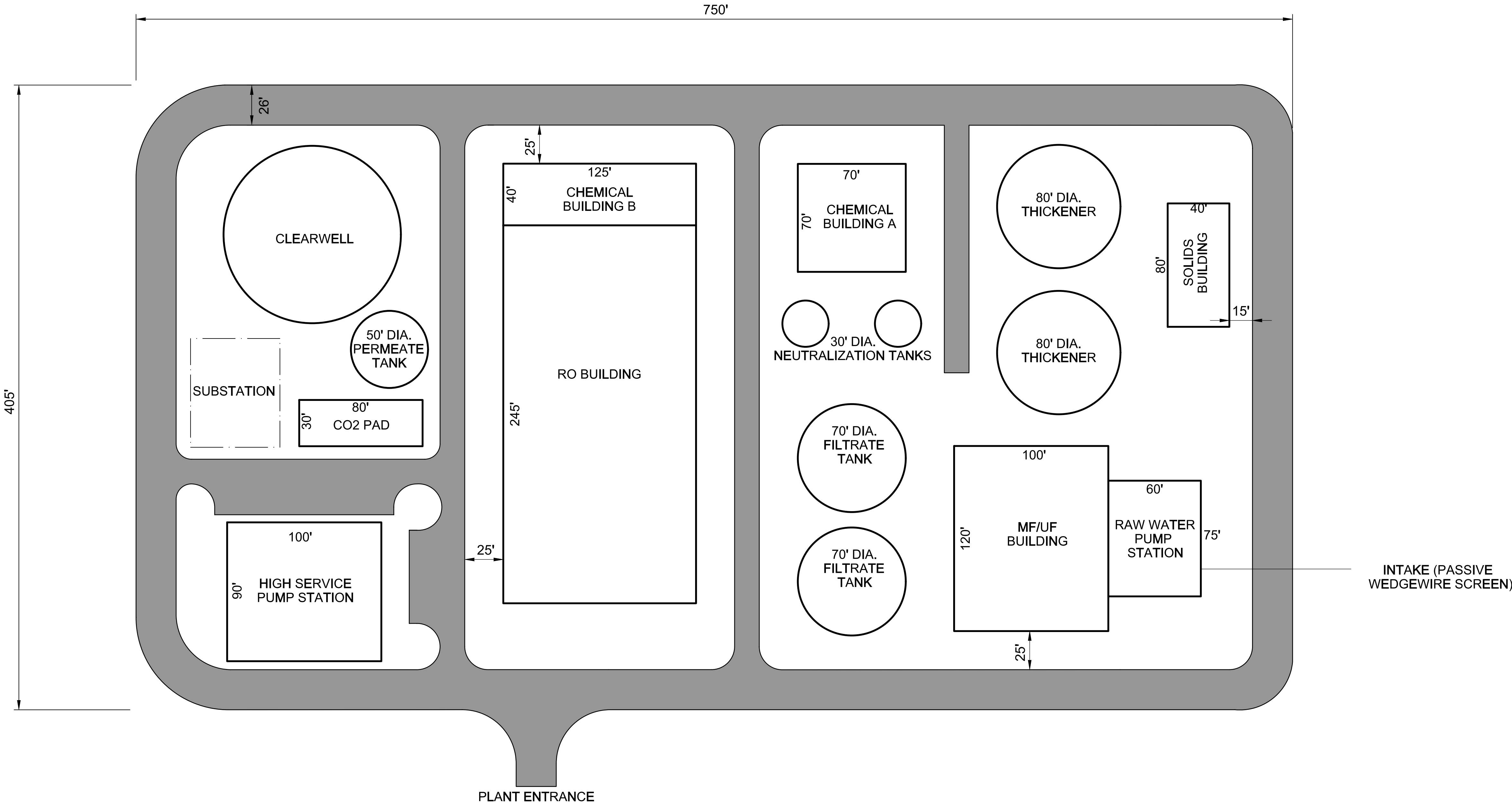
Energy Recovery in the form of a pressure exchanger mounted to each RO skid for capturing of waste energy present in RO concentrate.

Clearwell and High Service Pump Station which will store and convey treated water. The clearwell also provides a location for disinfecting the treated water.

Solids Handling System consisting of thickeners, centrifuges and sludge pumping equipment will receive backwash generated by the MF/UF system and separate solids for final disposal in a landfill or other appropriate location. Decant water from the thickeners (2 to 3 mgd) can potentially be recycled to the head of the plant, while centrate will need to be discharged to the sanitary sewer or other suitable alternatives.

Ancillary Systems include chemical storage and pumping, and common neutralization tanks for use by both the MF/UF and RO systems. Space will be required on-site for an electrical substation to serve the desalination power requirements.

It is assumed that operations functions will be incorporated into the RO Building.



File: site plan v2.dwg Model: Model ColorTable: bw.ctb DesignScript: MWH_Iplot_Pentable_v85.pen PlotScale: 80:1 User: Jennifer A Gelmini Plot Date: 01-DEC-2009 10:21

REV	DATE	BY	DESCRIPTION	SCALE 1"=40'	WARNING 0 ½ 1 IF THIS BAR DOES NOT MEASURE 1" THEN DRAWING IS NOT TO SCALE	DESIGNED C. BROMLEY DRAWN B. BEHNKE CHECKED G. BEHNKE	MWH AMERICAS, INC. 2121 N. CALIFORNIA BLVD WALNUT CREEK, CALIFORNIA 94596	MWH	BAY AREA REGIONAL DESALINATION PROJECT	FIGURE 1-6 CONCEPT SITE PLAN	SHEET

1.6 Unit Water Cost Estimate

Capital and annual operating costs were estimated based present worth analysis for the project. Alternative No. 1 offers a significant advantage with respect to both capital and annual costs over Alternative 2. It uses less desalination equipment and requires lower energy per unit of permeate to meet overall treated water production. Although the NF membranes furnished with Alternative No. 2 are energy efficient and require low operating pressures, this advantage is minimized when the parallel seawater membranes are utilized during high salinity periods as necessary to maintain treated water chloride goals. Based on historical dry year data, Alternative No. 2 actually presents a higher power requirement when it is compared to Alternative No. 1.

For developing the cost estimate for the proposed desalination facility, four scenarios were developed based on the two alternatives. Three implementation scenarios were developed using Alternative No. 1. Scenario 4 was developed using Alternative 2, i.e. using the NF and SW treatment train. The estimated costs for all four scenarios are presented in Table 1-5.

In ***Scenario No. 1***, the new desalination facility will be constructed near the Mallard Slough and will use the existing Mallard Slough Pump Station and intake to supply feed water. CCWD's existing source water transmission main will be converted for the conveyance of treated water from the site to the Multipurpose Pipeline. Structures include a new source water pump station, MF/UF building, RO building, chemical buildings (2), clearwell, high service pump station, filtrate tanks, neutralization tanks, thickeners, and solids handling building. Structures are assumed to be placed on pile foundations due to the poor subsurface conditions in this area. Scenario 1 unit water cost ($525/acre-feet) represents the water cost for a continuously operating plant.

Scenario No. 2 is similar to Scenario No. 1, with the exception that the desalination plant will be operated every third year, with minimal maintenance assumed to be performed during non-operational periods. It is assumed that in Scenario 2, the plant will be "moth-balled" during the wet years and will not produce any water ($0.5M was assumed for O&M costs during the wet years). Scenario 2 unit water cost ($1,020/acre-feet) represents the water cost for a plant operating only during droughts, i.e. 1 in 3 years on average.

In ***Scenario No. 3***, the new desalination facility will be constructed at an undetermined location away from the Mallard Slough and will require a new source water pump station and intake for supply of feed water. New structures also include an MF/UF building, RO building, chemical buildings (2), clearwell, high service pump station, filtrate tanks, neutralization tanks, thickeners, and centrifuge building. Structures will not require pile foundations since it is assumed that the site will not be subject to poor subsurface conditions characteristic of land parcels closer to the bay or to the Delta. Scenario 3 unit water cost ($540/acre-feet) represents the water cost for a continuously operating plant located in East Contra Costa but not use Mallard Slough.

In ***Scenario 4***, is based on Alternative No. 2 and is very similar to Scenario No 3: an undetermined location will be found away from the Mallard Slough and which will not require pile foundations. Scenario 4 unit water cost ($650/acre-feet) represents the water cost for a continuously operating plant in East Contra Costa but not use Mallard Slough and use Alternative 2 treatment process.

Table 1-5: Cost Summary

Cost Item	Scenario #1	Scenario #2	Scenario #3	Scenario #4
Estimated Capital Costs	$98,400,000	$98,400,000	$106,200,000	$114,400,000
Contingencies (20%)	$19,700,000	$19,700,000	$21,240,000	$22,880,000
Planning, Permitting, Engineering & Administrative Costs (25%)	$29,500,000	$29,500,000	$31,860,000	$34,320,000
Land Acquisition	$3,500,000	$3,500,000	$3,500,000	$3,500,000
Concentrate Discharge Permit	$1,000,000	$1,000,000	$1,000,000	$1,000,000
Total Capital Costs	**$152,100,000**	**$152,100,000**	**$163,800,000**	**$176,100,000**
Annual Costs	**$10,450,000** [1]	**$10,450,000** [1]	**$10,450,000** [1]	**$13,150,000**
Water produced (acre-feet) during the planning period	680,000	227,000	680,000	664,000
Present Worth Analysis				
Net Present Worth of Annual Costs - Continuous operation	$204,900,000		$204,900,000	$257,800,000
Net Present Worth of Annual Costs - Operation every 3 years		$79,000,000		
Total Present Worth Value	**$357,000,000**	**$231,100,000**	**$368,600,000**	**$433,900,000**
Net Present Worth per Acre-Foot				
O&M Only	$300	$350	$300	$390
Capital + O&M	$525	$1,020	$540	$650
Annual Worth Analysis				
Annual Worth of Capital and O&M Costs	$18,210,000/yr	$11,790,000/yr	$18,810,000/yr	$22,140,000/yr
Unit Cost of Water, Based on Annual Worth (Year 1), per Acre-Foot	$800	$1,560	$830	$1,000

Note 1: Annual cost during dry year operation. A dry year is assumed to occur once every three years.

The following assumptions were made to develop the cost estimates presented in Table 1-5.

1. All scenarios are based on 20 mgd production (25 mgd feedwater intake) during average TDS conditions from Table 1-4. Except for Scenario 2, the other scenarios assume continuous operation for all years. Scenario 2 assumes drought year operations, once every 3 years.
2. The estimated costs do not reflect fully developed cost of water to each agency. Water rights costs to supply water to a facility with no pre-existing water rights are not included in this estimate. The water rights costs along with other costs such as transmission, use of existing facilities, wheeling, and additional treatment (if needed) need to be negotiated as part of the inter-agency agreement to evaluate the full cost of water to each agency.
3. Annual operating costs include power, chemicals, periodic membrane replacement, labor, concentrate disposal, and disposal of dewatered solids into a landfill.
4. $1.0M was assumed for RO concentrate discharge permitting fees and $1.0M for upgrading unused pipelines to convey the RO concentrate to the discharge facilities.
5. Preliminary discussions regarding facility information and availability have been conducted with Mirant, Shell, and wastewater agencies. Costs were not discussed.
6. Power cost is $0.10/kwh.
7. Land cost is estimated to be $3.5M for 10 acres.
8. The present worth calculations are based on 30 year project life, 5.0% interest, 2.0% inflation rate, and the total amount of water produced over the 30 year project life.
9. All costs are in 2009 dollars.
10. This estimate is consistent with a Class 5 estimate as defined by the Association for the Advancement of Cost Estimation.

1.7 Conclusions and Recommendations

The following conclusions and recommendations should be considered by the four agencies as the BARDP moves beyond piloting and into a subsequent preliminary design phase:

1. Piloting at Mallard Slough provided data to suggest that a full-scale facility is viable in this location.

2. Piloting demonstrated that both types of pre-treatment systems can produce a suitable filtrate quality. The submerged UF system fouled during Run No. 2, and the membranes needed to be replaced for Run No. 3. The pressurized UF system operated within acceptable operational and cleaning parameters at 88% recovery and more than 40 gallons per square feet per day (gfd). Other pretreatment systems might provide higher recoveries at much lower flux and may be considered further during detailed design phases of this project. After the full-scale site is selected, any pre-treatment membrane will require additional pilot-scale activities to determine appropriate flux and recovery parameters, with the possible exception of Norit pressurized membranes installed on Mallard Slough feedwater.

3. A two-stage desalination process, consisting of brackish RO membranes in the first stage and seawater RO membranes in the second stage, was demonstrated in the pilot study to meet treated water quality goals with high recovery throughout the wide range of salinity variation expected for this project. Single stage seawater and NF RO membranes each provide acceptable operation as well, especially during certain periods depending on the salinity, i.e., seawater membranes in high salinity conditions and NF membranes during low salinity conditions. The former exhibits lower recovery and higher feed pressures, while the latter exhibits lower recovery and unacceptable water quality during higher TDS periods.

4. The full-scale evaluation relied on a Norit pressurized pretreatment membrane. There were two desalination alternatives evaluated due to the complex feedwater. Based upon the present worth evaluation, the design of a full-scale facility should be based on RO Train No. 1, a two-stage system with brackish water RO (BWRO) membranes in the first stage and seawater RO (SWRO) membranes in the second stage.

5. Several opportunities for managing desalination concentrate are available in the east Contra Costa region. The opportunity for mixing the concentrate with wastewater effluent produced by DDSD and/or the Central Contra Costa Sanitary District (CCCSD) may be explored. A variety of permits, including an NPDES permit, would be required for discharge of concentrate directly into Suisun Bay. Comingling with spent cooling water from the Mirant power plant, which is located east of the Mallard Slough Pump Station, or discharge into the power plant's intake itself, are both low cost and potentially acceptable options which the four agencies might consider. Operational status of the power plant, as well as its continued use of once-through cooling, are important and should be investigated.

   Another key environmental concern is the availability, at each potential desalination facility site, for a new source water intake. The Mallard Slough site, for example, offers an existing surface water intake of 40 mgd capacity which is already owned and operated by CCWD. Other surface or subsurface intake types as described in Technical Memorandum No. 2A would likely be considered if the site were to shift from Mallard Slough to another location.

6. Two methods for post-treatment stabilization were evaluated at the bench-scale using pilot plant permeate: liquid lime with carbon dioxide, and continuous flow through calcite bed filters. Both methods required sodium hydroxide to reach a suitable pH for the transmission systems. Both methods tested produced a stable product water which could be blended with EBMUD aqueduct water and CCWD multi-purpose pipeline water.

7. Site selection is very important for proper equipment sizing and selection. Source water obtained from Suisun Bay is subject to wide salinity variations, with lower maximum TDS values encountered further up the Sacramento River delta. This will have a direct bearing on total project cost and on operating costs. Water quality, project economics and technical application will likely vary at another site.

For this segment of the Suisun Bay, a maximum TDS design range between 11,500 mg/L and 12,000 mg/L is recommended. During historical dry years, the plant would normally operate between 500 mg/L and 5,500 mg/L. However, it is important to recognize that this pilot did not study water retrieved directly from the Suisun Bay but from the Mallard Slough, which is a dead-end, tidally influenced, narrow water body. The pilot intake was located at the end of the slough approximately 3,000-feet from the bay itself. The slough may have impacted water quality either by dampening the true impact of salinity variations induced by the bay, or by adding an organic carbon or algae load which is not present in the bay. While a new desalination facility designed at the Mallard Slough may be based on these pilot results, using sufficient safety factors for flux, coagulant feed, etc. is recommended. Additional pilot-scale activities are recommended if the final desalination process differs from what has been piloted or if the intake for the full-scale site is located elsewhere.

8. Discussions with all the regulatory agencies with jurisdiction over the proposed project should begin early in the site selection and preliminary design process to ensure the facility addresses regulatory concerns and to minimize rework associated with potential mitigation measures.

9. The four agencies have successfully brought this project forward from an initial concept and feasibility stage though piloting and identification of practical and viable technical solutions for removing salt from proposed source waters in the Suisun Bay water. As a greater level of detail is created, as environmental impact documents are developed, and as a firm site is identified, it will be useful for the four agencies to establish the necessary formal agreements for defining roles, responsibilities, and obligations among each agency. Key issues will eventually need to be resolved, such as ownership for the proposed desalination facility, intake and treated water conveyance facilities; operational responsibilities; and the transfer of treated water and other interagency water trades necessary to assure equity to each party.

2.0 INTRODUCTION AND BACKGROUND

The BARDP will address the regional water reliability needs (emergency, drought, planned outages, and/or long-term water supply needs) of the four regional agencies. The project is intended to minimize potential adverse environmental impacts associated with the construction of separate desalination plants in close proximity to one another and could also provide substantial cost savings to the four agencies by pooling resources and sharing costs. The proposed joint ownership, operation, and management of a single desalination facility will serve the needs of four major water providers and is a unique concept without precedent in California.

In order to properly evaluate the operations of a full-scale desalination plant located in an estuarine environment, a pilot plant study (PPS) was conducted. The pilot plant was located at CCWD's Mallard Slough Pumping Plant (MSPS) site near Pittsburg, California, adjacent to the Estuary at Suisun Bay, as identified in Figure 2-1. The pilot project allowed for data collection and evaluation of potential technologies that may be used in a full-scale plant to be located in the San Francisco Bay Area. Data obtained from the pilot plant study will also benefit others considering desalination in an estuarine environment.

Water diverted from the slough underwent treatment first by a self cleaning screen and pretreatment membrane system followed by passage through an RO or NF membrane. RO and NF permeate was discharged into an existing CCWD untreated water pipeline, and concentrate/reject streams were discharged into a nearby sanitary sewer.

As a general set of project goals, the PPS was conducted to:

- Minimize adverse environmental effects to aquatic organisms from the intake of source water,
- Confirm requirements for pre-screening upstream of the pretreatment systems,
- Make recommendations for a preferred pretreatment method,
- Test pretreatment residuals to evaluate disposal options,
- Test technologies and methods to maximize the efficiency of the plant (pretreatment and RO/NF configuration),
- Identify and test concentrate toxicity levels,
- Identify potential impacts of concentrate discharges, and
- Test treated water quality.

Figure 2-1: Pilot Plant Location

2.1 Study Objectives

The project consisted of a series of tasks: complete the pilot study, perform supplemental field work, and develop a full-scale treatment configuration, including a life cycle cost analysis. These objectives are discussed below.

2.1.1 UF System Performance

Two UF systems were tested to evaluate the effectiveness and cost benefits of UF membrane filtration pretreatment on RO productivity. The two systems represented different arrangements (pressurized vs. submerged), material types (polyvinylidene fluoride vs. polyethersulfone), flow direction (inside-out vs. outside-in) and backwashing strategies.

Each system was operated for the full pilot period, utilizing a coagulant as appropriate and recommended by the system manufacturer, and utilizing various flux and recovery rates to determine:

- water productivity,
- water quality,
- capital and operating costs, and
- cleaning efficiency and frequency.

2.1.2 RO and NF Performance

Using the combined filtrate from the pretreatment systems, three parallel systems were used to test NF and RO elements under varied conditions. RO Train No. 1 was a two stage system with brackish membranes in Stage 1 and seawater membranes in Stage 2, shown in Figure 2-2. RO Train No. 2 was a single stage RO with seawater membranes. The NF train, called NF Train No. 3, consisted of a single stage NF.

Figure 2-2: RO Train No. 1

All three trains operated in parallel. Operating pressure and flux were recorded to enable comparisons between energy requirements and flux. A range of flux and recovery were evaluated to identify:

- salt rejection,
- operating pressures,
- water productivity,
- water quality,
- chemical requirements,
- capital and operating costs, and
- cleaning efficiency and frequency.

2.1.3 Source Water Biological Impacts

Impingement and entrainment associated with the proposed pilot plant intake were studied by obtaining entrainment and source water samples during several seasons over a period of one year (six entrainment and four source water sampling events). Numbers of each species entrained into the intake system during operation of the pilot plant and full scale plant entrainment predictions are discussed in Section 5.2 of this report.

2.1.4 Treated Water Compatibility

A series of bench scale tests on the permeate produced by RO Train No. 1 was used to evaluate the compatibility of treated water with existing supplies in the EBMUD Mokelumne Aqueduct and the CCWD Multipurpose Pipeline. Post-treatment chemical requirements for pH and alkalinity adjustment were identified and recommendations were developed for the full-scale desalination plant. The details of this analysis are discussed in Section 5.1 of this report.

2.1.5 Concentrate Toxicity

One of the major potential issues associated with full-scale desalination operations is the discharge of the RO/NF concentrate, backwash and concentrate streams. Toxicity of the PPS concentrate was identified by initial testing. Sensitive species were determined and follow-up testing for both salinity and contaminant toxicity on those species was performed over a wide range of concentrations of the concentrate streams (0%, 2.5%, 5%, 10%, 25%, 50%, and 100%). Dry-season conditions represent highest ambient salinities, whereas wet-season conditions represent highest contaminant concentrations associated with storm runoff. The details of this analysis are discussed in Section 5.3 of this report.

2.1.6 Full Scale Treatment Application

Based on pilot results, a potential full-scale treatment configuration has been developed, complete with cost evaluations for two specific alternatives and four scenarios. Capital and annual average operating costs have been developed for implementation of the project in two stages:

- Initial conditions: With the water rights which CCWD currently maintains on the Mallard Slough, the proposed desalination plant would be initially capable of receiving up to 25 mgd of feed water. Treated water production would depend on overall system recovery and actual salinity, but during average salinity conditions, it is anticipated to be approximately 19 mgd. Water rights issues and restrictions due to potential fisheries impacts, however, will need to be evaluated and may limit feed water to less than 25 mgd

- Ultimate conditions: Previous reports state that the proposed regional desalination plant would need to produce 71 mgd of treated water. Expansion of intake capacity at Mallard Slough would require an increase in available water rights. The full-scale treatment application is discussed further in Section 6.0.

2.1.7 Marine Life Exclusion System Evaluation

An evaluation of potential screens, including aquatic filter barriers such as the Marine Life Exclusion System (MLES) as developed and manufactured by Gunderboom, Inc. was conducted and is included within Technical Memorandum No. 2A (**Appendix A**). Based on the results of the evaluation, and with DWR's concurrence, existing screens at the MSPS site were used for piloting. Using existing screens minimize potential adverse impacts associated with new intake construction.

2.2 Treated Water Quality Goals

To properly evaluate the RO technology, it is necessary to broadly consider the treated water quality which the BARDP will need to meet at full-scale. Table 2-1 summarizes the water quality goals for an RO system treating northern California Bay-Delta water.

Table 2-1: Desalination Water Quality Goals

Parameter	Water Quality Target
Disinfection	***Comply with Surface Water Treatment Rule (SWTR)***
Virus removal and inactivation	4 – 6 log reduction
Giardia removal and inactivation	3 – 5 log reduction
Cryptosporidium removal and inactivation	2 – 4 log reduction
Permeate Water Quality	***Meet all State and Federal Maximum Contaminant Levels (MCLs)***
Total Dissolved Solids (TDS)	< 500 mg/L
Chloride (selected target level)	< 100 mg/L
Bromide	<0.25 – 0.7 mg/L
Boron	<0.5 – 1.0 mg/L

Disinfection By-Products[1]	***Result in Stage 1 and Stage 2 Disinfection By-product (DBP) Rule Compliance***
Total Trihalomethanes (TTHM)	< 64 µg/L
Halo-Acetic Acids (HAA5)	< 48 µg/L

Note 1: TTHM and HHA5 values were selected based on a maximum of 80% of EPA drinking water requirements.

2.3 Summary of Work

The pilot plant study consisted of planning, developing, fabricating, installing, and operating all membrane pilot systems including feed water equipment, drainage equipment, and electrical power systems. Other facilities necessary to support and maintain pilot operations include preliminary screening, chemical feed equipment, tanks and pumps, pipelines, lighting and temporary structures.

The pilot plant is shown in the photograph included as Figure 2-3. A complete description can be found in Section 3 herein.

Figure 2-3: Pilot Plant Site

2.3.1 Pilot schedule

The pilot plant was operational from October 2008 through April 2009. Assessment of the UF, RO and NF membranes occurred in a series of three runs between November 2008 and April 2009, as described below. For each run, set points (the operation goals), were established and followed. Each run and set point is described in the following subsections.
Data collection for the assessment of source water biological impacts and concentrate toxicity occurred throughout pilot plant operations. Data collection for treated water compatibility occurred during Run 2.

2.3.1.1 Run 1: Baseline

Table 2-2 below is a summary of the set points for Run 1. Run 1 was a baseline run where fluxes and recoveries were based on manufacturer recommendations and Technical Memorandum No. 4A. This run occurred during dry weather conditions with feed conductivities ranging from 10-18 mS/cm. As discussed in Section 4.1.4, this can be converted to a TDS of approximately 6,000 – 11,000 mg/L based on a TDS to conductivity ratio of 0.61.

Table 2-2: Run 1 Set Points

Duration	Nov 6th 2009 - Dec 17th 2008
Submerged UF	32 gfd
Pressurized UF	40 gfd
RO Train No. 1 Flux	12 gfd
RO Train No. 1 Recovery	70%
RO Train No. 2 Flux	12.7 gfd
RO Train No. 2 Recovery	50%
RO Train No. 3 Flux	13.2 gfd
RO Train No. 3 Recovery	50%

2.3.1.2 Run 2: Mid-Challenge Run

After the conclusion of Run 1, the performance of each membrane system was evaluated and new, more challenging set points were established, as summarized in Table 2-3. Water quality during this run was similar to Run 1, with feed conductivity ranging from 8 – 18 mS/cm (approximately 5,000 – 11,000 mg/L TDS).

Table 2-3: Run 2 Set Points

Duration	Jan 8th 2009 - Feb 11th 2009
Submerged UF	32 gfd
Pressurized UF	44 gdf
RO Train No. 1 Flux	12 gfd
RO Train No. 1 Recovery	74%
RO Train No. 2 Flux	14.1 gfd
RO Train No. 2 Recovery	50%
RO Train No. 3 Flux	12.9 gfd
RO Train No. 3 Recovery	55%

2.3.1.3 Run 3: Challenge Run

In Run 3, fluxes and recoveries were increased to the maximums allowed by manufacturers without exceeding design warnings. Water quality was more representative of wet weather conditions during this run. Feed conductivities ranged from 0.5 – 5 mS/cm (approximately 300 – 3000 mg/L TDS).

Table 2-4: Run 3 Set Points

Duration	March 17th 2009 - Apr 23rd 2009
Submerged UF	41 gfd
Pressurized UF	55 gfd
RO Train No. 1 Flux	12 gfd
RO Train No. 1 Recovery	82%
RO Train No. 2 Flux	14.1 gfd
RO Train No. 2 Recovery	62%
RO Train No. 3 Flux	12.9 gfd
RO Train No. 3 Recovery	60%

2.3.2 Technical memoranda description

Initial studies were conducted during pilot development to:

- identify and resolve key issues associated with pilot operations,
- develop and gain feedback from the staff from the four agencies regarding the approach for conducting special studies during the course of the pilot period,
- summarize key permitting issues to be addressed before conducting the pilot study,
- evaluate various membrane technologies potentially suitable for pilot evaluation and testing,
- consider various intake configurations potentially applicable for use in the full-scale treatment facility.

All technical memoranda (TM) are provided in **Appendix A**.

2.3.2.1 TM No. 1A Discharge Characterization

TM 1A was developed to document the anticipated nature of the pilot plant discharge and the options for disposal, as the first step in resolving potential permitting and discharge constraints. The TM provides an estimate of quantity, chemicals added, and doses for normal pilot plant operations and cleaning discharges.

Three options were identified for disposal of the pilot waters, including CCWD's source water line, Delta Diablo Sewer District's (DDSD's) sewer line, and a hybrid approach in which permeate is discharged to CCWD and all other flows discharged to DDSD. For each option, flowrates, dilutions and TDS impacts were calculated during normal and worst-case scenarios.

Based on this evaluation, and after consultation with staff from the four agencies, it was decided that CCWD would receive the pilot plant permeate, while DDSD would receive all other waste discharges.

2.3.2.2 TM No. 1B Concentrate Toxicity Test Plan

One goal of the pilot study was to develop an understanding of the potential toxicity of the RO concentrate relative to species found in surrounding waters. TM 1B was prepared to document the proposed concentrate toxicity testing plan. Toxicity testing was designed to focus on the environmental extremes that were encountered during the pilot testing period in the dry season with higher ambient salinity and in the wet season with higher contaminant concentrations associated with storm water runoff.

The concentrate toxicity testing involved survival and growth testing of three estuarine test organisms (plant, crustacean, and vertebrae), with a determination of whether toxicity is due to salinity or potential contaminants. These species were selected for this study as being representative of aquatic organisms within the Delta. Based on test organism mortality, follow-up testing was to be conducted for the most sensitive species for both salinity and potential contaminant toxicity. Each test involved the blending of concentrate from the pilot plant with Lab Water Control to create various concentrate dilutions. QA/QC and reporting procedures are also defined in the TM.

2.3.2.3 TM No. 1C Environmental Assessment and Permitting Findings

Based upon TM 1A, it was decided that CCWD would receive the pilot plant permeate, while DDSD would receive all other flows including the cleaning wastes. The permitting requirements for the pilot plant project were revisited, and conclusions are documented in TM 1C.

Because no pilot discharges were being sent back to surface waters, and all of the activity was being conducted on an existing CCWD site, it was determined that all work associated with the projects could be covered by existing permits. A categorical Exemption was filed by the member agencies under CEQA for conducting the pilot test. No specific pilot study permits were issued.

2.3.2.4 TM No. 2A Intake Desktop Study

TM 2A was developed as a preliminary desktop evaluation of current intake screen technologies and current practices at local and global facilities. Screen technologies that were identified for potential use in the full scale plant include the following:

- Surface Water Intakes
  - Behavioral Barrier Systems
  - Velocity Caps
  - Passive Screens
  - Aquatic Filter Barriers
  - Fish Barrier Nets
  - Fish Mesh Stationary Screens
  - Traveling Water Screens

- Subsurface Intakes
  - Infiltration Galleries
  - Seabed Filtration Systems
  - Horizontal Collector Wells
  - Horizontal Directional Drilled Well Field
  - Slant Wells
  - Conventional Vertical Wells
  - Porous Dike

For the pilot study, it was recommended that the existing MSPS intake screens be utilized to draw water from the Mallard Slough and for diversion to pilot equipment.

A workshop was held with DWR early in the project in which the pros and cons of aquatic filter barrier testing were presented. Using the currently permitted MSPS intake was recommended and DWR subsequently approved the change.

2.3.2.5 TM No. 3A Feedwater Characterization

TM 3A was a pre-pilot characterization of water quality at Mallard Slough based on available data from ten years of CCWD sampling for some metals, salts, and general water quality indicators. Historical hourly conductivity and temperature data were also obtained from a water quality monitoring station located in Suisun Bay near the City of Pittsburg maintained by the California Department of Water Resources (DWR). The CCWD data showed that Mallard Slough TDS varied from approximately 150 to 7,000 mg/L (10^{th} to 95^{th} percentile of the data set), whereas the DWR data indicated that local TDS varied from approximately 150 to 10,600 mg/L.

The available water quality data did not show a complete representation of Mallard Slough water. Most of the available data was taken in the wet season to determine when the MSPS could supply source water to CCWD when TDS and chloride was low. Therefore, the data represented a specific season of the year. Furthermore, the water quality parameters normally tested do not cover the range of metals, salts and other parameters that are used to design an RO system. Therefore, additional water quality sampling was conducted in advance of the pilot study (December 2008), and the data is summarized in the TM 3A supplement. The additional water quality testing included grab samples at high tide and low tide for additional RO design parameters. Additionally, continuous data was collected for two days by a probe that recorded temperature, conductivity, pH, and oxidation reduction potential (ORP). Water quality exhibited a very slight correlation to tidal patterns, and salinity was relatively consistent with the nearby DWR station.

2.3.2.6 TM No. 3B Pretreatment Technology Evaluation

The purpose of TM 3B was to evaluate pretreatment technologies for their suitability to the pilot project. A questionnaire was sent to four established polymeric MF/UF membrane suppliers: Pall Corporation; Norit Americas, Inc.; GE/Zenon Water Process Technologies; and Siemens/Memcor Water Technologies Corporation. Ceramic membranes from NGK Insulators, Ltd. were also evaluated.

TM 3B includes a description of each technology and responses to questionnaires sent to each vendor to determine whether their system could meet our project requirements. The ceramic membrane system vendor was unable to furnish a pilot-scale system within the project schedule. However, all four of the polymeric membrane systems were suitable for the pilot study.

The conclusion of the TM was that Norit Seaguard and GE/Zenon ZeeWeed 1000 would be selected for the pilot study. However, subsequent discussions with all of the vendors led to several changes in the pretreatment technologies for the project. Norit reduced their efforts to get the Seaguard technology certified by the California Department Public Health (DPH) for drinking water, and so the project was built with Norit Aquaflex (SXL membranes in a vertical configuration). Siemens/Memcor decided to build a new S10V pilot unit for this project and was willing to provide it within the project schedule and at a lower pilot cost, so their technology was used in lieu of GE/Zenon ZeeWeed 1000.

2.3.2.7 TM No. 4A Reverse Osmosis Technology Evaluation

TM 4A documented the criteria and assumptions for the pilot-scale RO design. Treated water quality goals were defined, including salinity, disinfection, disinfection by-products (DBPs), and boron. Given the unusual feedwater quality and the anticipated range of feedwater salinity, other case studies with seasonal and tidal variations were considered. Four RO membrane suppliers were contacted and provided with feedwater quality data. A review of the supplier RO performance projections and discussions between project staff and supplier staff led to a recommendation to utilize two high pressure RO trains.

One train was recommended to be a two-stage system with brackish water membranes in the first stage and seawater membranes in the second stage, so that the benefits of a high recovery system could be evaluated in the pilot. In this TM, the second train was also initially recommended to be a two stage arrangement with seawater membranes only; however, it was eventually determined that both the second and third trains should be single stage. The membrane selections were further refined in TM 4B.

2.3.2.8 TM No. 4B Nanofiltration Technology Evaluation

Nanofiltration membrane technology was evaluated for implementation at the pilot scale. Performance projections for two separate NF elements were developed. As anticipated, the NF computerized performance models projected a lower operating pressure but a lower water recovery in order to meet the permeate salinity goals.

As a result of this evaluation, a third desalination train was added to the project, containing a single stage of NF elements. At the same time, the seawater membrane RO train identified in TM 4A was reduced to a single stage because of anticipated high periodic TDS in the Mallard Slough source water and resulting energy impacts.

3.0 PILOT PLANT DESCRIPTION

The pilot plant site was located at CCWD's existing MSPS in Pittsburg, California, and was fed with water from Mallard Slough. The pilot plant utilized the existing fish screen to minimize biological impacts to the feedwater. The pilot treatment process included a 100-µm self cleaning screen, followed by two parallel UF units. Combined filtrate was fed into three parallel RO/NF systems. Permeate was stabilized and pumped into the MSPS source water supply pipeline. Concentrate and other pilot plant wastes were collected and pumped to the nearby DDSD sewer.

Process chemicals included sodium hypochlorite, aqueous ammonia, ferric chloride (for pressurized UF only), sodium bisulfite, antiscalant, and caustic soda (permeate stabilization only). In addition to the chemicals listed, acids were also used for pretreatment cleaning procedures on a daily basis and for CIPs on all systems.

Water quality sampling and operational monitoring were conducted at regular intervals at various points in the treatment processes. Sampling and monitoring locations, frequencies, procedures, and standard methods were prescribed in the Experimental Plan (**Appendix B**), and results were documented on a daily basis. Operational monitoring included flows, pressures, temperatures, chemical tank levels, and periodic procedures such as silt density index (SDI) testing and instrument verifications/calibrations. Water quality analyses were conducted onsite and at external laboratories.

Data were downloaded from the pilot plant on a weekly basis and were populated into spreadsheet databases, where they were automatically normalized and plotted. Automatic calculations allowed for weekly evaluation of operational parameters, such as specific flux and net pressure, to monitor each system's performance.

This section provides a summary of the pilot plant site, the treatment processes, sampling and monitoring, data analysis, and QA/QC procedures.

3.1 Site Description

The Pilot Site was located in Pittsburg, California at CCWD's MSPS, adjacent to the Estuary at Suisun Bay. The existing MSPS intake has physical capacity of 40 mgd with a fish screen that was built and approved by the resource agencies in 2002. The fish screen's mesh size of 3/32 and low intake approach velocities are designed to eliminate the impingement of juvenile and adult fishes and to minimize the entrainment of larval fish. The performance of the new screen has been continuously monitored during pumping operations since 2002.

The MSPS site itself is isolated, located on an unpaved access road approximately half a mile long. Between the nearest public road and the site there were four locked gates and four sets of railroad tracks. The isolation of the site required additional security measures including housing all equipment inside locked steel shipping containers, as shown in Figure 3-3.

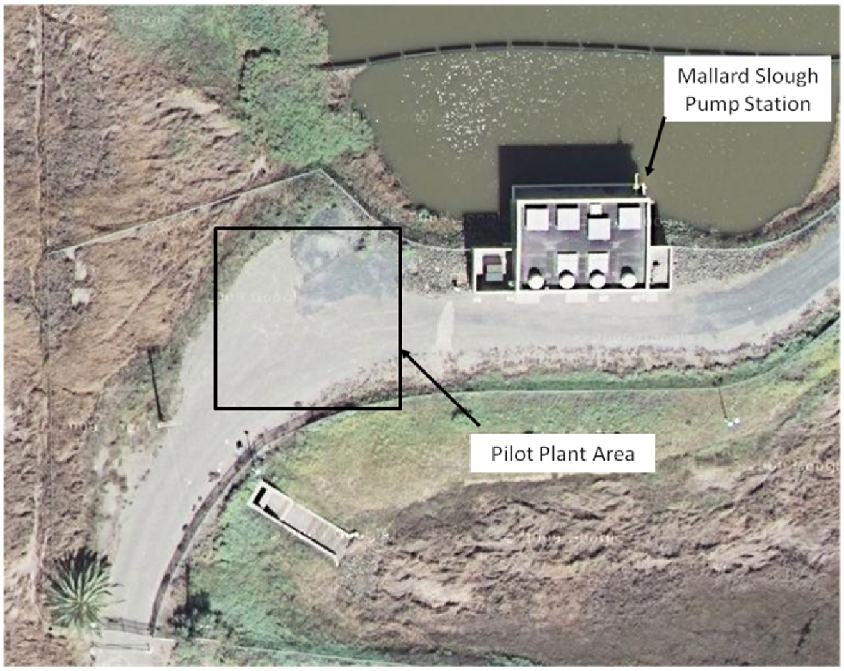


Figure 3-1: Pilot Plant Area

3.2 Pilot Plant Configuration

A schematic of the pilot plant is shown in Figure 3-2.

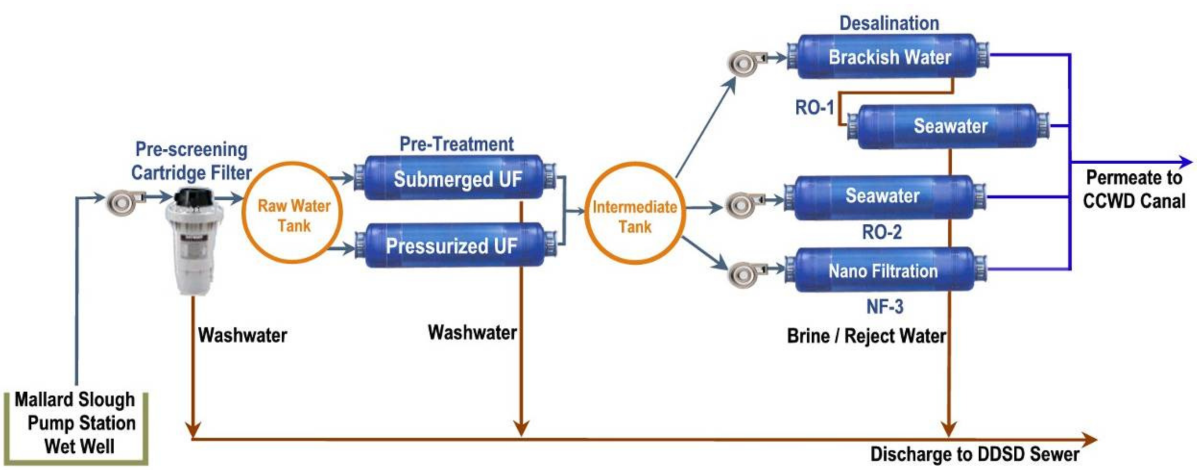


Figure 3-2: Pilot Plant Schematic

A submersible pump was installed in the MSPS wet well behind the existing fish screen, allowing the pilot plant to operate without using the Mallard Slough Pumps. Source water was pumped through a self-cleaning screen into the untreated water tank. Water was pumped from the untreated water tank to the two UF systems operating in parallel, and filtrate from those systems was sent to a combined filtrate tank. The filtrate tank provided water to three desalination membrane trains operating in parallel (two RO systems and one NF system). RO and NF permeate was sent to the permeate tank, which was pumped into CCWD's source water line that delivers Mallard Slough water from MSPS to the Contra Costa Canal and CCWD's untreated water system. Flushing waste from the self-cleaning screen, backwash waste from the UF membranes, concentrate from the RO and NF membranes, and tank overflows were sent to the waste tank, which was pumped into the DDSD sewer line.

The untreated water tank allowed for: untreated water balancing, contact time for chloramines, and a location for the untreated water quality testing. The filtrate tank allowed for storage of filtrate during the UF membrane backwashes and maintenance washes so that the RO systems would not have to be taken offline. The waste tank allowed for the neutralization of any cleaning chemicals (specifically during the cleans-in-place (CIPs)) so that no low or high pH water was discharged to the sewer. The permeate tank supplied clean water for the CIPs, permeate flushes for the RO and NF systems, and makeup water for chemicals that were diluted onsite.

Figure 3-3: View of Pilot Plant Site at Mallard Slough Pump Station.

3.2.1 Source Water Feed

Normal operations of the pilot system involved pumping source water from behind the existing 3/32-inch (2400 µm) MSPS fish screen at the south end of Mallard Slough. The submersible pump (Goulds WE2QH) delivered 85 gpm of slough water into the pilot facility. This source water pump was controlled by the level in the untreated water tank.

To protect the UF membranes from smaller debris and particles that passed through the fish screen, an inline 100 micron self-cleaning self cleaning screen was employed, shown in Figure 3-4. This type of pre-screen was required by all of the pretreatment membrane manufacturers contacted for this project to protect their membranes. The screen unit was model TAF750, manufactured by Amiad Filtration Systems. Motor driven backwashes were initiated by a predetermined time interval of 8 hours, a differential pressure of 7 psi, or by manual operator intervention. The duration of each backwash ranged from 15 to 30 seconds, with spent backwash water and trapped particles discharged ultimately to the DDSD sanitary sewer. A brief description of screen equipment is presented in Table 3-1.

Figure 3-4: Amiad Self Cleaning Screen

Table 3-1: Self Cleaning Screen Equipment

Parameter	Value
Manufacturer and unit	Amiad Super TAF-750
Screen size, micron	100
Flow rate, gpm max	110
Working pressure, psi	30 to 120
Minimum backpressure for backwash, psi	22
Filter area, sq. inches	110
Inlet & Outlet diameter, inches	2
Backwash time, seconds	16
Backwash flow, gpm	25
Backwash water use, gal	6.7

3.2.2 Membrane Pretreatment

The RO and NF processes require a high quality feed water to minimize fouling, maximize membrane life and operate efficiently. The principle objective of pretreatment is to reduce the concentration of fouling constituents in the feed water to a level that will produce long-term stable performance that prolongs the lifespan of the membranes.

The pilot study evaluated two UF membrane systems. One unit, a submerged membrane system, was manufactured and provided by Siemens/Memcor, pictured in Figure 3-5. The second unit, a pressurized membrane system, was provided by Layne Christensen Company and contained membranes manufactured by Norit Americas, Inc, pictured in Figure 3-6.

The terms submerged and pressurized are used throughout this report to differentiate between the two UF systems. However, all results and recommendations are for these specific systems as defined below and do not represent generalized results for either type of UF membrane arrangement. In addition, any discrepancies in the performance between the two membranes are not simply a result of the membrane arrangement (submerged vs. pressurized). There are a number of variables that differ between the two systems, including membrane material, flow direction through the lumen, and backwash flowrates.

Figure 3-5: Submerged UF System

Water was pumped from the untreated water tank to the two UF systems in parallel. The single pump (Goulds NPE/2ST) operated constantly, and both UF PLCs had a means of starting and stopping flow to its system (the submerged system utilized a valve; the pressurized system utilized an on-skid feed water tank in conjunction with an on-skid pump). If both systems were not calling for feedwater, the UF pump would shut down until water was required.

The submerged membrane system was a vacuum driven UF system that operated in an outside-in flow pattern. Polyvinylidene fluoride (PVDF) membrane modules were vertically immersed directly into an open process tank and connected to permeate collection headers and aeration hoses. Permeate pumps applied a slight vacuum to the end of each membrane fiber. The The submerged system did not require any coagulant. Cleaning procedures included an air-liquid backwash every 25-30 minutes, a maintenance wash every 0.5-1 days, and CIPs as discussed in Section 4.2.1.2.

In the submerged system, dummy modules were used to reach a desired flux at a given flow rate to minimize filtrate tank overflow. For example, in Run 3, two working modules and two dummy modules were used to achieve an average flux of 41 gfd and a flow rate of 17 gpm. Additional working modules would have increased the flow to more than was required for the RO and NF systems. Overflow from the filtrate tank would have been sent to waste tank and ultimately to the sewer, unnecessarily increasing the environmental impact of the pilot plant as well as increasing sewer fees. Using dummy modules avoided those unnecessary costs.

