

# **Assessment of Osmotic Mechanisms Pairing Desalination Concentrate and Wastewater Treatment**

## **Report**

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P.O. Box 13231, Capitol Station  
Austin, Texas 78711-3231  
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# **Texas Water Development Board**

## **Report 0804830852**

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## LIST OF ABBREVIATIONS

AACE	Association for the Advancement of Cost Engineering
ACS	American Chemical Society
AMTA	American Membrane Technology Association
avg	Average
AWWTP	Advanced Waste Water Treatment Plant
bar	Bar pressure
BPUB	Brownsville Public Utilities Board
BW	Brackish Water
CBOD	Carbonaceous Biological Oxygen Demand
cm	Centimeter
CP	Concentration Polarization
CPES	CH2M HILL's parametric cost estimating system
CSM	Colorado School of Mines
CTA	Cellulose Triacetate
d	Day
DI	Deionized water
DI-BW	Dionized water – brackish water
DI-SW	Deionized water – seawater
DO	Direct Osmosis
DOC	Dissolved organic carbon
dP	Differential pressure
DS	Draw Solution
ECP	External concentration polarization
ED	Electrodialysis
EDR	Electrodialysis Reversal
EDTA	Ethylenediaminetetraacetic acid
EPWU	El Paso Water Utilities
FH	Fred Harvey WWTP
FO	Forward Osmosis
FO-LT	Forward Osmosis – Low Temperature Distillation
FO-RO	Forward Osmosis – Reverse Osmosis
frac	Hydraulic fracturing of rock
ft	Foot
g	Gram
GE	General Electric
gfd	Gallons per square foot per day
g/L	Grams per liter
gpd	Gallons per day

gpm	Gallons per minute
GW	Groundwater
GWI	Global Water Intelligence
hr	Hour
HS	Haskell Street WWTP
HSD	Homogeneous solution diffusion
HTI	Hydration Technologies Incorporated
ICP	Internal Concentration Polarization
kPa	Kilopascal
kWh	Kilowatt-hour
lb	Pound
L	Liter
L/hr-m <sup>2</sup>	Liters per square meter per hour
LMH	Liters per square meter per hour
m	Meter
Max	Maximum
MBR	Membrane Bioreactor
MED	Multiple effect distillation
MF	Microfiltration
mgd	Million gallons per day
mg/L	Milligrams per liter
Min	Minimum
mS	Milli-Siemens
MSF	Multistage Flash Distillation
MUD	Municipal Utility District
NaCl	Sodium chloride
NASA	National Aerospace Agency
NDP	Net driving pressure
NF	Nanofiltration
NTU	Nephelometric Turbidity Unit
O&M	Operations and maintenance
ODMP	Osmotically Driven Membrane Process
PRO	Pressure Restrained Osmosis
psi	Pounds per square inch
psid	Pounds per square inch differential
psig	Pounds per square inch gauge
PVC	Polyvinyl chloride

R	Recovery
RB	Roberto Bustamante WWTP
RO	Reverse osmosis
ROSA	Reverse Osmosis Systems Analysis
S	Siemens (conductivity)
s	Second
SB-MBR	Sequencing batch – Membrane bioreactor
SBR	Sequencing batch reactor
SCADA	Supervisory Control and Data Acquisition
SecEff	Secondary effluent
SecEFF-BW	Secondary effluent – brackish water
SecEff-SW	Secondary effluent – seawater
SPLT	South Platte
SPLT-BW	South Platte – brackish water
SPLT-SW	South Platte – seawater water
SPMWD	San Patricio Municipal Water District
SRWA	Southmost Regional Water Authority
SurW	Surface water
SW	Seawater
SWRO	Seawater Reverse Osmosis
TDS	Total dissolved solids
TFPA	Thin-film polyamide
TOC	Total organic carbon
TSS	Total suspended solids
TWDB	Texas Water Development Board
UF	Ultrafiltration
USBR	United States Bureau of Reclamation
UVA	Ultraviolet Absorbance
UV-254	Ultraviolet Absorbance at a wavelength of 254 nanometers
WSC	Water Supply Corporation
WTP	Water treatment plant
WWTP	Wastewater treatment plant
ZLD	Zero-liquid discharge