Figure 3-6: Pressurized UF System

The pressurized system was a pressure driven inside-outside UF system with polyethersulfone (PES) hollow fiber membranes housed in 8-inch diameter pressure vessels assembled vertically on the skids. Filtration was dead-end in the pressure vessels. The pressurized system required the addition of a coagulant upstream of the membranes. Flows up to approximately 33 gpm were achieved. Cleaning procedures included a liquid-only backwash every 25-30 minutes, a chemically enhanced backwash every 0.5-1 days, and periodic CIPs as needed.

Table 3-2 provides a summary comparison of the two UF systems.

Table 3-2: Membrane UF Summary

Manufacturer	**Siemens/Memcor**	**Layne Christensen/Norit**
Product Name	S10V	SXL-225
Technology	UF membrane	UF membrane
Configuration	Submerged	Pressurized
Flow Direction	Outside-In	Inside-Out
Membrane Fiber Material	PVDF	PES
Terminal TMP, psi (max)	12	15
Fiber Dimensions, µm (OD/ID)	800/500	800 (ID only)
Pore size, µm (nominal)	0.04	0.025
Clean Water Permeability, gfd/psi	10 to 12	Not field tested
Dimensions Diameter, inches Length, inches Membrane Area, sf, all modules	 5.2 46.7 900 (3 modules) 600 (2 modules)	 8 60 860 (2 modules)
Membrane Type	Hollow Fiber	Hollow Fiber
Backwash Mechanism	Combined air & liquid	Liquid Only
Coagulant type	Not required	Ferric chloride
Coagulant Dose, mg/L	N/A	5

3.2.3 Membrane Desalination

Filtered water from the UF systems was pumped with low-lift pumps (Goulds NPE/2ST) from the filtrate storage tank and processed through 5-micron cartridge filters. Cartridge filters are commonly utilized as additional protection for the RO and NF membrane elements to capture any final particles of suspended solids that may enter the feed stream. From the cartridge filters, the filtered water was pumped into three different desalination trains, each with its own high-pressure pump with VFD.

During project design, historic water quality near the site demonstrated large seasonal variations in feed salinity (100 – 13,000 mg/L of TDS). This wide salinity range presented a challenge in selecting a desalination technology for this site. Coarse classifications of desalination technology are shown in Table 3-3, and there was no single technology that would be efficient in all seasons. Therefore, three different membrane technologies were used in order to determine which type of membrane system was most applicable for the site.

Table 3-3: Membrane Classifications

Membrane Classification	Log Salt Removal	Typical Feed Water Salinity to Achieve Efficiency
Nanofiltration	<1	<3,000 mg/L TDS
Brackish water	1-2	1,000 – 10,000 mg/L TDS
Seawater	>2	> 10,000 mg/L TDS

Selection of the RO membrane configurations is described in **Appendix A** (TM No. 4A *RO Technology Evaluation* and TM No. 4B *NF Technology Evaluation*).

3.2.3.1 Factors Affecting RO Performance

In reverse osmosis system, one indicator of performance is the specific flux, or permeability, of the membrane. Permeability is flow normalized by membrane surface area (flux) and also normalized by net pressure, and it is expressed in units of gallons per square foot per day per psi (gfd/psi). There are a number of factors that can affect permeability.

Osmotic Pressure: Given constant flux and recovery, as the RO systems were controlled, an increase in osmotic pressure (e.g. feed salinity) will cause the VFD to ramp up, increasing feed pressure to maintain the flow setpoints. This in turn will increase the net pressure, and decrease permeability. Likewise, a decrease in osmotic pressure will increase permeability.

Temperature: As temperatures decrease during winter months the water becomes more viscous, requiring additional pressure to move water through the semi-permeable membrane. In order to properly analyze data from the RO and NF systems, temperature fluctuations must be accounted for. All comparison data is normalized for temperature. This is discussed further in Section 3.4.

Fouling: Buildup of organic matter on the spacers and membrane surface can decrease cross-sectional flow area through the vessel and increase the feed-to-concentrate differential pressure. It can also physically block passage through the membrane. In order to maintain the flow setpoints, the feed pressure would increase, thus decreasing the permeability. This can be minimized by ensuring that there is no biological activity in the UF process or filtrate storage. The pilot plant was dosed continuously with chloramines, and bisulfite was added to the filtrate upstream of RO to prevent oxidation damage.

Scaling: Deposition of mineral scale onto the surface of the membrane can have the same effects as fouling. This can be minimized with the use of an antiscalant.

Recovery: Increased water recovery can affect performance in several ways. Raising recovery will drive up the operating pressure to push more water through the membrane. Because less flow is going to concentrate, the tail elements experience lower crossflow velocities combined with higher salt concentrations, and are more subject to material deposits. And finally, higher operating pressure can compress foulant deposits, making them more difficult to remove. Several different recovery setpoints were tested for each RO system, from lowest to highest.

In the case of scaling or fouling, chemical cleaning procedures can generally restore performance unless there is irreversible buildup. A high pH CIP targets organics, while a low pH CIP targets inorganics. CIPs are a routine part of normal membrane maintenance for pilot studies and at the full scale.

Three desalination membrane trains were designed, constructed, and pilot tested. Each is described in the following subsections. As noted in Section 6.2.1, the performance of the three desalination trains which were piloted during this study cannot be directly compared because of differences in staging and number of elements per vessel.

3.2.3.2 RO1 – 2:1 Brackish: Seawater

The RO1 system was a 2:1 array with brackish water membranes in the first stage (Dow BW30-4040) and seawater membranes in the second stage (Dow SW30HRLE-4040). Figure 3-7 shows a schematic of a 2:1 array. Brackish water membranes were determined to be most suitable for the anticipated feedwater quality and a second stage was used to increase recovery. Seawater membranes were selected for Stage 2 because of their higher rejection characteristics since the salt concentration in the Stage 2 feed was higher than the Stage 1 feed. Additional characteristic of the membrane elements can be found in Table 3-4.

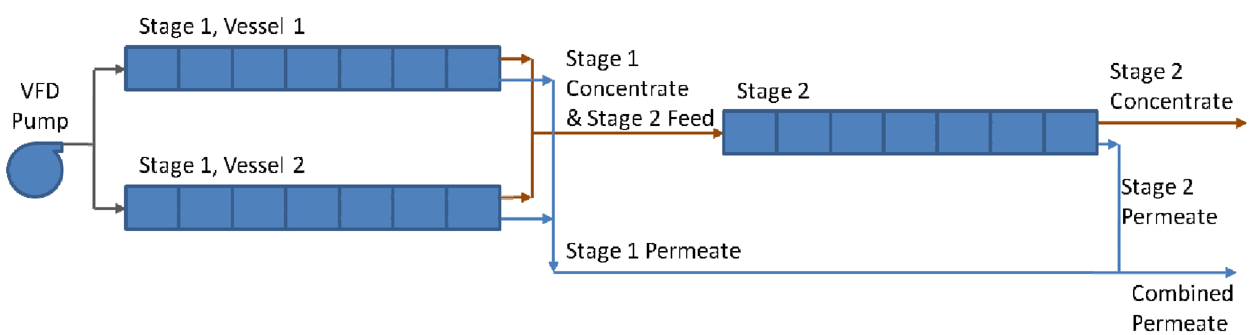


Figure 3-7: RO1 Schematic

3.2.3.3 RO2 – Seawater System

The RO2 system was a single vessel of seawater membranes (Dow SW30HRLE-4040). For full scale design, additional stages could be considered in order to enhance recovery, although at greater capital and operating cost. Figure 3-8 shows a schematic of the single vessel design of RO2 and NF3 systems. Seawater elements were tested because they were potentially suitable when the feedwater salinity was at its highest, but it was also desirable to understand how the seawater elements would perform given low salinity feedwater in the wet season. Table 3-4 summarizes additional element characteristics. Figure 3-9 is a photograph of the RO trailer, and the RO2 vessel is the top vessel. The piping and pump that can be seen are actually associated with NF3 but are similar to RO2.

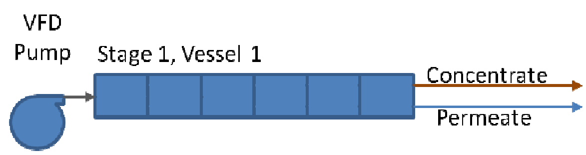


Figure 3-8: RO2 and NF3 Schematic

3.2.3.4 NF3 – Nanofiltration System

The NF3 system was a single vessel of nanofiltration membranes (Dow NF90-4040). Figure 3-8 shows a schematic of the single vessel design of RO2 and NF3 systems. Nanofiltration elements were tested because they were potentially suitable when the feedwater salinity was at its lowest, but it was also desirable to understand how the nanofiltration elements would perform given high salinity feedwater in the wet season. Table 3-4 summarizes additional element characteristics. Figure 3-9 is a photograph of the RO trailer, and the NF3 vessel is the fourth vessel down from the top. The piping and pump that can be seen are also associated with NF3.

Table 3-4: RO Skid Summary

	RO System No. 1	RO System No. 2	NF System No. 3
Vendor	Dow/Filmtec	Dow/Filmtec	Dow/Filmtec
First Stage			
Membrane Type (Manufacturer Model)	Brackish (BW30-4040)	Low pressure seawater (SW30HRLE-4040)	Nanofiltration (NF90-4040)
Array	2:1	Single	Single
Number of Vessels	3	1	1
Elements per Vessel	7	6	6
(Surface Area, sf)	78	85	82
Size, inch	4	4	4
Antiscalant Dose, mg/L	1.5	1.5	1.5
Second Stage			
Membrane Type (mfg model) [1]	Low pressure seawater (SW30HRLE-4040)	N/A	N/A
Element Number	7	--	--
(Surface Area, sf)	85	--	--
Size, inch	4	--	--
Antiscalant Dose, mg/L	None	--	--

Figure 3-9: Photograph of RO Trailer

3.2.3.5 RO System Controls

Each RO system had a dedicated feed pump with VFD, as well as motorized control valves and inline flow meters. Flow setpoints were changed manually in the PLC, and the VFD and control valves were automatically adjusted to meet the flow setpoints. In this manner, flux and recovery were held constant. Changes in osmotic pressure (e.g. feedwater salinity) were accommodated with changes in the pump speed (e.g. feed pressure) to maintain all the flow setpoints.

In RO1, Stage 1 recovery was maintained with a flow control valve on the Stage 1 permeate piping. Although this arrangement presents a small risk of back-pressuring the Stage 1 membranes, the flow control was strategically placed here to minimize headloss on the concentrate stream that feeds Stage 2. Recovery through Stage 2 was controlled with a flow control valve on the Stage 2 concentrate piping. This is illustrated in Figure 3-10.

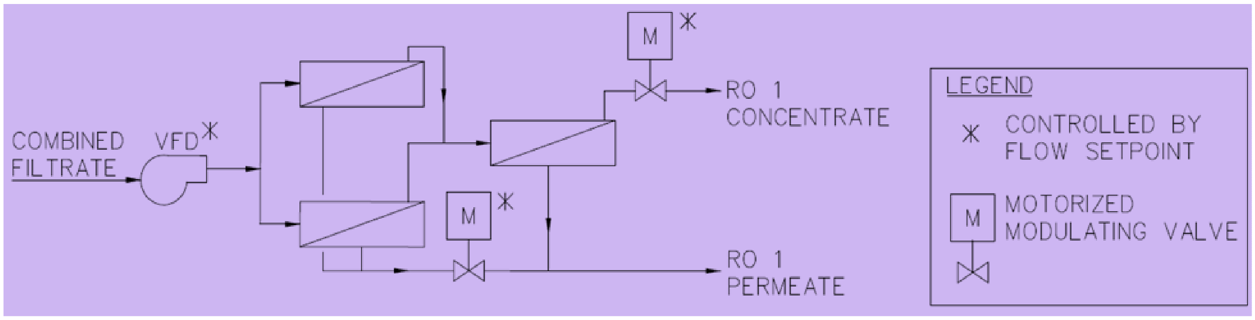


Figure 3-10: Control Schematic for RO1

In RO2 and NF3, recovery was controlled by defining a concentrate flow rate which was maintained via a flow control valve on the concentrate piping. This is illustrated in Figure 3-11

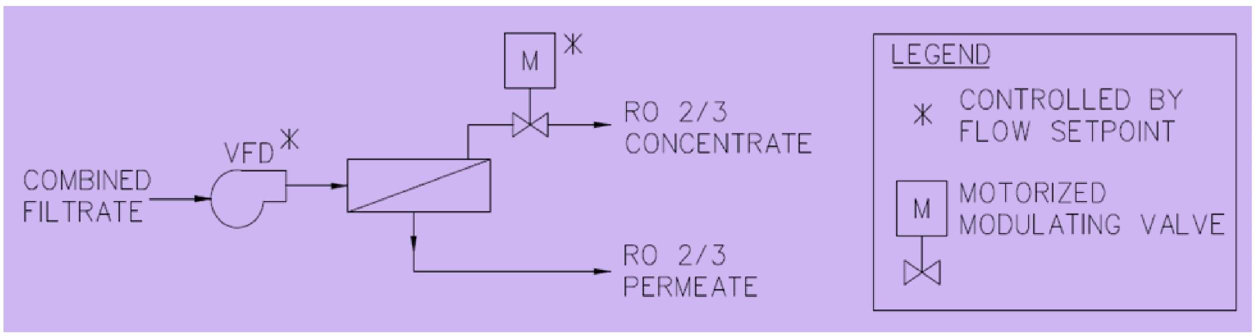


Figure 3-11: Control Schematic for RO2 and NF3

3.2.4 Ancillary Systems

In addition to the treatment technologies discussed above, there were a number of other systems that were integral to the operation of the pilot plant. Chemicals were used to adjust pH, to restrict microbial growth, to enhance treatment performance and to clean the systems. Permeate and waste discharge systems removed the final products from the pilot site.

3.2.4.1 Chemical Systems

There were a number of steps within the desalination process where chemicals were introduced to enhance performance or to protect the membranes. Some of the treatment processes required chemical addition to adjust water quality parameters to meet treatment goals, optimize performance and costs, and maintain process equipment. Process chemicals were continuously applied at various locations in the pilot plant. Chemical application points are shown in Figure 3-12 and doses are summarized in the subsections below.

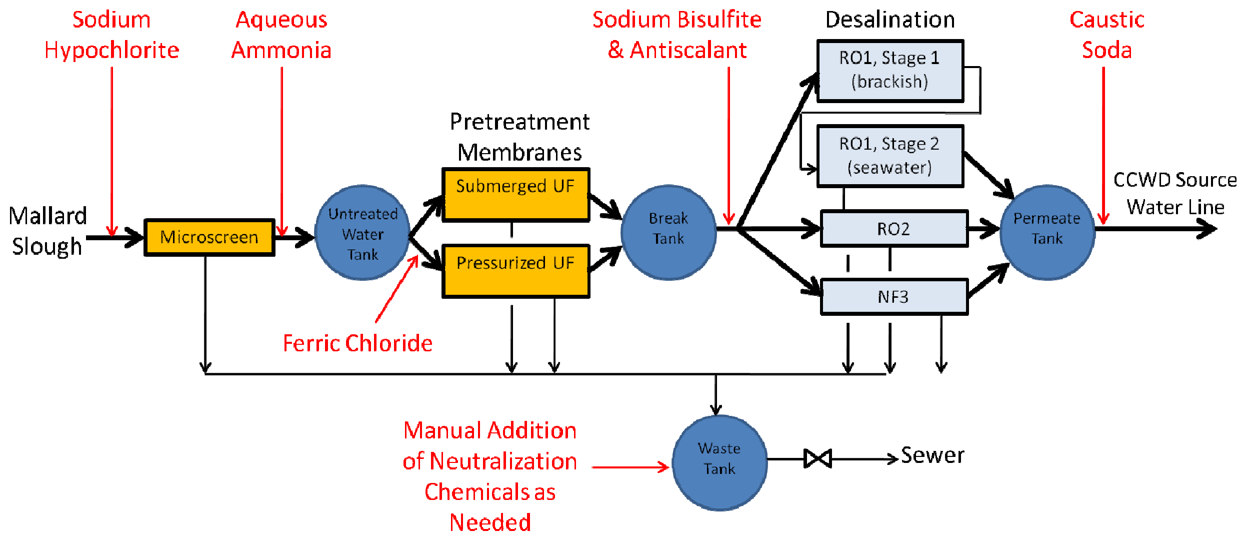


Figure 3-12: Schematic of Chemical Injection Points

Disinfection: Sodium hypochlorite was injected upstream of the Amiad TAF screen and was used to minimize biological growth on the self-cleaning filter, untreated water tank, UF membranes and filtrate tank. The dose was based on goal of 0.15 mg/L free chlorine upstream of the UF Membranes. Aqueous ammonia was added at a 4:1 ratio (chlorine to ammonia) downstream of the Amiad TAF to form chloramines, a chemical disinfectant that is less harmful to the RO membranes than free chlorine.

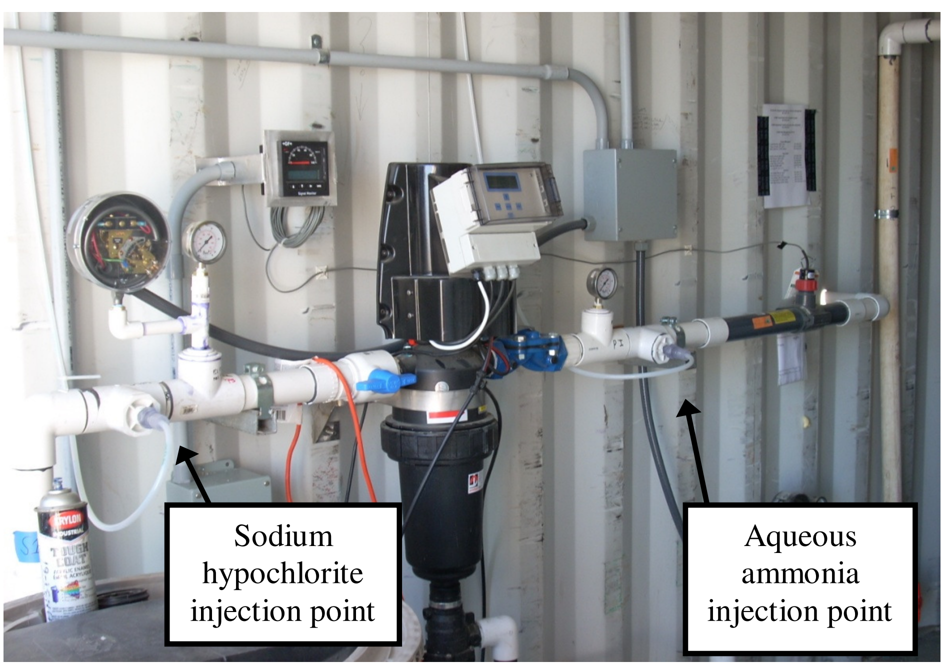


Figure 3-13: Disinfection Chemical Injection Points

Coagulation: Ferric chloride was used as a coagulant for the pressurized membranes and was added directly into the on-skid feed tank, in a location where the feedwater entering the tank would mix the chemical. A coagulant assessment performed during the pilot study determined that 5 mg/L was the optimum dose for this source water.

Sodium Bisulfite: Any free chlorine that was in the system after it was discharged from the filtrate tank was quenched with sodium bisulfite before being sent to the RO membranes, as chlorine and other strong oxidants can damage RO membranes. There was an inline ORP analyzer downstream of sodium bisulfite injection for monitoring, and the bisulfite dose was adjusted manually to keep ORP under 200 mV.

Antiscalant: Antiscalant was added to each membrane system to reduce the precipitation of sparingly soluble salts and other solids on the RO and NF membranes. The use of antiscalant increases the lifetime of the membranes and allows for longer periods between cleanings. The antiscalant used was Nalco's PermaTreat PC-1850T, and it was dosed at 1.5 mg/L.

Treated Water pH: Permeate from RO and NF systems generally has a depressed pH. Carbon dioxide as a dissolved gas can pass readily through the membranes while other aqueous species such as bicarbonate are rejected. When bicarbonate is removed from the system, a portion of the carbon dioxide gas will react to form carbonic acid, lowering the pH of the water. In order to ensure that the permeate does not negatively impact the CCWD source water pipeline, caustic soda was added to the permeate discharge line to raise the pH to 7.

Cleaning: A variety of chemicals were used for the MW/CEB of the UF membranes and for the periodic cleans-in-place (CIP) of the UF, RO and NF membranes. For UF daily cleans, chemical tanks and metering pumps supplied the required chemical doses to the backwash flow. For CIPs, operators made cleaning solutions in dedicated CIP tanks using filtrate or permeate for dilution water as indicated in Section 4.0. Table 3-5 summarizes the chemicals used in each type of cleaning.

Table 3-5: Summary of Cleaning Chemicals

	Siemens	Norit	RO1	RO2	NF3
MW/CEB Low pH	Muriatic Acid	Citric Acid	N/A	N/A	N/A
High pH	Hypochlorite	Hypochlorite	N/A	N/A	N/A
Clean-in-place Low pH	Citric Acid	Citric Acid	Citric Acid and Muriatic Acid	Citric Acid and Muriatic Acid	Citric Acid and Muriatic Acid
High pH	Hypochlorite	Hypochlorite and Caustic Soda	Caustic Soda	Caustic Soda	Caustic Soda

After CIPs were completed and cleaning solutions were sent to the waste tank, the waste tank was neutralized with acids or basis as appropriate until the pH was in the range of 6-9.

3.2.4.2 Permeate Discharges

Permeate was collected onsite in the Permeate Tank. The permeate was used onsite to dilute chemicals and to provide permeate for the RO and NF permeate flushes and CIPs. However, most permeate was not used onsite and was sent to the CCWD source water supply line. As mentioned above, caustic soda was added to the permeate to bring the pH of the water to 7. Permeate was pumped using a high-pressure pump (Goulds NPE/2ST) into the pipeline, where static pressure was approximately 115 feet.

3.2.4.3 Waste Discharges

The waste streams produced from the desalination process included:

- Backwash from the self-cleaning filter (typically once per 8 hours).
- Backwash water from UF membranes (typically twice per hour per UF system).
- RO and NF process concentrate (continuous, and largest contribution to waste by volume).
- Tank overflows
- Chemical cleaning waste from UF daily cleans (typically once or twice per day per UF system). Because the chemicals were fairly dilute and the volume was a small portion of the waste tank, it was not necessary to neutralize the waste tank prior to discharge
- Chemical cleaning waste from the UF, RO and NF CIPs (at the end of each Run). Due to chemical strength and volume, the waste tank valve was closed during CIPs, and all waste was neutralized prior to discharge to the sewer.

3.3 Sampling and Monitoring

In order to properly evaluate the performance and operation of the pilot plant throughout the pilot study a number of samples were collected through on-skid SCADA systems, onsite sampling and offsite laboratories. Samples were used to define feed water quality, determine rejection characteristics of the membranes, and analyze fouling potential. Samples were taken at a number of locations throughout the plant. Sampling locations are defined in Table 3-6.

Table 3-6: Sampling Locations

Sample Identifier	Sample Location	Comments
S1	Pilot Plant Feed	Also utilized for biological sampling
S2	Self Cleaning Screen Feed	
S3	UF Feed, common	
S4A	UF Feed, Submerged	
S4B	UF Feed, Pressurized	
S5A	UF Filtrate, Submerged	
S5B	UF Filtrate, Pressurized	
S6A	UF Backwash, Submerged	
S6B	UF Backwash, Pressurized	
S7	UF Filtrate, Combined	
S8-S9	Not Used	

Sample Identifier	Sample Location	Comments
S10	RO1 Feed	
S11	RO1 Stage 1 Permeate, Vessel No. 1	This was moved to the Stage 1 permeate piping on January 9th, 2009, as the sample was not representative of the entire vessel.
S12	RO1 Stage 1 Permeate, Vessel No. 2	This sampling port was removed on January 9th, 2009, as the sample was not representative of the entire vessel.
S13	RO1 Stage 2 Permeate	This was moved from the end cap to the Stage 2 permeate piping on January 9th, 2009, as the sample was not representative of the entire vessel.
S14	RO1 Stage 1 Concentrate, No. 1	
S15	RO1 Stage 1 Concentrate, No. 2	
S16	RO1 Stage 2 Concentrate	Also utilized for concentrate toxicity study
S17	RO1 Stage 2 Feed	
S18	RO1 Permeate, Combined	Also used for treated water compatibility study
S19	Not Used	
S20	RO2 Feed	
S21	RO2 Permeate	
S22-23	Not Used	
S24	RO2 Concentrate	
S30	NF3 Feed	
S31	NF3 Permeate	
S32-33	Not Used	
S34	NF3 Concentrate	

3.3.1 Parameters

As listed in Table 3-7 and Table 3-8, a large number of constituents were evaluated throughout the pilot study. In Run 3, to save on cost and because many parameters were consistently not detected or had very consistent results, selected components were reduced in the sampling schedule.

Some of the most relevant parameters are discussed below. A more detailed discussion of water quality can be found in Section 4.0.

3.3.1.1 Turbidity and Conductivity

Feed turbidity and conductivity were the main indicators used to determine the water quality of the UF and RO/NF membrane product water, respectively. Feed and filtrate turbidity were measured continuously via online meters on each of the UF skids. These values were field verified using handheld turbidimeters on a daily to weekly basis.

Feed conductivity was measured continuously with an online meter in the untreated water tank and by inline conductivity meters on RO2 and NF3. Permeate conductivity was measured with inline conductivity meters on all RO permeate piping. Conductivity data was field verified daily using a handheld conductivity meter. These values were further verified with offsite lab samples on a weekly basis.

3.3.1.2 Filtrate Turbidity and SDI

UF integrity was measured directly and indirectly as a performance indicator. Indirect methods included monitoring filtrate turbidity and the 15-minute silt density index (SDI). Success criteria require the filtrate turbidity not to exceed 0.15 NTU and maintain a 24 hour average less than 0.10 NTU. Filtrate SDI was to be less than 3.

3.3.1.3 Chlorine

The untreated water feed going into the UF systems was analyzed daily for free and total chlorine to ensure a sufficient disinfection of the water before entering the UF systems. Because free chlorine damages RO membranes, free and total chlorine were tested downstream of sodium bisulfite injection, which acts to quench free chlorine.

3.3.1.4 Metals

A subset of metals was tested in an offsite lab on a weekly bases. These included aluminum, barium, boron, manganese and iron. Once per Run, a much more extensive list of metals was analyzed. The onsite and offsite sampling schedule is summarized in Table 3-7 and Table 3-8, respectively.

Table 3-7: Onsite Sampling

Parameter	Sample Points	Frequency
Conductivity	S2, S10, S11, S13, S16, S24, S34	Daily
	S12	Daily until Jan 9 (sampling point removed)
	S14, S15	Daily to weekly
	S3, S4b, S7, S17, S18, S20, S30	Weekly
	S21, S31	Daily to weekly
pH	S2, S3, S17, S21, S24, S31, S34	Daily
	S3, s16, S18, S20, S30	Weekly
Free Chlorine	S3, S7	Daily
Total Chlorine	S3, S7	Daily
Turbidity	S10	Daily
	S18, S20, S21, S30, S31	Daily to weekly
	S3, S4B, S5A, S5B	Daily to weekly
Temperature	S2, S3	Daily
	S3, S4A, S4B, S10, S20, S30	Weekly
SDI	S5A, S5B, S10	Twice per week

Table 3-8: Offsite Sampling

Parameter	Sampling Points	Frequency
Alkalinity as CaCO3, Bicarbonate Alkalinity, Carbonate Alkalinity	S1, S7, S16, S17, S18, S21, S24, S31, S34	Weekly
Total Hardness as CaCO3	S1, S7, S16, S17, S18, S21, S24, S31, S34	Weekly
Total Cation/Anions (Ca, Mg, K, Na, SO4, ionic balance)	S1, S7, S16, S17, S18, S21, S24, S31, S34	Weekly
Weekly Metals (Al, Ba, B, Mn) - Total	S1, S7, S16, S17, S18, S21, S24, S31, S34	Weekly
Weekly Metals (Al, Ba, B, Mn) - Dissolved	S1, S16, S24, S34	Weekly
Iron – Total	S1, S4B, S5A, S4B, S7, S16, S17, S18, S21, S24, S31, S34	Weekly
Iron – Dissolved	S1, S4B, S16, S24, S34	Weekly
Minerals (Bromide, Chloride, Cyanide, Fluoride, Silica)	S1, S7, S16, S17, S18, S21, S24, S31, S34	Weekly
TSS, TDS, Conductivity	S3, S10, S16, S17, S18, S20, S21, S24, S30, S31, S34	Weekly
Nitrite, Nitrate (as Nitrogen)	S1, S18, S21, S31	Weekly
Total Phosphorus, Orthophosphate	S1, S18, S21, S31	Twice per month
TOC	S1, S5A, S5B, S7, S13, S18, S21, S31	Twice per week
UV254	S1, S5A, S5B, S7	Weekly
Algae	S1, S3, S18, S21, S31	Monthly
MTBE, Perchlorate	S1, S18, S21, S31	Monthly
Metals Primary List (Sb, As, B, Ba, Be, Cd, Cr total, Co, Cu, Pb, Hg, Mo, Se, Ni, S, Tl, V, Zn) - Total	S1, S7, S16, S17, S18, S21, S24, S31, S34	Once per Run
Metals Primary List (Sb, As, B, Ba, Be, Cd, Cr total, Co, Cu, Pb, Hg, Mo, Se, Ni, S, Tl, V, Zn) - Dissolved	S1, S16, S24, S34	Once per Run

Parameter	Sampling Points	Frequency
Radionuclides (alpha and beta particles, radium 226, radium 228, strontium 90, tritium, uranium)	S1, S16, S24, S34	Once

3.3.1.5 Operational Parameters

Operational monitoring included flows, pressures, temperatures, chemical tank levels. The operational daily and periodic checklists also included procedures such as instrument verifications and calibrations, flow tests, PDTs, instrument flushing/cleaning procedures, and general inspections

3.3.2 Analytical Methods

The analytic methods for offsite sampling are listed by constituent in Table 3-9. Seawater samples were analyzed at CRG Marine Laboratory with the exceptions of algae (MWH Labs) and UV254 (EBMUD and CCWD). Permeate sampling was performed at EBMUD for the first two Runs. For Run 3, permeate sampling was performed at CCWD.

Table 3-9: Analytical Methods

Parameter	Lab	Method	Detection Limit
Total Alkalinity	CRG EBMUD CCWD	SM 2320 B SM 2320 B	1 mg/L 5 mg/L
Bicarbonate Alkalinity	CRG EBMUD CCWD	SM 4500 CO2 D SM 4500 CO2 D	1 mg/L 5 mg/L
Carbonate Alkalinity	CRG EBMUD CCWD	SM 4500 CO2 D SM 4500 CO2 D	1 mg/L 0.1 mg/L
Hardness	CRG EBMUD CCWD	SM 2340 B SM 2340 C	1 mg/L 2 mg/L
Calcium (total)	CRG EBMUD CCWD	EPA 200.8m EPA 200.7	0.05 mg/L 0.022 mg/L
Magnesium (total)	CRG EBMUD CCWD	EPA 200.8m EPA 200.7	0.01 mg/L 0.0094 mg/L
Potassium (total)	CRG EBMUD CCWD	EPA 200.8m EPA 200.7	5 mg/L 0.057 mg/L

Parameter	Lab	Method	Detection Limit
Sodium (total)	CRG EBMUD CCWD	EPA 200.8m EPA 200.7	5 mg/L 0.021 mg/L
Sulfate	CRG EBMUD CCWD	EPA 300.0 EPA 300.0	0.01 mg/L 0.013 mg/L
Aluminum (total and dissolved)	CRG EBMUD	EPA 1640m EPA 200.7	3 mg/L 0.021 mg/L
Iron (total and dissolved)	CRG EBMUD	EPA 1640m EPA 200.7	0.5 mg/L 0.011 mg/L
Manganese (total and dissolved)	CRG EBMUD	EPA 1640m EPA 200.7	0.01 mg/L 0.624 µg/L
Barium (total and dissolved)	CRG EBMUD	EPA 200.8m EPA 200.7	0.2 mg/L 0.00104 mg/L
Boron (total and dissolved)	CRG EBMUD	EPA 200.8m EPA 200.7	1 mg/L 0.0104 mg/L
Bromide	CRG EBMUD CCWD	ICP-MS EPA 300.0	0.001 mg/L 0.0027 mg/L
Chloride	CRG EBMUD CCWD	EPA 300.0 EPA 300.0	0.01 mg/L 0.24 mg/L
Cyanide	CRG EBMUD	SM 4500-CN E SM 4500-CN F	0.005 mg/L 0.01 mg/L
Fluoride	CRG EBMUD CCWD	SM 4500-F D EPA 300.0	0.01 mg/L 0.00087 mg/L
Silica	CRG EBMUD	SM 4500-Si D EPA 200.7	0.1 mg/L 0.034 mg/L
Total Suspended Solids	CRG EBMUD CCWS	SM 2540 D SM 2540 D	0.5 mg/L 1 mg/L
Total Dissolved Solids	CRG EBMUD CCWD	SM 2540 C SM 2540 C	0.1 mg/L 12 mg/L
Conductivity	CRG EBMUD CCWD	SM 2510 SM 2510	0.001 mS/cm 0.0003 mS/cm
Nitrite	CRG EBMUD CCWD	SM 4500-NO2 B EPA 300.0	0.01 mg/L 0.00053 mg/L
Nitrate	CRG EBMUD CCWD	SM 4500-NO3 E EPA 300.0	0.05 mg/L 0.0028 mg/L

Parameter	Lab	Method	Detection Limit
Total Phosphate	CRG EBMUD CCWD	SM 4500-P E EPA 300.0	0.016 mg/L 0.01 mg/L
Orthophosphate	CRG EBMUD CCWD	SM 4500-P E EPA 300.0	0.01 mg/L 0.0033 mg/L
Total Organic Carbon	CRG EBMUD CCWD	SM 5310 B SM 5310 D	0.1 mg/L 0.023 mg/L
UV254	EBMUD CCWD	SM 5910	0.006 abs
Algae	MWH Labs	SM 10200 F	1 per mL
MTBE	CRG	unknown	1.0 mg/L
Perchlorate	CRG	EPA 331.0	0.2 µg/L
pH	EBMUD	SM 4500-H+	Not listed
Alpha Particles	CRG	SM 7110 C	0.934 piCi/L
Beta Particles	CRG	900.0	7.72 piCi/L
Radium 226	CRG	903.0	0.373 piCi/L
Radium 228	CRG	Ra-05	0.253 piCi/L
Strontium 90	CRG	905.0	0.596 piCi/L
Tritium	CRG	906.0	408 piCi/L
Uranium	CRG	908.0	0.305 piCi/L

3.4 Data Analysis

Data collected from instruments at various points in the treatment process were stored by PLCs for each individual pilot unit. Data were downloaded weekly. In order to make operational decisions on a weekly basis, online data was populated into spreadsheet databases, where they were automatically normalized and plotted. Automatic calculations allowed for weekly evaluation of operational parameters such as specific flux and net pressure to monitor each system's performance.

UF, RO and NF operational data were normalized to remove the effects of various factors that influence operation and performance of the membranes but in fact do not reflect degradation of membrane performance. Normalization is important in order to discern the true performance degradation, since parameters such as temperature would otherwise obscure a change.

Data can be normalized to reference data (i.e. temperature normalized to 25^{o}C) or to other operational parameters at actual conditions (i.e. flux normalized to actual pressure to show specific flux). By normalizing data to a reference temperature, it allows a more accurate comparison of a single membrane system operating at different times with different temperature conditions. By normalizing flux data to pressure, it allows a more accurate comparison of fouling over time and across membrane systems operating in parallel.

3.4.1 UF Data Normalization

UF data is typically normalized to a standard temperature (UF operation is not dependent on salinity). A temperature correction factor is applied to UF flux data to normalize it to 20°C. The equation below is used to account for variations in water viscosity with temperature.

$$J_{tmn} = \frac{Q_p}{SA} * TCF \qquad (1)$$

Where:
J_{tmn} = temperature corrected flux (gfd)
Q_p = actual filtrate flow (gpm)
S.A. = membrane surface area (sq ft)
TCF = UF Temperature Correction Factor normalized to 20°C

$$TCF = e^{-0.0239*(T-20)} \qquad (2)$$

Where:
T = actual temperature (deg C)

Specific flux is a normalization of flux to actual feed pressure. Temperature corrected UF specific flux is calculated from the temperature corrected flux and the actual feed pressure.

$$J_{spn} = \frac{J_{tmn}}{P_{feed}} \qquad (3)$$

Where:
J_{spn} = temperature corrected specific flux (gfd/psi)
P_{feed} = feed pressure (psi)

Transmembrane pressure (TMP) is the pressure differential across the membrane feed filtrate.

$$TMP = P_{feed} - P_{filtrate} \qquad (4)$$

Where:
TMP = transmembrane pressure (psi)
$P_{filtrate}$ = filtrate pressure (psi)

Feed water recovery identifies how much filtrate is available to feed the RO and NF systems. It can be calculated to varied degrees of accuracy depending on the time interval of interest. For example instantaneous recovery during normal operation is 100 percent because there is no waste. Alternatively waste can include waste from backwashes (hour time interval), backwashes and maintenance washes (24 hour time interval), or backwashes, maintenance washes and CIPs (30 day time interval). For the purpose of this study, the time interval is 24 hours.

$$Recovery = \frac{Volume_{Filtrate}}{Volume_{Feed}} \tag{5}$$

Where:
$Volume_{Filtrate}$ = feed minus backwash and maintenance wash filtrate consumption
$Volume_{Feed}$ = feed minus downtime for backwash, maintenance wash and pressure decay test

3.4.2 RO and NF Data Normalization

RO data is normalized at 25°C, which is different from the standard UF temperature of 20°C. There are several methods of RO data normalization. ASTM D4516-00 equation normalizes for a reference flow and pressure as well as temperature shown as (6). Standard flow and pressure are typically the day one or clean membrane operation values. Equation (6) is only useful if comparing operation of one type of membrane over time or one type of membrane side by side. Operation trends generated by operating different membrane types cannot be compared using equation (6) because standard flow and pressure may be different between different types of membranes.

$$\frac{Q_{pn}}{Q_{pa}} = \frac{P_{nets}}{P_{neta}} * \frac{TCF_s}{TCF_a} \tag{6}$$

Where:
Q_{pn} = normalized permeate flow (gpm)
Q_{pa} = actual permeate flow (gpm)
P_{nets} = net standard pressure (psi)
P_{neta} = net actual pressure (psi)
TCF_s = RO Temperature Correction Factor at standard conditions (equals 1 using equation (7) at 25°C)
TCF_a = RO Temperature Correction Factor at actual conditions normalized to 25°C

Rather, to effectively compare two different membrane products, specific flux presented in this report was normalized to temperature only, using equations (7) and (8).

$$TCF_a = 0.001 * T^2 - 0.0819 * T + 2.4273 \tag{7}$$

RO flux normalized to temperature only is shown in equation (8).

$$J_{tmn} = \frac{Q_{pa}}{SA} * TCF_a \tag{8}$$

Where:
SA = actual surface area provided by the manufacturer

Like UF, specific flux is a normalization of flux to the actual feed pressure. Temperature corrected RO specific flux is calculated from the temperature corrected flux and the actual net pressure.

$$J_{spn} = \frac{J_{tmn}}{P_{neta}} \tag{9}$$

Actual net pressure is calculated using several pressure variables.

$$P_{neta} = P_f - 0.5DP - P_p - \Pi_{f-c} + \Pi_p \tag{10}$$

Where:
P_f = Feed Pressure (psi)
DP = Differential Pressure (psi)
P_p = Permeate Pressure (assumed = 2 psi) (psi)
$\Pi_{f\text{-}c}$ = Feed-Concentrate Osmotic Pressure (psi)
Π_p = Permeate Osmotic Pressure (assumed negligible because the permeate concentration is two orders of magnitude less than the feed-concentrate concentration) (psi)

$$DP = P_f - P_c \tag{11}$$

Where:
P_c = Concentrate Pressure (psi)

$$\Pi_{f-c} = C_{f-c}\left(\frac{mg}{L}\right) * \frac{11.6\ psi}{1000\frac{mg}{L}NaCl_{soln}} \tag{12}$$

Where:
$C_{f\text{-}c}$ = Feed- Concentrate Average Concentration of TDS (mg/L)

$$C_{f-c}\left(\frac{mg}{L}\right) = Feed_{Conductivity}\left(\frac{\mu S}{cm}\right) * 0.61 * IAF \tag{13}$$

Where:
IAF = Log Mean Average

$$IAF = \frac{\ln\left(\frac{1}{1-\%Recovery}\right)}{\%Recovery} \tag{14}$$

Where:
Recovery = Feed Water Recovery

$$\%Recovery = \left(1 - \frac{Q_{concentrate}}{Q_{feed}}\right) * 100\% \qquad (15)$$

Where:
$Q_{Concentrate}$ = Concentrate flow (gpm)
Q_{feed} = Feed flow (gpm)

Global rejection or Percent Removal is a calculation that determines the removal of a constituent based on the feed and permeate concentrations. Another method can be used which incorporates the feed-concentrate average values to define rejection.

$$Removal = 1 - \frac{C_p}{C_f} \qquad (16)$$

Where:
C_p = Permeate concentration
C_f = Feed concentration

3.5 Quality Control and Quality Assurance

QA/QC measures were implemented throughout the pilot operation. Refer to the Experimental Plan in **Appendix B** for more detail. Instrumentation and flow rates were regularly calibrated where possible and verified with calibrated instruments where possible. Several instruments such as pressure transmitters and flow meters could not be calibrated or directly verified due to plumbing or space restrictions which are inherent to pilot testing restraints.

The Experimental Plan laid out a schedule to calibrate temperature and pH on a weekly basis and turbidimeters and ORP meter on a monthly basis. In addition, turbidimeters were verified on a weekly basis, flow meters on the UF systems on a bimonthly basis and pressures transmitters on a monthly basis. There were significant issues with pressure transmitters on RO2 throughout the PPS.

4.0 PILOT RESULTS

The pilot plant was constructed in 2008 and operated for approximately six months in 2008 and 2009, spanning the annual shift between the two local water quality seasons.

Construction	September 2008
Shakedown/Startup	October 2008
Stable Operations	November 2008 through April 2009

During the stable operations, the systems ran steadily. Each process unit was independently controlled by a local programmable logic controller (PLC) with hardwired switches for automatic shutdown in prescribed circumstances. Short plant interruptions were caused by automatic shutdowns caused by equipment failure on any given skid, power outage, waste drainage problems, etc. Longer shutdowns were planned between runs for cleaning procedures. In March, the ambient salinity fell to less than 500 mg/L TDS due to normal seasonal changes, and the pilot plant was suspended for one month while CCWD ran its intake pumps for water supply purposes.

4.1 Feed Water Quality

The pilot plant was situated at the end of Mallard Slough, a narrow channel that is approximately 100-feet wide and 3,000-feet long. Mallard Slough is fed by Suisun Bay, in an area that is influenced predominantly by tides during the dry season, and predominantly by rain and snowmelt runoff during the wet season.

4.1.1 Data Sources

There are three data sources that are considered in this feedwater quality analysis:

- Source water quality collected during pilot operations
- Historical data summary from CCWD. Mean, minimum, and maximum values are available for water quality parameters for two five-year periods. These data have been collected typically when MSPS is operating, which generally occurs during the spring rains when salinity is at its lowest. In order to get additional information for the season when salinity is typically high and to collect data for parameters that had not been tested previously, a grab sample was taken before the pilot testing began. These pre-pilot data are presented separately in **Appendix A**, (TM 3A *Feedwater Quality Characterization*).
- Online real-time data from Suisun Bay. Historical hourly conductivity and temperature data were also obtained from a water quality monitoring station located in Suisun Bay near the City of Pittsburg maintained by the California Department of Water Resources (DWR).

The feed water quality analysis in this section incorporates data from the pilot plant feed water quality over the duration of the pilot test, as well as data from the other sources where appropriate. Sampling conducted during the pilot study represents a coherent snapshot at reliable sampling intervals and spans the two local water quality seasons. Pre-pilot data is less emphasized in this section because it is heavily weighted toward the low-salinity season when MSPS typically operates.

4.1.2 Water Quality Seasons

The pilot's stable operating period began in November 2008, when ambient air temperatures were warm and little to no precipitation had fallen since the summer. As a result, the freshwater flow from the Sierra Nevada snowmelt was minimal, and the seawater influence was high. The first major rainstorms occurred in early February 2009 and were severe enough to bring Mallard Slough chloride below 100 ppm as the rainfall runoff flushed the seawater downstream. At this trigger, CCWD began operation of their MSPS pumps, and the pilot operation was ceased for approximately one month. When the pilot plant restarted, the continued freshwater influence was evident in the lower salinity levels.

The suspension of pilot operations in February coincided with the break between Runs 2 and 3 in the study and is a natural reference point to define a seasonal shift, as demonstrated in the sections below.

Therefore, the dry season for this pilot study is defined as the period up to and including Run 2 which ended on February 11, 2009. The wet season for this pilot study is defined as the period of Run 3 which began on March 17, 2009. A source water sample taken on February 26, 2009 is also included in the wet season data.

4.1.3 Conductivity

With tidal and seasonal mixing of freshwater and seawater in Suisun Bay, conductivity (as a surrogate for TDS or chlorides) is a good indicator of changes in the character of the water. Pilot plant feedwater conductivity is shown in Figure 4-1, as well as a comparison from the DWR station that collects continuous conductivity data on Suisun Bay.

As shown, conductivity over the duration of the pilot plant demonstrated two distinct seasons. From October to February (dry season), the conductivity was higher and exhibited regular fluctuations that are tied to the two-week lunar cycle. The highest conductivity observed in the feedwater was approximately 20,000 µS/cm, which occurred twice. In March and April (wet season), the conductivity was significantly lower and fluctuated with rainstorms. During heavy rains, conductivity was 1,000-2,000 µS/cm for weeks at a time.

The approximate TDS equivalent values are shown on the right axis of the graph. The conversion from conductivity to TDS is described in Section 4.1.4.

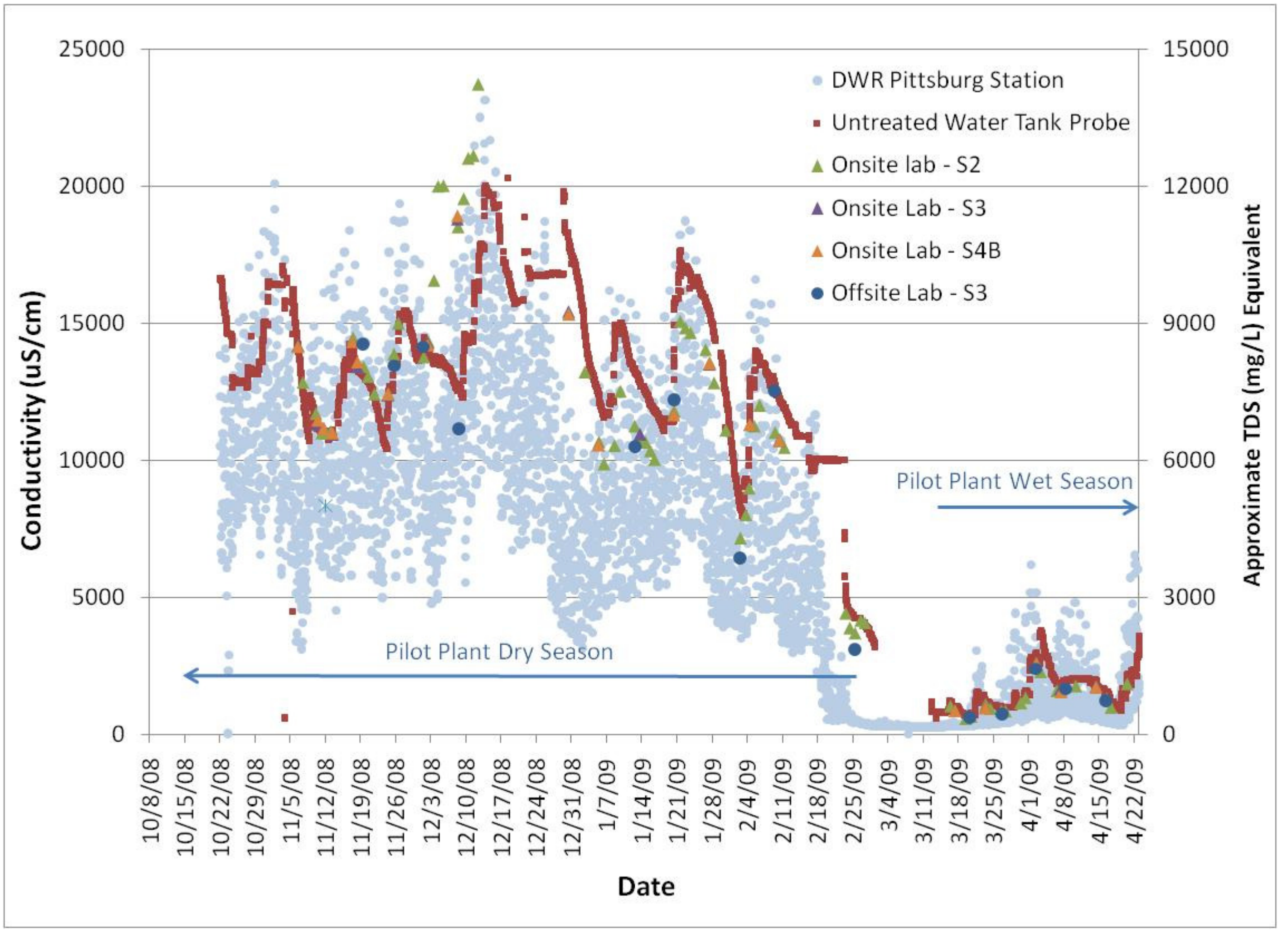


Figure 4-1: Feedwater Conductivity, October through April, Seasonal Changes

The conductivity data from the DWR station in Suisun Bay is shown in a background blue color on Figure 4-1. Although the general trends between Suisun Bay conductivity and pilot plant feedwater conductivity align with each other, Suisun Bay exhibits a wider range of values on a daily basis. This is because Suisun Bay experiences much greater daily fluctuations in salinity due to the tides. The travel time in Mallard Slough dampens the extremes in conductivity and potentially provides for evapotranspiration so that the conductivity at the end of Mallard Slough is on the high end of the range of conductivity in Suisun Bay. Section 6.1.1 provides a statistical summary of historical DWR conductivity data from Suisun Bay which is used as the basis for the RO full-scale analysis.

The range of historical salinity data from CCWD from 1996-2005 is 70 to 7,130 mg/L TDS, with an average value of 2,600 mg/L TDS. This translates to a conductivity range of approximately 115-11,700 µS/cm and an average value of approximately 4,300 µS/cm (see Section 4.1.4 below for TDS-conductivity conversion). The high values accumulated over 10 years by CCWD are lower than the normal dry season values collected during the pilot test. This is likely because CCWD data is collected in the wet season when MSPS could potentially operate and salinity is low.

Overall, the two week lunar cycle dominated changes in salinity, with the conductivity peaks being close to the new or full moons. Rises in conductivity associated with an upcoming full or new moon tended to occur over short periods on a daily basis, associated with the high tides (decreases in conductivity tended to be more continuous). Figure 4-2 shows a two-week period during which the salinity rose and fell with the lunar cycle. High tides and full/new moons are also mapped out. As shown, the dominant influence in conductivity change was the lunar cycle, but conductivity increases occurred in daily step-changes coinciding with high tides.

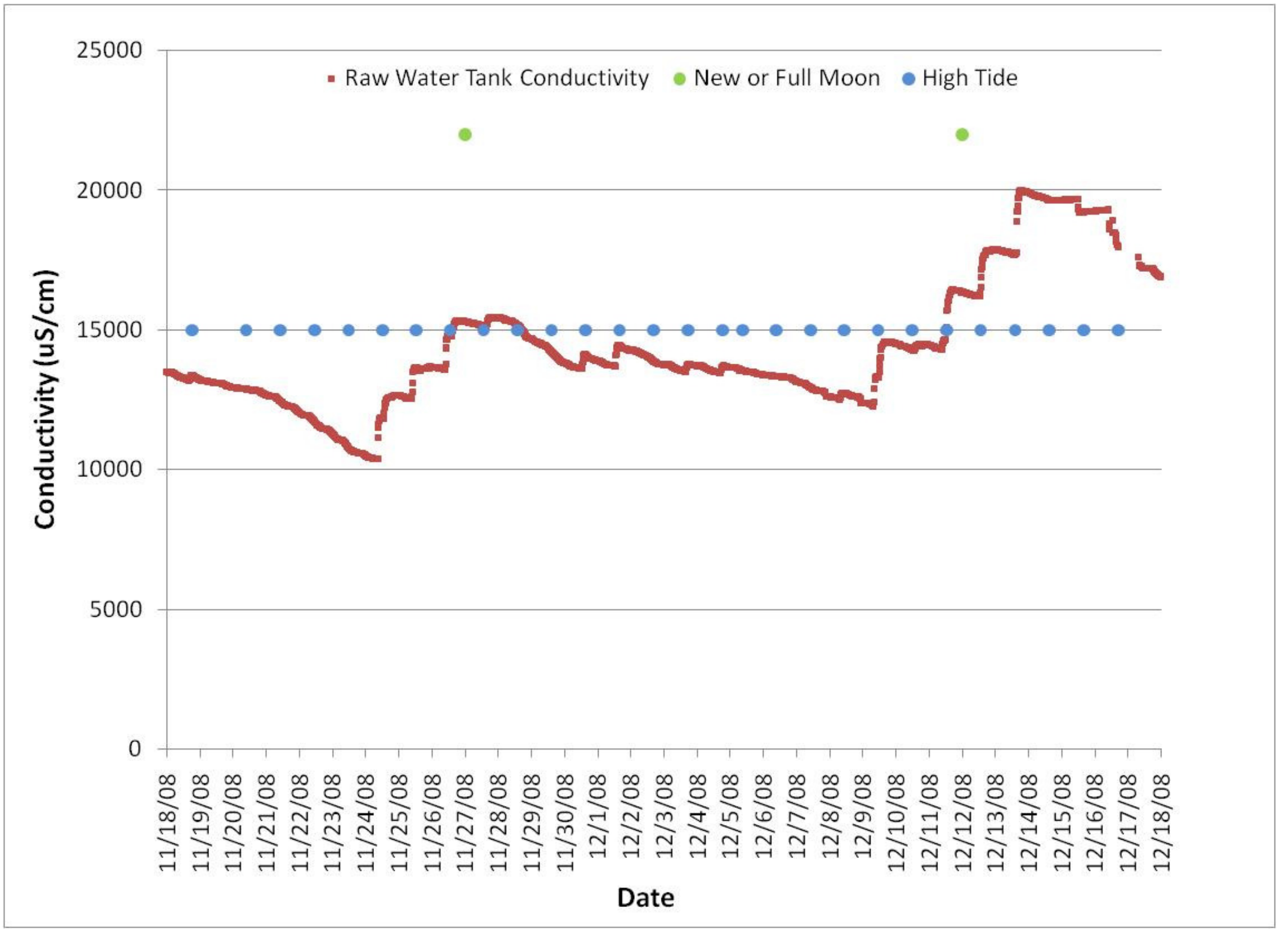


Figure 4-2: Feedwater Conductivity, Two Weeks Showing Lunar and Tidal Changes

4.1.4 TDS Correlation to Conductivity

When external laboratories analyzed for TDS in any given sample, they also analyzed for conductivity. This is important because conductivity is easily monitored continuously while TDS requires manual grab samples. A strong correlation can provide a basis for using conductivity as a surrogate for TDS.

Based on empirical lab results, TDS and conductivity are correlated, but the TDS:Conductivity ratio varies between waters, as shown:

Pilot Plant Stream	TDS:Conductivity Mean Ratio (with Standard Deviation)
Brackish Water (pre-RO)	0.61 (SD = 0.05)
NF/RO Concentrate	0.71 (SD = 0.05)
NF/RO Permeate	0.49 (SD = 0.06)

The ratio values were consistently different between the streams of water. The brackish water TDS:Conductivity ratio is consistent with other published values for the Delta and local waterways. The permeate ratio is lower and the concentrate ratio is higher due to the electrochemistry of low- and high-salinity solutions.

Based on the empirical brackish water TDS:Conductivity ratio and the conductivity data presented in Section 4.1.3, the salinity encountered during the pilot testing ranged up to 12,000 mg/L TDS in the dry season, and was 500-1,000 mg/L TDS during the wet season. The grab samples for TDS analyzed in the laboratory do not reflect these extremes due to the intermittent nature of grab samples, but the conversion provides a means for determining TDS peaks.

It should be noted that when the final site is selected, the surrogacy ratio between any parameters will need to be reestablished.

4.1.5 Turbidity

In addition to conductivity, turbidity changes drastically between seasons. Pilot plant feedwater turbidity is shown in Figure 4-3.

Feedwater turbidity, like conductivity, also demonstrated two distinct seasons over the duration of the pilot plant. From October to February (dry weather), the turbidity was lower and relatively consistent between 5 and 15 NTU. In March and April (wet weather), the turbidity was significantly higher and spanned a much greater range, possibly tied to rainstorm runoff events, spiking to 40 NTU and higher.

The range of historical turbidity data from CCWD from 1996-2005 is 4 to 146 NTU, with an average value of 26 NTU. The turbidity data accumulated over 10 years by CCWD are substantially higher than the normal dry season values collected during the pilot test. This is likely because CCWD data is collected when MSPS is operating, and turbidity is naturally higher. Furthermore, when the pump station is operating, additional sediment may be suspended along the length of Mallard Slough.

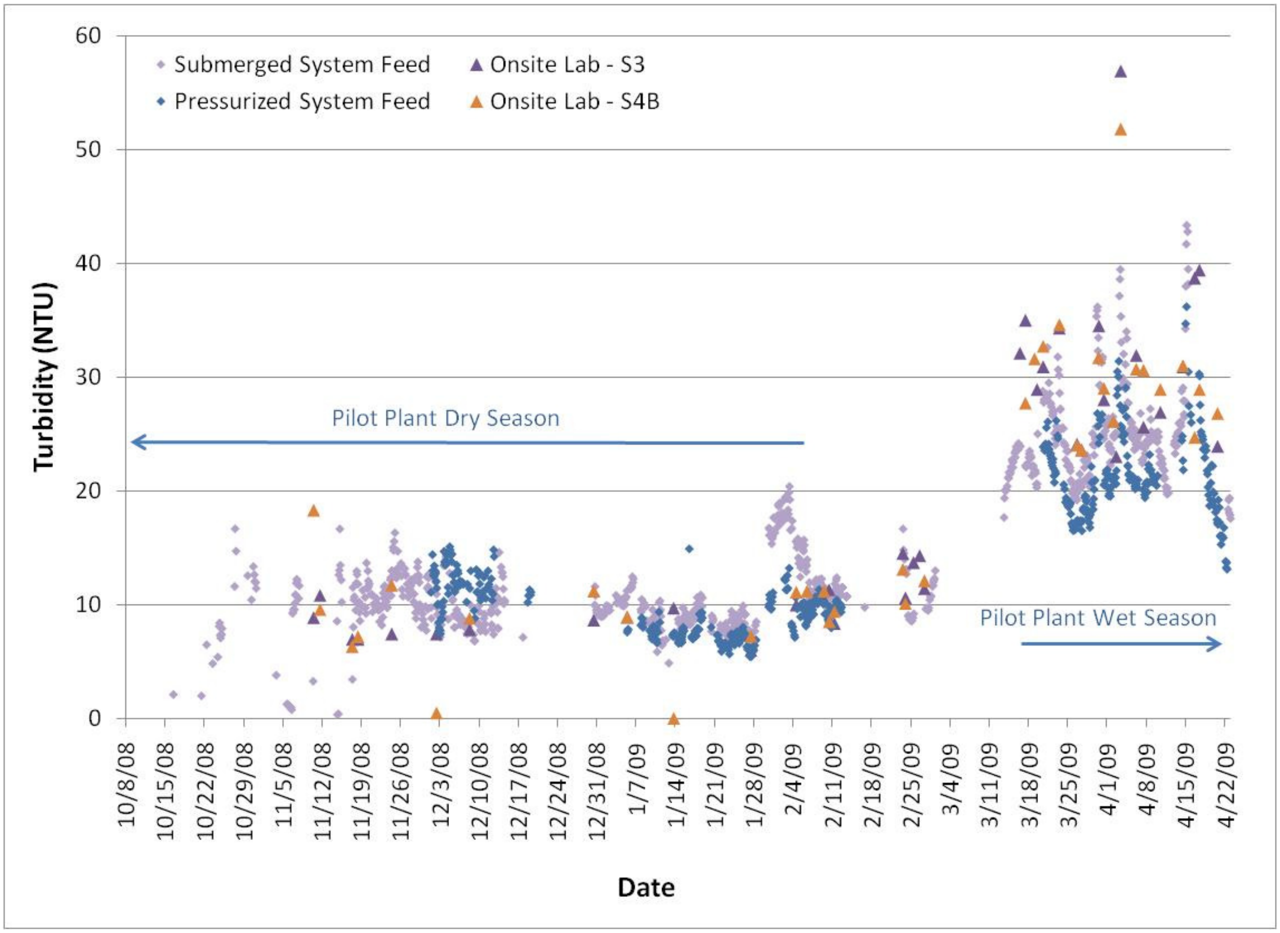


Figure 4-3: Feedwater Turbidity, October through April, Seasonal Changes

4.1.6 Temperature

Feedwater temperature exhibited seasonal patterns also, with colder temperatures during the winter and warmer temperatures in the fall and spring. However, temperature did not mirror the same seasonal shift as seen in turbidity and conductivity in March. Instead, temperature was more heavily influenced by the ambient climate, as the source waters are shallow surface waters. The feedwater temperature and maximum daily air temperature are shown in Figure 4-4.

As illustrated, pilot feedwater temperature spanned a wide range over the course of the pilot test. The lowest temperatures occurred in early January, around 8 deg C. It is anticipated that the seasonal water temperature outside of the piloting period would generally be warmer than what was observed during the test period because the test period did not include a summer, or a peak temperature, season.

The temperature data from the DWR station in Suisun Bay is shown in a background blue color on Figure 4-4. The temperatures in Suisun Bay and the pilot plant feedwater align with each other because the two waters are both shallow and subject to the same ambient air temperatures.

Section 6.1.1 provides a statistical summary of historical DWR temperature data from Suisun Bay which is used as the basis for the RO full-scale analysis.

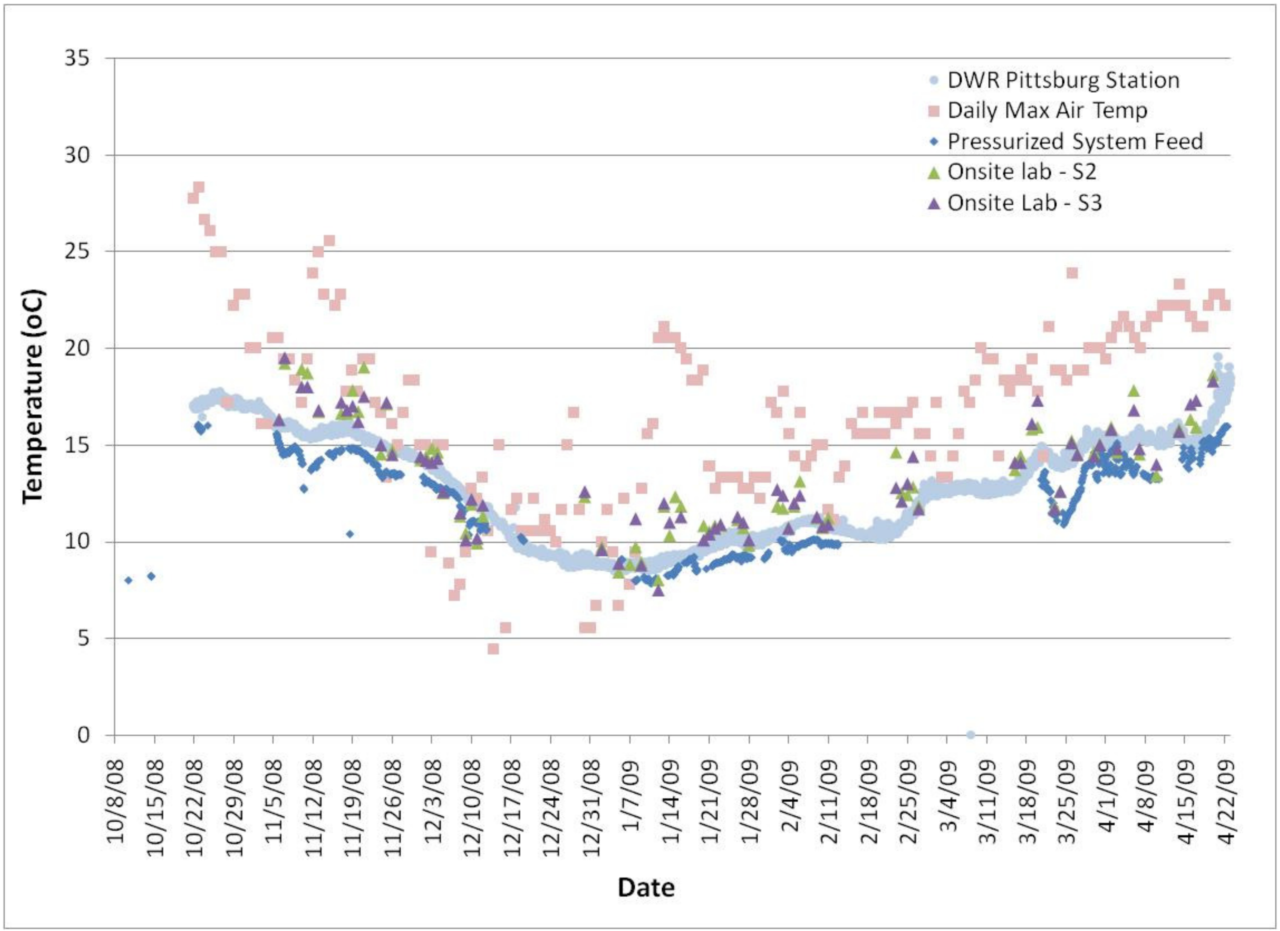


Figure 4-4: Feedwater Temperature, October through April, Seasonal Changes

4.1.7 Other Water Quality Parameters

In addition to conductivity and turbidity, which were measured with continuous instruments, other feedwater quality parameters were analyzed at offsite laboratories regularly in grab samples collected at the plant. A seasonal comparison of the feedwater quality is summarized in Table 4-1, with data separated into the wet season and dry season for comparison. For each pilot study parameter, the mean value, number of samples, and standard deviation is provided. In the calculations of average and standard deviation, any non-detect values were conservatively assumed to be at the detection limit. A complete data set is provided in **Appendix C** (Pilot Plant Data).

Where available, the mean of the 10-year historical data collected by CCWD is also shown in Table 4-1.

Table 4-1: Feedwater Quality Summary, Offsite Laboratory Results

Analyte:		Units:	Dry Season Mean	Dry Season N	Dry Season Std Dev	Wet Season Mean	Wet Season N	Wet Season Std Dev	CCWD[1] Mean
			S1 - Pilot Plant Feed						
Total Alkalinity as CaCO3		mg/L	99.3	11	27.3	85.6	5	12.8	69
Total Hardness as CaCO3		mg/L	1109	10	431.2	164	5	71.8	320
Bicarbonate Alkalinity		mg/L	99.5	11	27.1	85.6	5	12.8	
Carbonate Alkalinity		mg/L	8.5	11	25.0	1.0	5	0.0	
Total Calcium (Ca)		mg/L	56.5	10	18.6	15.0	5	4.0	34.1
Total Magnesium (Mg)		mg/L	235	10	93.5	30.7	5	15.0	76
Total Potassium (K)		mg/L	71.6	10	27.4	9.1	5	4.57	19.8
Total Sodium (Na)		mg/L	1944	10	808	198	5	118	523
Sulfate (SO4)		mg/L	458	11	188	52.3	5	25.2	85.3
Iron (Fe)	Total	µg/L	406	9	139	1040	5	247	
	Diss	µg/L	5.3	8	7.0	37.5	5	23.7	
Aluminum (Al)	Total	µg/L	249	10	81.2	493	5	110	
	Diss	µg/L	4.5	10	4.8	5.2	5	3.0	
Barium (Ba)	Total	µg/L	48.5	9	11.8	31.5	5	5.1	
	Diss	µg/L	49.1	10	12.2	25.7	5	6.4	
Boron (B)	Total	µg/L	986	10	361	201	5	65.0	
	Diss	µg/L	944	10	333	200	5	68.3	
Manganese (Mn)	Total	µg/L	36.3	9	25.5	43.1	4	12.6	
	Diss	µg/L	25.1	9	24.1	15.5	4	11.8	
Bromide		mg/L	14.9	9	4.4	1.3	5	0.66	
Chloride		mg/L	3259	11	1350	311	5	198	558
Fluoride		mg/L	0.66	11	0.19	0.11	2	0.014	
Silica		mg/L	18.2	11	4.7	36.1	5	3.6	17
Total Suspended Solids		mg/L	10.1	2	6.0				
Nitrite as Nitrogen		mg/L	0.017	11	0.0047	0.02	1		
Nitrate as Nitrogen		mg/L	0.56	11	0.066	0.40	1		0.46
Orthophosphate as P SM4500-P/C		mg/L	0.096	6	0.042	0.10	1		<0.2
TOC		mg/L	1.8	21	0.81	2.4	9	1.3	2.7
UV254		abs	0.11	12	0.017	0.19	5	0.028	
Algae Count		#/mL	96.0	3	56.2				
MTBE		µg/L	1.0	1					
Perchlorate		µg/L	0.20	1					
			S3 - Following Untreated Water Tank						
Total Dissolved Solids[2]		mg/L	9726	9	11449	758	5	390	2293
Total Suspended Solids		mg/L	9.4	9	1.5	21.0	5	3.8	
Algae Count		#/mL	38	2	7.1				

Note 1: CCWD historical data from 1996-2005 was taken primarily in the wet season when MSPS could operate.

Note 2: TDS range in this table is based on several grab samples; a more complete characterization is provided using conductivity as a surrogate (see Section 4.1.4).

There were fewer samples taken in the wet season because a greater percentage of the testing period occurred during the dry season. The apparent seasonal differences are distinct. The rains brought significantly lower hardness and lower concentrations of many minerals including magnesium, calcium, and potassium. Some metals were higher in the wet weather, including iron and aluminum. Among the parameters that are generally of concern in desalination, boron was higher when seawater dominates; and TOC was higher when freshwater dominates.

The CCWD mean data is more representative of the wet season sampling, as CCWD data is collected when MSPS is operating during wet weather.

Also shown in Table 4-1 are the results of samples taken at Sample Point 3 (S3), just downstream of the untreated water tank. These results indicate that the self-cleaning filter provided some benefit in removing TSS and algae from the pilot plant feedwater.

4.2 UF Pretreatment Results

This section summarizes the results from the two UF systems that operated in parallel for the duration of the pilot study. A submerged system from Siemens/Memcor and a pressurized system from Norit were installed and tested in the pilot plant.

Flux was the setpoint prescribed for each system for each run. Because the temperature of the water couldn't be predicted, the flux setpoints were absolute (not temperature-corrected), and each system operated at a constant flux. Temperature-correction was applied to the specific flux calculation. Flux data presented in this report is not temperature corrected to demonstrate that it matches the run setpoints, but specific flux is temperature corrected.

4.2.1 Submerged UF System Performance

The submerged UF pilot skid operated continuously during the three runs. The initially-installed submerged membranes were the manufacturer's new formulation with a better retention of small particles and mechanically stronger. Clean water flux testing was performed on the first set of membranes by the manufacturer's staff that started up the unit. In the first two runs, irreversible fouling occurred, and cleaning procedures were unable to recover membrane permeability. As a result, a new set of membranes with the older formulation was installed for Run 3; clean water flux testing was performed by MWH staff for the second set of membranes. Discussion of membrane operations and performance is provided in this section.

4.2.1.1 Flux and Recovery

For each run, flux and recovery were held constant. The flux and recovery setpoints are summarized in Table 4-2.

Table 4-2: Submerged UF Flux and Recovery by Run

	Run 1	**Run 2**	**Run 3**
Flux (gfd)	32	32	41
Recovery	91.2%	91.2%	94.6%

Flux targets for each run were developed based on conversations between MWH membrane experts and manufacturer's staff. Run 1 targeted a baseline flux, Run 2 a mid-challenge flux, and Run 3 a challenge condition. The actual flux in Run 2 was lower than initially planned due to unanticipated hydraulic constraints of the pilot plant feed system.

Recovery is calculated based on filtrate production, filtrate use for system cleaning, and drain/waste flows. At times during pilot testing, one or two modules were removed from the membrane tank to minimize filtrate tank overflow. However, while filtrate production decreases with fewer modules, feed and filtrate flow requirements for backwash (BW) and maintenance wash (MW) are the same regardless of how many membranes are in the tank. For this reason, the actual pilot-scale recovery depends on the number of modules installed. Therefore, in order to compare runs, the recovery in the above table was calculated assuming that all four membranes were installed in the tank.

The factors that impacted recovery across the runs were the backwash interval and the maintenance wash interval. BW and MW procedures were initially set, and then adjusted between runs as deemed appropriate by the pilot team for the given run conditions, as summarized below. As reflected in Table 4-3, the changes in BW and MW cleaning procedures were not enough to have a significant impact on the overall system recovery.

Table 4-3: Submerged UF Backwash and Maintenance Wash

	Run 1	Run 2	Run 3
Backwash Interval (min)	25 min	25 min	30 min
Hypochlorite MW Interval	24 hrs	24 hrs	36 hrs
Hypochlorite MW Conc.	200 ppm	50 ppm	200 ppm
Muriatic Acid MW Interval	24 hrs	24 hrs	36-48 hrs
Muriatic Acid MW Conc.	0 ppm[1]	100 ppm	500 ppm

Note 1: An equipment malfunction inhibited the acid maintenance washes during Run 1.

As an indicator of membrane performance, permeability (also known as specific flux) is the flux divided by trans-membrane pressure (TMP) and normalized with temperature and is plotted in Figure 4-5.

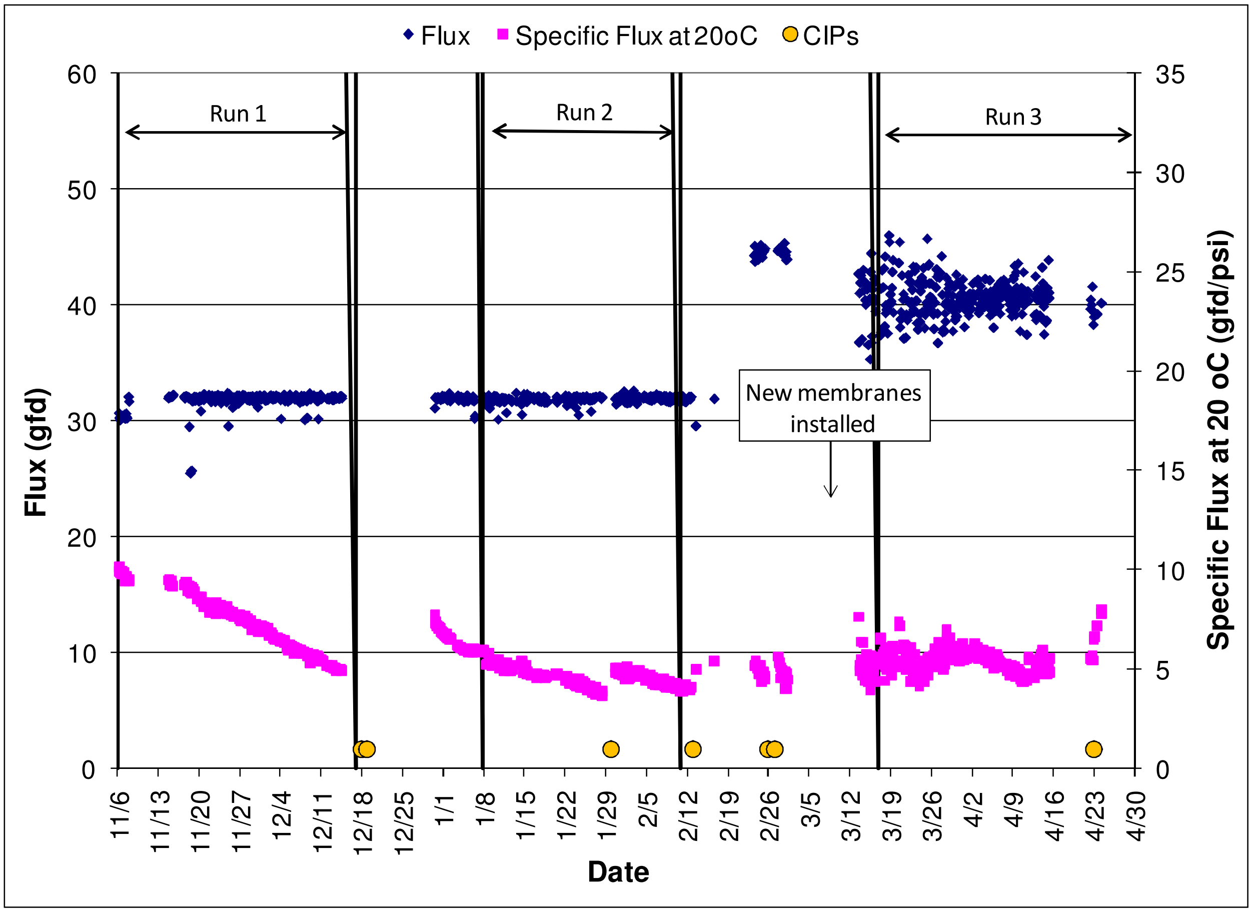


Figure 4-5: Submerged UF Flux

The system was running where data is plotted. Operational periods between the official runs can be explained as follows:

- Runtime between Runs 1 and 2: Although all pilot systems were started on January 2, 2009, the submerged UF system was the only unit that operated with minimal interruption for the first week. The other systems experienced some operational difficulties and were not able to start until January 8, so January 8 was named as the nominal start date of Run 2.
- Runtime between Runs 2 and 3: All pilot systems were intended to be started on February 26, 2009, but there were corrosion problems with the pressurized UF skid's pneumatic system since it had been sitting idle for several weeks. It turned out to be beneficial for the submerged UF system because the flux decline was so rapid upon restart that another CIP was performed immediately, and again the flux decline was rapid. It was decided on March 2 to order and replace the membranes, and they were installed on March 12. After clean water flux testing, all three units were started up, and Run 3 officially began.
- After the end of Run 3, a CIP was performed, and the pilot system was run for the remaining two days before site demolition.

As shown, although the flux was constant, the rising TMP and falling temperatures made the temperature-corrected permeability decrease rapidly in Run 1. This was not recoverable by the CIP procedures employed, and cleaning efficiency of the submerged membranes is discussed further in Section 4.2.1.2. The critical flux for this system (generally considered to be the highest flux value where TMP is constant over time - above the critical flux value, TMPbegins to vary and is no longer steady with time could not be identified due to the fouling issues and the change of membranes. The flux of 41 gfd in Run 3 appeared to be sustainable, but it also represented the inaugural run for the new membranes and thus would need to be repeated for verification.

Regardless of the flux and recovery setpoints, the submerged membranes performed suitably, producing a filtrate turbidity always less than 0.15 NTU and an SDI generally below the limit of 3.0, as shown below in Figure 4-6.

The SDI values were always below the goal of 3.0, with the exception of one data point. None of the SDI values were greater than the typical RO membrane warranty SDI of 5.0.

The step down in filtrate turbidity on January 29, 2009, can be explained by the fact that the operations staff experienced considerable trouble calibrating the filtrate turbidimeter on the submerged unit until the end of January. From February onward, turbidimeter calibrations did not cause any trouble. It is likely that prior to February, the turbidimeter was reading a higher value. However, even if the filtrate turbidimeter was reading high, all turbidity data were acceptable for RO feed.

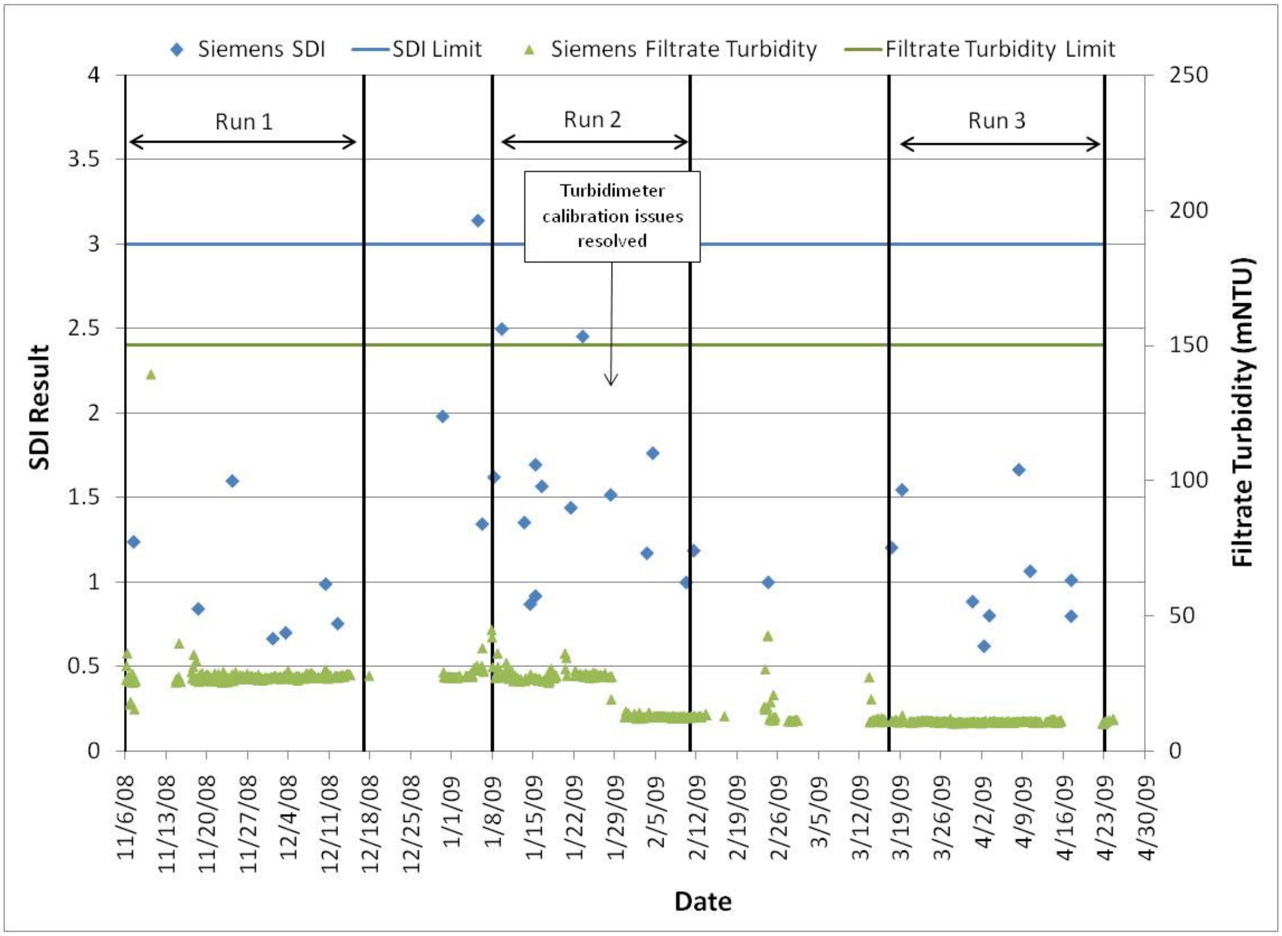


Figure 4-6: Submerged UF Turbidity and SDI

4.2.1.2 Cleaning Efficiency

The submerged UF system experienced some trouble with permeability recovery from the CIPs. Initially, the heating unit did not function and was replaced. Therefore, the initial several CIPs were conducted at a cold water temperature (9 to 12 deg C). The specific flux before, between, and after each CIP procedure, is illustrated in Figure 4-7. As shown, the specific flux during clean water flux testing was 8.2 gfd/psi. During normal membrane operations (between CIPs), specific flux declined. No single CIP was able to restore the membranes to their original performance. After an unsuccessful CIP #3, an aggressive CIP was undertaken, with little impact. The membranes were replaced, and Run 3 was conducted with a new set of membranes which recovered nicely after their first CIP.

The initially-installed submerged membranes were the manufacturer's new formulation with a better retention of small particles and mechanically stronger. At the time of installation, the membranes were in the final stages of testing to receive recognition by DPH as a drinking water approved technology. After these membranes were irreversibly fouled in the pilot testing, they were replaced with the older formulation.

Descriptions of the individual CIP procedures are provided in Table 4-4. Heat was introduced in the system as noted. Acid CIPs were run at 2.5% citric acid. Chlorine CIPs were run at 0.05% NaOCl and solution was made up with permeate instead of filtrate, and without pH adjustment (pH of the solution was 9.2 to 9.6). CIP procedures were automated to have nine cycles of 10-minute soak followed by 30-second aeration. Therefore, the soak period was approximately 90 minutes for each CIP.

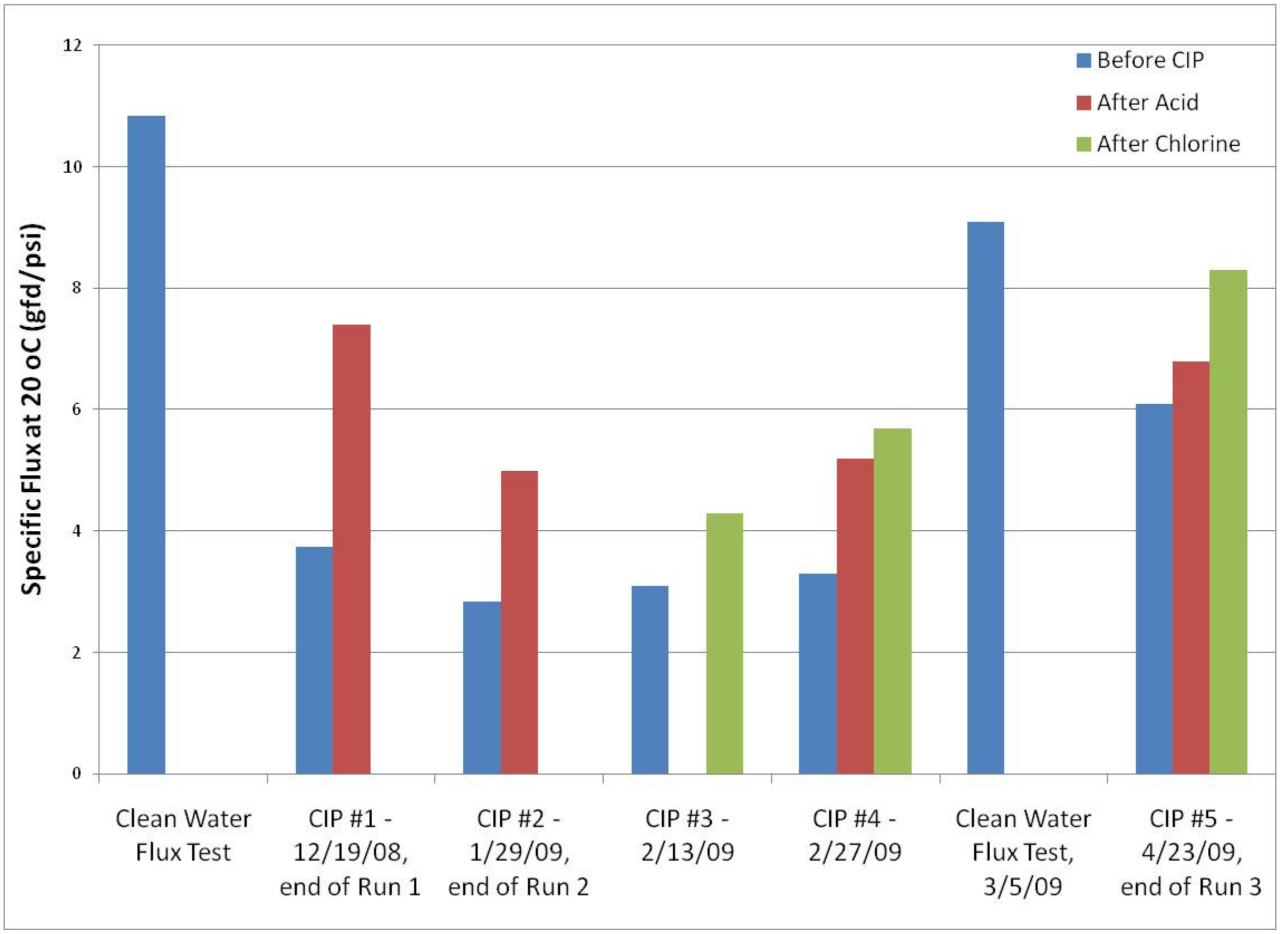


Figure 4-7: Submerged UF CIP Efficiency – Specific Flux

Table 4-4: Submerged UF CIPs

CIP	Date	Purpose	Description
1	12/18-19/08	Followed Run 1; heater not working.	Partial acid CIP (heater discovered not working), followed by full acid CIP without heat.
2	1/29/09	Acid with heat, after heater was fixed.	Acid CIP with heat (38 deg C)
3	2/13/09	Followed Run 2	Acid CIP with heat (38 deg C), followed by chlorine CIP with heat (20 deg C)
4	2/26-27/09	CIP #3 failed to recover permeability or drop TMP, and a more aggressive clean was attempted. More specific flux decline. Changed membranes.	Chlorine CIP with heat (20 deg C) followed by overnight chlorine soak, followed by acid CIP with heat (38 deg C)
5	4/23/09	Followed Run 3	Acid CIP with heat (38 deg C), followed by chlorine CIP with heat (20 deg C)

4.2.1.3 Membrane Integrity

Membrane integrity was tested daily when the system was running with pressure decay tests (PDTs) that were automatically programmed and recorded. The PDT results rose slightly over the course of the pilot study, but was always well below the limit of 0.725 psi/min.

4.2.1.4 Membrane Fouling Investigation

Due to the irreversible fouling on the first set of membranes, an autopsy was requested from the manufacturer to identify potential foulants and to gain a more thorough understanding of design ramifications. Unfortunately, because of the proprietary nature of the membrane material and its associated manufacturing process, the submerged membrane vendor was unwilling to allow the fouled membrane to be autopsied. As a result, membrane fouling cause and any further details are unknown at this time.