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## LIST OF SYMBOLS

$A_{\text{mem}}$	Area, membrane
$A, K_w$	Mass transfer coefficient, water; also, water permeability coefficient
$B, K_s$	Mass transfer coefficient, solute (salt); also, solute (salt) permeability coefficient
$^{\circ}\text{C}$	Degree centigrade
$C$	Concentration
$C_f$	Concentration, feed stream
$C_{i,\text{in}}$	Concentration, impaired water stream pumped into FO cell
$C_{i,\text{out}}$	Concentration, impaired water stream coming out of FO cell
$C_m$	Concentration at the membrane surface
$C_p$	Concentration, permeate stream
$C_c$	Concentration, concentrate stream
CF	Concentration Factor
$\Delta\pi$	Difference between osmotic pressure, typically across the membrane
$\Delta C$	Difference in concentration, typically across the membrane
$\Delta P$	Difference in pressure; typically applied differential pressure across the membrane
dP	Differential pressure, feed-concentrate flow stream
$J_s$	Flux solute (salt)
$J_w$	Flux water
M	Molar concentration of dissolved species
$M_{\text{DI}}$	Mass of deionized water
$M_{\text{DS}}$	Solute mass of concentrated draw solution
NAP	Net Applied Pressure
Q	Flow
$Q_f$	Flow, feed stream
$Q_p$	Flow, permeate stream
$Q_c$	Flow, concentrate stream
R	Ideal gas constant (Equation 2-1)
R	Recovery rate (Equation 2-7)
r	Rejection
$R^2$	Correlation coefficient
sg	Specific gravity
t	Time, duration
T	temperature $^{\circ}\text{Kelvin}$
TMP	Trans-membrane Pressure
$\rho$	Density of concentrate
$\rho_f$	Density of feedwater
$\rho_p$	Density of permeate
$\sigma$	Reflection coefficient
$\mu$	Viscosity
$\pi$	Osmotic pressure
$\pi_{\text{D}}$	Osmotic pressure of the draw solution
$\pi_{\text{f}}$	Osmotic pressure of the impaired water stream

Subscripts

<i>D</i>	<i>Draw solution</i>
<i>c</i>	<i>Concentrate</i>
<i>f</i>	<i>Feed solution (impaired water)</i>
<i>I, i</i>	<i>Impaired water</i>
<i>p</i>	<i>Permeate</i>
<i>Poten</i>	<i>Potential</i>
<i>s</i>	<i>Solute</i>
<i>w</i>	<i>Water</i>
<i>ww</i>	<i>Wastewater</i>

## **Executive Summary**

With the population of Texas expected to nearly double in the next 50 years, Texans will be searching for new, sustainable, and drought-proof water resources (TWDB, 2010). Planning agencies and municipalities continue to rely on the expanded use of existing renewable and non-renewable water resources including fresh groundwater and surface water. Existing resources are limited in terms of their geographic reach as well as their ability to sustain population growth well into the future. In addition, these water resources face an uncertain regulatory environment. An underutilized resource in the state is treated wastewater effluent as a source of reclaimed water for industrial and commercial uses as well as landscape irrigation. Cities such as San Antonio and El Paso have substantial reclaimed water systems that supplement their water portfolios, facilitating their mission to meet ever-growing water demands.

Disposal of desalination concentrate is a major cost component of water desalination, especially in inland communities where the alternatives for disposal of the concentrate are limited and costly. Identifying and implementing applications for desalination concentrate (beneficial reuse) can significantly reduce the overall cost of a water desalination project as well as augment existing water supplies. One potential beneficial application is to take advantage of the high salinity levels of desalination plant concentrate by using it as a “draw solution” in a forward osmosis process. This report describes work conducted to evaluate a hybrid forward osmosis/reverse osmosis process to recover water from treated wastewater effluent for beneficial use.

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To accomplish these objectives, data from four Texas desalination plants and several wastewater treatment plants were analyzed to determine the osmotic potential for FO applications. Bench-scale pilot testing was completed to evaluate water fluxes and reverse-direction solute fluxes. Extensive pilot scale testing was also conducted at the Colorado School of Mines to examine the ability of simulated concentrate to act as a draw solution to recover treated wastewater. The data indicated that simulated concentrate of up to 60 grams per liter (g/L) can effectively extract water from secondary wastewater produced from a membrane bioreactor plant.

Cost modeling was also conducted, which indicated that the use of FO-RO is not cost competitive when compared to tertiary treatment of wastewater using an advanced treatment

process including membrane filtration following by RO and UV disinfection. However, assuming that membranes can be commercially produced at a reasonable price point, it is anticipated that use of FO-RO may be viable at some point in the future. Finally, in conducting this study, several recommendations were developed for continuing the development of the hybrid FO-RO process.

## 1. Introduction

Texas Water Development Board (TWDB) has identified disposal of desalination concentrate as a major cost component of water desalination (TWDB, 2008). Current practices for concentrate management have been identified by Mickley (2001) and typically include evaporation ponds, deep well injection, surface discharge, sewer discharge and zero liquid discharge, depending upon a project's location and concentrate volume. Useful applications for desalination concentrate can significantly reduce the cost of water desalination. Research efforts are currently underway by a number of parties attempting to commercialize recovery of various salts from concentrate, including companies such as GeoProcessors Inc., New Sky Energy, and Calera.

Another potential beneficial application for concentrate is to take advantage of the high salinity level of desalination concentrate, and its osmotic pressure, by using it as a draw solution in a forward osmosis process to remove water from wastewater or other impaired water for volume minimization.

Forward osmosis (FO) is an emerging membrane treatment process that belongs to the class of osmotically-driven membrane processes. Unlike pressure-driven membrane processes, FO is driven by the concentration gradient created across a semi-permeable membrane by two flowing streams of different concentrations, and therefore different osmotic pressures. Water flows (diffuses) from the stream with lower osmotic pressure (feed or impaired solution) to the one with higher osmotic pressure (draw solution). The higher the osmotic pressure differential across the membrane, the higher the flow of purified water across the membrane. When this flow of water is normalized by the membrane surface area, it is referred to as water flux. Past and current research has shown that FO membranes are good barriers to a broad range of contaminants, including bacteria, protozoa, viruses, and other dissolved organic and inorganic constituents in contaminated water (Cath and others, 2006).

In order to beneficially use water extracted from the impaired solution using FO, another separation process must be used to recover water from the now diluted draw solution exiting the FO membrane unit. While a number of different draw solutions have been tested by researchers, each with distinct separation process associated with it including magnetic molecules (electromagnetic reconstitution), large macro-molecules (UF reconstitution), and thermally degradable salts (distillation reconstitution), few of these draw solutions possess high enough osmotic pressures to be of value in a practical desalination application (Adham and others, 2006). The reconstitution step to recover beneficial water typically utilizes significantly more energy than the FO process itself. When salt solutions, or reverse osmosis (RO) waste streams (concentrate) are used, the coupling of FO with RO has many potential advantages. One of such potential advantages is that the concentrate generated during RO membrane desalination can be reused as a high osmotic pressure draw solution to drive an FO process, thereby simultaneously reducing the cost of concentrate (or brine) disposal (either through dilution or volume reduction, depending upon the application) and producing desalted water. While the RO process is relatively well known, there is still need to explore many aspects of the FO process. In order to successfully implement the FO-RO hybrid processes, further research is needed of the FO process on its own (that is, investigate the performance of FO membranes with regard to water

flux and rejection of solutes and solids), recovery of water from the draw solution (using RO or other methods), and investigation of the performance of the combined FO-RO process.

Many issues related to the implementation of FO and FO-RO hybrid systems in Texas are not well understood. These include technical issues associated with technology commercialization, legal issues related to permitting and water rights, and implementation issues associated with evaluation of potential suitable locations for deployment of the technology and how it fits with local, regional, and statewide water planning efforts.

In order to evaluate the feasibility of recovering wastewater using an FO process with concentrate from a RO system, TWDB contracted a team consisting of CH2M HILL and Colorado School of Mines to conduct a combination laboratory/feasibility study to evaluate the use of RO concentrate as a draw solution in a FO application. This document is the product of a research study funded by the TWDB to develop information necessary for planners and designers to:

- Examine the mechanics of a hybrid FO and RO water treatment system,
- Assess the feasibility of using high salinity streams to extract water out of treated wastewater streams,
- Determine the characteristics required for cost-effective application of FO to wastewater treatment with desalination concentrate as a draw solution, and
- Identify the costs associated with implementation of FO and FO-RO systems throughout the state of Texas.