It is speculated that the new formulation membranes which were used at the start of this study had not been fully tested and vetted by the manufacturer prior to installation at Mallard Slough.

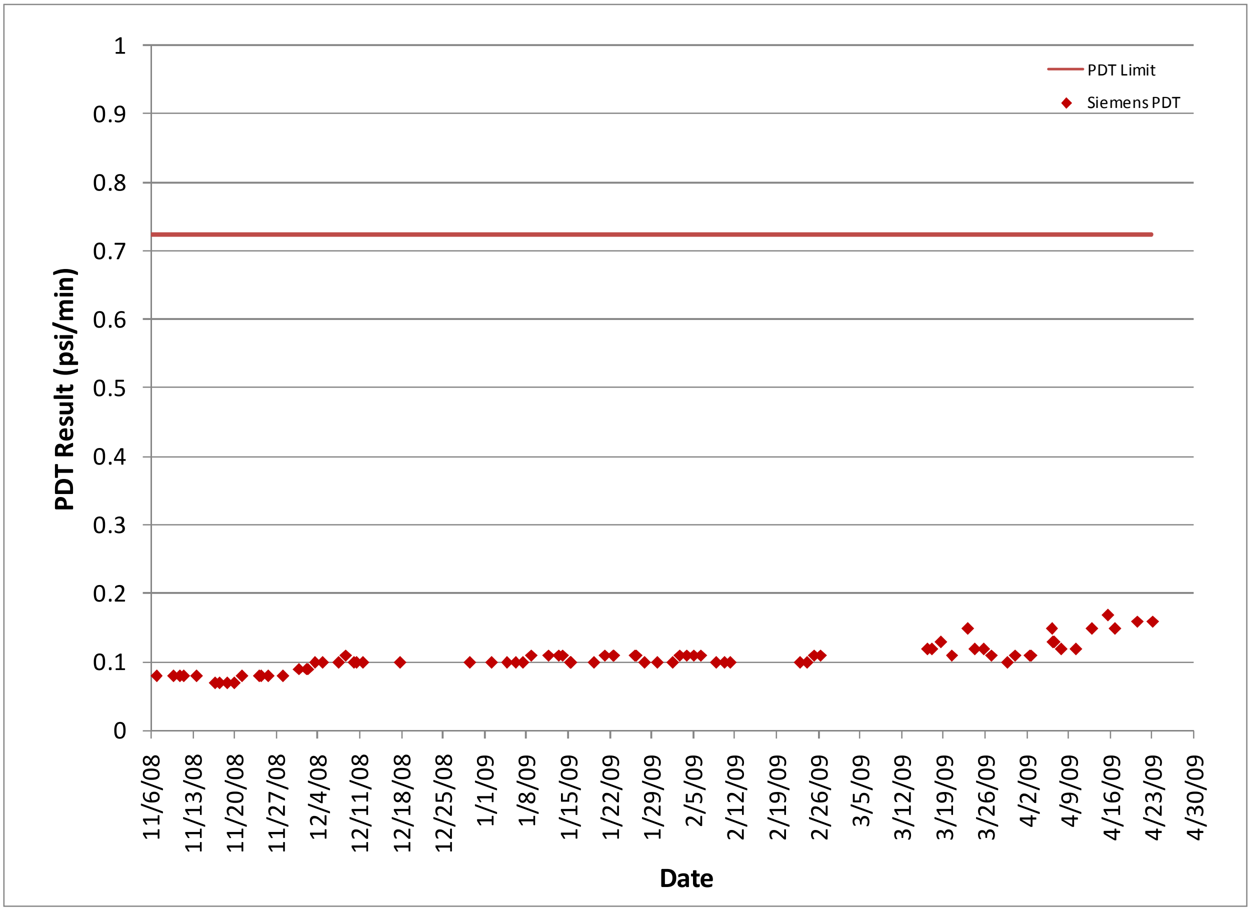


Figure 4-8: Submerged UF Membrane Integrity

4.2.2 Pressurized UF System Performance

The pressurized UF pilot skid operated continuously during the three runs. It should be noted that clean water flux testing was not conducted by the manufacturer's staff that started up the unit. Discussion of membrane operations and performance is provided in this section.

4.2.2.1 Coagulation Assessment

The pressurized UF membranes required the use of a coagulant. Based upon the feedwater quality collected before the pilot plant started, the manufacturer specified a ferric chloride dose of 5 mg/L (1.7 mg/L as iron), and did not require rigorous jar testing. This dose of 5 mg/L is consistent with numerous other successful projects using the same membranes. A visual coagulant assessment was performed by pilot operations staff to confirm the dose recommendation, and it determined that adding ferric chloride is beneficial, but within the range of 3-7 mg/L, the specific dose is not significant. Therefore, 5 mg/L ferric was targeted in the feedwater for the duration of the pilot study.

4.2.2.2 Flux and Recovery

For each run, flux and recovery were held constant. The flux and recovery setpoints are summarized in Table 4-5.

Table 4-5: Pressurized UF Flux and Recovery by Run

	Run 1	Run 2	Run 3
Flux (gfd)	40	44	55
Recovery	86.3%	86.1%	89.4%

Flux targets for each run were developed based on conversations between MWH membrane experts and the manufacturer's staff. Run 1 targeted a baseline flux, Run 2 a mid-challenge flux, and Run 3 a challenge condition.

Factors that typically impact recovery include backwash interval and the chemically enhanced backwash (CEB) interval. BW and CEB procedures were initially set, and then adjusted between pilot runs as deemed appropriate for the given run conditions, as summarized in Table 4-6.

Table 4-6: Pressurized UF Backwash and Chemically Enhanced Backwash

	Run 1	Run 2	Run 3
Backwash Interval (min)	30 min	30 min	30 min
Hypochlorite CEB Interval	48 hrs	24 hrs	36 hrs
Hypochlorite CEB Conc	200 ppm	200 ppm	200 ppm
Citric Acid CEB Interval	48 hrs	24 hrs	36 hrs
Citric Acid CEB Conc	1800 ppm	1800 ppm	1800 ppm

Recovery was calculated as:

Filtrate produced / (Filtrate used for BWs + Filtrate used for CEBs)

During Run 2, the filtrate produced increased, but the filtrate used for CEB also increased due to the more frequent procedures, leading to a similar recovery between the two runs. Although changes to the cleaning setpoints were minor in Run 3, the significant increase in flux increased the recovery.

As an indicator of membrane performance, permeability (also known as specific flux) is the flux divided by TMP and normalized with temperature, and is plotted in Figure 4-9.

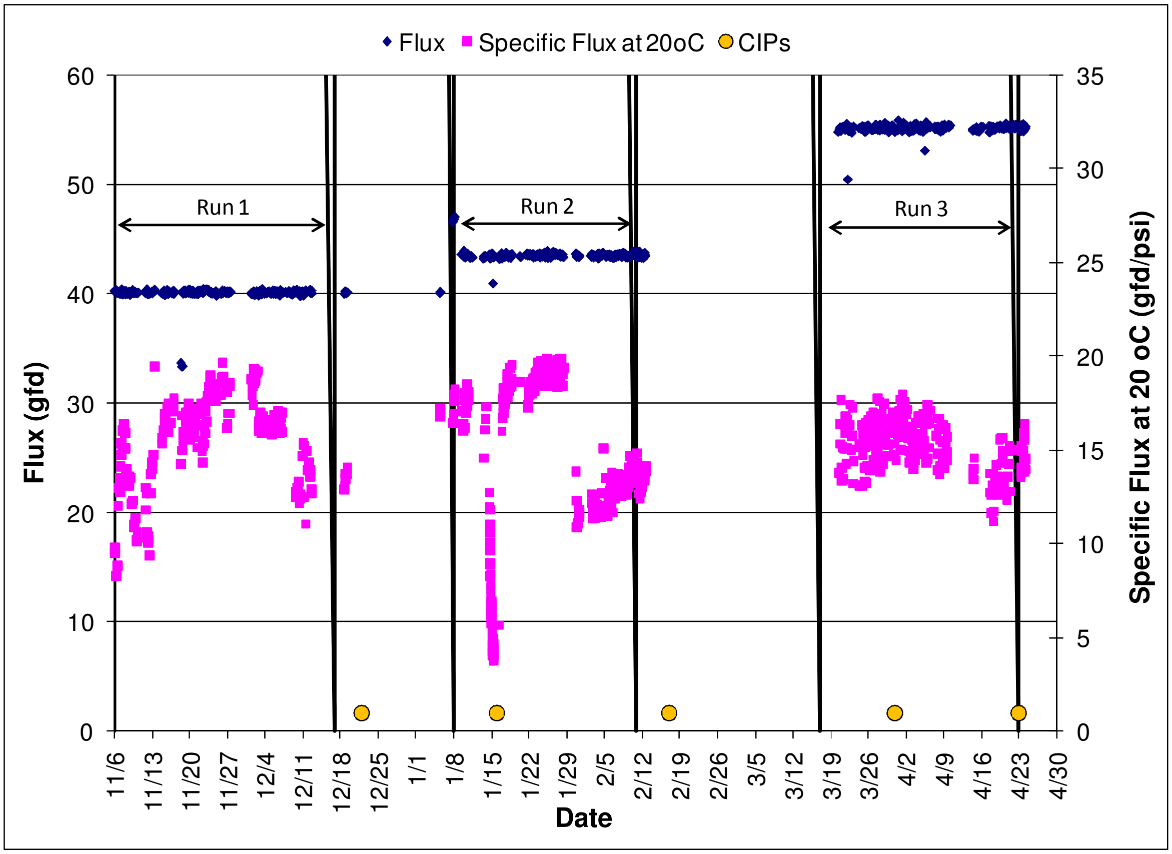


Figure 4-9: Pressurized UF Flux

As shown, the permeability was consistent and the pressurized UF system did not experience significant fouling, regardless of the flux. In the instance where the membranes did foul due to a suspected ferric overdose into the on-skid feed water tank in mid January, they fouled quickly and completely, and all of the permeability was recovered with the CIP. Cleaning efficiency of the pressurized membranes is discussed further in Section 4.2.2.3. The reason for the sharp drop in specific flux observed at the end of January is unknown; the pressurized UF system was shut down for approximately two days to repair unrelated equipment, and upon restart, the flux was lower. Two day outages were not uncommon, and no other similar outage had the same result. The critical flux for this system was not identified because flux decline did not occur, but it is known that the flux of 55 gfd was sustainable during the pilot Run 3 conditions (i.e. higher turbidity and lower salinity)

The pressurized UF membranes performed suitably, producing a filtrate turbidity always less than 0.15 NTU and an SDI generally below the limit of 3.0, as shown below in Figure 4-10.

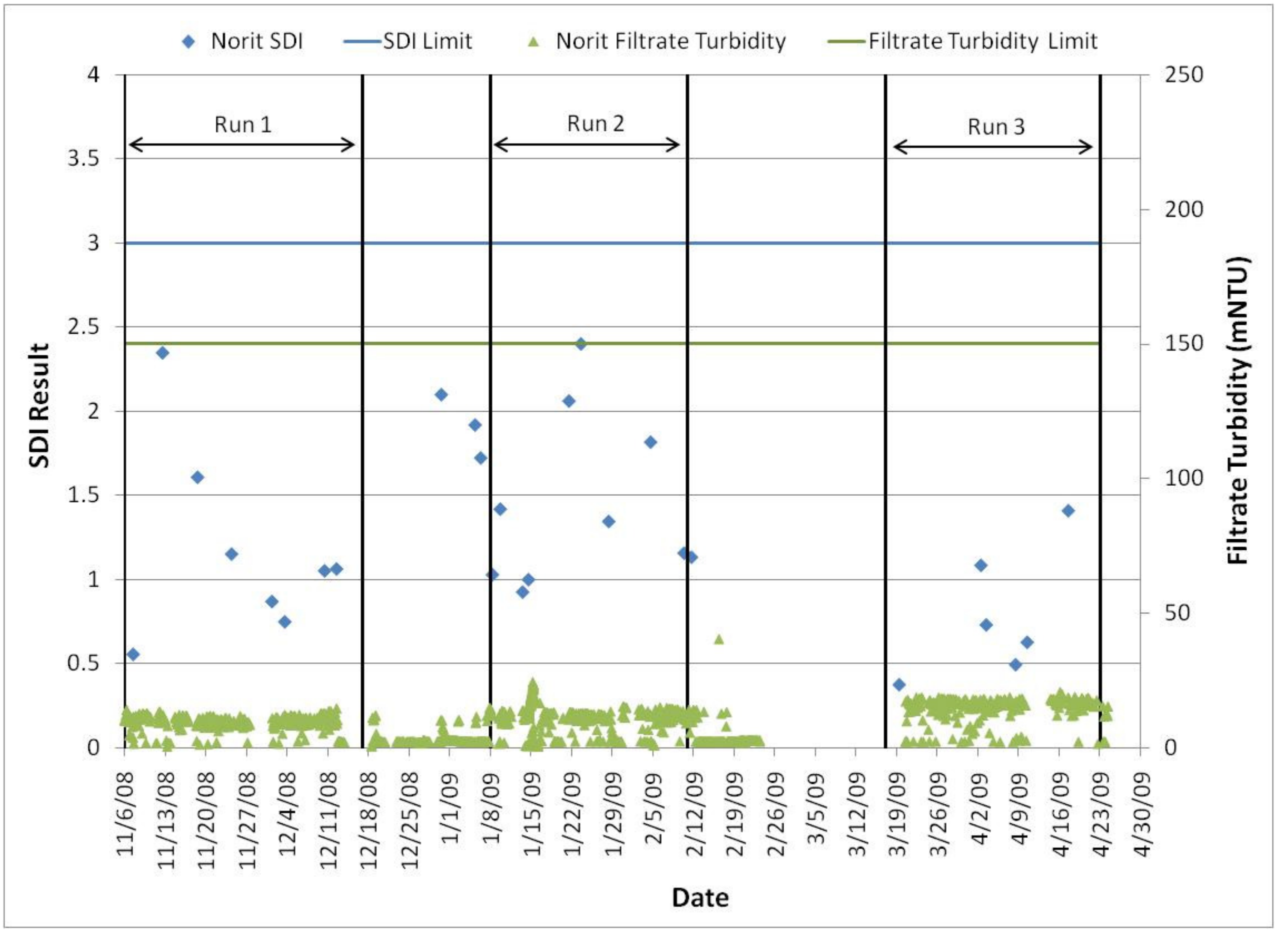


Figure 4-10: Pressurized UF Turbidity and SDI

The SDI values were consistently below the goal of 3.0, which is well below the typical RO membrane warranty SDI of 5.0.

4.2.2.3 Cleaning Efficiency

The pressurized UF system did not foul sufficiently to understand the efficiency of CIPs, with the exception of CIP #2, a suspected ferric overdose. In this case, the CIP following the incident was able to recover permeability and decrease the TMP to new-membrane levels. The specific flux before, between, and after CIP each procedure is illustrated in Figure 4-11. As shown, each individual CIP was able to successfully attain the specific flux of the initial run. The water for CIPs was not heated.

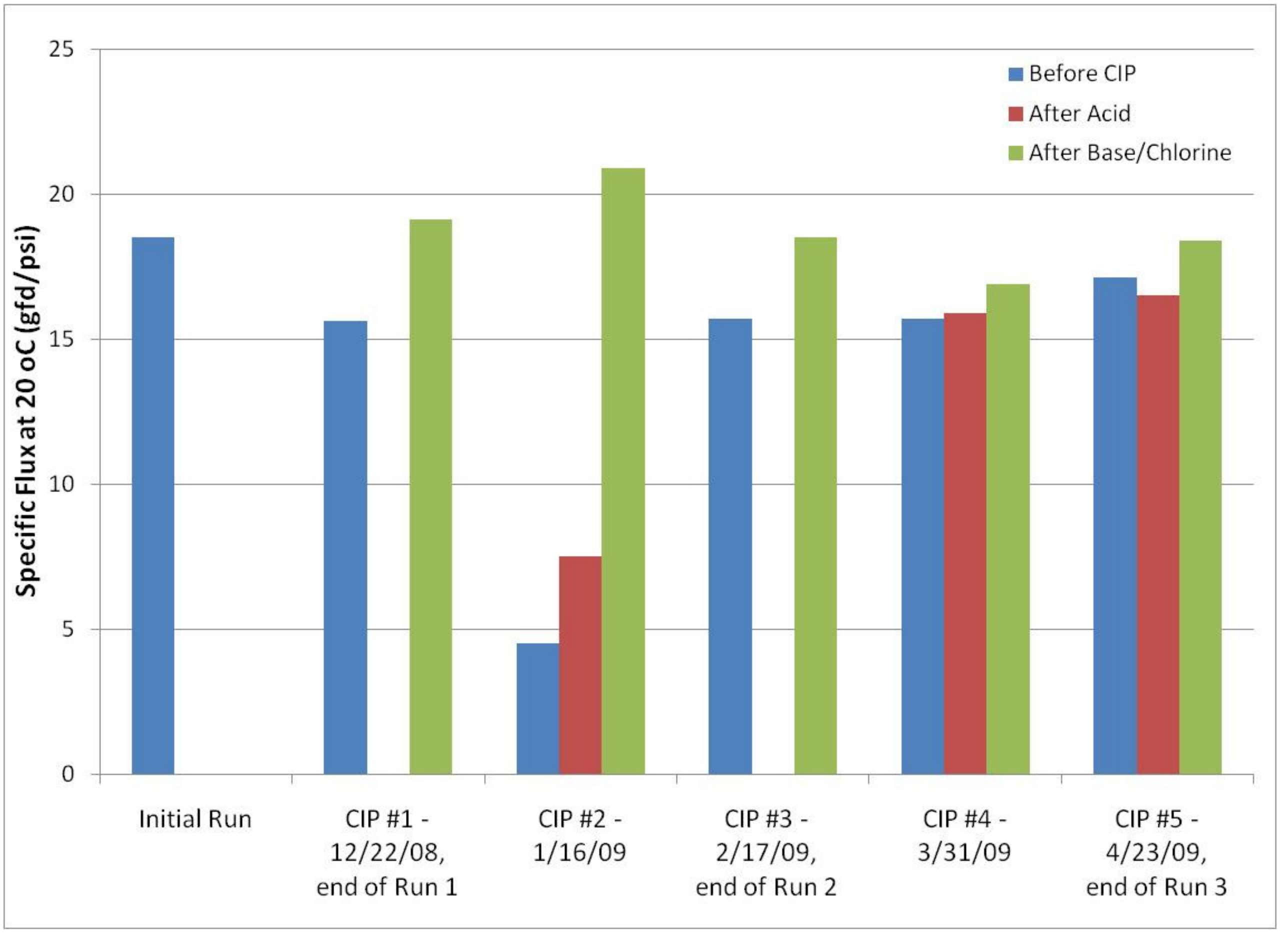


Figure 4-11: Pressurized UF CIP Efficiency – Specific Flux

Descriptions of the CIP procedures are provided. No heat was introduced in the pressurized system.

Acid CIPs were conducted using the following procedure:

- Cleaning solution: filtrate with approximately 2% citric acid solution (at or near pH 2)
- 10-minute recirculation at a flow rate of 15 gpm/module
- 20-minute soak
- 15-minute recirculation at 15 gpm/module. If pH was stable, then the CIP was done, otherwise soak and recirculation were repeated. Number of recirculation cycles is noted in Table 4-7.
- 15-minute recirculation with permeate valve open.

Caustic/chlorine CIPs were conducted using the following procedure:

- Cleaning solution: filtrate with approximately 0.5% sodium hydroxide (at or near pH 12) with 200 ppm chlorine
- 20- to 30-minute recirculation at a flow rate of 15 gpm/module
- 20-minute soak

- 15- to 20-minute recirculation at 15 gpm/module. If pH and chlorine concentration were stable, then the CIP was done. If not, soak and recirculation were repeated. Number of recirculation cycles is noted in Table 4-7.
- 15-minute recirculation with permeate valve open.

Table 4-7: Pressurized UF CIPs

CIP	Date	Purpose	Description
1	12/22/08	Followed Run 1	Acid CIP (1 recirc cycle) followed by caustic/ chlorine (3 recirc cycles)
2	1/16/09	Fouling event, likely caused by ferric overdose when system went down briefly and ferric feed to on-skid feed water tank stayed on; after restart, filtrate was red.	Acid CIP (2 recirc cycles) followed by 2 caustic/ chlorine CIPs (1 and 3 recirc cycles, respectively)
3	2/17/09	Followed Run 2	Acid CIP (1 recirc cycle) followed by caustic/ chlorine (3 recirc cycles)
4	3/31/09	Wide range of flux after Run 3 startup (TMP difference between top and bottom feed was 8 psi, previously 4 psi). Suspected foulant deposited at bottom of vessels during one-month downtime while CCWD was running its pump station. CIP did not affect the TMP range.	Acid CIP (1 recirc cycle) followed by caustic/ chlorine (2 recirc cycles)
5	4/23/09	Followed Run 3	Acid CIP (1 recirc cycle) followed by caustic/ chlorine (3 recirc cycles)

4.2.2.4 Membrane Integrity

Membrane integrity was tested approximately daily when the system was running. Performing a PDT required manually stopping filtration, performing a PDT, and restarting the system. Results were recorded on logsheets.

The pneumatic system on the Norit skid controlling the PDT procedures had numerous problems during the pilot test. At various times during the pilot study, the air compressor was replaced three times, the air valving was changed out, and each individual pneumatic actuator was replaced at least once. All of the old system components had salt crystals, indicating a breach in the water-air interface (the PDT piping offered the only interface). Therefore, any unusually high PDT values may have been related to air escaping through the PDT system instead of a membrane breach. The PDT results rose slightly over the course of the pilot study, but was always well below the limit of 0.725 psi/min.

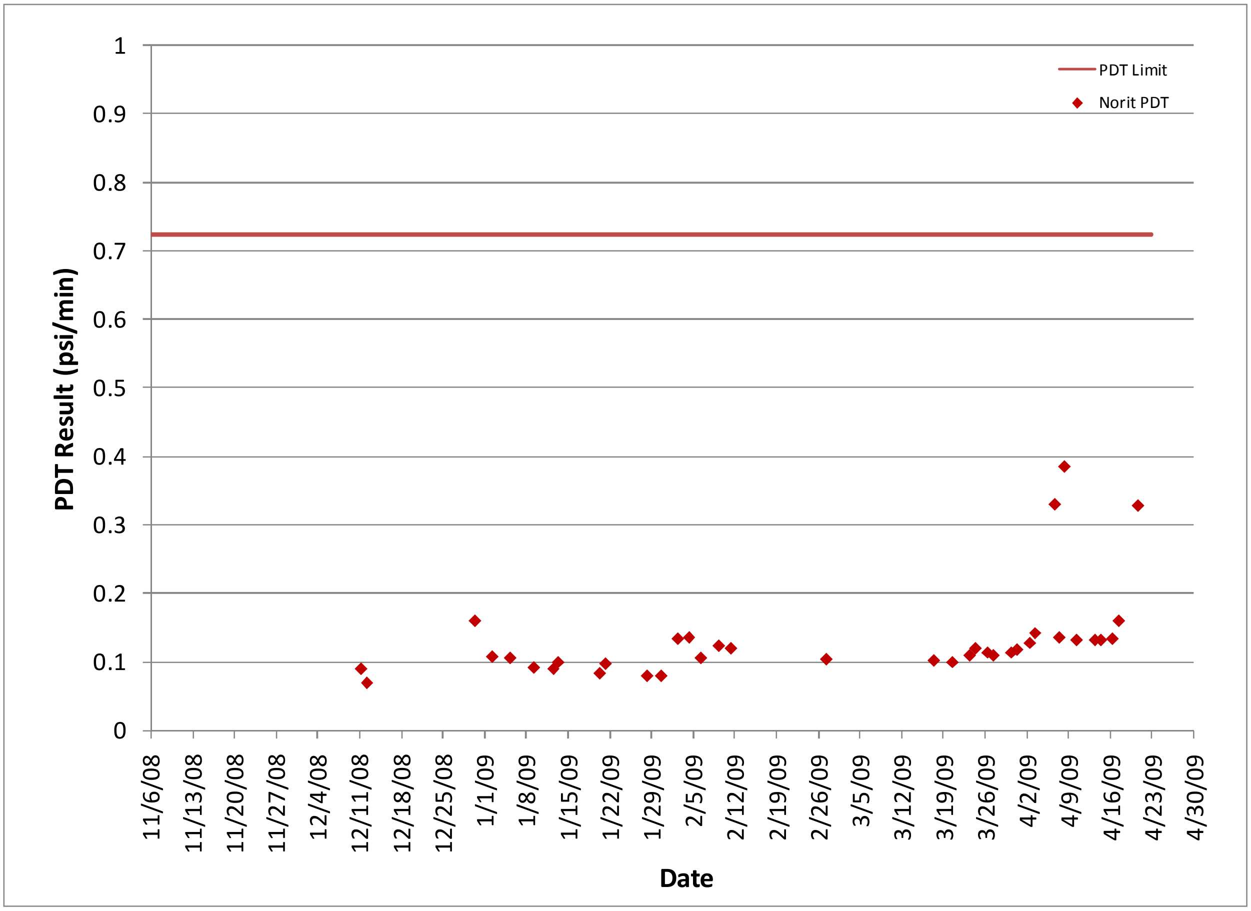


Figure 4-12: Pressurized UF Membrane Integrity

4.2.3 Residuals Analysis

Samples of the UF backwash waste were collected and sent to manufacturers of solids handling equipment who had agreed to conduct bench-scale testing to develop design criteria for a potential full-scale solids handling facility. Backwash samples were sent to Andritz (manufacturer of belt press and centrifuge), Ashbrook (manufacturer of belt press), and Alfa Laval (manufacturer of centrifuge).

All three of the vendors originally thought that they could work with the dilute, unthickened backwash water. However, upon receiving the samples, they had insufficient volume of solids to do so. They recommended the use of a gravity thickener with polymer addition upstream of the centrifuge or belt press.

In order to conduct the full-scale evaluation, the quantity of solids to be produced was estimated. Table 4-8 is a summary of calculated solids characteristics. Total solids in the backwash includes solids from influent turbidity (assuming a worst case condition of 20 mg/L) as well as solids from the use of ferric chlorine as a coagulant (5 mg/L). Production of dry solids per day is based on a 95% capture from the gravity thickener.

Table 4-8: Residual Solids Production Assumptions

Parameter	Value	Unit
Backwash water flow	2.5	mgd
Total solids in backwash	4670	ppd
Solids concentration in backwash	224	mg/L
Design Solids Concentration in Backwash	450	mg/L
Dry Solids produced per day	9,000	ppd

4.2.4 System Comparisons and Recommendations

The specific flux that the pressurized system was able to sustain was significantly higher in value than that of the submerged system, indicating that more water could be produced per membrane area with a lower TMP. The trade-off for the higher specific flux was that the pressurized system requires the use of ferric chloride as a coagulant, adding iron to the backwash waste, and the pressurized system had a lower water recovery over the pilot study, meaning that more feedwater is drawn into the system to produce the same amount of filtrate. It should be noted that the pilot study was not of sufficient duration to fully optimize all operating parameters.

Both the pressurized and submerged systems produced a suitable feedwater for the RO systems in terms of turbidity and SDI. The two systems are compared in Table 4-9.

Table 4-9: Summary comparison of UF Systems

	Submerged UF	Pressurized UF
Filtrate turbidity	0.010 – 0.013 NTU[1]	0.010 – 0.020 NTU
SDI	Median 1.20	Median 1.14
Flux	32-41 gfd	40-55 gfd
Specific Flux	4-10 gfd/psi	10-20 gfd/psi
Recovery	90-95%	85-90%
Chemical use	No coagulant. 1-2 MW/day (hypo 50-200 ppm; citric 0-500 ppm)	Ferric chloride required for coagulation. 1-2 CEB/day (hypo 200 ppm; citric 1800 ppm)
Operational Performance	Irreversible fouling and inefficient CIPs	Complete permeability recovery on each CIP
CIP interval observations	Irrecoverable fouling on first set of membranes (CIP interval of 4 weeks to a few days). Stable operations on second set of membranes too little runtime to project CIP intervals.	Operations were very stable; no evident fouling to project CIP intervals

Note 1: Data after turbidimeter calibration issue was resolved.

The target CIP interval for the pilot plant was met by the pressurized system and by the submerged system in the final run. The submerged UF membranes experienced irrecoverable fouling on the first set of membranes. Although the second set of membranes operated stably during their inaugural run, there was not sufficient runtime to verify their performance. The pressurized membranes were stable throughout the duration of the pilot project without any fouling, so the CIP interval would likely be longer than the 40 days if the run was extended. For the purpose of the full-scale evaluation, a CIP interval of 40 days was assumed for the pretreatment system.

For the purpose of the evaluation in this report, the pressurized UF system is considered in the full-scale evaluation. This is primarily because the initial submerged membranes experienced irreversible fouling and needed to be replaced, and the replacement membranes did not have sufficient runtime to prove out their performance.

4.3 RO Results

This section summarizes the results from the three RO/NF desalination systems that operated in parallel for the duration of the pilot study.

4.3.1 RO1 System

RO1 was a 2:1 array with seven brackish water membranes per vessel in the first stage (Dow/Filmtec BW40-3030) and seven seawater membranes per vessel in the second stage (Dow/Filmtec SW30HRLE-4040). This RO system achieved the highest recovery since it was two stages, with a medium operating pressure and a medium water quality that met all treated water goals.

4.3.1.1 Flux and Recovery

For each run, flux and recovery were held constant. For all three of the RO systems, the Run 1 targeted baseline conditions; Run 2 represented a mid-challenge (with higher flux and/or recovery); and Run 3 was the challenge condition whose recovery was set as high as the membrane models would allow without getting a design warning for low concentrate flow in the second stage (first stage low flow concentrate was allowed). The flux and recovery are summarized in Table 4-10.

Table 4-10: RO1 Flux and Recovery by Run

	Run 1	Run 2	Run 3
Flux (gfd)	12	12	12
Recovery	70%	74%	82%

In order to maintain the constant flux and recovery values given the variations in feedwater operating salinity, the RO1 system PLC adjusted the speed on the feed pump VFD to provide sufficient pressure. Higher salinity required higher feed pressure. In the highest salinity season (Run 1), the operating pressure for RO1 was typically 250-300 psi. When feedwater conductivity was at its lowest (Run 3), the operating pressure for RO1 was typically 130-150 psi.

As an indicator of membrane performance, permeability (also known as specific flux) is simply the flux divided by net pressure (a function of osmotic pressure and feed pressure) and normalized with temperature. Temperature-corrected specific flux ranged from 0.06-0.13 gfd/psi during the dry season, likely affected by the wide salinity variations and the changing temperature, and was constant at 0.1 during the wet season.

Feed pressure and specific flux are plotted in Figure 4-13. The calculations used for normalization are presented in Section 3.4.2.

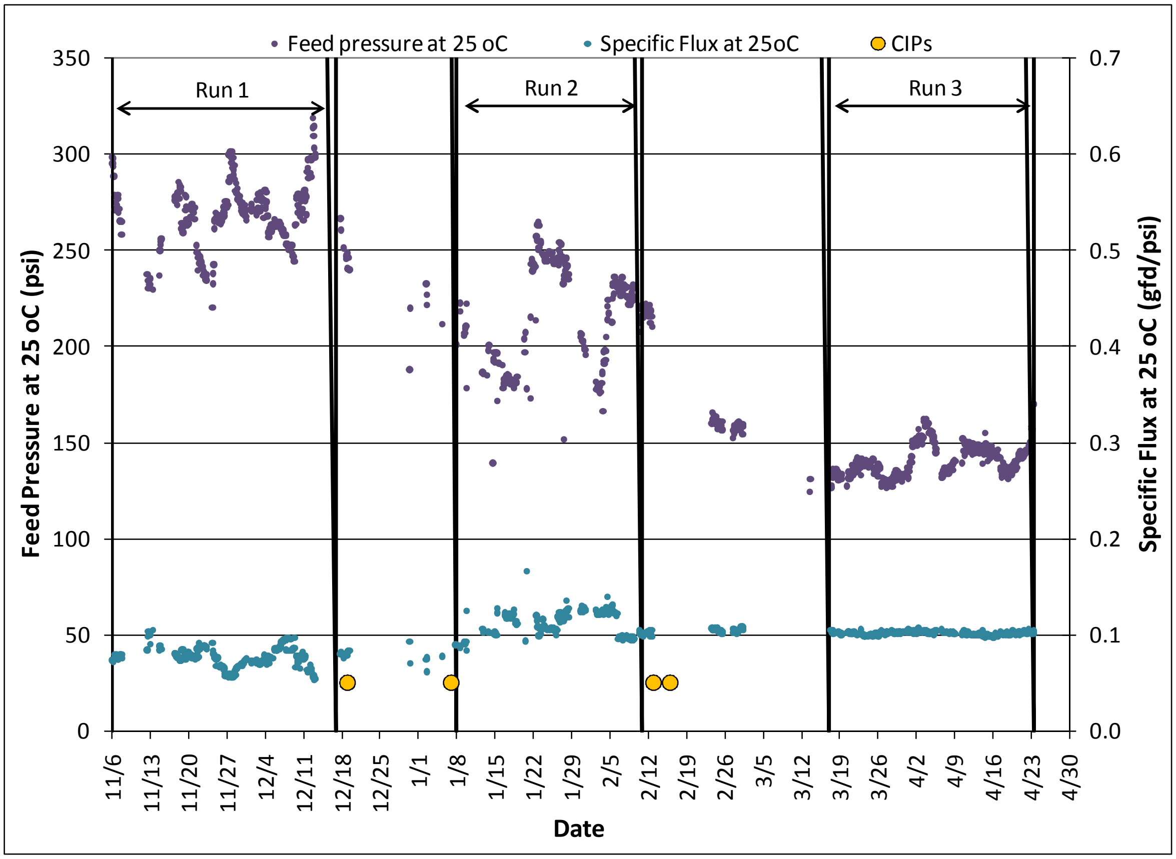


Figure 4-13: RO1 Feed Pressure and Specific Flux

Where data is not shown in the figure, the system was not operational. Operational time that occured between Runs is explained in Section 4.2.1.1. When only one UF system was operating, only half of the RO feedwater was supplied, so RO1 could operate, or the other two single-array RO/NF systems could operate. When the submerged UF system ran in late February, RO1 was also operated.

As shown, the permeability was consistent and the RO1 system did not experience significant scaling or fouling, regardless of the flux and recovery. Cleaning efficiency of the RO1 membranes is discussed further in Section 4.3.1.2. The critical flux for this system was not identified because flux decline did not occur, but it is known that the challenge condition (Run 3) was sustainable during the wet season runoff conditions (i.e. lower salinity).

4.3.1.2 Cleaning Efficiency

The RO1 system did not scale/foul sufficiently to understand the efficiency of CIPs. The specific flux before, between, and after CIP each procedure, is illustrated in Figure 4-14. As shown, each individual CIP did not have a significant impact on the permeability of the system.

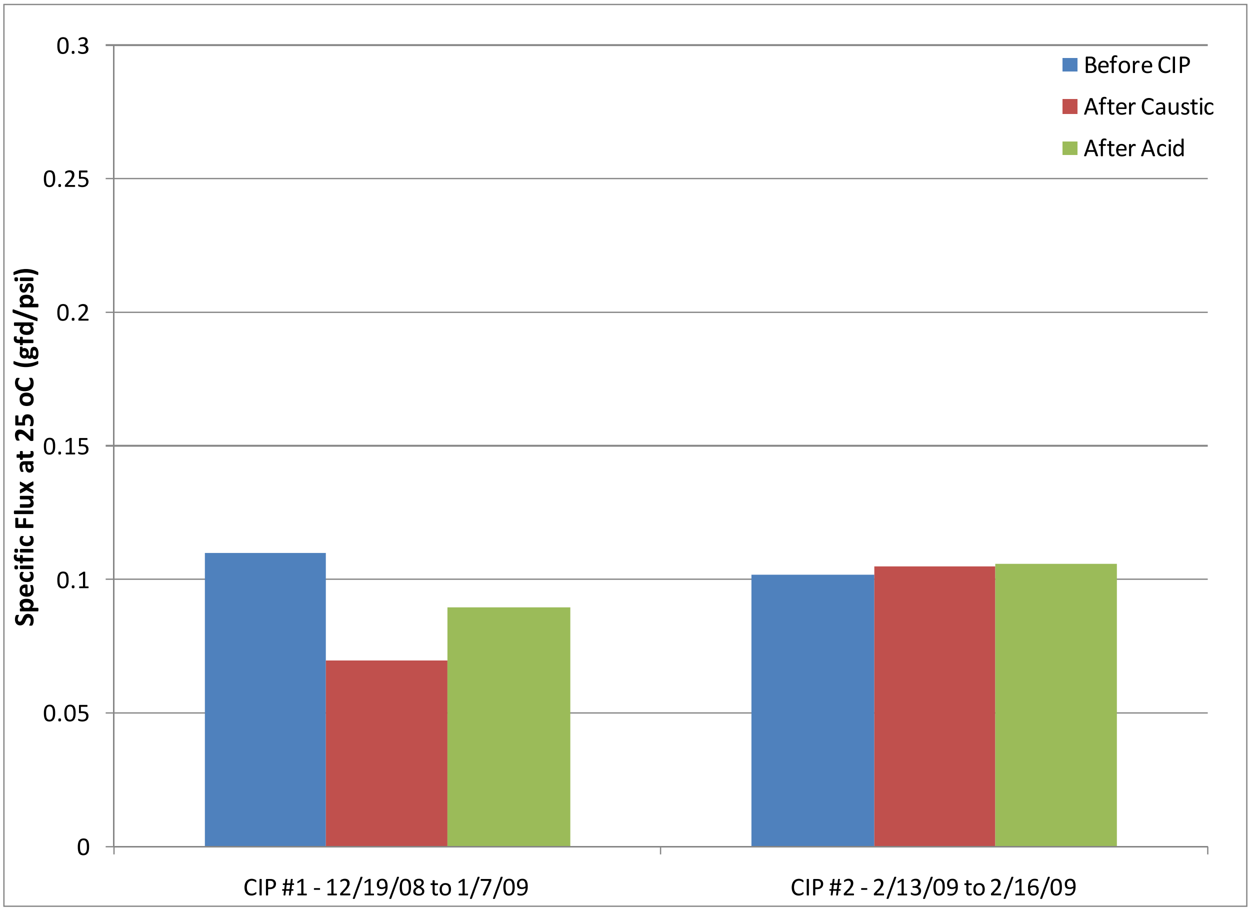


Figure 4-14: RO1 CIP Efficiency – Specific Flux

Each CIP procedure consisted of a high pH clean with caustic soda at pH 11.5-12, followed by a low pH clean with citric (primary) and muriatic (trimming) acid at pH 2-2.5. When RO1 was cleaned, Stage 1 was cleaned prior to Stage 2 for each chemical to prevent the passage of Stage 1 foulants through Stage 2.

During the CIPs that followed Run 1, no heat was used, and the CIPs were conducted at a temperature of 9-11 deg C. During the CIPs that followed Run 2, the CIP solutions were heated so that caustic CIPs were run at 30 deg C and acid CIPs were run at 20 deg C.

After CIP #1, procedures were discussed in detail with the manufacturer to determine whether the CIP efficiency data were significant given the erratic nature of the specific flux during Run 1, and the apparent drop in permeability in CIP #1 following Run 1. It was determined by the manufacturer that the CIP procedures followed were suitable, and the noise in the specific flux data likely represented the brand new membranes experiencing normal inconsistencies during their inaugural run.

4.3.1.3 Treated Water Quality

RO1 permeate met all state and federal MCLs that were analyzed for with the exception of the SMCL for pH (6.5-8.5). This is expected in RO permeate, and a normal full-scale facility would have post-treatment for stabilization and pH elevation to protect the distribution system. Therefore, at full-scale, the treated water pH would meet the SMCL.

Project goals for TDS, chloride, and boron were met, and results are presented in Section 4.3.4.3.

4.3.1.4 Membrane Autopsy

Because all three RO systems were subject to the same feedwater scaling potential, only one element was sent for autopsy at the close of the project. Avista Technologies, Ltd. performed the autopsy on the tail element of the second stage of the RO1 system, since it saw the greatest concentration of potential scalants/foulants over the duration of the pilot study.

Avista found a very limited amount of a tan foulant, which was determined to be clay with trace amounts of polysaccharides, proteins, and carbohydrates. There was insufficient fouling for Avista to gather any more information on the foulant.

Avista observed membrane compaction as part of the autopsy. The suspected cause was a period in the pilot test when O-rings ruptured approximately once a day for about a week. During this time, RO1 was subject to mechanical stresses in one vessel due to the inadequate placement of elements, spacers, and O-rings; whenever an O-ring failed, the system was subject to a jolt of pressure. The issue was resolved by reloading the vessel, but may have led to the compaction identified in the autopsy.

No oxidation damage was found during the autopsy. A complete autopsy report is attached in **Appendix C**.

4.3.2 RO2 System

RO2 was a single-vessel system with six seawater membranes (Dow/Filmtec SW30HRLE-4040). This RO system achieved the best water quality since the seawater technology is the "tightest" membrane, with a relatively high operating pressure and somewhat low recovery with a single stage.

4.3.2.1 Flux and Recovery

For each run, flux and recovery were held constant. For all three of the RO systems, the Run 1 targeted baseline conditions; Run 2 represented a mid-challenge (with higher flux and/or recovery); and Run 3 was the challenge condition whose recovery was set as high as the membrane models would allow without a design warning for low concentrate flow in the second stage (first stage low flow concentrate was allowed). The flux and recovery are summarized in Table 4-11.

Table 4-11: RO2 Flux and Recovery by Run

	Run 1	Run 2	Run 3
Flux (gfd)	12.7	14.1	14.1
Recovery	50	50	62

The significance of feed pressure and specific flux is explained in Section 4.3.1.1. Feed pressure and specific flux are plotted in Figure 4-15. The calculations used for normalization are presented in Section 3.4.2.

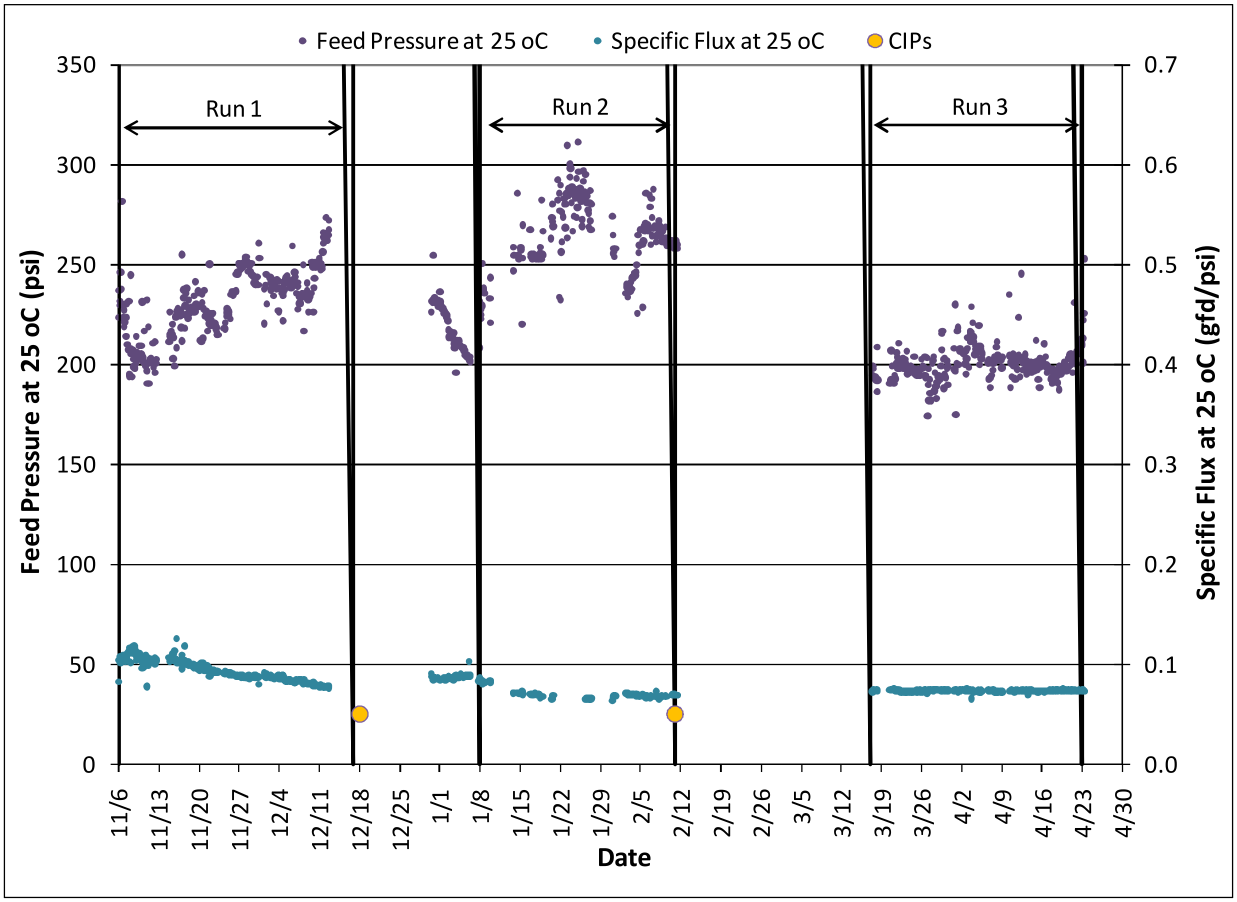


Figure 4-15: RO2 Feed Pressure and Specific Flux

Where data is not shown in the figure, the system was not operational. Operational time that occurs between Runs is explained in Section 4.2.1.1. When only one UF system was operating, only half of the RO feedwater was supplied, so RO1 could operate, or the other two single-array RO/NF systems could operate. When the submerged UF system ran in early January, RO/NF 2 and 3 were also operated.