The information provided in this document was developed by completing the following research tasks:

- Perform a literature search to identify and document the current state of technology.
- Survey several existing desalination plants and wastewater treatment plants to characterize the available water quality of concentrate and wastewater streams.
- Conduct a screening and selection of different water quality combinations for laboratory and treatability testing.
- Conduct bench-scale and pilot testing of a coupled hybrid FO-RO system.
- Perform system and process modeling of a coupled hybrid FO-RO system.
- Produce cost modeling of a coupled FO-RO system.
- Prepare a final report summarizing the work completed under the contract.

## 1.1. Guide to this Document

The use of concentrate to recover wastewater using osmotic mechanisms is currently a developing technology, with no known full-scale facilities currently in operation. The performance, planning aspects, order-of-magnitude costs, and other factors involved in applying this technology are currently poorly understood. This document examines some of the factors involved in the use of osmotic mechanisms to recover high quality water from wastewater effluents, as well as provides initial guidance on performance and costs that may be associated with future implementation of the technology.

This report is organized into the following chapters:

- Executive Summary. This section provides a brief overview of the report.
- Introduction (Chapter 1). This section provides background information on this study, including the scope of work.
- Osmotic Mechanisms in Water and Wastewater Treatment (Chapter 2). A brief literature review on current applications for FO and key equations used to model FO and RO applications are included in this section.
- Survey of Water Categories and Quality (Chapter 3). A review of existing desalination plant concentrate and wastewater characterization for plants in Texas is provided. An initial screening of potential viable combinations of feed water and draw solution is presented in this section using FO flux data from existing literature sources, to limit the number of options tested during bench-scale and field piloting.
- Performance of Forward Osmosis Membranes (Chapter 4). This section describes the experimental protocol, results, and analysis for bench-scale and pilot testing of the hybrid FO-RO membrane system. This work was completed at the Colorado School of Mines during 2009 and 2010.
- System and Process Modeling for Forward Osmosis (Chapter 5). This section provides modeling examples of a hybrid FO-RO membrane system as well as very preliminary optimization analysis.
- Development of a FO-RO System Cost Model (Chapter 6). This section describes the development of cost estimating tools for a hybrid FO-RO membrane system and provides a simple model for use by others.
- Conclusions (Chapter 7). This section summarizes the key conclusions developed as part of this study
- Recommendations (Chapter 8). This section provides key recommendations for additional future work on FO.
- Acknowledgements (Chapter 9). This section acknowledges key contributions made to the completion of the study.
- References (Chapter 10). This section provides a listing of all sources cited within the report.

For quick reference, the Scope of Work for the project is cross-referenced to the relevant report sections in Table 1.1.



**Table 1.1 Summary of Work Elements in the Scope and Corresponding Report Sections**

<b>Scope Item</b>	<b>Description</b>	<b>Report section</b>	<b>Section Title</b>
Task 1	Survey of Water Categories and Quality.	3 – 3.3	Survey of Water Categories and Quality
Task 2	Screening and Selection of Hybrid Forward Osmosis System Configurations	3.4	Characterization of Osmotic Pressure of Various Waters in Texas
Task 3	Testing of the Novel Forward Osmosis Spiral-wound Membrane Element.	4	Performance of Forward Osmosis Membranes
Task 4	System and Process Modeling	5	System and Process Modeling for Forward Osmosis
Task 5	Cost Modeling and Feasibility Analysis	6	Development of a hybrid Forward Osmosis/Reverse Osmosis System Cost Model

## 2. Osmotic Mechanisms in Water and Wastewater Treatment

Osmosis is defined as the net movement of water across a selectively permeable membrane driven by a difference in osmotic pressure across the membrane. A selectively permeable membrane allows passage of water, but rejects solute molecules or ions. The use of osmosis as a separation mechanism through an engineered membrane has been designated FO or direct osmosis (DO). RO, a process which produces high quality water from a saline source, is the pressure-driven passage of solvent (water) through a semi-permeable membrane.

In the TWDB assessment of FO, the specific pairing of desalination plant concentrate and treated wastewater treatment streams is of interest. While a review of literature and key concepts are not specifically part of the scope of this study, this additional material provides background information on FO and RO that is used later in the report. The following sub-tasks are contained in this chapter:

- Introduction to key concepts in FO
- Historical context of FO
- Introduction to key equations describing mass transfer in FO and RO applications

### 2.1. Fundamentals of Forward Osmosis

The osmotic pressure is a physical thermodynamic property that describes the flow of pure water across a semi-permeable membrane into a solution possessing higher solute concentration until equilibrium is reached. The resulting difference in height is the osmotic pressure of the solution. In reverse osmosis, pressure is applied to the concentrated solution, and a low salinity permeate is produced. Osmotic pressure ( $\pi$ ) is calculated using the Van't Hoff equation:

$$\pi = MRT \quad (2-1)$$

where:

$M$  is the molar concentration of dissolved species.