As shown, the permeability declined over the first two runs and was not recovered during CIP procedures. Cleaning efficiency of the RO2 membranes is discussed further in Section 4.3.2.2. It was determined by the manufacturer that the initial decline in specific flux, especially in Run 1, likely represented the brand new membranes experiencing normal inconsistencies during their inaugural run. The manufacturer indicated that it could take 30-60 days to achieve steady operation for certain new membranes.

The critical flux for this system was not identified because the flux decline experienced was deemed to be an inaugural occurrence and operational fouling or scaling was not apparent. However, it is known that the challenge condition (Run 3) was sustainable during the wet season runoff conditions (i.e. lower salinity).

4.3.2.2 Cleaning Efficiency

The RO2 system did not scale/foul sufficiently to understand the efficiency of CIPs. The specific flux before, between, and after CIP each procedure, is illustrated in Figure 4-16. As shown, each individual CIP did not have a significant impact on the permeability of the system.

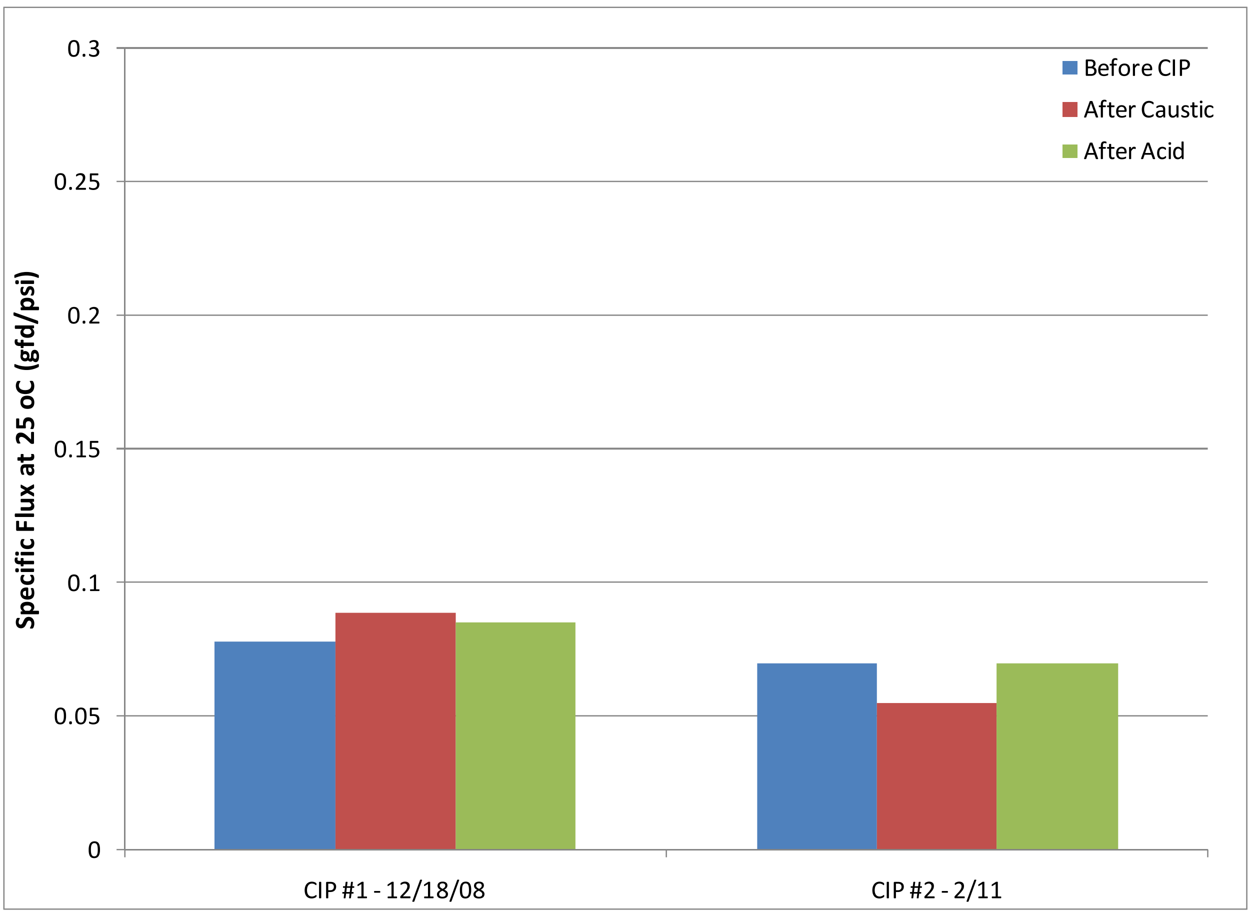


Figure 4-16: RO2 CIP Efficiency – Specific Flux

CIP procedures for RO2 were the same as RO1, and are discussed in Section 4.3.1.2. It was determined by the manufacturer that the noise in the specific flux recovery data likely represented the brand new membranes experiencing normal inconsistencies during their break-in period, which could take 30-60 days for certain new membranes.

4.3.2.3 Treated Water Quality

RO2 permeate met all state and federal MCLs that were analyzed for with the exception of the SMCL for pH (6.5-8.5). This is expected in RO permeate, and a normal full-scale facility would have post-treatment for stabilization and pH elevation to protect the distribution system. Therefore, at full-scale, the treated water pH would meet the SMCL.

Project goals for TDS, chloride, and boron were met, and results are presented in Section 4.3.4.3.

4.3.3 NF3 System

NF3 was a single-vessel system with six nanofiltration membranes (Dow/Filmtec NF90-4040). This RO system achieved the lowest operating pressure, but permeate did not meet the project's chloride goal at all times, and the recovery was low with a single stage.

4.3.3.1 Flux and Recovery

For each run, flux and recovery were held constant. For all three of the RO systems, the Run 1 targeted baseline conditions; Run 2 represented a mid-challenge (with higher flux and/or recovery); and Run 3 was the challenge condition whose recovery was set as high as the membrane models would allow without a design warning for low concentrate flow in the second stage (first stage low flow concentrate was allowed). The flux and recovery are summarized in Table 4-12.

Table 4-12: NF3 Flux and Recovery by Run

	Run 1	Run 2	Run 3
Flux (gfd)	13.2	12.9	12.9
Recovery	50	56	63

The significance of feed pressure and specific flux is explained in Section 4.3.1.1. Feed pressure and specific flux are plotted in Figure 4-17. The calculations used for normalization are presented in Section 3.4.2.

Where data is not shown in the figure, the system was not operational. Operational time that occurs between Runs is explained in Section 4.2.1.1. When only one UF system was operating, only half of the RO feedwater was supplied, so RO1 could operate, or the other two single-array RO/NF systems could operate. When the submerged UF system ran in early January, RO/NF 2 and 3 were also operated.

As shown, the permeability was relatively stable, showing a slight decline in the first run, and it was speculated by the manufacturers that this represented the brand new membranes experiencing normal inconsistencies during their inaugural run. However, the feedwater salinity in Run 3 represented the range of salinity in which the nanofiltration technology is designed to operate most efficiently, and the corresponding specific flux was slightly higher. It is not known what caused the drop in specific flux at the end of Run 3, and the pilot plant shutdown date precluded further investigation to see if the specific flux decline would continue, and/or whether a CIP could recover the permeability. Cleaning efficiency of the NF3 membranes is discussed further in Section 4.3.3.2.

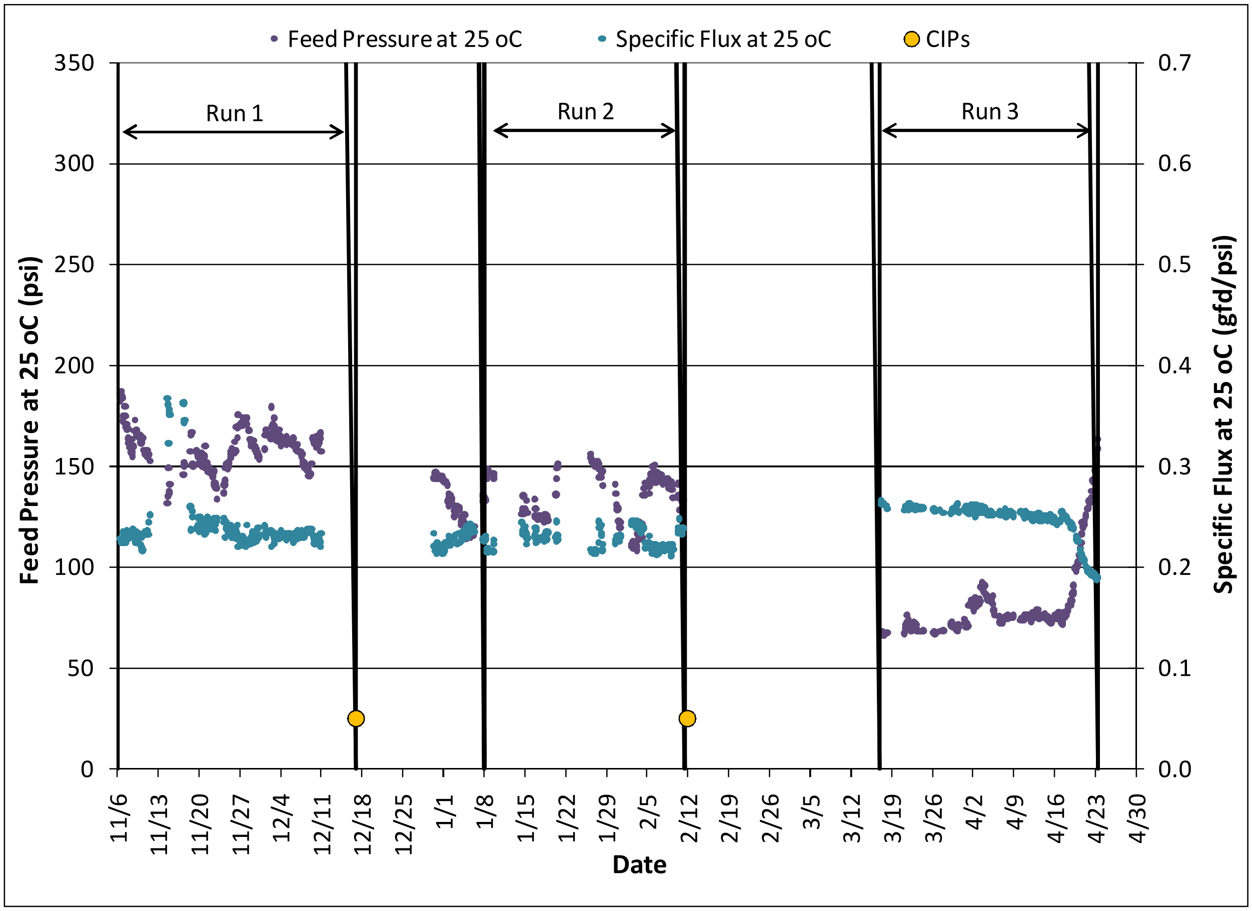


Figure 4-17: NF3 Feed Pressure and Specific Flux

The critical flux for this system was not identified because the flux decline experienced was deemed to be an inaugural occurrence, and operational fouling or scaling did not otherwise occur. However, it is known that the challenge condition (Run 3) was sustainable during the wet season runoff conditions (i.e. lower salinity).

4.3.3.2 Cleaning Efficiency

The NF3 system did not scale/foul sufficiently to understand the efficiency of CIPs. The specific flux before, between, and after CIP each procedure, is illustrated in Figure 4-18. As shown, each individual CIP did not have a significant impact on the permeability of the system.

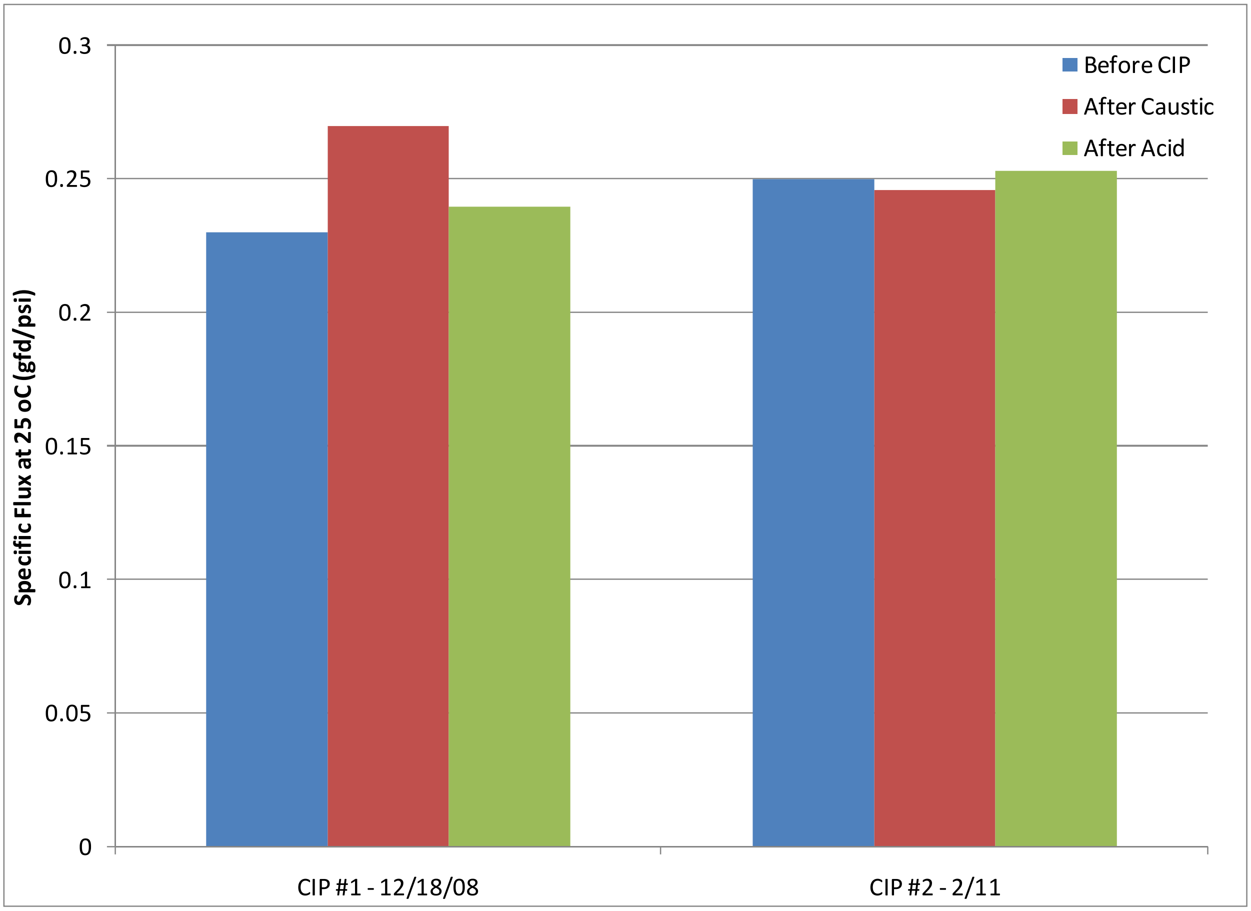


Figure 4-18: NF3 CIP Efficiency – Specific Flux

CIP procedures for NF3 were the same as RO1 and RO2, and are discussed in Section 4.3.1.2. It was determined by the manufacturer that the noise in the specific flux recovery data likely represented the brand new membranes experiencing normal inconsistencies during their break-in period, which could take 30-60 days for certain new membranes.

4.3.3.3 Treated Water Quality

NF3 permeate met all state and federal MCLs that were analyzed for with the exception of the SMCL for pH (6.5-8.5). This is expected in RO permeate, and a normal full-scale facility would have post-treatment for stabilization and pH elevation to protect the distribution system. Therefore, at full-scale, the treated water pH would meet the SMCL.

Project goals for TDS and boron were met, and results are presented in Section 4.3.4.3. However, NF3 had several exceedances above the project treated water chloride goal of 100 ppm. As shown in Section 4.3.4.3, NF3 produced chlorides at or above the goal of 100 ppm for most of Runs 1 and 2.

4.3.4 Comparisons and Recommendations

Due to the wide variation in desalination technologies as discussed in Section 3.2.3, the three RO systems performed quite differently. Table 4-13 provides a summary of the three RO systems, including the flux/recovery, permeability, and salt passage.

Table 4-13: RO System Comparison

	Run 1	**Run 2**	**Run 3**
Description	Baseline run: baseline flux and recovery for each system	Mid-challenge run: increase flux and/or recovery slightly to push each system	Challenge run: push recoveries to max. allowed in membrane models
Duration	11/6-12/17	1/8-2/11	3/13-4/23
Feed Conductivity	10 - 17 mS/cm	8 - 20 mS/cm	0.5 – 3.7 mS/cm
Feed TDS (grab samples)	6.9 – 9.5 g/L	4.1 – 8.6 g/L	0.4 – 2.0 g/L
RO1 – 2:1 Array, Brackish:Seawater			
Flux	12 gfd	12 gfd	12 gfd
Recovery	70%	74%	82%
Feed Pressure (25 deg C)	230 - 320 psi	170 - 270 psi	120 - 170 psi
Permeability (gfd/psi)	0.07-0.10	0.09-0.13	0.10
Permeate TDS	100 mg/L	65 mg/L	13 mg/L
Permeate chloride	55 mg/L	30 mg/L	5 mg/L
Salt passage (25 deg C)	1.5 - 3%	0.5 - 2%	1 - 1.5%
RO2 – Single Stage, Seawater			
Flux	12.7 gfd	14.1 gfd	14.1 gfd
Recovery, single stage	50%	50%	62%
Feed Pressure (25 deg C)	190 - 280 psi	200 - 320 psi	175 - 225 psi
Permeability (gfd/psi)	0.08-0.11	0.07-0.09	0.075
Permeate TDS	18 mg/L	15 mg/L	< 10 mg/L
Permeate chloride	9.2 mg/L	6.9 mg/L	< 4 mg/L
Salt passage (25 deg C)	0.4 - 0.5%	0.4 - 0.5%	0.5 - 1.5%
NF3 – Single Stage, Nanofiltration			
Flux	13.2 gfd	12.9 gfd	12.9 gfd
Recovery, single stage	50%	55%	60%
Feed Pressure (25 deg C)	130 - 190 psi	110 - 160 psi	60 - 160 psi
Permeability (gfd/psi)	0.22-0.25	0.22-0.24	0.19-0.26
Permeate TDS	190 mg/L	170 mg/L	16 mg/L
Permeate chloride	100 mg/L	85 mg/L	20 mg/L
Salt passage (25 deg C)	4 - 6%	4 - 6%	3 - 5%

Each RO system has its own operational advantage, as shown in Table 4-14.

Table 4-14: Desalination System Performance Comparison

	Goal	RO Train No. 1	RO Train No. 2	NF Train No. 3
Description		Two Stage Brackish and Seawater Membranes	Single Stage Seawater Membranes	Single Stage Nanofiltration Membranes
Recovery	High	70-82%	50-62%	50-60%
Specific Flux, gfd/psi	High	Typically 0.1	0.07-0.075	0.19-0.26
Permeate TDS, mg/L	< 500	<10-120	<10-27	<10-220
Permeate Boron, mg/L	< 0.5	0.06-0.48	<0.05-0.2	0.08-0.69
Permeate Chloride, mg/L	< 100	<4-67	<4-11	5-130
Permeate Sodium, mg/L		<2-43	<1-7.8	<2-82
Permeate Turbidity, NTU		<0.05	<0.05	<0.05
Permeate TOC, mg/L		<0.1 – 0.5	<0.1 – 0.5	<0.1 – 0.5
Permeate Iron, mg/L		< 0.011	< 0.011	< 0.011
Permeate Aluminum, mg/L		< 0.021	< 0.021	< 0.021

As noted in Section 6.2.1, the performance of the three desalination trains which were piloted during this study cannot be directly compared because of differences in staging and number of elements per vessel.

The general trends shown in Table 4-14 were hypothesized before testing began, based on the results of membrane models. However, running the systems at pilot scale over the diverse feedwater conditions in the testing period provides specific performance data that can be used to project full-scale capital and operational expenditures, provided in Section 6.0.

4.3.4.1 Comparison of Feed Pressure

The feed pressure is a surrogate for energy use, as the feed pumps are the single biggest energy sink in a desalination plant.

As noted in Section 3.2.3.1, feed pressure of the membranes responded to changing feedwater salinity to maintain the flux and recovery setpoints. Since all three systems received the same feedwater, feed pressure is a basis upon which the systems can be compared. Figure 4-19 shows the feed pressure of each RO system during the three runs.

As shown, Run 3 had generally lower feed pressure because the feed salinity was considerably lower and the spring temperatures were warmer. NF3 had the lowest feed pressure under all conditions. RO2 had a high feed pressure that did not decrease significantly during Run 3. Although RO1 had a high feed pressure in the high-salinity season, it benefitted significantly by the lower salinity in Run 3.

When comparing feed pressure, it is important to note that the recovery on the RO1 system was higher than the other two systems, so the energy used for pumping RO1 feedwater leads to the production of a greater amount of permeate.

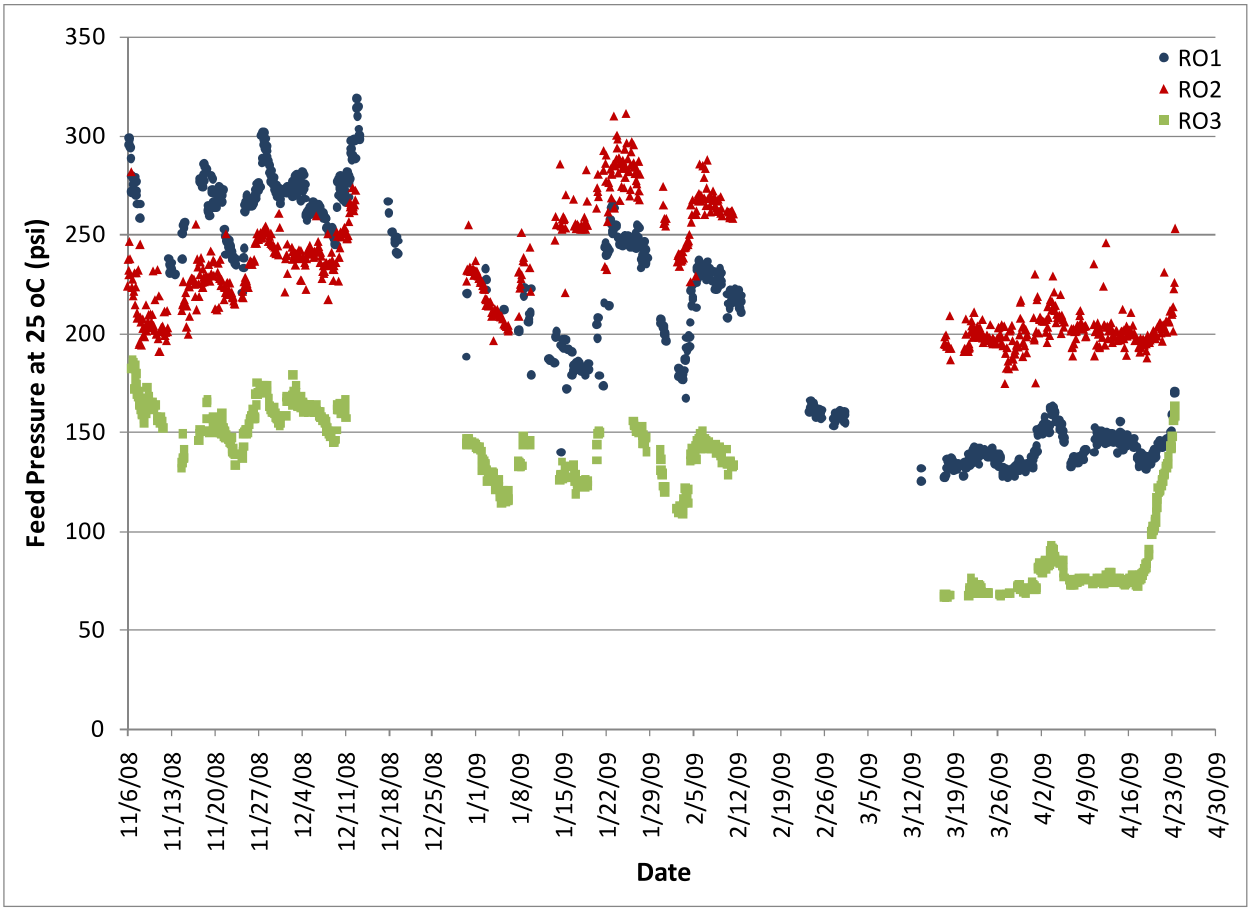


Figure 4-19: RO System Comparison – Operating Pressure

4.3.4.2 Comparison of Permeate Conductivity

Permeate conductivity is a surrogate for dissolved solids in the permeate, as monovalent salts such as chloride or sodium often govern the membrane technology selection and design criteria of a full-scale desalination plant.

Since all three RO/NF systems received the same feedwater, permeate conductivity is a basis upon which the systems can be compared. Figure 4-20 shows the permeate conductivity of each RO/NF system during the three runs plotted on a log scale.

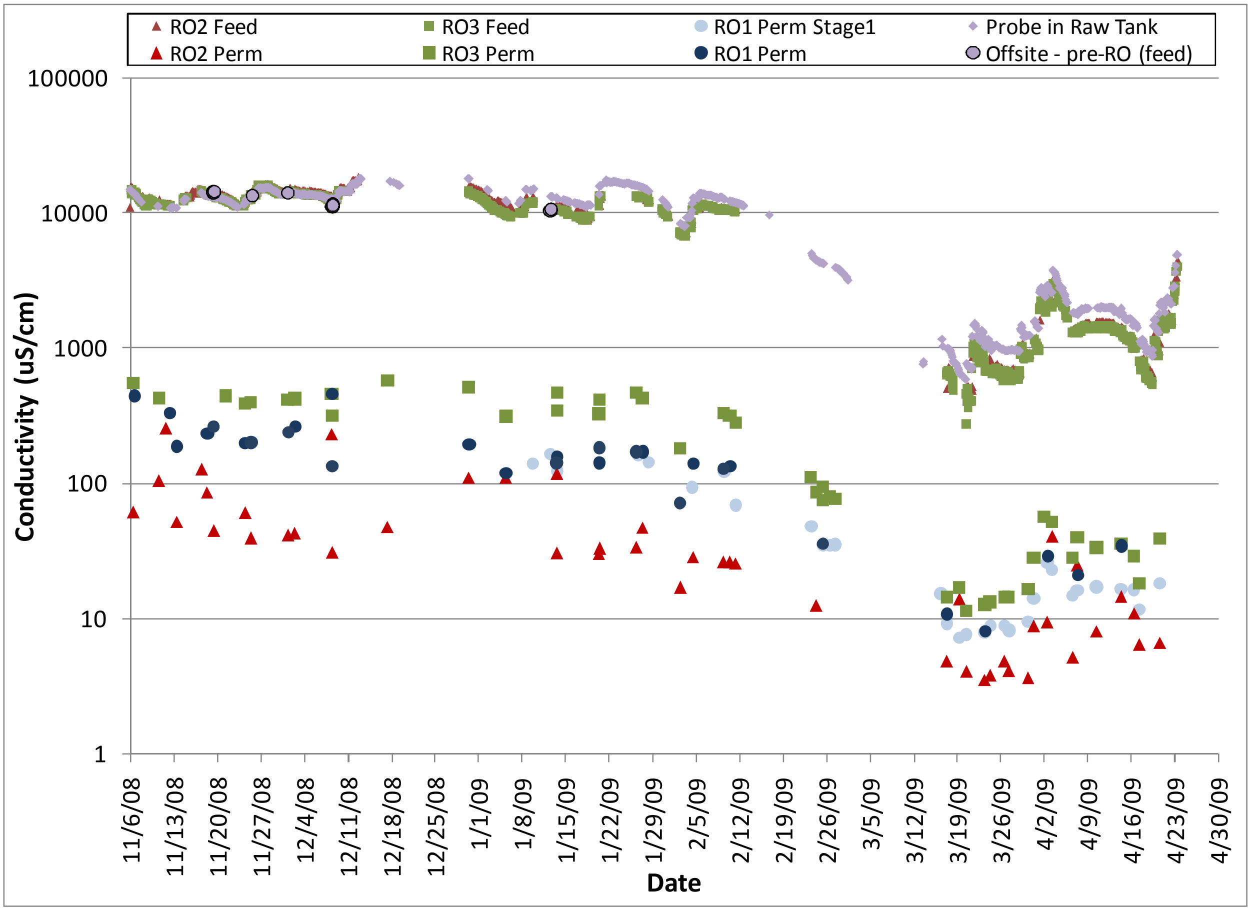


Figure 4-20: RO System Comparison – Conductivity

As shown, Run 3 had generally lower feed conductivity. RO2 had the lowest permeate conductivity under all conditions, consistent between onsite sampling and offsite sampling. NF3 had the highest permeate conductivity at all times, about an order of magnitude greater than RO2. RO1 conductivity was between RO2 and NF3 at all times. And RO1 Stage 1 had a slightly lower conductivity than the combined RO permeate.

All these trends are predictable, since salt passage is a function of the membrane technology, and seawater membranes have lower salt passage than NF, with brackish water in between.

4.3.4.3 Comparison with Treated Water Quality Goals

The treated water quality goals for the project included:

- Meeting all federal and state MCLs for drinking water
- TDS < 500 mg/L
- Chloride < 100 mg/L
- Boron < 0.5-1.0 mg/L

Based upon the offsite sampling that was conducted as part of this pilot study (outlined in the Study Plan), the permeate of all three desalination systems did not exceed any federal or state

MCLs, with the exception of pH which would normally be adjusted in post-treatment at full scale. Furthermore, pH is a secondary or aesthetic standard and is not health based. Detailed data collected during the pilot study is included in **Appendix C** (Pilot Analytical and Operational Data).

Permeate TDS, chloride, and boron levels for each RO system during each run are shown in Figure 4-21, Figure 4-22, and Figure 4-23, respectively. The data is from samples collected on a weekly basis and sent to offsite laboratory for analysis.

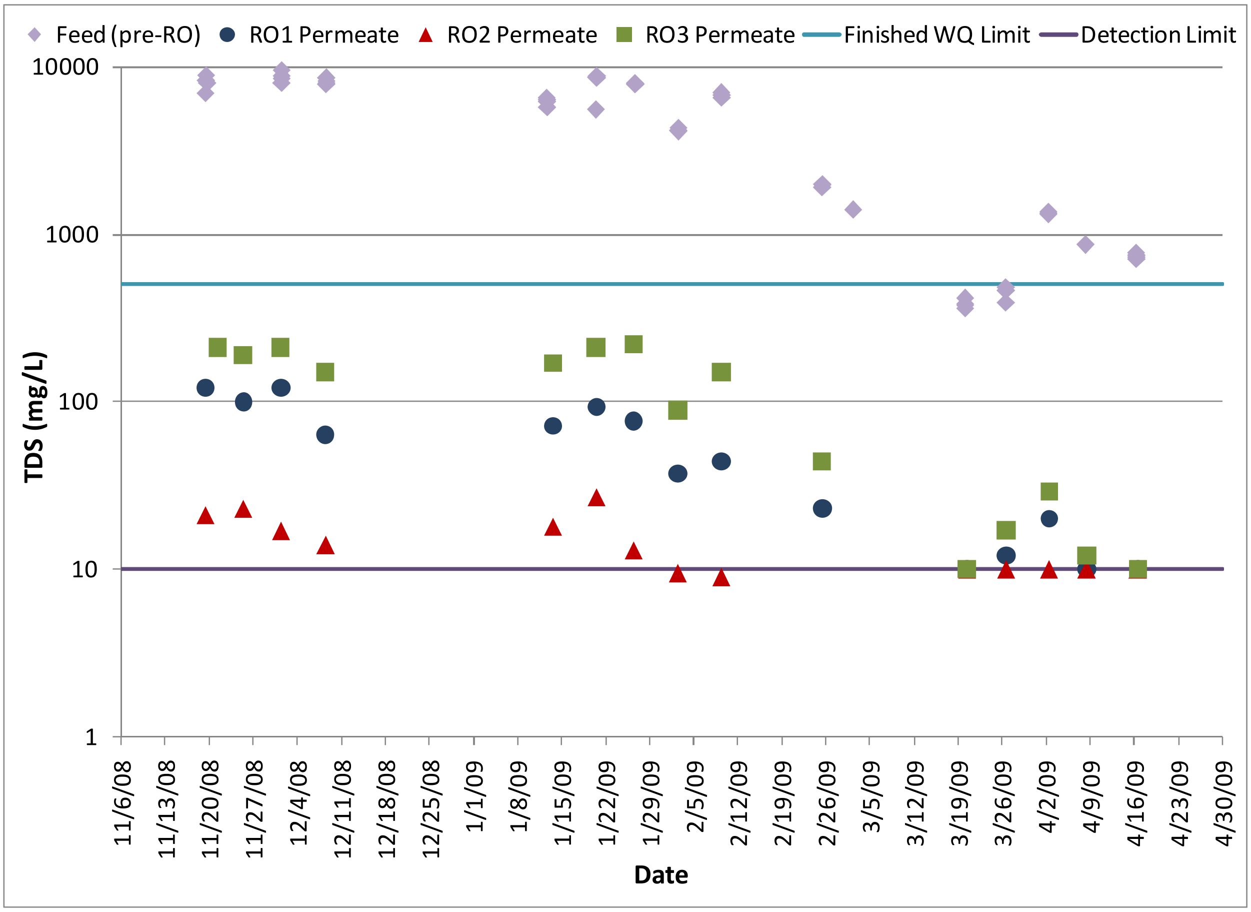


Figure 4-21: RO System Comparison – Permeate TDS

Feed TDS ranged from 6900 to 9400 mg/L during Run 1. During Run 2, TDS in the feedwater was slightly lower, ranging from 4100 to 8700 mg/L. In early- to mid-March, feedwater TDS was less than 500 mg/L, which was the trigger for CCWD to turn on its Mallard Slough Pump Station. By April, when the rains were minimal, the feedwater TDS was creeping back up to around 1000 mg/L. TDS in the permeate was lowest in RO2 and highest in NF3, but met the project goal of <500 mg/L.

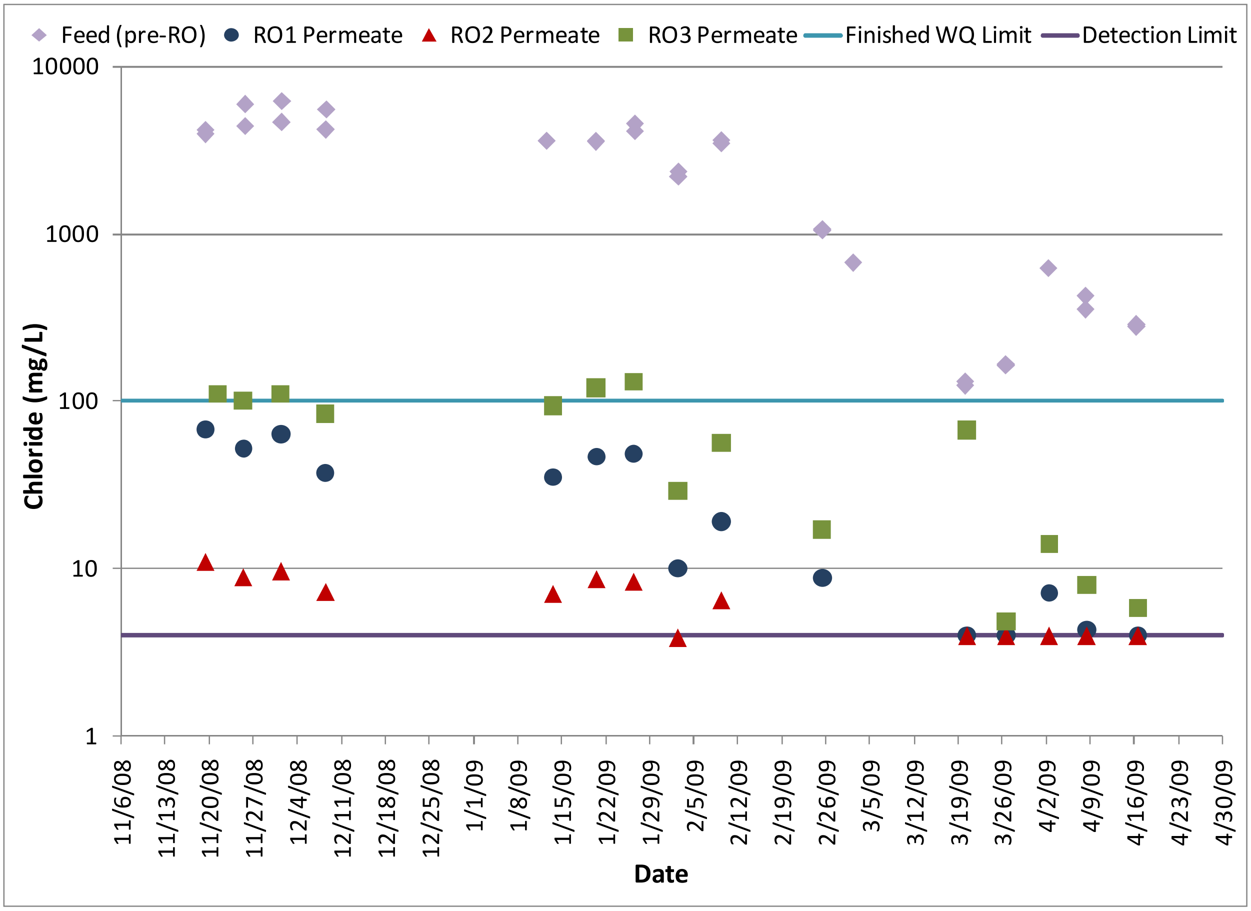


Figure 4-22: RO System Comparison – Permeate Chloride

Feed chloride ranged from 3900 to 6200 mg/L during Run 1. During Run 2, chloride in the feedwater was slightly lower, ranging from 2200 to 4500 mg/L. In early- to mid-March, feedwater chloride was less than 200 mg/L, and by April, when the rains were minimal, the chloride was creeping back up to around 300-400 mg/L.

Chloride in the permeate always met the project goal of <100 mg/L in RO1 and RO2. However, in the dry season (Runs 1 and 2), chloride in NF3 exceeded 100 mg/L on several occasions. Therefore, as a stand-alone system, NF3 would not be an acceptable technology for the full-scale plant.

Feed boron ranged from 1 to 1.3 mg/L during Runs 1 and 2. In early- to mid-March, feedwater boron was approximately 0.15 mg/L, and by April, when the rains were minimal, the boron was creeping back up to around 0.2 mg/L.

0.5 mg/L is the level above which plant species can be affected by boron, whereas 1 mg/L is the limit under consideration by the State of California. Boron in the permeate always met the project goal of 0.5 mg/L in RO1 and RO2. However, in the dry season (Runs 1 and 2), boron in NF3 permeate almost always exceeded 0.5 mg/L. Therefore, as a stand-alone system, NF3 would not be an acceptable technology for the full-scale plant to meet a boron goal of 0.5 mg/L. All three RO systems were able to meet the higher 1.0 mg/L goal during all conditions piloted.

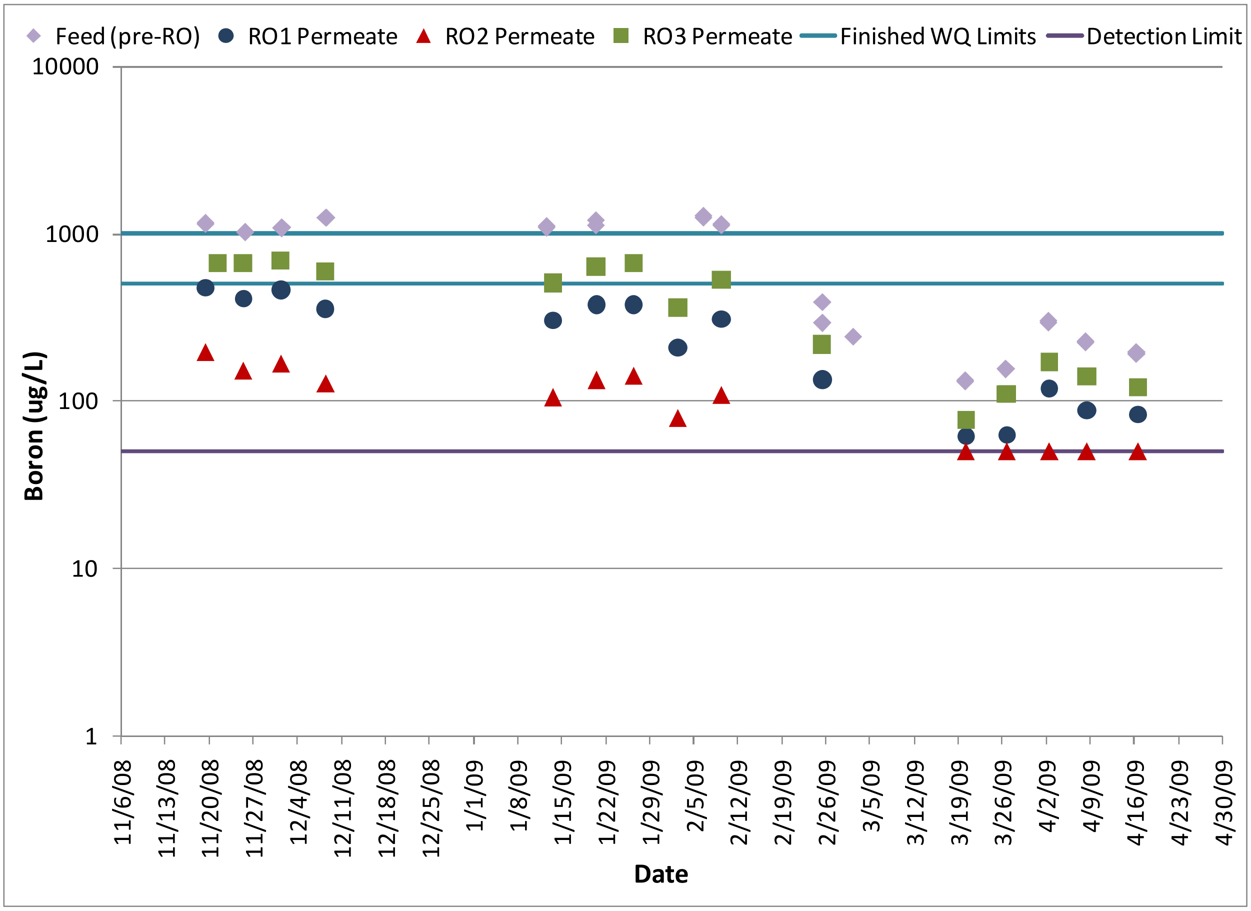


Figure 4-23: RO System Comparison – Permeate Boron

4.3.4.4 Transferability of Pilot Plant Data

At the present time, the final site selection has not occurred. The Feasibility Study (July 2007) indicated that the East Contra Costa Site was ranked highest, above the Ocean Site and the Near Bay Bridge Site. As referenced in the Feasibility Study, however, The East Contra Costa Site is not one specific land parcel but in concept comprises the region between the Suisun Bay to the San Joaquin River, and extending from Mallard Slough to Antioch. Pilot testing was performed at CCWD's existing pump station on Mallard Slough, representing the East Contra Costa Site. It is recognized that the proposed desalination plant could potentially be installed at a variety of specific locations within the East Contra Costa Site. Each of these could present a different water quality.

The San Joaquin River and Suisun Bay flow from east to west through the East Contra Costa Site, although the site is subject to heavy tidal influence from the San Francisco Bay during dry weather. The Mallard Slough location represents the most saline location along the East Contra Costa Site, as the tidal influence diminishes to the east.

Pilot testing was conducted in order to develop preliminary design criteria for a full-scale facility at Mallard Slough and to test viability of the proposed treatment processes for water which might be representative of this extended East Contra Costa Site. However, some of the data collected during the pilot may be transferrable to other potential locations for full-scale applications within the extended East Contra Costa Site, at the Near Bay Bridge Site, or at the Ocean Site.

While a regulatory investigation was not conducted, Table 4-15 has been compiled to provide an estimate of what data could potentially be transferred. Obviously water quality will change significantly as the proposed desalination plant location being considered moves further from Mallard Slough.

Table 4-15: Transferability of Pilot Plant Data

Data Collected in BARDP Pilot Plant Study	Needed to Meet Pilot Plant Goals	Is Data Transferrable to:			Comments
		Full-Scale System at Mallard Slough	Full-Scale System at East CC Site	Full-Scale System at Bay/Ocean Site	
Pilot Plant Permitting Studies	YES	MAYBE/ PARTIAL	NO	NO	Regulatory permits were not required for the pilot plant. Each location will be confronted w/ unique permitting issues and challenges. Data collected during pilot test will be used to the extent possible for applying for permits.
Source Water Characteriz-ation	YES	YES	NO	NO	Water quality at other East Contra Costa locations will vary due to influence of the freshwater vs. tidal effects. Bay and Ocean sites are unlike the East Contra Costa brackish conditions.
Pretreatment Flux/ Recovery Testing	YES	YES	NO	NO	Water quality at other East Contra Costa locations will vary due to influence of freshwater vs. tidal effects (seasonal turbidity indicates the relative influence of very different feedwaters). Bay and Ocean sites are unlike the East Contra brackish conditions.
Pretreatment solids characteriz-ation	YES	YES	LIKELY	NO	East Contra Costa locations will likely exhibit similar feed water TOC and turbidity as Mallard Slough, but the Bay/Ocean sites would not
RO/NF Flux/ Recovery Testing	YES	YES	NO	NO	Water quality at other East Contra Costa locations will vary due to influence of freshwater vs. tidal effects (seasonal salinity indicates the relative influence of very different feedwaters). Bay and Ocean sites are unlike the East Contra brackish conditions.