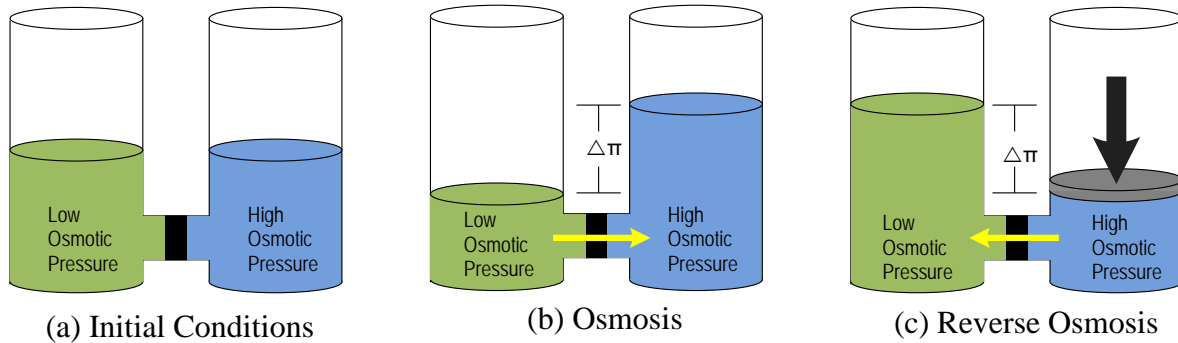
$R$  is the ideal gas constant.

$T$  is the temperature on the Kelvin scale.

In FO, a semi-permeable membrane separates a high osmotic pressure 'draw' solution from a feed water stream with relatively lower salinity and osmotic pressure. Water is drawn across the membrane by natural osmosis, rejecting salts at the membrane surface and subsequently concentrating the feed water stream and diluting the draw solution. Early scientific literature on FO was first presented by Loeb and Sourirajan (1963).

Figure 2.1 graphically illustrates the concepts of FO in comparison to RO. When equal volumes of a dilute feed solution and a concentrated draw solution are separated by a semi-permeable membrane, water flows into the concentrated draw solution, which has a higher osmotic pressure. Water flow continues until chemical equilibrium is reached. The increase in water column height in the high osmotic pressure chamber at equilibrium equates to the difference in osmotic pressure between the dilute and concentrated solutions. When pressure is applied to the

concentrated solution that is greater than the osmotic pressure difference, water flow reverses through the membrane, and is referred to as RO.



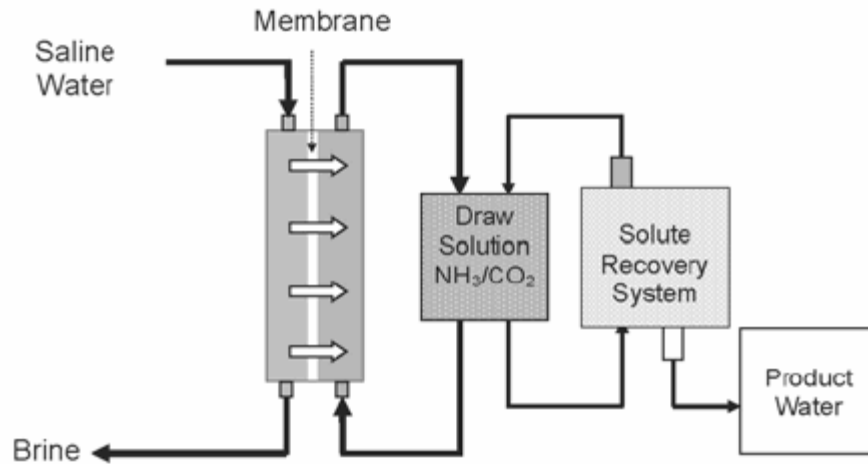
**Figure 2.1 Simplified Schematic of Forward Osmosis (Cath and others, 2006)**