Data Collected in BARDP Pilot Plant Study	Needed to Meet Pilot Plant Goals	Is Data Transferrable to: Full-Scale System at Mallard Slough	Full-Scale System at East CC Site	Full-Scale System at Bay/Ocean Site	Comments
Cleaning Efficiency	YES	YES	MAYBE	NO	Cleaning efficiencies depend on foulants in the feedwater. Since the feedwater quality varies significantly between sites, it is not likely that the cleaning efficiency data is transferrable.
Treated Water Quality	YES	YES	MAYBE	NO	Treated water quality is dependent on feedwater quality and membrane rejection properties. Since feedwater quality varies significantly between sites, it is not likely that the treated water quality is transferrable.
Intake Evaluation	YES	YES	LIKELY	LIKELY	Intake types are generally applicable at each site, with the exception of subsurface intakes, which depend on local geologic conditions.
Source Water Biological Impacts	YES	YES	MAYBE	NO	Highly dependent on intake system type and the local biota from site to site.
Concentrate Toxicity	YES	YES	LIKELY	MAYBE	Highly dependent on receiving water quality and native species.
Treated Water Compatibility	YES	YES	LIKELY	MAYBE	Post treatment chemical addition can adjust the permeate to meet most treated water quality conditions. Site specific data would be useful, particularly for an ocean water desalination site.
Full-Scale Cost Estimates	N/A	YES	NO	NO	Differing water qualities will significantly impact capital and operating cost estimates.

Actual transferability of data needs to be verified through discussions with each individual regulatory agency that will be involved in permitting a full-scale facility. Additional pilot-scale activities are likely to be required if the full-scale facility is designed at any site other than Mallard Slough. The additional pilot-scale activities would target collection of data that is deemed to be non-transferrable.

4.3.4.5 Recommended Configuration for Full Scale

Based upon the pilot data, RO1 appears to be the most suitable system for full-scale because it achieves a high recovery, while meeting water quality goals. RO2 has the best permeate water quality, but permeate produced by RO1 is adequate for this project and meets agency goals. NF3, although having the lowest energy use, does not meet the project's goal for chloride or boron in the dry season.

Given the widely varying salinity of the source water as encountered during the study, the single stage pilot systems are probably not representative of a full-scale system which could be optimized for this project. Recovery and energy efficiency will likely be improved by utilizing multiple stages, as demonstrated with RO1, or by combining single stage membrane vessels to achieve low energy use during low salinity months and maximum water quality during dry months. Optimizing a single stage system for both high and low TDS would be quite challenging and would result in inefficiencies at extreme operational ranges.

Therefore, the full-scale evaluation focuses on two options:

Alternative No. 1: RO Train No. 1, as piloted.

- 2:1 array
- Brackish water RO membranes in Stage 1 and seawater RO membranes in Stage 2. All feed water is pumped first to the brackish water membranes, with only concentrate being directed to the second stage seawater membranes.
- Interstage boosting with VFD.
- Approximately 70% total system recovery during high TDS conditions.

Alternative No. 2: A hybrid plant that has independent trains for RO Train No. 2 and NF Train No. 3.

- Two single-stage membrane systems operating in parallel: NF RO membranes in one train, and seawater membranes in the second train.
- Feed water split between the membrane systems to meet treated water goals, with the seawater RO train handling approximately 70% of the feed flow during the dry seasonal periods due to its better water quality, and the NF train treating nearly 100% of the feed flow during lower TDS periods due to its lower feed water pressure requirement.
- NF concentrate partially recovered by blending with filtrate being pumped to the seawater RO train.
- Approximately 58% total system recovery during high TDS conditions.

The full-scale evaluation is presented in Section 6.0.

4.4 Pathogen Removal of the Treatment Systems

Pathogen removal was not evaluated for any of the pretreatment or desalination systems as part of the pilot study, primarily because Mallard Slough has not been established as the source for the proposed desalination plant. Full-scale disinfection requirements will depend on the final site selection.

For the removal of microbial contaminants in drinking water treatment, the DPH currently requires a multi-barrier approach that achieves 4-log removal or inactivation of virus and 3-log for *Giardia*. The Long Term 2 Enhanced Surface Water Treatment Rule determines the log-removal requirements for Cryptosporidium based upon the source water quality. It is likely that the source water quality at Mallard Slough would fall into Bin 2 under this rule, requiring 3-log *Cryptosporidium* inactivation. However, comprehensive monitoring (watershed sanitary survey) must be performed once the full-scale site is selected to establish the microbiological water quality of the source water. Several years of data may be requested, so this testing should begin as soon as possible. It should be noted that the DPH reserves the right to increase the pathogen removal and inactivation requirements based on source water quality.

The DPH accepts alternative filtration technologies as part of their Drinking Water Program, and prescribes log-removal credits for *Giardia*, *Cryptosporidium*, and virus based upon third party testing for any accepted technology. Both pretreatment technologies that were pilot tested have been accepted by the Drinking Water Program, and testing has established the following pathogen removal credits for the two pretreatment systems which were piloted as part of this project:

	Log Removal Credits		
	Virus	*Giardia*	*Cryptosporidium*
Siemens S10V (submerged)	1.5	4	4
Norit SXL225 (pressurized)	4	4	4

In addition to the pretreatment credits, the DPH also gives credit for RO systems based on TDS removal. If an RO system removes 99% or more of the TDS in the feed water, then a 2-log removal credit is given for virus, *Giardia* and *Cryptosporidium*. With this removal credit, up to 6-log removal of these three pathogens can be obtained. However, even though the total removal credits could be greater than those required, the DPH still requires a minimum 0.5-log inactivation of *Giardia* and 2-log inactivation of virus (the same as required for conventional treatment). Under special circumstances, the 0.5-log *Giardia* inactivation requirement has been waived in the past, but the DPH, as a matter of policy established in April 2000, will not waive the 2-log virus inactivation requirement. For the purposes of this report, it is assumed that 0.5-log *Giardia* and 2-log virus inactivation will be required.

It should be noted that any treatment credit will be negated by DPH if the treated water is blended with raw or partially treated water, as is being considered for this project. At the point of downstream treatment, the pathogen removal and inactivation requirements of the source(s), with which the desalination plant product water is blended, must be met.

Pathogen goals are listed in Table 2-1. Additional disinfection is required at the full scale to fulfill the remaining pathogen removal requirements not provided by the membrane systems and to achieve the multi-barrier approach. The combined membrane pretreatment system and RO process will be capable of meeting the multi-barrier regulatory requirement. Post-treatment disinfection using free-chlorine, and establishing a secondary disinfectant residual, will assist in meeting disinfection goals.

5.0 RELATED STUDIES

Supplemental studies in support of the pilot study were conducted by the MWH team to investigate key issues expected to have a significant effect on project success. These studies are included with this pilot report in the following appendices:

Appendix E Treated Water Compatibility Analysis

Appendix F Biological Sampling and Impingement/Entrainment Analysis

Appendix G Concentrate Toxicity Testing Analysis

5.1 Assess Treated Water Compatibility

Treated water served to customers in the San Francisco Bay region originates from several different sources, including Sierra snowmelt, Sacramento River delta, and local runoff. A new regional desalination facility will introduce yet another supply, one which is more saline than current sources and representing a different water chemistry than the agencies have encountered in the past.

Permeate from the desalination facility will be quite low in alkalinity and mineral content and is poorly buffered against pH changes. It must be chemically adjusted prior to blending with existing waters within existing transmission and distribution systems to assure compatibility and to avoid potential corrosion and unacceptable aesthetic impacts for customers of the four agencies
.
It is anticipated that the BARDP treated water will be pumped to the EBMUD Mokelumne Aqueduct (untreated), the CCWD Multipurpose Pipeline (treated), or the Contra Cost Canal (untreated). NF and RO permeate is characteristically low in pH, with strongly negative values of Calcium Carbonate Precipitation Potential (CCPP) and Langelier Saturation Index (LSI), indicating that the RO permeate is very corrosive and will require post-treatment stabilization.

5.1.1 Study Objectives

A series of tests were conducted during the pilot study to evaluate potential benefits of two common post-treatment stabilization techniques: lime plus carbon dioxide addition, and a calcite filter. Testing also served to verify that the stabilized permeate can be rendered compatible with existing water supplies and would be able of maintaining a disinfectant residual. Specific objectives were as follows:

1. Identify and test two post-treatment options stabilization effectiveness as determined by LSI, CCPP, alkalinity, calcium content, pH, and turbidity.

2. Verify compatibility of the stabilized RO permeate with EBMUD Mokelumne Aqueduct and CCWD Multipurpose Pipeline water.

3. Assess disinfection within stabilized RO permeate containing chloramines and in water blended with chloraminated CCWD water. DBP formation was also evaluated.

4. Assess disinfection within stabilized RO permeate containing free chlorine and in water blended with chlorinated EBMUD water. DBP formation was also evaluated.

5.1.2 Approach

RO permeate samples were collected from the RO Train No.1, from the EBMUD Aqueduct #2, and from CCWD Multiple Purpose Pipeline. All samples were analyzed for water quality parameters (pH, turbidity, conductivity, alkalinity, etc.). CCPP and LSI values were calculated from the measured parameters.

Tests were conducted on the RO permeate sample using liquid lime to increase alkalinity, hardness and pH levels. Carbon dioxide gas was added simultaneously and sequentially to reduce the pH and achieve targeted alkalinity and corrosion index parameters.

Calcite filter tests were conducted permeate feed as well utilizing a continuous flow-through calcite filter. Acid feed upstream of the filter was used to reduce pH to as low as 5.0 to promote sufficient calcite dissolution. Figure 5-1 depicts the calcite filter system used in the tests.

Figure 5-1: Calcite Filter Pilot System

Filter effluent samples were collected and analyzed for pH, turbidity, conductivity, alkalinity, hardness, and other parameters. Downstream of the calcite filter, caustic soda solution was added to increase the filter effluent pH to achieve positive CCPP and LSI values.

The following blending ratios were assumed for this project based on consultations with the staff from the four agencies:

- Permeate: EBMUD Mokelumne Aqueduct water 1:2 blend ratio
- Permeate: CCWD Multipurpose Pipeline water 1:1 blend ratio

Stabilized RO permeate was chlorinated or chloraminated to match the disinfectant residuals in the EBMUD and CCWD water. Disinfected RO permeate was subsequently blended with water from these two sources. Disinfectant stability tests were then performed as described herein.

5.1.3 Compatibility Test Findings

Initial CCPP and LSI values based on the field measurements were found to be highly negative, indicating that RO permeate would be very corrosive to many pipeline materials. EBMUD Aqueduct water and CCWD Multipurpose Pipeline water also exhibited moderate corrosive characteristics based on CCPP and LSI calculations as evidenced in Table 5-1.

Table 5-1: Calculated Corrosivity Parameters

Parameter	RO Permeate	EBMUD Aqueduct #2	CCWD Multi-Purpose Pipeline
pH	5.76[1]	7.83	7.71
Total Dissolved Solids[3] (mg/L)	24[2]	48	380
CCPP (mg/L) at 25 °C	-14.4	-3.7	-3.1
LSI at 25 °C	-5.6	-1.2	-0.3

Note 1. RO permeate pH during the calcite filter test on January 30' 2009 and January 31, 2009 was lower, ranging from 5.50 to 5.65.
Note 2. Lab analyses were conducted for RO permeate on a sample collected January 31, 2009.
Note 3. Lab analyses; for CCPP and LSI calculations, field analyses were used where available because they comprised a complete data set for each water tested.

5.1.3.1 Liquid Lime Tests

For the liquid lime tests, stabilized RO permeate was produced with alkalinity and calcium hardness values of 50 mg/L as $CaCO_3$, for a liquid lime dose of 40 mg/L (with CO_2 addition). Based on the measured parameters, CCPP and LSI values were 4.2 and 0.5, respectively, which met the desired targets and suggested that the liquid lime stabilized RO permeate would not be corrosive. A pH decrease occurred within three hours after CO_2 addition was finished. Caustic soda was added to the RO permeate to restore pH.

For the calcite filter tests with and without upstream sulfuric acid for pH adjustment, the stabilized RO permeate alkalinity and calcium hardness ranged between 35 mg/L and 40 mg/L as $CaCO_3$, with slightly lower values being observed where pH adjustment was not employed. In both cases, calcite filter effluent needed additional pH adjustment to achieve positive CCPP and LSI values. Stabilized RO permeate with and without acid had similar CCPP and LSI values as long as pH was increased after the calcite filter. Data may be found in **Appendix E**.

Stabilized RO permeate blending could not compensate for the negative CCPP and LSI values of EBMUD and CCWD waters as observed during this bench test study. It is expected that blending of these waters with stabilized permeate would drive the LSI and CCPP values to somewhat less corrosive values; however, this was not attempted during the pilot study and should be subjected to further investigation.

5.1.3.2 Disinfectant Tests

Stabilized RO permeate in liquid lime tests was chlorinated and chloraminated to match the disinfectant in EBMUD Aqueduct #2 and CCWD Multiple Purpose Pipeline waters, respectively. Decay plots are illustrated in Figure 5-2 and Figure 5-3.

Chlorination tests of stabilized RO permeate showed that a chlorine dose of 0.3 mg/L resulted in a free chlorine residual of 0.16 mg/L within two hours. Residual did not reach complete stabilization during the 2 ½ hour test duration, as indicated in the figures.

Initial chloramination results of stabilized RO permeate indicate that a chlorine dose between 2.5 mg/L and 4.5 mg/L would be needed to reach a target total chlorine residual of 2.5 mg/L. Further testing is necessary to determine a better approximation of total chlorine dose to avoid breakpoint chlorination. Additionally, significant decreases in chlorine and total ammonia concentrations were observed when CCWD water was blended with chloraminated RO permeate during the bench-scale study, which should be investigated further to observe whether this behavior would be duplicated at full scale.

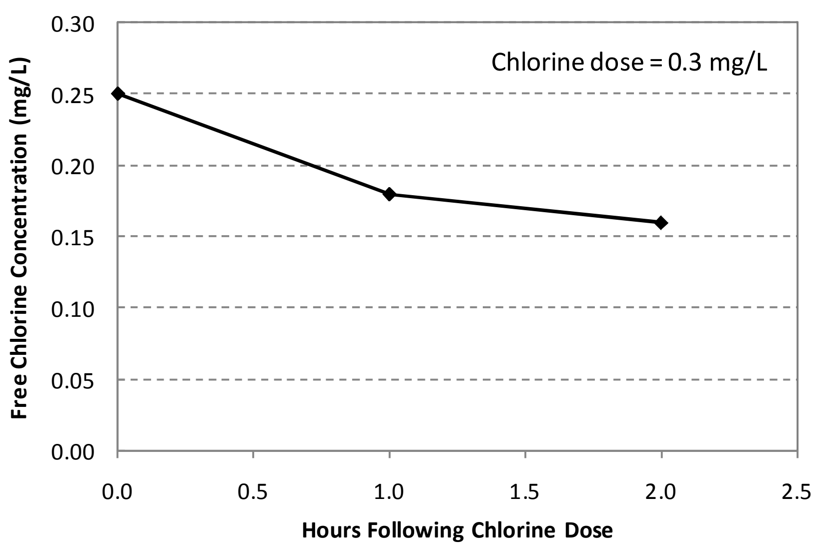


Figure 5-2: Free Chlorine Degradation Curve for Stabilized RO Permeate

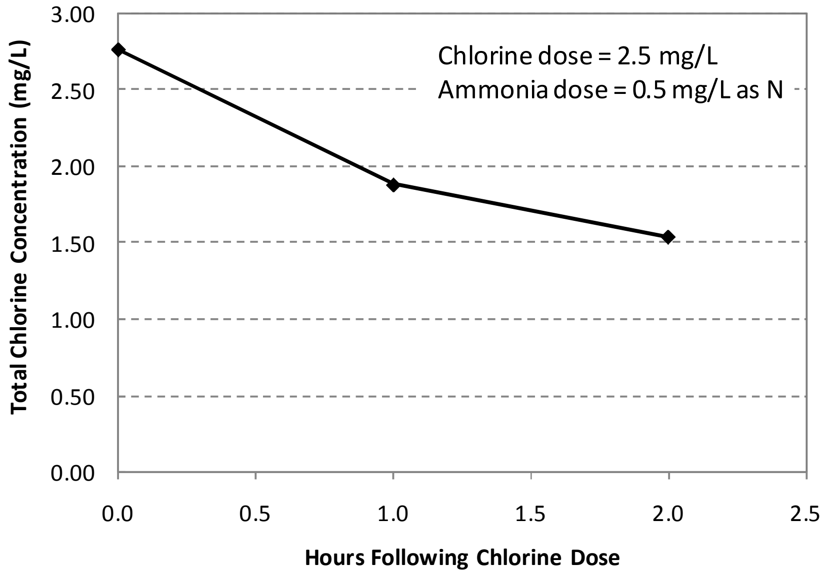


Figure 5-3: Total Chlorine Degradation Curve for Stabilized RO Permeate

5.1.3.3 DBP Tests

For the time period (24 hrs) and residuals tested, RO permeate blending with EBMUD water (having 0.3 mg/L free chlorine) caused a slight increase in DBPs. When blended with CCWD water (having 2.5 mg/L chloramines), however, RO permeate caused a decrease in DBPs.

These conditions are illustrated in Figure 5-4 and Figure 5-5. Overall, DBPs in the blends were well below the primary MCLs for TTHM and HAA5. The following species are referenced in these graphs:

- $CHCl_2Br$ – Bromodichloromethane
- $CHCl_3$ – Chloroform
- $CHBr_3$ – Bromoform;
- $CHClBr_2$ – Dibromochloromethane
- TCAA - Trichloroacetic Acid
- MCAA - Monochloroacetic Acid
- MBAA - Monobromoacetic Acid
- DCAA - Dichloroacetic Acid
- DBAA - Dibromoacetic Acid

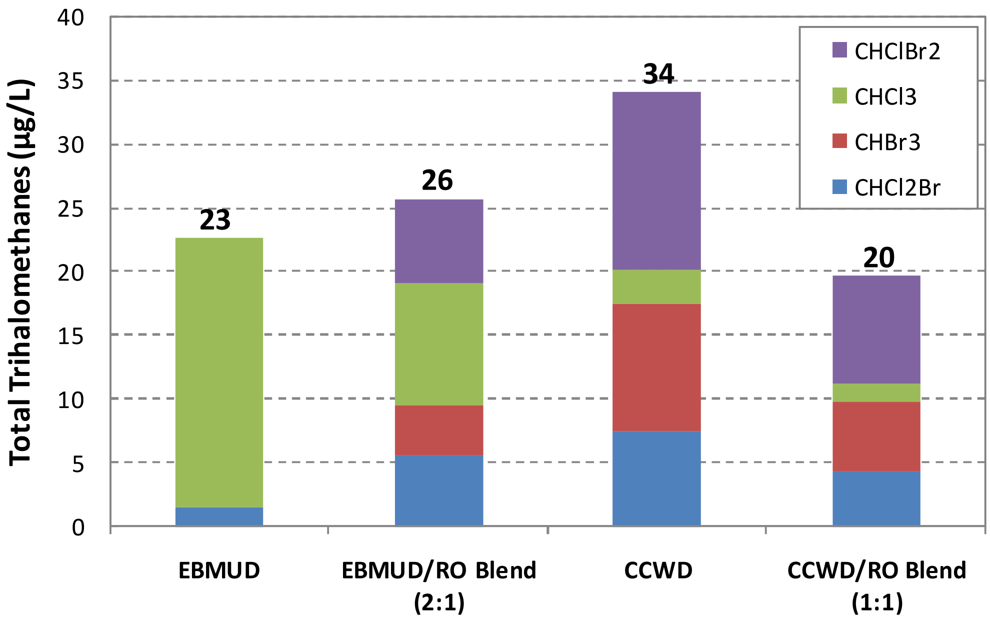


Figure 5-4: TTHM Formation

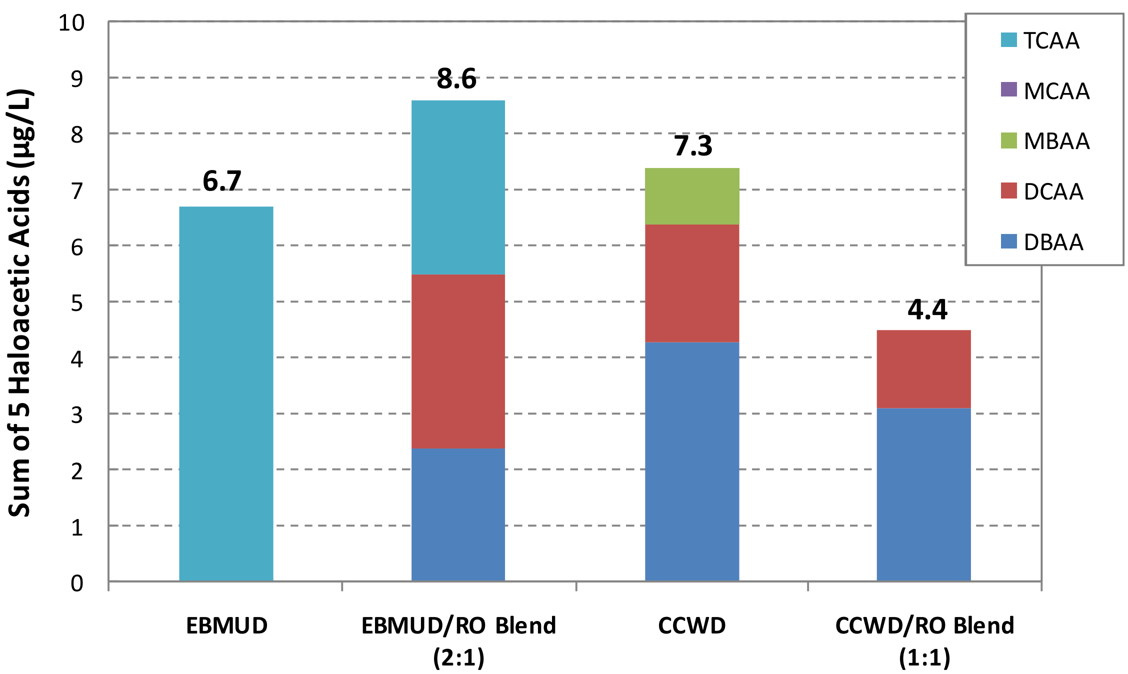


Figure 5-5: HHA5 Formation

5.2 Assess Source Water Biological Impacts

A treatment facility relying on a source water surface intake will impact native aquatic species due to impingement onto intake screens and entrainment within the treatment works. Biological sampling was completed while the pilot study field operations were being conducted, and at other periods after the pilot plant was decommissioned, to identify:

- The species composition and abundance of larval fishes and fish eggs entrained by the pilot plant.
- The local species composition and abundance of entrainable larval fishes and fish eggs in the Mallard Slough source water.
- The potential impacts of entrainment losses on larval fish and fish eggs due to operation of a full-scale feedwater intake system.

Entrainment occurs when organisms smaller than the openings in the 3/32-inch intake screens (e.g., larval fishes) are drawn into the feedwater intake system. Entrainment sampling was conducted behind the intake screens and source water sampling was conducted in Mallard Slough. The entrainment and source water studies focused on larval fishes and fish eggs whose adult populations might be affected by operation of the feedwater intake system.

Figure 5-6: Entrainment Sampling Equipment

Six entrainment surveys and four source water surveys were conducted over a 12-month period (November 5, 2008 to October 9, 2009). Sampling occurred at least once during each run as summarized:

Pilot Plant Run	Season	Entrainment Survey	Source Survey
Run 1	Dry Season	November 5, 2008	November 5, 2008
Run 1	Dry Season	December 16, 2008	December 16, 2008
Run 2	Dry Season	February 20, 2009	
Run 3	Wet Season	March 6, 2009	
Not operating	Summer Conditions	July 16-17, 2009	July 16-17, 2009
Not operating	Fall Conditions	October 8-9, 2009	October 8-9, 2009

Runs 2 and 3 occurred during the sensitive fish period (January through June) when access to the slough for source survey sampling is limited. As a result only one entrainment survey was completed for each run and no source water surveys were completed. Two additional entrainment and source water surveys were conducted in July and October during times when the pilot plant was not operating. The purpose of these surveys was to gather information from the summer and fall time periods.

The results of six entrainment surveys for larval fish and fish eggs show the following:

- Three taxa of larval fishes were collected during entrainment sampling: prickly sculpin, longfin/delta smelts, and bluegill/redear sunfishes. Prickly sculpin are an abundant native species. Bluegill/redear sunfishes are abundant introduced species. Both longfin smelt and delta smelt are listed species under the California Endangered Species Act. Fish were only detected during the sensitive fish period of January through June, as shown in Figure 5-7. (Note that inverted triangles indicate that no fish were collected.)
- No fish eggs were collected in entrainment or source water samples during the entire study.

The species composition of larval fishes collected during the entrainment and source water sampling was consistent with published life history information for species found in Suisun Bay, along with documented collections from other studies conducted in Suisun Bay (PG&E 1981, Moyle 2002, Tenera 2009, IEP/CDFG survey results). The estimated small annual loss of adult prickly sculpin and bluegill/redear sunfishes is unlikely to affect adult populations. However, longfin and delta smelts are listed species. Delta smelt is listed as a threatened species under the Federal Endangered Species Act and is listed as endangered under the California Endangered Species Act. Longfin smelt is listed as a threatened species under the California Endangered Species Act. The spawning times of both species vary between years depending on water temperature and salinity. January through June has typically been designated as the sensitive period, although longfin smelt larvae may be present during December of some years.

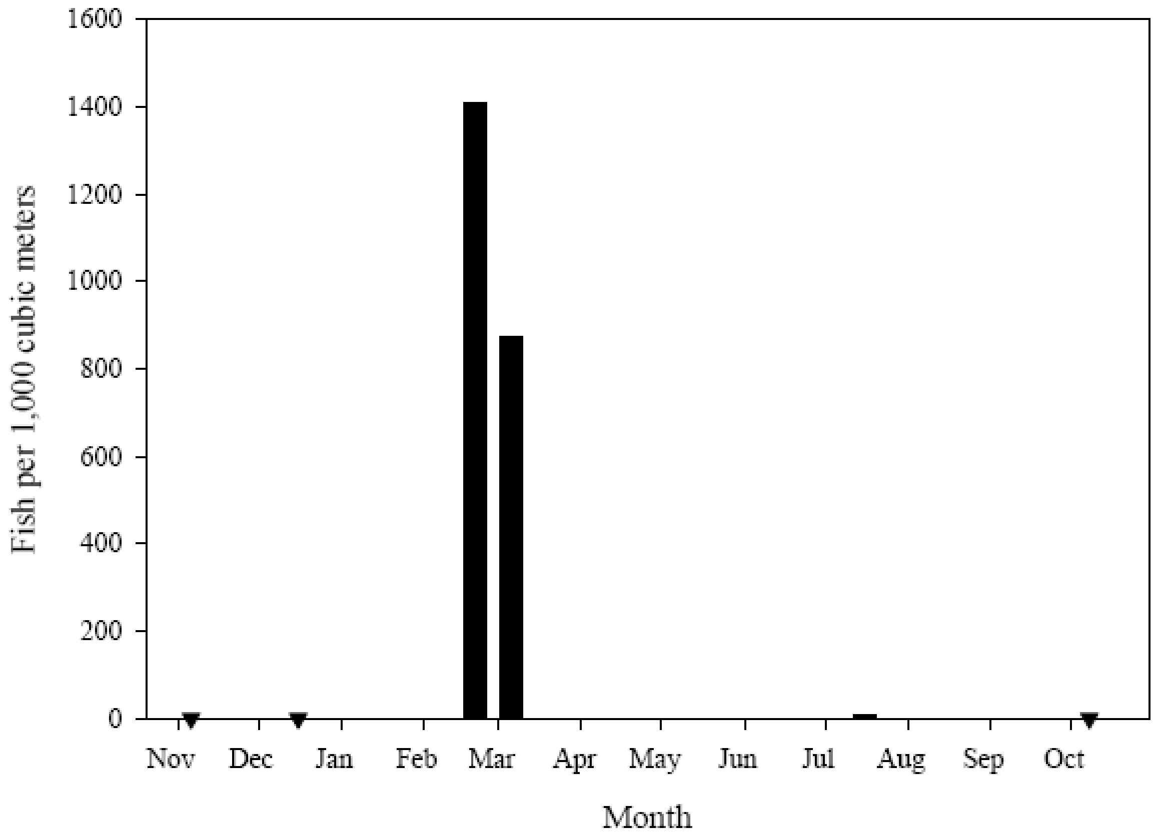


Figure 5-7: Total Fish Collected[1]

A summary of the estimated entrainment effects for all the entrained fishes for a 25 mgd intake, is provided in Table 5-2. These values are based on analyses using the Fecundity Hindcast (*FH*) model and the Adult Equivalent Loss (*AEL*) model. These models require species-specific estimates of age, growth, fecundity, and survivorship of various life stages. Demographic data were available to allow at least one of the two modeling approaches to be applied to two of the three fish taxa. AEL values for prickly sculpin could not be computed due to the absence of any published larval mortality information. The estimated low number of entrained bluegill/redear sunfishes (n=1,771) was less than the fecundity values of bluegill; therefore, FH and AEL values were not computed. Proportional entrainment estimates (PE) could not be calculated because this calculation requires both entrainment and source data. Entrained fish were only detected during the sensitive fish period, when source water samples could not be collected.

[1] Inverted triangles indicate that no fish were collected.

Table 5-2: Potential Larval Entrainment and Equivalent Losses of Adult Fish for a 25 mgd Facility

Taxa	Estimated Annual Larval Entrainment	Adult Equivalent Losses	
		2FH[2] Estimate	*AEL* Estimate
Prickly sculpin	990,605	510	Note 1
Longfin/Delta smelts	36,777	39	25
Bluegill/Redear sunfishes	1,771	Note 1	Note 1

Note 1. Unavailable information or value that could not be computed.
Note 2. 2*FH* (number of estimated females x 2) values are presented to provide comparison to *AEL* estimates, which include both males and females.

A full-scale desalination facility at the MSPS may require preparation of a Biological Assessment, which would be reviewed by the United States Fish and Wildlife Service (USFWS) and the California Department of Fish and Game (CDFG). USFWS may issue a Biological opinion and CDFG may have permit requirements. The Biological Assessment should contain the following information:

- A description of the project including operations,
- The listed species potentially affected,
- An analysis of impacts to species,
- Proposed minimization and mitigation measures, and
- Plans to monitor compliance.

Due to the presence of listed salmonids and green sturgeon in Suisun Bay, consultation would be required with National Marine Fisheries Service (NMFS). Since entrainment impacts are highly unlikely to occur with these species it may be possible to pursue a “not likely to adversely affect” determination. However, if any dredging is required, it would need to occur within the established dredging work windows, and it would be necessary to consult with NMFS on the presence of green sturgeon and listed salmonids. In addition, the MSPS is surrounded by essential fish habitat for all Pacific Coast salmon, starry flounder, and northern anchovy, as designated by the NMFS. An assessment of full-scale MSPS operations on essential fish habitat would be required by NMFS.

5.3 Assess Concentrate Toxicity

One of the major potential issues associated with potential full-scale desalination operations is the discharge of the RO and NF backwash and concentrate streams which are produced during normal operation. While backwash can be thickened and dewatered, RO and NF concentrate cannot be treated and must be disposed of in an environmentally acceptable manner.

Since concentrate is often directed to a receiving water as a point source discharge through the NPDES permitting process, identifying the potential toxicity of this stream was evaluated during the pilot study. Toxicity of the pilot plant concentrate was identified by initial testing using a series of sensitive aquatic species. This study was conducted for dry-season conditions representing highest ambient salinity, and for wet-season conditions representing highest contaminant concentrations associated with storm runoff.

5.3.1 Study Objectives

Specific concentrate toxicity objectives:

- Evaluate concentrate produced by the pilot plant during dry period (high salinity, low contaminant concentration) and wet period (low salinity, high contaminant concentration) conditions.
- Assess algal growth toxicity (*Thalassiosira pseudonana*).
- Assess crustacean survival and growth toxicity (*Americamysis bahia*) (see Figure 5-8).
- Assess fish survival and growth toxicity (*Menidia beryllina*) (see Figure 5-7).

Figure 5-8. Photo of *Americamysis bahia*[2]

Figure 5-9: Photo of *Menidia beryllina*[3]

These species were selected for this study as being representative of aquatic organisms within the Delta.

[2] Photo courtesy of Marinco Bioassay Laboratory, Inc.
[3] Photo courtesy of Marinco Bioassay Laboratory, Inc.

5.3.2 Approach

Desalination concentrates both salts and contaminants into the system concentrate. Toxicity is consequently influenced by seawater salinity, due to the proximity of the pilot plant to Suisun Bay, and by concentration of contaminants originating in local runoff. Concentrate testing focused on evaluating its potential toxicity to several test organisms during extremes in salinity and contaminant input from source water to differentiate the separate effects.

Thalassiosira pseudonana was exposed to varied treatment concentrations of concentrate for 96 hours (2.5%, 5%, 10%, 25%, 50%, and 100%). Effects on cell growth were assessed to identify concentrate-induced toxicity. A reference toxicity test was performed to determine the sensitivity of the diatoms to toxic stress, by exposing *Thalassiosira* to varied concentrations of potassium chloride (KCl), also for 96 hours. Data were analyzed to determine key dose-response point estimates.

Americamysis bahia was exposed to a series of concentrate dilutions for seven days. After the test, effects on survival and growth were assessed to determine concentrate-induced impairment. A reference toxicity test was performed to determine the sensitivity of the mysids to toxic stress, by exposing *Americamysis* to serial dilutions of a toxicant for seven days. Data were analyzed to determine key dose-response point estimates.

Chromium was used as the toxicant for the dry season sample. For safety reasons, KCl was used for the wet season sample. Using different toxicants in the reference tests has no effect on sample tests.

Menidia beryllina larvae were also exposed to a series of concentrate dilutions for seven days. After the test, effects on survival and growth were similarly evaluated. A reference toxicity test was performed to determine the sensitivity of the fish to toxic stress, by exposing *Menidia* to serial dilutions of KCl. The test response data were analyzed to determine key dose-response point estimates.

Specific test procedures are described in the Technical Memorandum in **Appendix G**.

5.3.3 Concentrate Toxicity Findings

Concentrate samples are described in Table 5-3. Concentrate collected in the “dry” season at high tide had a salinity of 17,700 mg/L and conductivity of 29.35 mS/cm, whereas the “wet” season sample at low tide had salinity of 7,000 mg/L and 12.42 mS/cm.

Table 5-3: Concentrate Sample Description

Season[1]	Target Salinity	Target Contam.	Date	Time	Tidal Height[2]	pH	DO (mg/L)	Salinity (ppt)	Cond. (mS/cm)	Total Ammonia (mg/L N)
"Dry"	High	Low	11/14/08	1444-1451	High (+4.8 ft)	7.55	7.3	17.7	29.35	<1.0
"Wet"	Low	High	2/25/09	0915-0935	Low (+1.0 ft)	7.55	5.5	7.0	12.42	<1.0

Note 1. Representative, based on precipitation and river discharge
Note 2. Estimated at Mallard Island Ferry Wharf, Suisun Bay (38° 02.6' N, 121° 55.1' W)

Toxicity tests of desalination concentrate collected during the "dry" season (salinity-dominant scenario) and "wet" season (contaminant-dominant scenario) from the pilot plant in Mallard Slough showed:

- no significant effects on the survival or growth of algal organisms,
- no significant effects on the survival or growth of invertebrate organisms, and
- no significant effects on the survival or growth of fish test organisms.

Assuming that concentrate samples tested in this study are representative of those produced by an operational desalination plant at Mallard Slough, there would be no expected toxic effects of the concentrate on biota were the concentrate to be discharged into the Delta. Where source water TDS is higher than that experienced during this study, concentrate salinity will be commensurately higher and toxicity effects may differ.

6.0 IMPLICATIONS FOR FULL SCALE AT MALLARD SLOUGH

Based on information developed in the pilot study, an evaluation was conducted for full scale application of membrane technology at the Mallard Slough site. Recommendations for the treatment processes and operating parameters, and associated life cycle cost estimates, were developed and are presented within this section.

6.1 Treated Water Quality and Production Goals

Treated water quality goals for the pilot plant have been identified in Technical Memorandum No. 4A and in Section 2 herein. Membrane design issues are noted in Table 6-1**.**

Table 6-1: Desalination Water Quality Objectives

Parameter	Water Quality Target
Disinfection	***Comply with SWTR***
Virus	4 – 6 log reduction
Giardia	3 – 5 log reduction
Cryptosporidium	2 – 4 log reduction
Disinfection goals will be met by the proposed membrane systems, including post-treatment chemical addition. The proposed desalination plant will be required to maintain a secondary disinfectant if finished water will be served to customers directly and not co-mingled with Mokelumne Aqueduct water.	
Permeate Water Quality	***Meet all State and Federal MCLs***
Total Dissolved Solids (TDS)	< 500 mg/L
Chloride (selected target level)	< 100 mg/L
Bromide	<0.25 – 0.7 mg/L
Boron	<0.5 – 1.0 mg/L
Member agencies are generally characterized by serving finished water with TDS and chloride levels far below these levels, particularly during seasonally wet conditions and periods of abundant runoff and river flow. CCWD, for example, seeks to maintain a chloride level of 65 mg/L except during dry months when chlorides may increase to 100 mg/L. Lower goals are possible but will affect project costs dramatically. Based on pilot data, bromide and boron do not apparently present a design issue for the proposed desalination plant.	
Disinfection By-Products	***Result in Stage 1 and Stage 2 DBP Rule Compliance***
Total Trihalomethanes (TTHM)	< 64 μg/L
Halo-Acetic Acids (HAA5)	< 48 μg/L
Although not measured during the pilot study, DBPs are not anticipated to present an issue for the proposed desalination plant.	

The four agencies have established production goals for the proposed desalination plant:

- As established by member agency representatives, the full-scale plant will be subject to the current CCWD water rights at Mallard Slough, with a production goal of approximately 20 mgd, based on seasonal water quality, from a 25 mgd facility if achievable based on seasonal water quality. Actual production might be less depending on membrane system performance. Water potentially available under CCWD's existing water right varies depending on time of year and demand. Plant production will depend upon the timing of using the available water rights. Expanding production beyond what is covered in the existing water rights would require the acquisition of additional water rights. CCWD documentation for Mallard Slough water rights is included in **Appendix J**.

- Ultimate production to meet the identified demand from the four member agencies is 71 mgd of treated water, in keeping with previous planning and feasibility studies which were developed based on anticipated agency need. Ultimate production may be achieved at one or more physical locations.

6.1.1 Feed Water Quality

While the pilot study was conducted using source water obtained from Mallard Slough, it is anticipated that feedwater diverted to a full scale plant in this vicinity would resemble water quality in the Suisun Bay with respect to organic and inorganic constituents. A water quality monitoring station located in Suisun Bay near the City of Pittsburg and maintained by the California Department of Water Resources (DWR) has produced historical conductivity and temperature data which are very helpful in determining parameters to be used in developing a full scale membrane facility.

The Pittsburg station (Identifier designation PTS) is located at Latitude 38.0330°N, Longitude 121.8830°W near the New York Slough approximately one mile east of the Mallard Slough, as shown in Figure 6-1. For this analysis, daily minimum water temperatures have been averaged together for each month beginning in August 2005 and ending in July 2009. Results are illustrated in Figure 6-2. The lowest temperature recorded at the Pittsburg station is 44 deg F. Based on Figure 6-2, a minimum design temperature of 46 deg F is recommended for use in this scale-up analysis. Since feed pressures increase with colder temperatures, designing for the minimum temperature range will insure feed pumps are suitably sized.

Figure 6-1: Location of Pittsburg Station

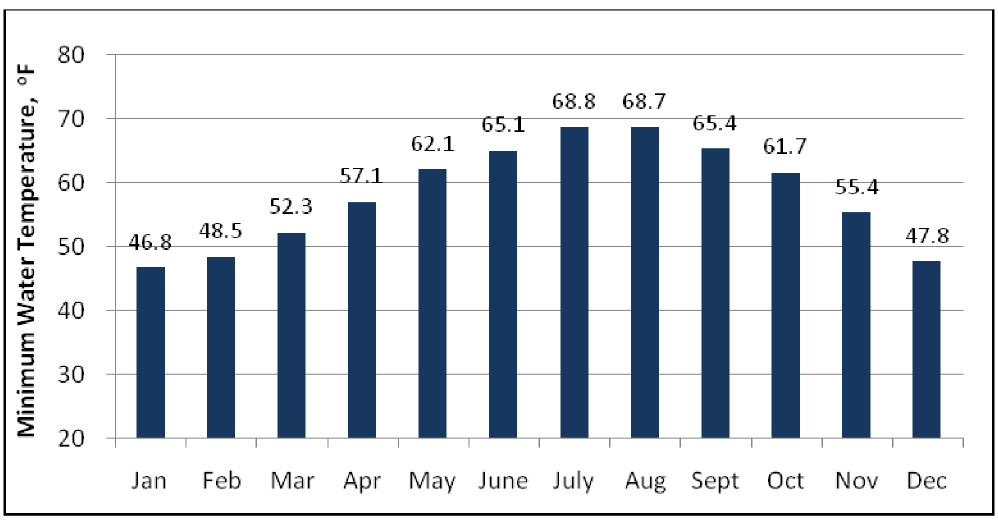


Figure 6-2: Suisun Bay Minimum Water Temperature, Aug 2005 to July 2009

Historical monthly TDS values for Suisun Bay at the Pittsburg station are shown in Figure 6-3 for the period from October 2006 through August 2009. This period was chosen because the 2007 and 2008 water years have been classified as a dry year type or as a critical year type by DWR for the Sacramento River and San Joaquin River systems based upon published Water Year Hydrologic Classification indices. At the time of this analysis, data for September 2009 was not yet available.

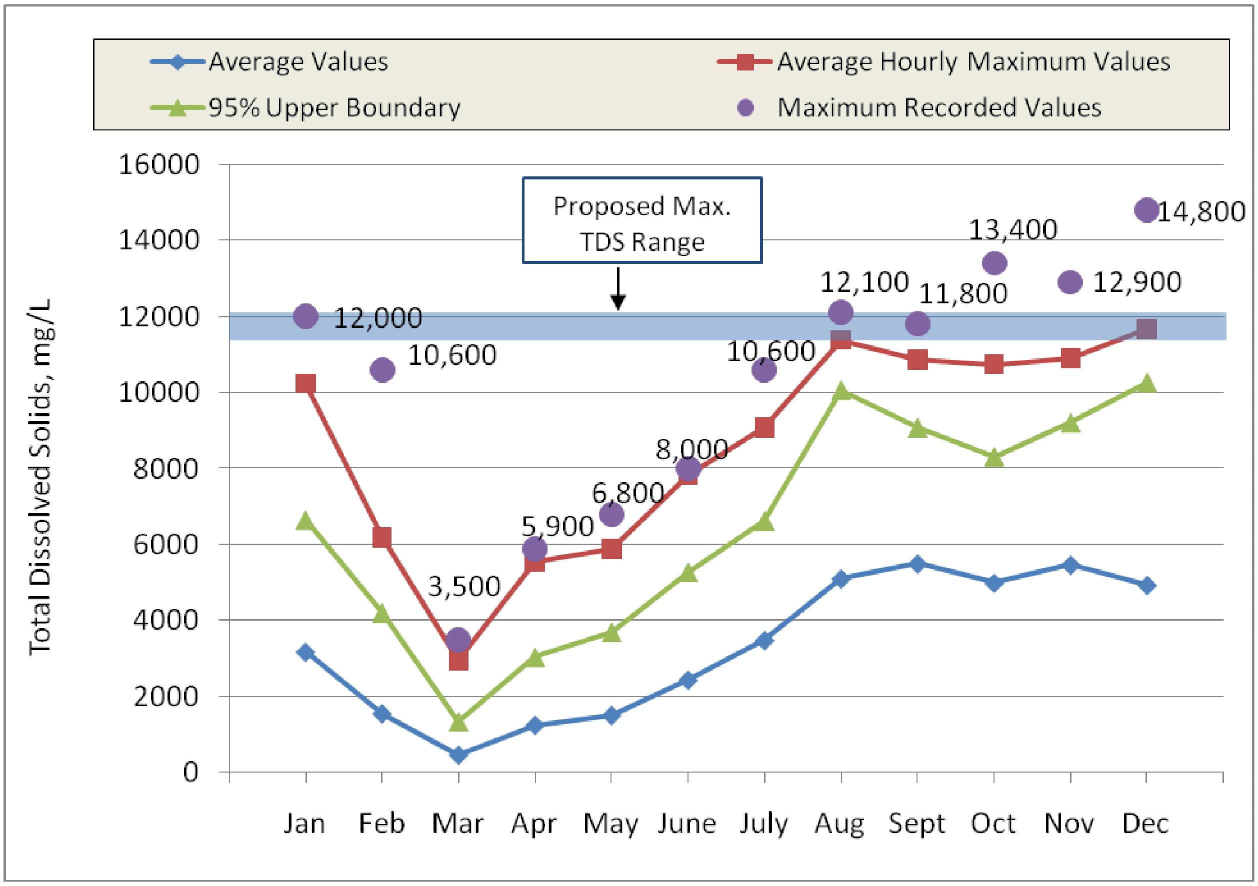


Figure 6-3: Suisun Bay TDS, Oct 2006 to Aug 2009

A water year is defined as the period from October 1 through September 30 annually. This period would consequently represent the specific type of hydrologic conditions for which the proposed treatment facility would be intended.

The graph is based on hourly recordings at the Pittsburg Station and illustrates the following:

- Average monthly TDS values, based on all of the available hourly data.
- The average maximum hourly TDS values. This is derived by identifying the maximum hourly recording for each individual month during the 35 month period and taking an average of these values.
- The upper 95% confidence interval of all the monthly data, for each month, derived by calculating the average and standard deviation for the data set.
- The maximum hourly individually recorded values for each month.

Average TDS values are observed to follow a wide range, from a low of approximately 500 mg/L in March, to a high of 5,500 mg/L in September. For 95% of the available data, TDS at the Pittsburg station is determined to be less than 10,000 mg/L in the dry season, and below 6,000 mg/L from February to June. Extreme hourly values above 12,000 mg/L were observed on five occasions, particularly during the August through December period.

Maximum TDS events are the result of hourly salinity spikes related to the simultaneous occurrence of several factors:

1. Seasonal dry conditions. Dry months typically occur from August through January, although this can vary significantly from year to year. Sacramento River flow is lessened due to the lack of snowmelt and local rainfall, resulting in the movement of San Francisco Bay water into the Suisun Bay.

2. A peak in the monthly lunar cycle resulting from either a new moon or full moon with a resulting peak gravitation force drawing saline San Francisco Bay water into Suisun Bay. Maximums are evident on a 2-week basis.

3. Daily high tides. Two tidal events occur daily, generally over a one to two hour interval each, also drawing the saline San Francisco Bay water into Suisun Bay.

The lunar cycle and daily tidal events have great effect on TDS in the bay, and this effect is particularly pronounced during seasonal dry periods. When all three conditions occur together, hourly peak TDS values have historically exceeded normal ranges as shown in the graph.

6.1.2 Selection of TDS Value for Conceptual Design Purposes

As indicated in Figure 6-3, average maximum hourly TDS values during the dry months, based an average of the hourly peaks during the past 3 years, are observed to fluctuate between approximately 10,500 mg/L and 11,500 mg/L, depending on the specific month. The highest individual hourly peaks are observed to reach as high as 15,000 mg/L.

One project goal established by the four agencies is for the proposed desalination plant to serve as a drought-resistant water supply for this region. The proposed plant would need to operate during the worst water quality and highest saline conditions for indefinite periods. Consequently, the plant should be designed to handle maximum feed water TDS which occurs during the dry late summer, fall, and early winter months.

Also shown in Figure 6-3 is the 95% confidence limit, which is a statistical parameter calculated using the mean and standard deviation of all the hourly data recorded each month. The variability of the data is indicated by the significant difference between the maximum values and the 95% confidence limit. This limit has two peak values, in August and December, when it reaches approximately 10,000 mg/L.

Based on this analysis, and in consultation with representatives from the four agencies, a maximum design TDS range between 11,500 mg/L and 12,000 mg/L has been selected for this project. This range will allow full plant production based on an average of the maximum TDS observations recorded during recent dry and critically dry years (refer to the previous section of this report for a description of these dry and critically dry years).

As indicated in the figure, feed water may have occasional TDS hourly peaks which are greater than 12,000 mg/L; however, these appear to be rare and of short duration and would have little effect on overall production. On the other hand, designing for lower salinity might not be acceptable during a drought or extended dry period, when maximizing regional water supplies with this new desalination source would be particularly important.

For the brackish feedwater, a TDS:Conductivity ratio equal to 0.61 (mg/L TDS to mS/cm conductivity) has been confirmed by pilot data. Similar ratios as high as 0.64 have been used in the past for surface waters in this geographic region.

6.2 System Recommendations

Based on the various systems which were piloted during the course of this project, the following assessment has been prepared to identify potential desalination systems which can provide a high level of recovery, meet water quality goals, and offer low operating costs.

6.2.1 Pilot Assessment

Pilot results are described in greater detail in Section 4 of this report. Data indicate that a two-stage system is capable of operating effectively over the salinity range encountered during the pilot study. Recovery will be highest for the two-stage system, which will assist in minimizing source water impacts, reducing volume of concentrate to be discharged into the environment, and reducing cost per gallon of permeate produced. Piloting demonstrated that a two-stage brackish-seawater system can routinely achieve between 70% and 82% recovery for the desalination membranes, at an operational flux of 12 gfd.

As shown in the pilot, a single stage seawater system is capable of excellent water quality. Permeate quality will significantly exceed standards illustrated Table 6-1, enabling plant staff to control operating costs by bypassing a portion of the filtrate around the seawater membranes during certain periods of the year, although such an operating scheme would cause loss of virus removal credits. Nevertheless, high operating pressures and lower comparative recovery will negatively affect overall system economics. Piloting demonstrated that a single stage seawater system can routinely achieve between 50% and 63% recovery at between 12.9 gfd and 13.2 gfd (flux is not temperature corrected).

A single stage NF system offers lower operating pressures than either the two-stage or the single seawater system. During wet weather months when feedwater TDS is low, an NF system can operate quite effectively. Higher TDS periods, however, are somewhat challenging with respect to salt passage. Although the pilot consistently achieved better than 97% chloride reduction, which is higher than that predicted by the Dow/Filmtec projection model, it was not sufficient to achieve treated water chloride goals when feedwater chlorides are above 3,500 mg/L, as is often the case for Mallard Slough water from August through January. Consequently, a single stage NF system is not suitable by itself unless modified with a second pass or in combination with other membranes. In either case, total system recovery is expected to be limited.

Other options, such as installing an NF system and turning it off during the worse water quality periods, are not favorable as this approach would compromise the overall goal of creating a drought-proof water supply.

By combining an NF train with a seawater train, it is possible that a hybrid system would allow the four agencies to take advantage of low NF feed pressures, with water quality enhanced by passing a portion of the filtrate through parallel seawater membranes, at a total life cycle cost which might prove to be lower than with a two stage arrangement. A parallel seawater-NF system would be suitable for low and high TDS conditions, without need to reduce throughput during dry months.

It should be noted that the performance of the three desalination trains which were piloted during this study cannot be directly compared because of differences in staging and number of elements per vessel. RO projection modeling is necessary to fully evaluate the capabilities of each pilot system to produce a permeate with suitable water quality balanced with high recovery and low energy use.

As demonstrated in the pilot study, UF membranes are capable of providing the water quality necessary for operation of a membrane desalination system at the Mallard Slough. The pressurized UF system tested features inside-out flow configuration and provided the highest specific flux and smooth operation. All permeability loss which was observed with this system due to fouling was recoverable, as compared to the submerged system. Consequently the pressurized UF membrane arrangement will be carried forward into this scale-up analysis.

6.2.2 Description of Proposed Desalination Alternatives

The two recommended alternatives to be carried forward into a life cycle analysis for the proposed desalination plant have been evaluated with the Reverse Osmosis System Analysis (ROSA) computer projection model as published by The Dow Chemical Company, version 6.1.5. This evaluation has been conducted using recommendations from Dow/Filmtec regarding specific 8-inch elements which are suitable for the proposed installation and which may be scaled from the 4-inch elements utilized for piloting purposes.

Characteristics of the two proposed desalination processes are summarized in Table 6-2 and Table 6-3. Performance is based on ROSA projections using targeted flux values as developed in the pilot study. Feed water quality is based on pilot data and extrapolations from the DWR Pittsburg water quality station. Projection results are provided in **Appendix H**.

1. Alternative No. 1: A two-stage brackish and seawater desalination system which will achieve high recovery of feed water for improved system economics and lessened impacts on source waters. Calculations indicate that the two-stage system will average approximately 82% recovery throughout a typical dry year hydrologic condition, for the desalination membranes themselves. Alternative No. 1 is based on the membrane characteristics shown in Table 6-2. A 100 psi interstage boost pump is furnished, which

enables Stage 2 to be fed with higher pressures than could be provided by the RO feed pump alone.

At maximum 12,000 mg/L TDS, it is estimated that the system will require 590 psi feed pressure and will operate at an average flux of 11 gfd. During these relatively short high TDS events, total recovery is 70%.

At low TDS, net system average recovery is calculated to be 83%. Feed pressure will decrease to approximately 240 psi.

Table 6-2: Alternative No. 1 Membrane Characteristics

	First Stage	Second Stage
Membrane Type[1]	Model BW30-440i	Model SW30XLE-400i
Available area per element	440 sf	400 sf
Design feed flow per element	11,500 gpd	9,000 gpd
Salt Passage	99.5%	99.7%

Note 1. As manufactured by Dow/Filmtec

2. Alternative No. 2: Two parallel single stage RO systems consisting of NF and seawater membranes, which together are intended to meet treated water quality goals while taking advantage of low feed pressures characteristic of NF membranes. Recovery will vary from 58% to greater than 79% during high and low TDS periods, respectively, for the desalination membranes themselves. Recovery is enhanced by diverting NF concentrate to the seawater train. Alternative No. 2 is based on the membrane characteristics shown in Table 6-3.

Table 6-3: Alternative No. 2 Membrane Characteristics

	NF Membranes Train A	Seawater Membranes Train B
Membrane Type[1]	Model NF90-400	Model SW30XLE-400i
Available area per element	400 sf	400 sf
Design feed flow per element	10,000 gpd	9,000 gpd
Salt Passage	97.0%	99.7%

Note 1. As manufactured by Dow/Filmtec

During low TDS periods, 100% of the total feed will be diverted to Train A. To increase recovery, concentrate from the NF membranes will be sent to the seawater train. Calculations indicate that the NF membranes can easily meet water quality goals at low feed water TDS. It is calculated that the NF membranes will require 110 psi feed pressure and will operate at an average flux of 16.3 gfd.

Bypassing a sidestream of MF/UF filtrate around the RO membranes can be considered to improve total recovery and system economics during extremely low TDS periods. Bypassing is most appropriate when TDS is consistently less than 1,000 mg/L; higher values will cause treated water quality goals to be exceeded. Bypassing consequently would have minimal influence on the life cycle analysis being conducted for this project.

Under high TDS dry weather conditions, when the NF membranes are unable to meet the 100 mg/L permeate chloride goal, a portion of the feed water will be diverted to the seawater membranes. Based on recovery and salt rejection observed in the pilot study for the two single stage seawater and NF trains, a 70/30 flow split (seawater/NF) is recommended so that the 100 mg/L chloride goal can be met with a reasonable safety factor during high TDS conditions. As shown in illustrated in **Appendix H**, feed pressure for the Train B seawater membranes at maximum TDS condition is projected to be 832 psi.

By combining NF concentrate with the 70% feed being pumped to the seawater train, overall recovery can be increased to 58% for both trains at maximum TDS. Calculations indicate that the parallel NF and seawater system will average approximately 79% recovery throughout a typical dry year hydrologic condition, for the desalination membranes themselves (not including MF/UF recovery).

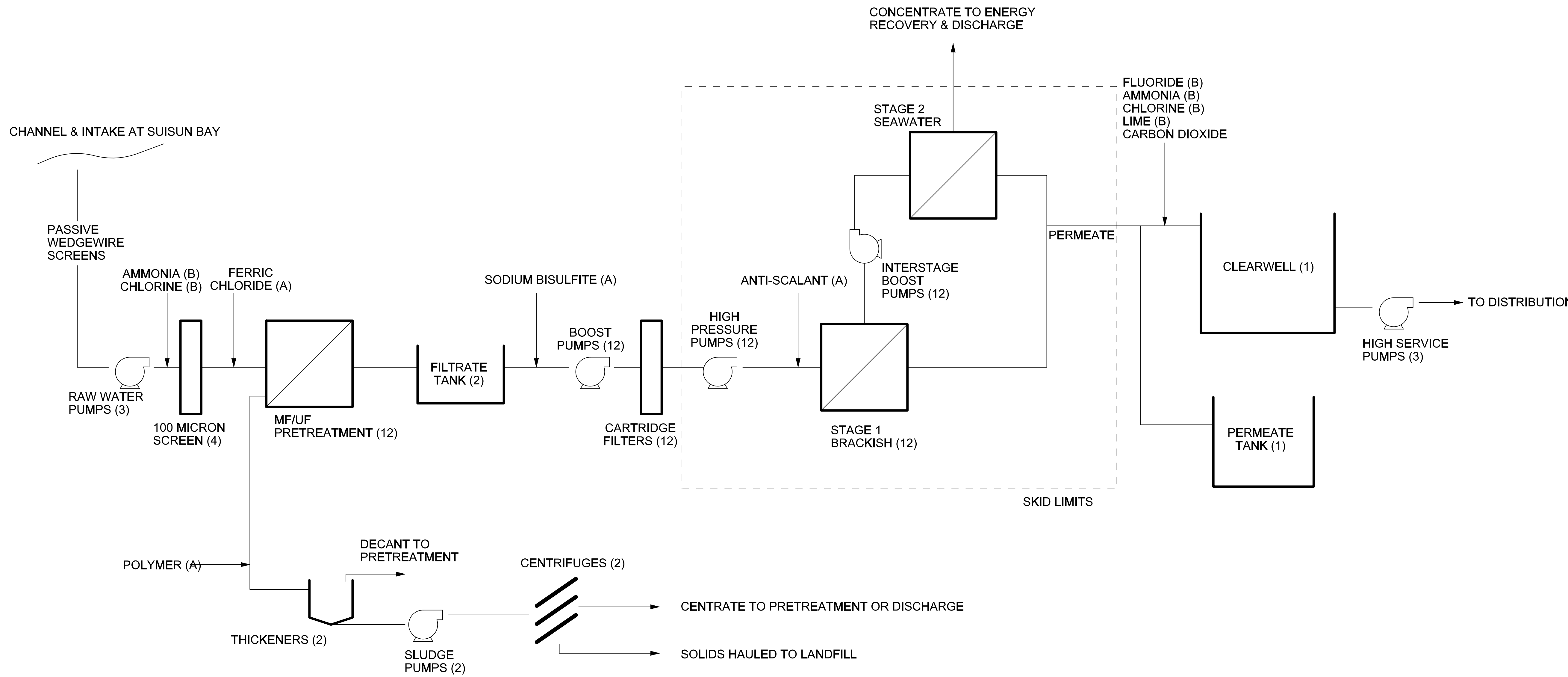
It should be noted that use of 8-inch diameter membrane elements has been assumed for this investigation, which represents a standard commercially available design. Larger 16-inch or 18-inch diameter elements will likely become more commonplace should this project move forward into subsequent design and construction phases in the next several years, and may represent a total net cost savings. Important considerations include ability to substitute elements from other manufacturers, which currently does not necessarily exist with these large elements, track record and performance history, and overall cost.

It should also be noted that a lower finished chloride goal (below 100 mg/L) will have limited impact to Alternative No. 1, since the two-stage system already provides very low permeate chloride levels under all source water quality conditions. Alternative No. 2, however, would require greater diversion of source water to the seawater membranes, increasing project costs accordingly.

6.3 Conceptual Plant Design

Additional components are necessary for a comprehensive treatment facility to support the desalination process and to provide the four agencies with a suitable treated water supply. Schematic diagrams for both alternatives are provided in Figure 6-4 and Figure 6-5.

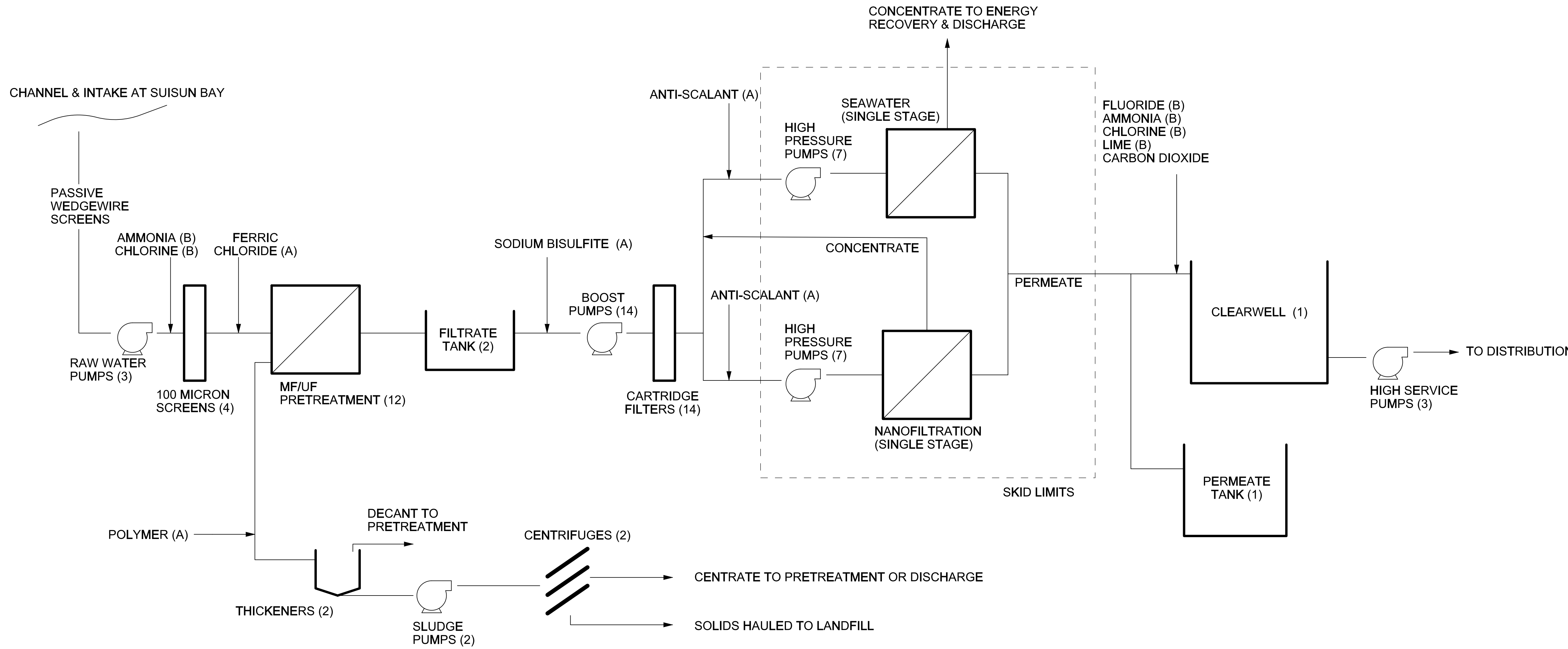
Headworks. Various styles and types of intakes are available for consideration as described in Technical Memorandum No. 2A. For the purposes of this analysis, a passive wedgewire screen intake has been included, to be installed onto a pipeline extended into the waterway. Source water drawn through the intake will be pumped through the 100 micron self cleaning screens to the MF/UF system. These will serve to protect the membranes by removing large particulates, sand and shells, stringy material, and other debris which might cause premature membrane fouling.



NOTES

1. NUMBER OF UNITS SHOWN IN ( )
2. MEMBRANE CLEANING SYSTEMS NOT SHOWN
3. (A) INDICATES CHEMICALS ARE STORED IN CHEMICAL BUILDING A.
4. (B) INDICATES CHEMICALS ARE STORED IN CHEMICAL BUILDING B.

REV	DATE	BY	DESCRIPTION	SCALE	WARNING					SHEET
				NONE	0 ½ 1 IF THIS BAR DOES NOT MEASURE 1" THEN DRAWING IS NOT TO SCALE	DESIGNED C. BROMLEY DRAWN B. BEHNKE CHECKED G. BEHNKE	MWH AMERICAS, INC. 2121 N. CALIFORNIA BLVD WALNUT CREEK, CALIF. 94596	MWH	BAY AREA REGIONAL DESALINATION PROJECT	FIGURE 6-4 ALTERNATE NO. 1 SCHEMATIC DIAGRAM 2-STAGE BRACKISH-SEAWATER MEMBRANES



NOTES

1. NUMBER OF UNITS SHOWN IN ( )
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3. (A) INDICATES CHEMICALS ARE STORED IN CHEMICAL BUILDING A.
4. (B) INDICATES CHEMICALS ARE STORED IN CHEMICAL BUILDING B.

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DESIGNED C. BROMLEY
DRAWN B. BEHNKE
CHECKED G. BEHNKE

MWH AMERICAS, INC.
2121 N. CALIFORNIA BLVD
WALNUT CREEK, CALIF. 94596

MWH

BAY AREA REGIONAL DESALINATION PROJECT

FIGURE 6-5
ALTERNATE NO. 2 SCHEMATIC DIAGRAM
PARALLEL SEAWATER AND NF MEMBRANES

SHEET

File: Alt 2 schematic.dgn Model: Layout1 ColorTable: bw.ctb DesignScript: MWH_Iplot_Pentable_v85.pen PlotScale: 80:1 User: Jennifer A Gelmini Plot Date: 01-DEC-2009 14:38

The pilot study investigated a 100-micron self cleaning screen, with size range based on pretreatment membrane vendor recommendations. The self-cleaning feature will be a highly desirable aspect of the full-scale installation due to source water algae and other particulate content. The pilot demonstrated that this screen size is suitable for removing materials harmful to the UF membranes. The pilot also demonstrated effectiveness of the self-cleaning operation, which was found to result in less than 1% water loss. Consequently the full-scale analysis will be based on the self-cleaning screen equipment as piloted,

Chemicals applied to source water are sodium hypochlorite and ammonia to form chloramines residual for control of biogrowth and potential fouling of the self cleaning screens.

It is recognized that the agencies will conduct a thorough evaluation of potential intakes, including subsurface intakes based on local geology, once a firm site is selected for the proposed desalination plant. If the Mallard Slough Pump Station were to be converted for use as an intake and source water pumping facility for the proposed desalination plant, then a new intake and pump station would not be required. The station itself would need to be retrofitted with VFDs and other components. Pending further review during the design phase of this project, at this time its discharge pressure appears to be adequate based on a review of pump curves provided by CCWD.

Pretreatment. Based on the pilot plant results, a pressurized membrane system is included with this analysis. The system will be furnished complete with clean-in-place equipment, compressed air system, backwash pumps, and related components. Piloting demonstrated that the pressurized system could achieve flux in the range of 40 to 44 gfd in the dry season and 55 gfd in the wet season. For this report, it is assumed that the flux will be 44 gfd.

Piloting also demonstrated that membrane pretreatment recovery ranged from 85% to 88% using the pressurized system, depending upon cleaning procedures. This analysis utilizes recovery equal to 88%; however, it is recognized that a full-scale design could use a higher figure as determined by the vendor to meet specified process goals and conditions.

A budgetary proposal was obtained for the proposed UF system and included 12 skids, each with 102 UF modules. One of the skids would be fully redundant allowing for CIP or other service interruptions. The UF equipment costs also include the Motor Control Center (MCC) and controls, CIP systems (tanks and pumps), backwash pumps, and compressed air system for membrane integrity and pneumatic process control. Unlike a submerged membrane system, the piloted pressurized system does not require air scour during backwash.

A low dose of ferric chloride (approximately 5 mg/L) is necessary for the pressurized system for coagulation of source water solids and improved membrane performance. Pilot experience demonstrated the importance for metal salt coagulant addition to achieve consistent performance and maintaining high flux for the piloted membranes. This is consistent with manufacturer recommendations.

Filtrate produced by the pretreatment system will be stored in ground level storage tanks prior to pumping to desalination, and will also be used for backwashing of the MF/UF membranes.

Volume is assumed to be equal to one hour of filtrate production, which must be verified during the detailed design phase of this project. Sodium bisulfite will be fed at the outlet from the filtrate tanks to remove chlorine residual prior to the UF filtrate entering the RO membranes.

Using a metal salt coagulant as required for success of the piloted pressurized system is an important issue and will affect project costs due to chemical cost and higher coagulant solids to be handled. Furthermore, it is unlikely that sludges, particularly those produced using metal salt coagulants, may be discharged as a point source into the Suisun Bay. Regulators instead will likely require a solids handling system to thicken, dewater and make final disposal (e.g. landfill).

In future stages of this project and once a final site for the desalination plant is established, the four agencies may wish to investigate an alternate membrane pretreatment system which does not require chemical coagulant. Other MF and UF manufacturers may be consulted and other systems possibly piloted to demonstrate ability to pretreat the source water without fouling and without use of coagulant. While the tradeoff may be a reduced design flux, greater membrane area, and higher membrane system capital costs, the potential savings in chemical storage and feed equipment, solids handling facilities, and solids hauling and disposal costs would be significant.

Desalination. The desalination trains were designed according to the two options described in Section 6.2.2. A budgetary proposal was provided by Biwater AEWT, Inc., including a desalination skid layout and cost analysis for the two desalination systems under consideration. Membrane vessels, piping, valves, meters, and related components will be furnished on skids. Each skid is provided with a cartridge filter and a high pressure pump. Other components included with the desalination membranes include low pressure filtrate boost pumps, and CIP pumps and tanks.

Sodium bisulfite is provided to preserve the membranes during extended periods of downtime. Citric acid and caustic soda (sodium hydroxide) are both furnished for cleaning. Anti-scalant will be required to mitigate impact of highly concentrated salts on the feed side of the membranes to prevent membrane scaling. A dedicated permeate tank is furnished for storing permeate to be used for CIP and flushing purposes. Water in this tank is to remain free of any post-treatment chemicals to avoid damage to the desalination membranes.

For the purposes of this project, it is assumed that RO concentrate will be discharged to Suisun Bay. Other discharge options will be investigated by the four agencies, including the potential of combining the concentrate with treated effluent produced by the local sanitary sewer district (DDSD or CCCSD) or with nearby power plant cooling water. Economics and choice of disposal option will depend significantly on final site selection. It should be noted that antiscalant chemicals used in the desalination process will affect the ability of the four agencies to use surface discharge for concentrate disposal. Costs have been included in this analysis to account for capital facilities necessary for future disposal.

Clearwell and High Service Pumping. Permeate produced by desalination will be treated with carbon dioxide and lime for stability, to restore alkalinity, and to meet treated water compatibility requirements. Stabilized permeate will be stored in a circular clearwell prior to off-site distribution. The clearwell can be either welded steel or prestressed concrete. The initial phase of the project will require a single 1.6 million gallon clearwells to provide sufficient hydraulic residence time for disinfection and inactivation of virus and *Giardia* at 50% capacity.

Prior to pumping, chlorine (and ammonia, depending upon which system receives the water) will be added to establish a secondary residual disinfectant; and fluorosilicic acid to meet any fluoridation requirement.

High service pumping is required to furnish approximately 500-feet of total lift in order to meet existing pressures in the EBMUD Mokelumne Aqueduct. Approximately 450-feet of lift would be used to deliver treated water to the CCWD Multi-Purpose Pipeline. Pumps will suction directly from the clearwell. These pressures include friction losses in the conveyance pipeline and will need to be verified during future design phases of this project.

Solids Handling. For the purposes of this project, it is assumed that water produced by backwashing of the MF/UF membranes, along with other water potentially recovered from the various treatment processes, will be directed to a series of gravity thickeners to increase solids content prior to dewatering. Backwash will represent a very dilute solids concentration of approximately 350 mg/L. Thickened sludge is estimated to be approximately 1.5%, resulting in approximately 68,000 gpd of wet sludge sent to the centrifuges. Thickener supernatant will be recovered and sent to the plant inlet for reuse, increasing overall recovery of the UF pretreatment process because water losses are subsequently limited only to what is contained in thickened sludge being centrifuged.

Based on information provided by centrifuge manufacturers, it is anticipated that the centrifuges will produce approximately 20% solids to be trucked to nearby landfills or otherwise transported to an off-site disposal location. Centrate will be blended and discharged with RO concentrate or returned to the plant inlet for reuse.

The four agencies may wish to consider alternatives in lieu of thickening, dewatering and landfilling of these solids, thereby avoiding costs of thickening, dewatering, or both. The lowest cost option is discharge of spent backwash water containing dilute solids directly into the bay; however, this may not be considered to be a viable option by local regulators. Co-locating the proposed desalination plant with, or pumping thickened solids to, a nearby drinking water treatment plant operated by CCWD (e.g. the Bollman WTP) or other agency might also be a viable solution. In this case, thickened solids from both the new desalination plant and the existing desalination plant could be co-mingled and handled together, providing an economy of scale for both facilities and reducing overall cost for the new desalination plant.

Ancillary Facilities. The concept design has made allowance for operations functions to be co-located with the RO Building, which would include a control room with SCADA; operator's laboratory for performing routine plant analyses; kitchen, showers, restrooms and convenience facilities; conference room and training room; and equipment maintenance.

For the purposes of this analysis, it is assumed that two chemical buildings are required for bulk storage and transfer. Process chemicals are located in Chemical Building A (caustic soda, citric acid, ferric chloride, antiscalant, bisulfite). Chemicals generally used for source water conditioning and post treatment addition (aqua ammonia, sodium hypochlorite, fluorosilicic acid, and lime) are contained in Chemical Building B.

The desalination plant site plan will need to accommodate space for a new electric utility substation and switchgear for serving the proposed desalination plant. The four agencies will work with the local electric utility to identify requirements, incoming power feed parameters, and associated costs.

Energy Recovery. An important aspect of the proposed desalination plant design is recovery and reuse of energy available in the concentrate being rejected from the RO or NF membranes. Based on current technologies and modeling performed by Energy Recovery Inc., one or more pressure exchange devices would be provided for each of the RO systems and would serve to transfer waste energy from the concentrate into the feed water at the inlet to each RO skid. Fluctuation in flows and pressures during the year caused by the widely variable source water salinity and temperature will present a considerable challenge. Separate devices may be required to specifically handle low TDS and high TDS periods. Capital costs for the energy recovery units are included with this analysis. Operational savings are also included.

Siting. While a siting study and final selection has yet to be performed by the four agencies, it is assumed that the proposed desalination plant will be placed somewhere in the Pittsburg-Antioch and adjacent Suisun Bay area, either near Mallard Slough or in the vicinity of the neighboring Mirant power plant. A conceptual site plan suitable for either alternative has been developed for this report to assist with the life cycle cost analysis and to furnish the agencies with an understanding of potential layouts. This conceptual plan is illustrated in Figure 6-6 and is representative of a campus-style layout where dedicated buildings are furnished for each key function.

The existing Mallard Slough Pump Station is constructed on concrete piles due to the poor nature of soils in this region. It is similarly anticipated that piles will be needed for any new plant structure in the vicinity of the MSPS. Since groundwater is quite shallow, the concept site plan is based on installation of all structures above-ground. No new below-ground tankage or substructures would be provided. All structures are placed on slabs located at or near original grade, with the possible exception of source water and high service pumping which may be placed at a lower elevation to ensure flooded pump suctions.

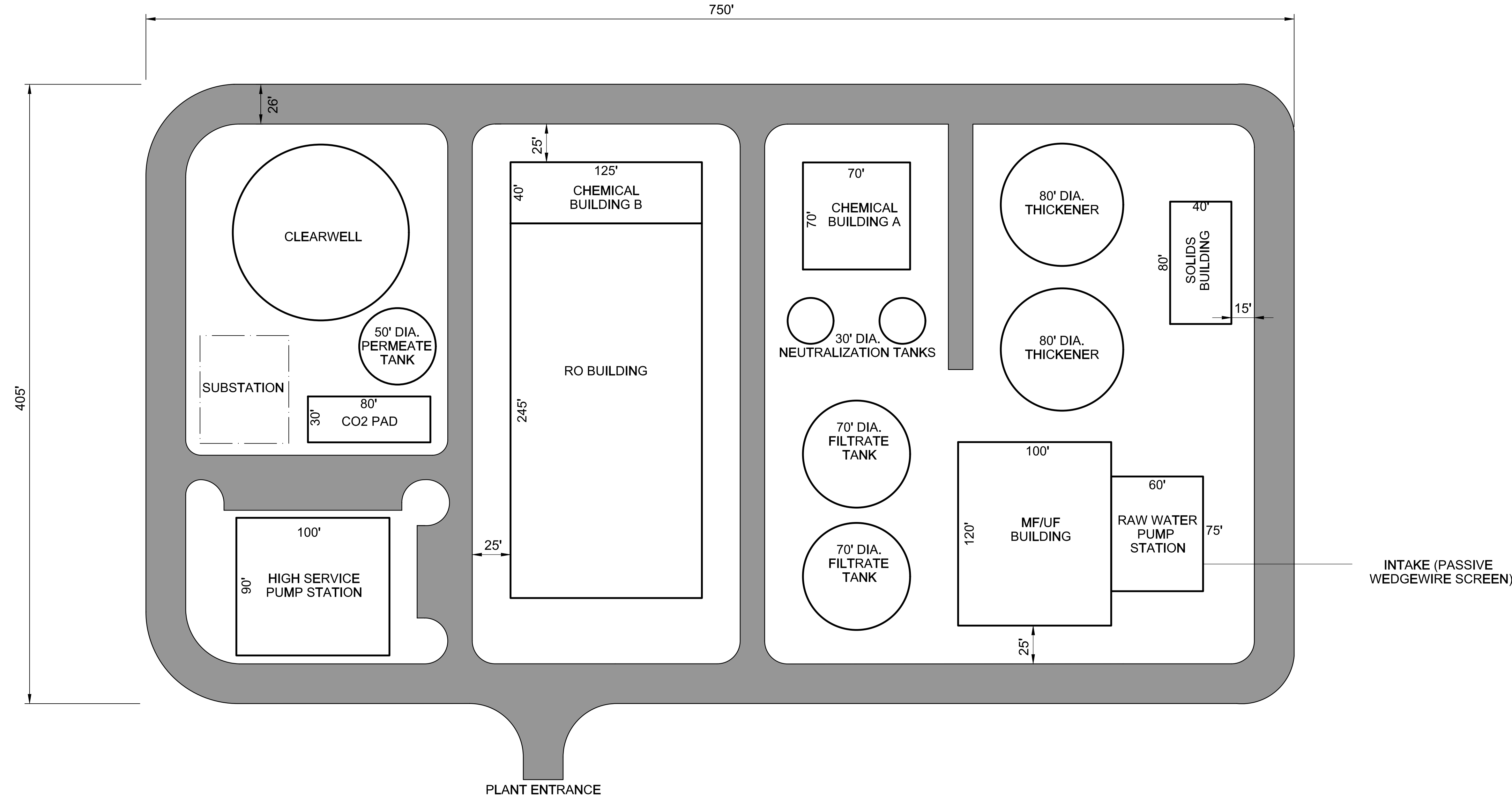
Filtrate tanks, neutralization tanks, and the clearwell are assumed to be circular welded steel ground storage tanks. For the purposes of this report, it is assumed that all buildings are constructed of prefabricated steel.

Actual layout will depend on the final site for the proposed desalination plant. The perimeter road is assumed to be 26-feet in width to accommodate fire and emergency vehicles. Interior roads are narrower and are intended for use in delivering chemicals, installing or removing equipment, and accessing all systems for routine operation and maintenance. Total footprint to accommodate the initial 25 mgd feed water capacity is approximately 7 acres, not including any perimeter buffer zone.

6.3.1 Preliminary Design Criteria

Preliminary design criteria for each of the two alternatives are contained in **Appendix H**, along with the RO system performance projections. For the initial 25 mgd feed flow and the ultimate 71 mgd treated water flow, initial sizing data are provided for source water, treated water, sludge pumping; self cleaning screens; UF, NF and RO processes; chemical storage and feed systems; permeate and neutralization tanks and clearwell; and solids handling systems.

It is assumed that a single location would be used for initial and final plant production. If multiple sites are eventually utilized to achieve the ultimate 71 mgd treated water capacity, these criteria will require revision.



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SCALE 1"=40'

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DESIGNED C. BROMLEY
DRAWN B. BEHNKE
CHECKED G. BEHNKE

MWH AMERICAS, INC.
2121 N. CALIFORNIA BLVD
WALNUT CREEK, CALIFORNIA 94596

MWH

BAY AREA REGIONAL DESALINATION PROJECT

FIGURE 6-6
CONCEPT SITE PLAN

SHEET

File: site plan v2.dwg Model: Model ColorTable: bw.ctb DesignScript: MWH_Iplot_Pentable_v65.pen PlotScale: 80:1
User: Jennifer A Gelmini
Plot Date: 02-DEC-2009 09:29

6.4 Concept Study Life Cycle Cost Analysis

Life cycle costs are documented at a Class 5 Conceptual Design Level Estimate according to American Association of Cost Engineers (AACE) International Cost Estimate Classification. These costs were prepared based on limited engineering information, and are intended for planning purposes and for comparing various similar alternatives. This analysis will assist the four agencies in gaining an understanding of approximate full-scale facility capital and operational requirements. Approximate accuracy range is -20% to -50% on the low side, and +30% to +50% on the high side.

Life cycle costs are intended to represent the approximate cost of a project by incorporating both capital and annual costs into a single present worth value. Annual costs are converted to a net present value by taking into account the approximate cost of funds (inflation and interest as represented by the discount rate) and the anticipated project period. For this cost analysis the net present value calculations are based on 3% discount rate and a 30 year project planning period.

For comparison purposes, an annual worth analysis is also presented, in which a series of equal annual payments are derived from estimated capital and operating costs, utilizing the same factors as described above.

6.4.1 Capital Costs

Capital cost estimates were prepared by the MWH team and detailed documentation can be found in **Appendix I**. Costs are in current dollars (November 2009) and intended to represent an estimated value for construction and do not represent potential low bid costs. Capital costs were based on system criteria as listed in **Appendix H** and were marked up to address three cost components:

1. Contractor mobilization and compliance with General Conditions of the contract at 5% of the installed estimated price.
2. Bonds and insurance at 2% of the estimated price plus mobilization.
3. General contractor's overhead and profit at 5% of the summation of the estimate price mobilization, and bonds and insurance.

For the initial 25 mgd feed water condition, estimated capital costs for the two desalination alternatives are shown in Table 6-4. These are subdivided into 16 major categories:

1. Sitework – site clearing, miscellaneous demolition, paving and grading, fencing, entry control, site security, yard piping, and building pads.
2. Intake and Source Water Pump Station – new wedgewire intake, source water pumps, forebay and building.
3. Brine Disposal – for this study it is assumed that up to $1.2M will be required to divert membrane concentrate to an off-site location.

4. Filtrate Tank – circular tanks required to store filtrate to be used for UF membrane backwashing and as feed supply to the RO membranes.
5. MF/UF building - pretreatment screens and membranes, cleaning systems, and building.
6. RO Building – RO equipment, cartridge filters, boost and high pressure pumps, operator facilities, and building.
7. Permeate Tank – circular tank required to store permeate to be used in chemical cleaning and flushing of the RO membranes.
8. Clearwell – circular tank required to store treated water prior to distribution and for CT credit for disinfection purposes.
9. High Service Pump Station - treated water pumps and building.
10. Neutralization Tanks - circular tanks required for storing and neutralizing of acid and base membrane cleaning chemicals prior to disposal.
11. Chemical Building A – chemical systems necessary for storing and feeding of such chemicals as bisulfite, ferric chloride, antiscalant, and polymer.
12. Chemical Building B – chemical systems necessary for storing and feeding of such chemicals as ammonia, hypochlorite, fluorosilicic acid, lime, and carbon dioxide.
13. Solids Handling Facilities - thickeners, sludge pump station centrifuges, and centrifuge building.
14. Pile Foundations – in the event that this project is constructed in poor soil conditions near the Mallard Slough, costs for pile foundations have been broken out separately.
15. Transmission Main – 30-inch pipeline necessary to convey treated water off-site to either the EBMUD Mokelumne Aqueduct or the CCWD Multipurpose Pipeline.
16. Site Electrical Systems – facilities necessary to receive a commercial power feed and distribute to the on-site processes. Does not include cost of bringing a commercial power feeder to the site itself.

Table 6-4: Life Cycle Analysis Summary

	Alternative No. 1	Alternative No. 2
Capital Costs		
1. Sitework	$4,200,000	$4,200,000
2. Intake and Source Water Pump Station	$3,100,000	$3,100,000
3. Brine Disposal	$1,100,000	$1,100,000
4. MF/UF Facilities	$18,300,000	$18,300,000
5. Filtrate Tanks	$1,100,000	$1,100,000
6. RO Facilities	$44,100,000	$51,300,000
7. Permeate Tank	$500,000	$500,000
8. Clearwell	$1,900,000	$1,800,000
9. High Service Pumping Station	$4,400,000	$4,400,000
10. Neutralization Tanks	$400,000	$400,000
11. Chemical Building A	$1,900,000	$1,900,000
12. Chemical Building B	$2,300,000	$2,300,000
13. Solids Handling Facilities	$9,900,000	$10,600,000
14. Pile Foundations	$3,100,000	$3,300,000

	Alternative No. 1	Alternative No. 2
15. Transmission Main	$7,800,000	$7,800,000
16. Site Electrical Systems	$5,200,000	$5,600,000
Subtotal	$109,300,000	$117,700,000
Contingencies (20%)	$21,900,000	$23,500,000
Planning, Permitting, Engr, Admin Costs (25%)	$32,800,000	$35,300,000
Land Acquisition	$3,500,000	$3,500,000
Concentrate Discharge Permit & Connection Fee	$1,000,000	$1,000,000
Total Capital Cost	$168,500,000	$181,000,000
Annual Costs		
1. Power Requirements	$5,400,000	$7,900,000
2. Chemical Costs	$1,400,000	$1,300,000
3. Equipment Replacement Cost	$1,400,000	$1,700,000
4. Staffing Costs	$900,000	$900,000
5. Outside Services (sludge disposal)	$1,350,000	$1,350,000
Total Annual Cost	$10,450,000	$13,150,000
Present Worth of Annual Costs	**$204,900,000**	**$257,800,000**
Annual Worth of Capital Costs	**$8,600,000**	**$9,300,000**
TOTAL PRESENT WORTH VALUE	**$373,400,000**	**$438,800,000**
Net Present Worth, per acre-foot	**$550/acre-foot**	**$660/acre-foot**
TOTAL ANNUAL WORTH VALUE	**$19,050,000/year**	**$22,450,000/year**
Unit Cost of Water, based on Annual Worth (Year 1), per acre-foot	**$840/yr/acre-foot**	**$1,010/yr/acre-foot**

Graphical representations of the capital cost breakdowns are show in Figure 6-7 and Figure 6-8. In both alternatives the largest contributors to capital cost are the MF/UF facilities and the RO facilities.

The chief difference in capital cost between Alternative 1 and Alternative 2 is the cost of the RO Facilities. In Alternative No. 1 the system is determined to require 540 RO vessels installed onto twelve skids, compared to 637 vessels installed onto 14 skids for Alternative No. 2. This affects the cost of purchasing the membranes, associated piping, number of pumps, and number of energy recovery devices. The building must also be larger to accommodate the extra skids and associated equipment. Note that UF facility costs are the same for each alternative, since each will handle the same quantity of feed water and filtrate.

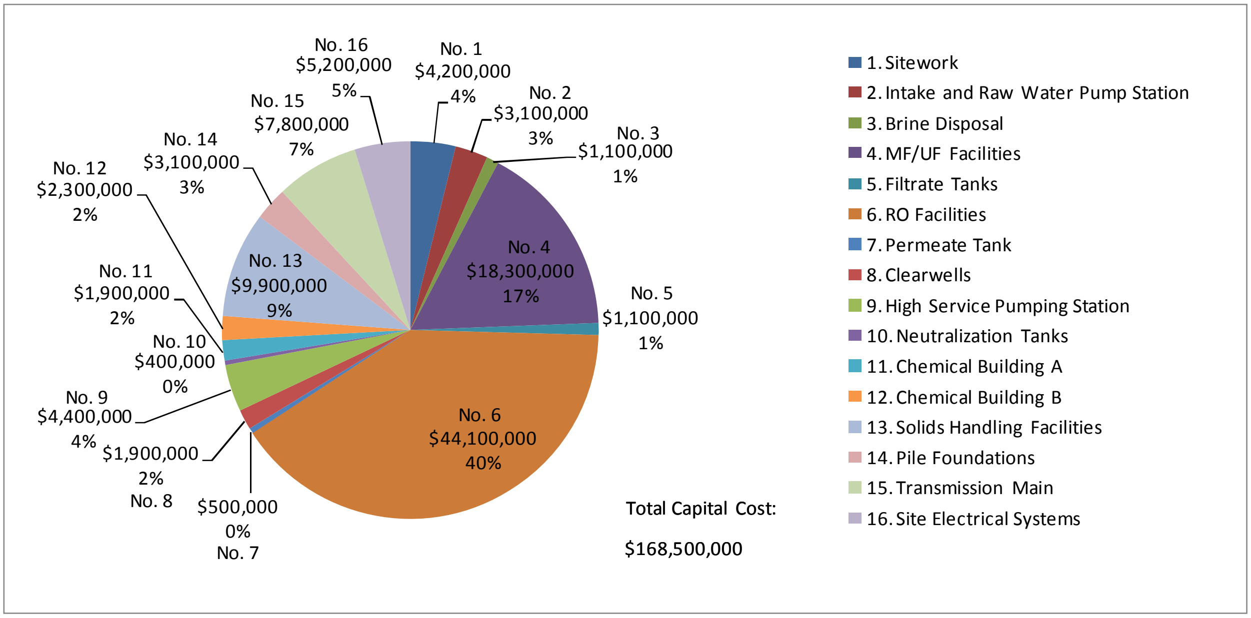


Figure 6-7: Alternative No. 1 Capital Costs

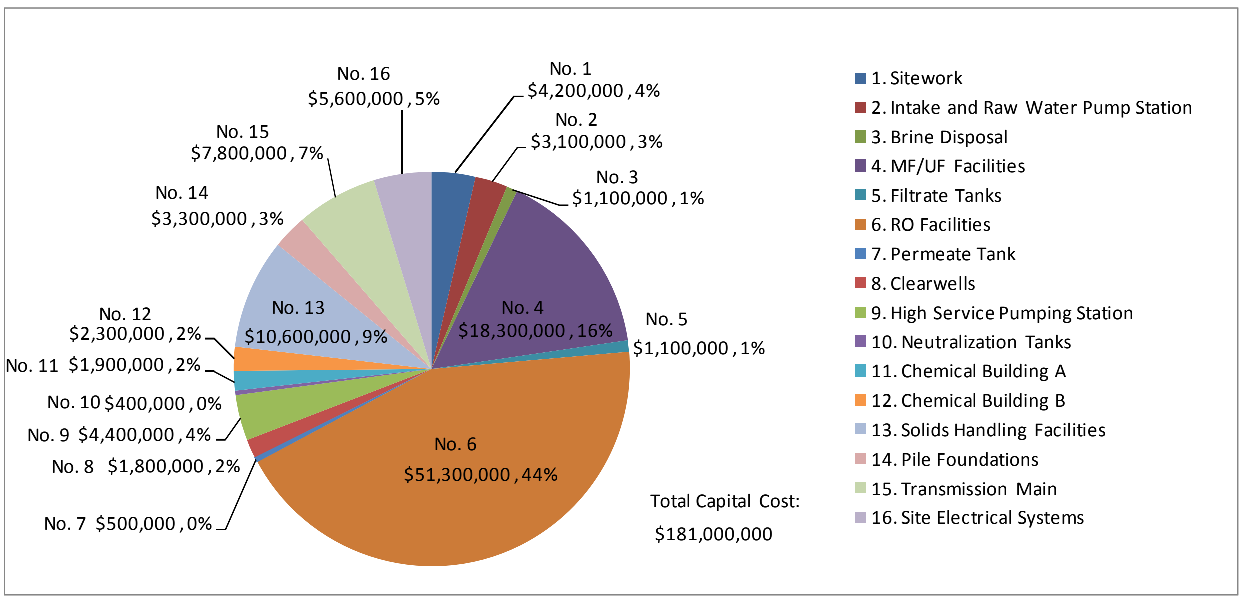


Figure 6-8: Alternative No. 2 Capital Costs

6.4.2 Annual Costs

Annual costs contribute significantly to the life cycle cost of a project. For each alternative, annual costs are presented in Table 6-4 for the following categories:

- Power requirements – for major systems, including source water pumps, RO boost pumps, RO high pressure pumps, RO interstage pumps, treated water pumps, and solids handling centrifuge.
- Chemical costs – source water disinfection, pretreatment coagulation, treated water pH and alkalinity adjustment, treated water disinfection and fluoridation, sludge conditioning, and antiscalant addition. Membrane cleaning chemical requirements are also included with this analysis. It should be noted that blending with source water in the Multipurpose Pipeline will result in less chemical use and the elimination or discontinuance of some of these systems.
- Equipment replacement costs – for major systems, including self cleaning screens, MF/UF membranes, cartridge filters, RO and/or NF membranes.
- Staff costs – consisting of two operators, two technicians, two maintenance staff, and one administrator. It is anticipated that the desalination plant would not be occupied full-time, although staffing must be considered more thoroughly by the four agencies during later stages of this project.
- Solids disposal – hauling and tipping fees for disposal of centrifuge solids in a landfill or other appropriate destination.

Graphical representations of the annual cost breakdowns are shown in Figure 6-9 and Figure 6-10. As expected, the power requirement is the largest contributor to annual costs for both alternatives. It was originally anticipated that the power requirements for Alternative No. 2 would be lower than Alternative No. 1 because of the lower pressure requirements of the NF membranes. The seawater membranes, however, contribute significantly to the total system energy usage because of the inability of the NF membranes to meet treated water quality during high TDS periods. Even in low TDS periods, the seawater membranes are used to recover NF concentrate and consequently have a measureable contribution to overall energy use.

6.4.2.1 Energy Consumption

A significant portion of energy consumed by each of the two treatment alternatives under consideration is required to feed the high pressure membranes. For the dry year period, it is projected that Alternative No. 1 will require 28.1 million kWhr per year to produce an average of 19.8 mgd finished water as shown in **Appendix H**. Similarly, Alternative No. 2 will require 53.4 million kWhr per year to produce an average of 19.2 mgd finished water, also as shown in **Appendix H**.

These values include energy recovery. As previously noted herein, recovery of energy present in the membrane concentrate may be achieved in the range of 70% to 83% through the application of specialized equipment. Factors which affect recovery include source water quality (e.g. salinity) and temperature, and such operating conditions as pressure and flow.

Figure 6-9 and Figure 6-10 show the annual cost breakdown for Alternatives No. 1 and 2, respectively. The power cost used for these calculations was $0.10/kwh.

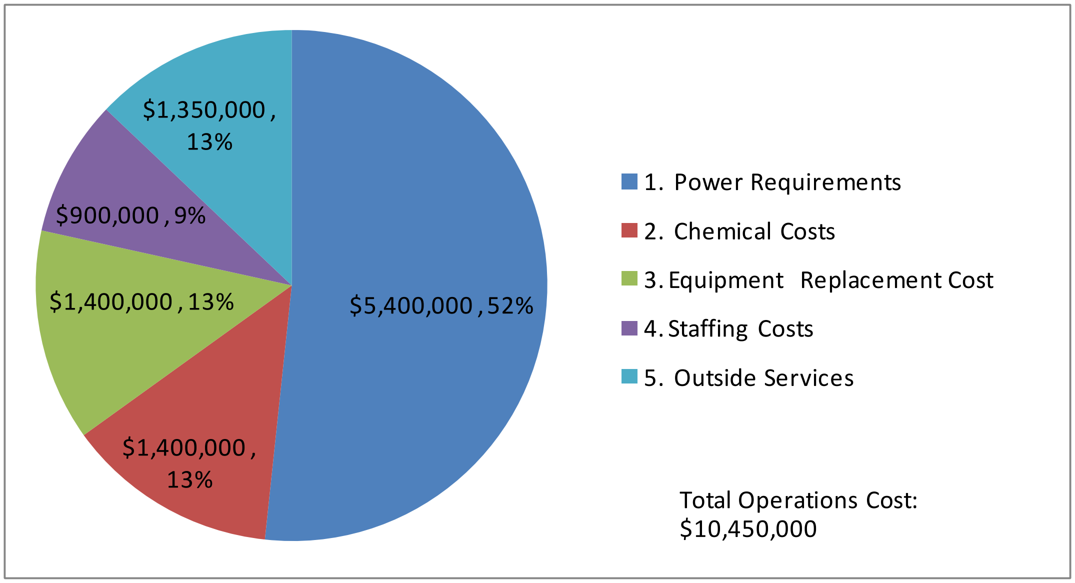


Figure 6-9: Alternative No. 1 Annual Cost Breakdown

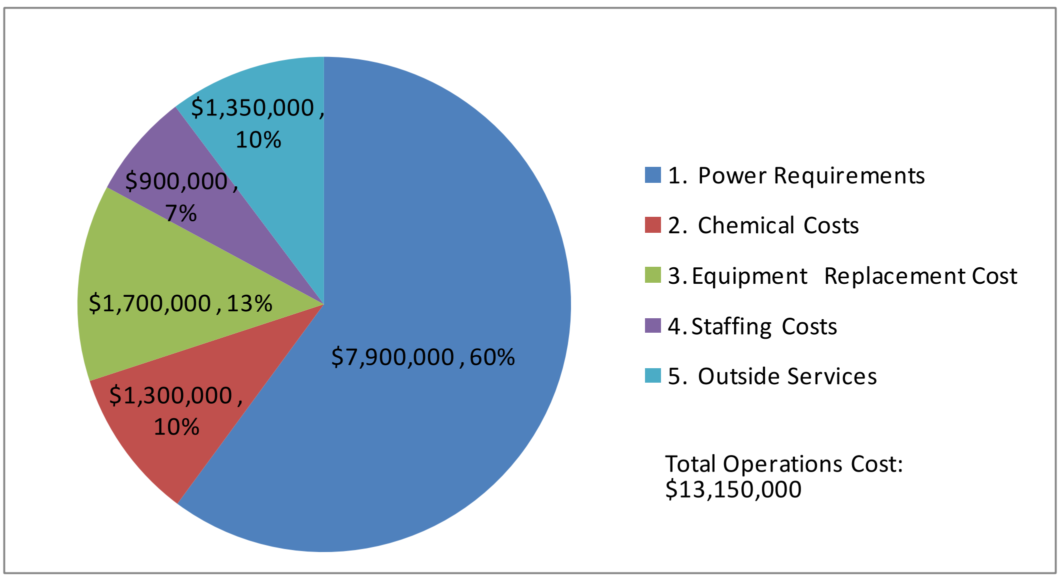


Figure 6-10: Alternative No. 2 Annual Cost Breakdown

6.4.3 Life Cycle Analysis

In order to evaluate the true cost of any project, both the capital and annual costs of the project should be evaluated. Annual costs are converted to a present worth value which takes into account inflation, interest, and the lifetime of the project. The present value of annual costs plus the capital costs is referred to as the Present Worth Value. Similarly, the annual value of capital cost plus the annual operating cost is referred to as the Annual Worth Value.

As indicated in Table 6-5, Alternative No. 1 presents the best combination of capital and annual operating costs as evidenced by its lower net present value.

Table 6-5: Cost Analysis Summary

	Alternative No. 1	Alternative No. 2
Present Worth Analysis		
Total Present Worth Value	$373,400,000	$438,800,000
Net Present Worth, per acre-foot	$550/acre foot	$660/acre foot
Annual Worth Analysis		
Annual Worth Value	$19,050,000/year	$22,450,000/year
Unit Cost of Water, based on Annual Worth, per acre-foot	$840/year/acre-foot	$1,010/year/acre-foot

Another useful means of comparing water projects is to consider the unit cost of water, which is the lifecycle cost of the project divided by the volume of water produced during the lifetime of the plant. Also shown in Table 6-5, Alternative 1 saves approximately $110 per acre-foot of water compared to Alternative 2, based on the Present Worth Analysis method, while Alternative 1 saves approximately $170 per year per acre-foot of water. These values are fairly representative of a highly brackish water source.

6.4.4 Capital Cost of a 71 mgd Desalination Plant

Costs prepared for this pilot study are based on the 25 mgd of feed water, which is currently available from the Mallard Slough based on CCWD's current water entitlement. As indicated in the Design Criteria of **Appendix H**, under this condition, Alternative No. 1 and Alternative No. 2 will provide an average of 19.8 mgd and 19.2 mgd of treated water on an annual basis during dry years. Using the capital costs of Table 6-4, installation costs per unit of treated water produced is determined to be:

- Alternative No. 1 $8.50 capital cost per gpd capacity
- Alternative No. 2 $9.40 capital cost per gpd capacity

If the four agencies were to consider a treatment facility producing 71 mgd of treated water, and using these unit cost values, total capital cost can be estimated to range from approximately $603 million to $667 million, depending on which alternative is selected.

This analysis is very simplistic and excludes economies of scale which would serve to drive the project cost lower. It is presented herein to give an order of magnitude understanding of the capital cost for such a project. If this were to be considered by the four agencies, greater study and evaluation would be required.

6.4.5 Implementation Scenarios

Based on Alternative No. 1, three implementation scenarios have been developed for the proposed desalination facility. In Scenario No. 1, the new desalination facility will be constructed near the Mallard Slough and will use the existing Mallard Slough Pump Station and intake to supply feed water. CCWD's existing source water transmission main will be converted for the conveyance of treated water from the site to the Multipurpose Pipeline. Structures include a new source water pump station, MF/UF building, RO building, chemical buildings (2), clearwell, high service pump station, filtrate tanks, neutralization tanks, thickeners, and solids handling building. Structures are assumed to be placed on pile foundations due to the poor subsurface conditions in this area.

Scenario No. 2 is similar to Scenario No. 1, with the exception that the desalination plant will be operated every third year, with minimal maintenance assumed to be performed during non-operational periods.

In Scenario No. 3, the new desalination facility will be constructed at an undetermined location away from the Mallard Slough and will require a new source water pump station and intake for supply of feed water. Structures also include an MF/UF building, RO building, chemical buildings (2), clearwell, high service pump station, filtrate tanks, neutralization tanks, thickeners, and centrifuge building. Structures will not require pile foundations since it is assumed that the site will not be subject to poor subsurface conditions characteristic of parcels closer to the bay or to the Delta.

A fourth scenario has also been developed and is based on Alternative No. 2. It is very similar to Scenario No 3: an undetermined location will be found away from the Mallard Slough and which will not require pile foundations.

Capital, operating, and net present worth costs for each scenario are illustrated in Table 6-6. Scenario No. 2 is observed to offer the lowest present worth and annual worth value; however, it also produces approximately one-third of the finished water compared to the other alternatives.

Table 6-6: Scenario Cost Summary

	Scenario #1	Scenario #2	Scenario #3	Scenario #4
Capital Costs	**$152,100,000**	**$152,100,000**	**$163,800,000**	**$176,100,000**
1. Sitework	$4,200,000	$4,200,000	$4,200,000	$4,200,000
2. Intake & Raw Water P.S.			$3,100,000	$3,100,000
3. Brine Disposal	$1,100,000	$1,100,000	$1,100,000	$1,100,000
4. MF/UF Facilities	$18,300,000	$18,300,000	$18,300,000	$18,300,000
5. Filtrate Tanks	$1,100,000	$1,100,000	$1,100,000	$1,100,000
6. RO Facilities	$44,100,000	$44,100,000	$44,100,000	$51,300,000
7. Permeate Tank	$500,000	$500,000	$500,000	$500,000
8. Clearwells	$1,900,000	$1,900,000	$1,900,000	$1,800,000
9. High Service Pumping Station	$4,400,000	$4,400,000	$4,400,000	$4,400,000
10. Neutralization Tanks	$400,000	$400,000	$400,000	$400,000
11. Chemical Building A	$1,900,000	$1,900,000	$1,900,000	$1,900,000

12. Chemical Building B	$2,300,000	$2,300,000	$2,300,000	$2,300,000
13. Solids Handling Facilities	$9,900,000	$9,900,000	$9,900,000	$10,600,000
14. Pile Foundations	$3,100,000	$3,100,000		
15. Transmission Main			$7,800,000	$7,800,000
16. Site Electrical Systems	$5,200,000	$5,200,000	$5,200,000	$5,600,000
Subtotal	**$98,400,000**	**$98,400,000**	**$106,200,000**	**$114,400,000**
Contingencies	$19,700,000	$19,700,000	$21,200,000	$22,900,000
Planning, Permit, Eng & Admin	$29,500,000	$29,500,000	$31,900,000	$34,300,000
Land Acquisition	$3,500,000	$3,500,000	$3,500,000	$3,500,000
Concentrate Discharge Permit &	$1,000,000	$1,000,000	$1,000,000	$1,000,000
Annual Costs	**$10,450,000** [1]	**$10,450,000** [1]	**$10,450,000** [1]	**$13,150,000**
1. Power Requirements	$5,400,000	$5,400,000	$5,400,000	$7,900,000
2. Chemical Costs	$1,400,000	$1,400,000	$1,400,000	$1,300,000
3. Equipment Replacement Cost	$1,400,000	$1,400,000	$1,400,000	$1,700,000
4. Staffing Costs	$900,000	$900,000	$900,000	$900,000
5. Outside Services	$1,350,000	$1,350,000	$1,350,000	$1,350,000
Net Present Worth of Annual Costs				
Continuous operation	**$204,900,000**		**$204,900,000**	**$257,800,000**
Operation every 3 years		**$79,000,000**		
PRESENT WORTH VALUE	**$357,000,000**	**$231,100,000**	**$368,600,000**	**$433,900,000**
Annual Worth Value	**$18,210,000/yr**	**$11,790,000/yr**	**$18,810,000/yr**	**$22,140,000/yr**

Note 1: Annual cost during dry year operation. A dry year is assumed to occur once every three years.

6.5 Preliminary Plant Hydraulics

To provide a preliminary understanding of treatment pressures and water surface elevations for the proposed desalination facility, a preliminary hydraulic profile for Alternative No. 1 is furnished in Figure 6-11. Key aspects are as follows:

1. Initial feed pressure is approximately 57 psi, which is necessary to direct feed water through the self-cleaning screens and UF system to the downstream atmospheric Filtrate Tank. This value includes pressure loss through partially clogged intake fish screens and pre-screens, transmembrane pressure losses through the UF membrane, piping head losses, and allowance for backpressure necessary for operation of the self-cleaning screens.

2. Booster pumps will elevate filtrate to approximately 50 psi upstream of the cartridge filters and serve to pressurize the inlet to the RO high pressure pumps.

3. RO high pressure pumps will operate based on membrane osmotic pressure, which is a function of several factors, including salinity, temperature, and membrane type. RO feed pressures are estimate to range from approximately 217 psi to 590 psi.

4. An interstage boost system is furnished and will add 100 psi to the Stage 1 concentrate pressure being fed to the Stage 2 membranes.

5. As indicated in the Feasibility Report and confirmed by agency personnel, the treated water pump station must generate sufficient energy to overcome residual pressure in either the CCWD Multipurpose Pipeline or the EBMUD Mokelumne Aqueduct, as well as friction losses in the 3 mile transmission main. It is estimated that approximately 190 psi will be required to discharge treatment plant treated water to the Multipurpose Pipeline and 240 psi will be required for discharging to the Mokelumne Aqueduct.

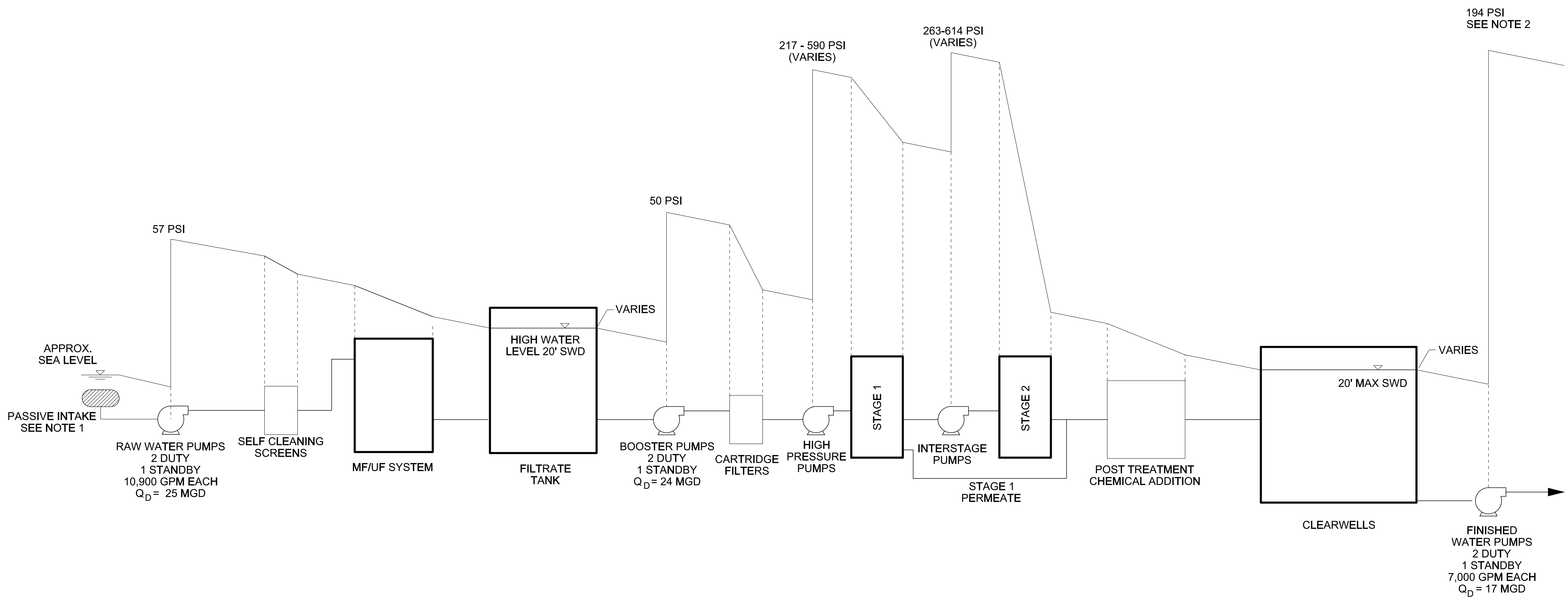
All values shown in the figure are for preliminary planning purposes only and must be verified during subsequent design phases of this project.

NOTES:
1. EXISTING SCREENS AT THE MALLARD SLOUGH PUMP STATION COULD POTENTIALLY BE USED IN LIEU OF A NEW INTAKE.

2. BASED ON 124 PSI PRESSURE AT THE CCWD MULTIPURPOSE PIPELINE PLUS ENERGY REQUIRED FOR TRANSMISSION OF FINISHED WATER TO THE AQUEDUCT AND STATIC LIFT.

Q_D: DESIGN FLOW
SWD: SIDE WATER DEPTH

ALTERNATIVE 2 IS SIMILAR.



REV	DATE	BY	DESCRIPTION	SCALE: NONE	WARNING: 0 ½ 1 IF THIS BAR DOES NOT MEASURE 1" THEN DRAWING IS NOT TO SCALE	DESIGNED / DRAWN / CHECKED	MWH AMERICAS, INC. 2121 N. CALIFORNIA BLVD WALNUT CREEK, CALIF. 94596	MWH	BAY AREA REGIONAL DESALINATION PROJECT	FIGURE 6-11 ALTERNATIVE 1 PRELIMINARY HYDRAULIC PROFILE	SHEET

File: HYDRAULIC PROFILE.dgn Model: Layout1 ColorTable: bw.ctb DesignScript: MWH_Iplot_Pentable_v85.pen PlotScale: 40:1 User: Jennifer A Gelmini Plot Date: 02-DEC-2009 09:10

6.6 Full Scale Project Implementation

A preliminary implementation diagram is furnished in Figure 6-12, including estimated durations for each main activity. Once the pilot study is finalized, the four agencies will need to develop the necessary interagency agreements which will serve to define roles, responsibilities, and obligations among and between the various parties. Key issues will also need to be resolved, such as ownership for the proposed desalination plant, intake and treated water conveyance facilities; operational responsibilities; costs and benefits sharing; and the transfer of treated water and other interagency water transfers necessary to assure equity to each party. These latter issues may be confronted while subsequent tasks are being finalized.

Once the agencies are in agreement with an overall framework for moving forward, the next important activity is site identification and selection. From a technical standpoint, site selection is very important for proper equipment sizing and selection. Feed water obtained from Suisun Bay is subject to wide salinity variations, with lower maximum TDS values encountered further up the Sacramento River delta. This will have a direct bearing on total project cost and on operating costs.

Piloting at Mallard Slough provided data to suggest that a full-scale facility is quite viable in this location; however, project economics and technical application will likely vary at another site. A siting study will also need to address the social and community issues related to a new desalination plant, with final site to be selected during the EIR phase of this project. Appropriate parcels may be available at locations remote to Mallard Slough and may offer advantages relative to access, subsurface conditions, cost, or proximity to piping infrastructure.

Once the site is identified, or while the site selection is ongoing, the agencies may begin to conduct several activities simultaneously. Intake assessment, field survey, preliminary design and geotechnical studies may all be performed. Information developed in each activity will be utilized to complete environmental assessments and for NEPA and CEQA compliance. Land acquisitions may also be initiated, including easements as well as direct purchase as necessary.

With 18 months set aside for preparing detailed design and construction documents, it is estimated that the initial planning and design will require approximately 48 months for completion. Construction and commissioning of the new desalination plant will require an additional 30 months, depending on intake, treatment, and treated water interconnection objectives.

6.7 Water Rights on Mallard Slough

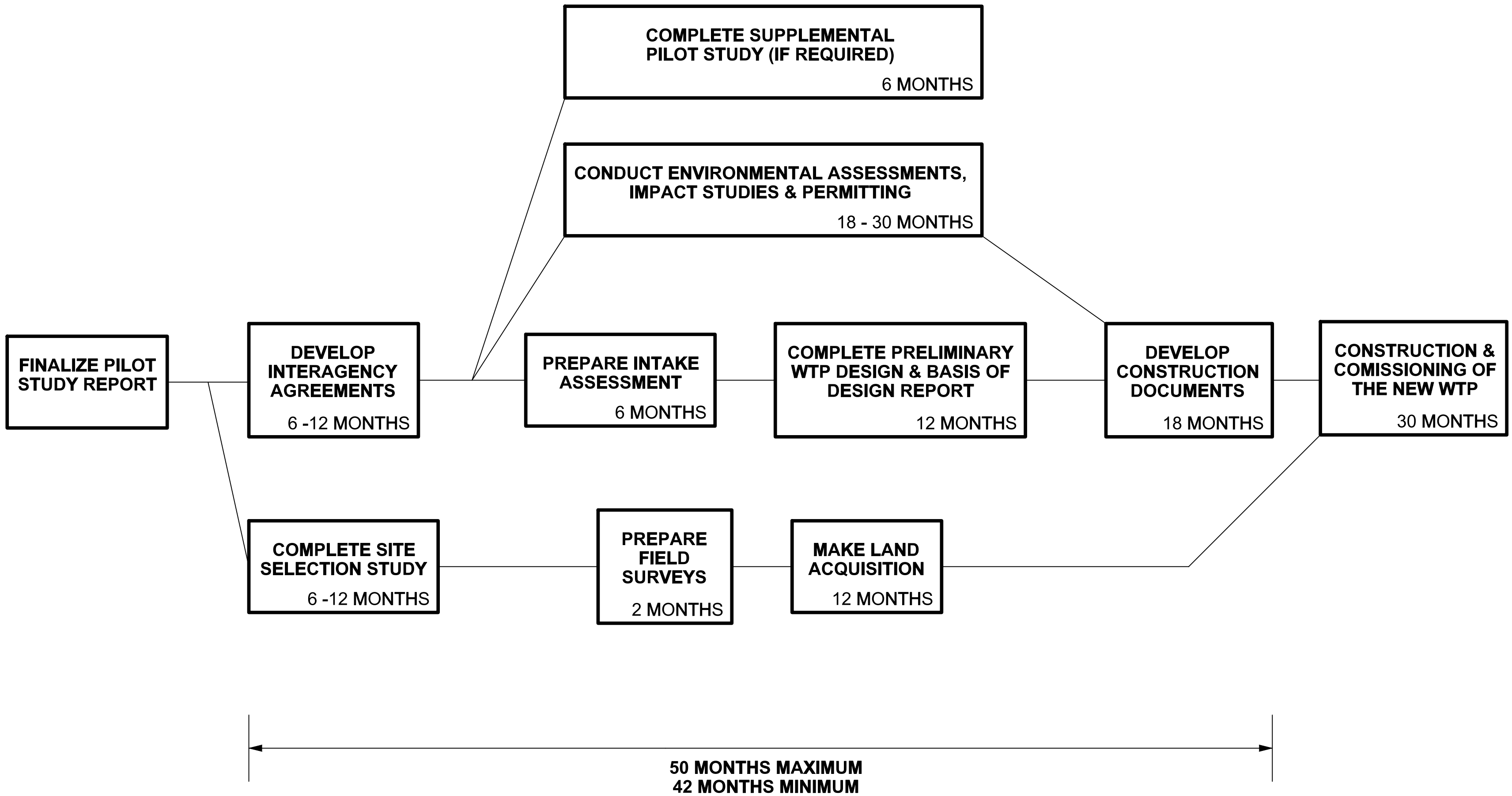
As indicated by the documentation in **Appendix J**, CCWD currently maintains two water rights at the MSPS: a license and a permit. The license was applied for in 1928 and perfected in 1971. It allows for year-round withdrawals with a maximum direct diversion rate of 39.3 cfs, a beneficial use not to exceed 13,690 acre-feet, and a maximum total to storage of 3,780 acre-feet per calendar year. The combined maximum total direct diversion and storage from the license is 14,880 acre-feet per year.

The permit was applied for in 1983, and allows for withdrawals within a 5 month period from August 1 through December 31, with a maximum direct diversion rate of 39.3 cfs and a maximum total diversion of 11,900 acre-feet per year. Due to the variations in annual diversions at MSPS a consistent pattern of use has yet to be established, therefore CCWD has yet to perfect the permit.

MSPS is potentially subject to a District-wide 30-day no-diversion period, the default timing of which is the month of April, but which is subject to revision based on other regulatory permits and operating negotiations. If diversions from the license are distributed across 11 months (excludes April), the license could supply approximately 14.5 mgd. Alternatively, if the license withdrawals are distributed over 6 months (January through July, excluding April), the license could supply approximately 26.6 mgd. Similarly, if withdrawals from the permit are distributed across 5 months (August through December) the permit could supply 25.34 mgd.

The permit and license provide numerous possible withdrawal combinations on a monthly and seasonal basis. For example if water from the license is withdrawn over 11 months the supply is 14.47 mgd from January through July (excluding April) and 39.81 mgd from August through December when water from the permit can also be withdrawn.

For the purposes of this report, it is assumed that the 25 mgd would be utilized for the full-scale desalination facility, and future increase in plant production would require expansion of existing diversion rights.



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BAY AREA REGIONAL DESALINATION PROJECT

FIGURE 6-12
BARDP IMPLEMENTATION DIAGRAM

7.0 PUBLIC OUTREACH

The four member agencies jointly developed an outreach program to inform the public about the pilot test and the BARDP. Elements of the program included targeted outreach to interested groups, workshops, and information dissemination using fact sheets, media releases, and postings on agency and project websites. Several speaking engagements and conference presentations were conducted. Public input received during public outreach has been incorporated into the Pilot Project Engineering Report.

7.1 Open House Events

Two public open house events were conducted, one during the pilot testing and the other after pilot test was completed and the data analyzed. The open houses were held as follows:

- Open house, San Francisco, SFPUC headquarters, December 16, 2008.
- Open house, EBMUD headquarters, Oakland, December 9, 2009.

The open houses were open to members of the public and were advertized in local newspapers (included in **Appendix D**). Factsheets and poster boards were developed for the open houses (included in **Appendix D**). Figure 7-1 shows the open house held at EBMUD office in Oakland on December 9, 2009.

Figure 7-1: Open House at EBMUD, Oakland, California

7.2 Pilot Test Video

The pilot test site was located in remote area of Pittsburg, California. Due to security concerns, the pilot test equipment was placed in secured containers which were locked when staff was not present. The pilot test site area was constrained physically with limited access and parking options. Due to the above-mentioned reasons, public site tours were not offered. Instead, the four agencies developed a video to present the pilot testing site and equipment. The video has been displayed to the public and has been received well (included in the **Appendix D**). Key personnel for each agency were interviewed as well as agency project managers and onsite operations staff. The video provided the public with an opportunity to observe the pilot plant in action and proved to be a suitable substitute for public site tours.

7.3 Presentations and Public Speaking

Several presentation and public speaking events were conducted by project staff. The following public outreach events were conducted during the course of this project.

- Presentation to Contra Costa Council Water Task Force, Walnut Creek, June 17, 2008.
- Presentation to Richmond-Pinole Lion’s Club, San Pablo, February 18, 2009.
- Presentation to Walnut Creek Lion’s Club, Walnut Creek, April 8, 2009.
- Presentation to SIR-51 meeting, Los Altos Hills, January 6, 2009.

In all cases, the audiences were very interested and engaged in the discussion.

7.4 Conference Presentations

Project staff also attended National and State conferences. The following presentations were given, or are scheduled:

- AMTA Technology Transfer Workshop, Santa Rosa CA, March 12-13, 2008
- AMTA Annual Conference and Exhibition, Austin TX, July 13-16, 2009
- AMTA Annual Conference and Exhibition, San Diego CA, July 12-15, 2010
- WateReuse Chapter meeting, San Francisco CA, December 4, 2009
- 2008 National Salinity Summit, Las Vegas NV, January 17, 2008
- AWWA Annual Conference, San Diego CA, June 14-18, 2009
- AWWA Annual Conference, Atlanta GA, June 8-12, 2008
- AWWA Membrane Technology Conference, Memphis TN, March 15-19, 2009
- AWWA California-Nevada Section Spring Conference, Hollywood CA, March 21-April 1, 2010

7.5 Website Development

A website was created to facilitate dissemination of information to the public. The website address is www.regionaldesal.com. Current information regarding the project may be accessed from the site.

Public outreach efforts and activities such as presentations as guest speakers, media releases, fact sheets dissemination, and website updates will be continued by the member agencies.

8.0 SUMMARY AND RECOMMENDATIONS

The following sections represent major findings of the pilot study and how the data gathered can be used as the BARDP moves forward into design and beyond.

8.1 A Desalination Facility is Technically Feasible at Mallard Slough

Piloting at Mallard Slough provided data to suggest that a full-scale facility is viable in this location; however, project economics and technical application will likely vary at another site. Site selection is very important for proper equipment sizing and selection. Source water obtained from Suisun Bay is subject to wide salinity variations, with lower maximum TDS values encountered further up the Sacramento River delta. This will have a direct bearing on total project cost and on operating costs.

A key aspect of site selection is the availability of options at each potential desalination plant site for a new source water intake. The Mallard Slough site, for example, offers an existing surface water intake which is already owned and operated by CCWD. Other surface or subsurface intake types as described in Technical Memorandum No. 2A would likely be considered if the site were to shift from Mallard Slough to another location.

It is important to recognize that this pilot did not study water from the Suisun Bay, but from a semi-stagnant water body which may have impacted water quality either by dampening the true impact of salinity variations induced by the bay, or by adding a TOC or algae load which is not present in the bay. The pilot intake was located at the end of this slough, approximately 3,000-feet from the bay itself. While a new desalination plant designed on the Mallard Slough may be based on these pilot results, additional pilot-scale activities will be needed if the full-scale site is located elsewhere.

8.2 Pressurized Pretreatment System Was Successful in Pilot Trials

Both of the pretreatment systems tested produced a suitable feedwater for the RO in the pilot testing period. However, the long-term viability of the two systems did not appear to be the same based upon the short pilot duration.

The submerged membranes experienced irreversible fouling that was not recoverable by CIPs, even after modifying the CIP procedures for more aggressive cleaning. The initial permeability was 10.9 gfd/psi, and after three acid CIPs, it subsequently declined to 7.4, 5, and 4.3 gfd/psi. A fourth, more aggressive CIP was undertaken including an acid CIP, an overnight chlorine soak, followed by another acid CIP, and the resulting permeability was 5.7 gfd/psi – an improvement, but not close to the initial 10.9 gfd/psi. During the first run, the acid MW procedures were disabled, which could explain some of the fouling issues; however, the continued permeability decline through Run 2 is not explained by this.

The initially-installed submerged membranes were the manufacturer's new formulation with a better retention of small particles and mechanically stronger. At the time of installation, the membranes were in the final stages of testing to receive recognition by DPH for acceptance on their list of alternative filtration technologies. After these membranes were irreversibly fouled in the pilot testing, they were replaced with the older formulation. The flux in Run 3 appeared to be sustainable, and the CIP was more successful, but it also represented the inaugural run for the new membranes during low TDS conditions, and thus would need to be repeated for verification. There was not sufficient operating time to do this during the pilot test.

The pressurized membranes, on the other hand, experienced consistent permeability and no significant fouling, regardless of the flux. In the instance where the membranes did foul due to a suspected ferric overdose into the on-skid feed water tank, they fouled quickly and completely, and all of the permeability was recovered with the CIP.

Therefore, because the long-term operability of the second set of submerged membranes is not clear, and because the pressurized membranes ran smoothly, the pressurized inside-out membranes were used for the scale-up evaluation. The budgetary capital cost estimate for the full-scale UF system is $18.1M for a system that can produce 24 mgd of filtrate, and $31.2M for a system that can produce 87 mgd of filtrate, not including installation costs or sales taxes. Submerged membrane capital costs are expected to range from 40% to 50% higher due to greater membrane area requirements. Additional capital will be needed for concrete or steel basins necessary to house the submerged membranes, although higher recovery will allow a reduction in solids handling systems.

The advantages of the pressurized membranes are proven operability and higher specific flux than most membrane manufacturers. The disadvantages of using Norit membranes at the full scale are coagulant requirement (requiring solids processing) and lower system recovery.

After the full-scale site is selected, any membrane will require additional pilot-scale activities to determine appropriate flux and recovery parameters, with the possible exception of Norit membranes installed on Mallard Slough feedwater. Submerged membranes could potentially be reconsidered at that time.

8.3 RO1 System Achieved Balance between Recovery, Water Quality, and Efficiency

Each treatment train had distinct operational advantages and disadvantages which are broadly summarized in Table 8-1.

These general trends were anticipated before testing began due to membrane evaluation using performance projection models. However, running the systems at pilot scale over the diverse feedwater conditions in the testing period provided specific performance data that can be used to project full-scale capital and operational expenditures.

None of the three desalination systems experienced quantifiable fouling, and it is projected that all three systems would be able to achieve a 90-day CIP interval. Therefore, the design of the full-scale facility should be based upon the system performance to meet water quality goals, and the present worth cost estimate based upon operating parameters determined in pilot testing such as recovery and feed pressure.

Table 8-1: Desalination System Performance Comparison

	Goal	RO Train No. 1	RO Train No. 2	NF Train No. 3
Description		Two Stage Brackish and Seawater Membranes	Single Stage Seawater Membranes	Single Stage Nanofiltration Membranes
Recovery	High	70-82%	50-62%	50-60%
Specific Flux, gfd/psi	High	Typically 0.1	0.07-0.075	0.19-0.26
Permeate TDS, mg/L	< 500	<10-120	<10-27	<10-220
Permeate Boron, mg/L	< 0.5	0.06-0.48	<0.05-0.2	0.08-0.69
Permeate Chloride, mg/L	< 100	<4-67	<4-11	5-130
Permeate Sodium, mg/L		<2-43	<1-7.8	<2-82
Permeate Turbidity, NTU		<0.05	<0.05	<0.05
Permeate TOC, mg/L		<0.1 – 0.5	<0.1 – 0.5	<0.1 – 0.5
Permeate Iron, mg/L		< 0.011	< 0.011	< 0.011
Permeate Aluminum, mg/L		< 0.021	< 0.021	< 0.021

Based upon the pilot data, RO1 appears to be the most suitable system for full-scale because it achieves a high recovery, while meeting water quality goals. RO2 has the best permeate water quality, but permeate produced by RO1 is adequate for this project and meets agency goals. NF3, although having the lowest energy use, does not meet the project's goal for chloride or boron in the dry season.

Given the widely varying salinity of the source water as encountered during the study, the single stage pilot systems are probably not representative of a full-scale system which could be optimized for this project. Recovery and energy efficiency will likely be improved by utilizing multiple stages, as demonstrated with RO1, or by combining single stage membrane vessels to achieve low energy use during low salinity months and maximum water quality during dry months. Optimizing a single stage system for both high and low TDS would be quite challenging and would result in inefficiencies at extreme operational ranges.

Therefore, the full-scale evaluation focuses on two options:

Alternative No. 1: RO Train No. 1, as piloted.

- 2:1 array
- Brackish water RO membranes in Stage 1 and seawater RO membranes in Stage 2. All feed water is pumped first to the brackish water membranes, with only concentrate being directed to the second stage seawater membranes.

- Interstage boosting with VFD.
- An average of 70% total system recovery during dry conditions.

Alternative No. 2: A hybrid plant that has independent trains for RO Train No. 2 and NF Train No. 3.

- Two single-stage membrane systems operating in parallel: NF RO membranes in one train, and seawater membranes in the second train.
- Feed water split between the membrane systems to meet treated water goals, with the seawater RO train handling approximately 70% of the feed flow during the dry seasonal periods due to its better water quality, and the NF train treating nearly 100% of the feed flow during lower TDS periods due to its lower feed water pressure requirement.
- NF concentrate partially recovered by blending with filtrate being pumped to the seawater RO train.
- An average of 58% total system recovery during dry conditions.

Both of these options are projected to meet the water quality requirements of the project in all seasons, and are expected to be viable desalination systems for the full-scale plant. Either option could be carried forward into the design phase. A present worth evaluation in the following section provides a more specific recommendation.

8.4 Pilot Data Provide a Basis for Full-Scale Cost Estimates for a Mallard Slough Plant

The full-scale evaluation relied on a Norit pressurized pretreatment membrane. There were two desalination alternatives evaluated due to the complex feedwater. Therefore, in order to gain an understanding of the differences in costs between these different desalination alternatives, capital an annual operating cost estimates were developed. In the two evaluations, the full-scale facilities were the same except for the desalination system. The results of the cost evaluation are summarized below in Table 8-2.

Table 8-2: Present Worth Evaluation Summary

	Alternative 1	Alternative 2
Total Capital Cost	$168,500,000	$181,000,000
Annual Cost	$10,450,000	$13,150,000

Based upon both capital and operating costs, which favor Alternative No. 1, the design of a full-scale facility should be based on RO1, a two-stage system with brackish water RO (BWRO) membranes in the first stage and seawater RO (SWRO) membranes in the second stage.

8.5 Biological Impacts Due to Intake and Concentrate were Minimal

The species composition of larval fishes collected during entrainment and source water sampling around the pilot plant intake was consistent with published life history information for species found in Suisun Bay. The estimated small annual loss of adult prickly sculpin and bluegill/redear sunfishes is unlikely to affect adult populations. Entrainment of longfin/delta smelts occurred during the sensitive fish period of January through June when these larvae are normally present in the vicinity of Mallard Slough. Entrainment of these listed species at a full-scale desalination facility would require Endangered Species Act consultation with USFWS and CDFG for delta smelt and CDFG for longfin smelt.

The concentrate toxicity testing conducted as part of this project indicated that no significant growth toxicity of the desalination concentrate was found for the algae, and no significant survival or growth toxicity was found for the invertebrate or fish test organisms for either the dry weather sample or the wet weather sample. Because neither salinity- nor contaminant-related toxicity was found, it was not possible to distinguish the relative effects of each. The toxicity results suggest that, if the concentrate samples tested are representative of those at an operational full-scale desalination plant at the Mallard Slough location, then there would be no expected toxic effects of the effluent on biota.

Biological impact results from the pilot study showing little to no impact are not a guarantee that a full-scale facility will be permitted. The information collected during piloting should be used as a starting point to open discussions with all the regulatory agencies with jurisdiction over the proposed project. This process should begin early in the site selection and preliminary design process to ensure the facility addresses regulatory concerns and to minimize rework associated with potential mitigation measures.

8.6 Treated Water Can Be Compatible with Existing Transmission Systems

Two methods for post-treatment stabilization were evaluated at the bench-scale using pilot plant permeate:

1) A liquid lime dose of 40 mg/L with a carbon dioxide dose of 30-40 mg/L resulted in a stable permeate from RO1. The resulting post-treated permeate also required a 2 mg/L dose of sodium hydroxide to reach a suitable pH for the transmission systems.
2) Continuous flow through calcite bed filters at a loading rate of 3 gpm/sf and an empty bed contact time of 10 minutes. The resulting post-treated permeate also required a 1.5-2 mg/L dose of sodium hydroxide to reach a suitable pH for the transmission systems

Both methods tested produced a stable product water which could be blended with EBMUD aqueduct water and CCWD pipeline water. The cost estimate in this report is based on post-treatment with lime in combination with carbon dioxide.

The waters from EBMUD's Aqueduct #2 and CCWD's Multipurpose Pipeline, however, appear to exhibit corrosive tendencies by themselves based on Langelier Saturation Index (LSI) and Calcium Carbonate Precipitation Potential (CCPP). Blend ratios of permeate from a full-scale desalination facility with these waters needs to be carefully considered in the future, as the blends evaluated in the bench-scale test were not sufficient to drive the blended water to positive LSI or CCPP values.

Once the full scale is selected and the receiving agency(ies) have determined where the treated water will be added into the local distribution system(s), additional studies should be completed for further water quality compatibility and corrosion assessment. Pipe loop or material coupon studies would provide a more complete impacts of blended water on the agency(ies) infrastructure.

8.7 Site Selection is an Important Next Step

As discussed in Section 4.3.4.4, the final site for a full-scale facility has not yet been selected. Site selection is very important for proper equipment sizing and selection. Source water obtained from Suisun Bay is subject to wide salinity variations, having a direct bearing on total project cost and on operating costs. Water quality, project economics, solids disposal, and technical application will likely vary at another site.

A maximum TDS design range between 11,500 mg/L and 12,000 mg/L is recommended for this site. During historical dry years, the plant would normally operate between 500 mg/L and 5,500 mg/L. This maximum TDS is based on historical data from Suisun Bay. Additional pilot-scale activities may be needed if the full-scale site is located elsewhere. Potential impacts from upstream storage facilities could also affect the range of salinity, especially if patterns are defined in drought years when the full scale facility would be operating. These and other potential feedwater quality impacts should be evaluated once the full scale site is selected.

A key aspect of site selection is the availability of options at each potential desalination facility site for a new source water intake. The Mallard Slough site, for example, offers an existing screened source water intake with two 25 mgd pumps and which is already owned and operated by CCWD. Other surface or subsurface intake types as described in Technical Memorandum No. 2A would likely be considered if the site were to shift from Mallard Slough to another location.

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