Early applications of FO include the concentration of fruit juices, where the concentrated draw solution becomes the product. In another early commercial application, Hydration Technology Innovations used FO to prepare a high-energy potable drink from contaminated water supplies (Water Desalination Report, 2011). Several researchers have investigated the use of FO as a potential energy source using the pressure restrained osmosis (PRO) process. In PRO, a semi-permeable membrane is used to separate fresh water and saline water. Through FO, water passes from the fresh water side of the membrane to the saline side, resulting in an increase in pressure. This pressure is released through a Francis turbine, generating electricity. PRO represents a subset of the applications of FO. A PRO demonstration plant was commissioned in Norway in 2009 by Statkraft (Øystein Skråmestø Sandvik and others, 2009).

Other applications demonstrated to date include treatment of highly degraded waste streams. In 1998, CH2M HILL studied the use of FO to treat and concentrate landfill leachate at Coffin Butte Landfill near Corvallis, Oregon (York and others, 1999). A sodium chloride solution was used as the draw solution; the diluted NaCl solution was then treated with RO to recover water. Concentrated leachate (5 – 10% stream) was then treated for disposal. More recently, NASA has investigated use of FO in closed-system water recycling and the dewatering of WWTP centrate has been demonstrated (Cath and others, 2006).

When used in desalination, where the objective is to recover a potable water stream, two different classifications of draw solutions are being investigated. In the first, a draw solution such as NaCl is used and then processed through an additional membrane step to recover product water. The second classification of draw solutions readily undergoes thermal decomposition, such as ammonium bicarbonate. Developed at Yale University, the Forward Osmosis–Low Temperature (FO-LT) distillation process uses low quality heat to separate water from the draw solution. The resultant decomposition gases (ammonia and CO<sub>2</sub>) are reconstituted for reuse (Figure 2.2). Recently, two demonstration plants have been constructed. The first is a pilot plant constructed by Yale University that investigates the use of an ammonium bicarbonate draw solution (McGinnis and Elimelech, 2007). The Yale process is currently being commercialized by the company Oasys ([www.oasyswater.com](http://www.oasyswater.com)). The second plant, announced on March 3, 2008,

will be a 25,000 gallons per day (gpd) facility developed by Modern Water, PLc located in Gibraltar (Water Desalination Report, 2008).



**Figure 2.2 Forward Osmosis Desalination Process Using Ammonium Bicarbonate Draw Solution (Cath and others, 2006)**

Several advantages exist in the use of FO processes:

- The process typically operates at pressure of less than 5 atmospheres, since osmotic pressure and not hydraulic pressure provides the predominant driving force, resulting in relatively low energy consumption.
- Membrane compaction is typically not an issue.
- The hydrophilic<sup>1</sup> membrane has a high organic fouling resistance.
- Minimum pretreatment requirements (100 micron strainers).

There are, however, a number of technical challenges to overcome, including:

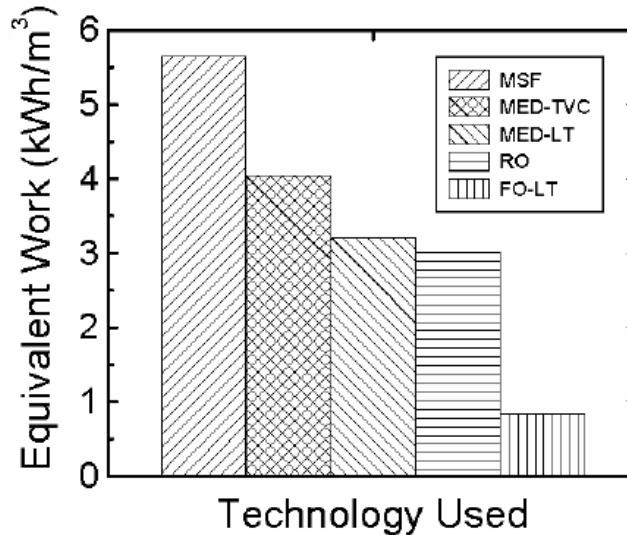
- Development of new and additional membrane sources. Currently, only one commercially available membrane is on the market that uses a cellulose triacetate membrane prone to hydrolysis. Other researchers and manufacturers are working on membranes specific to FO and PRO power generation.
- Addressing mass transfer limitations due to concentration polarization within the membrane support layer.
- Developing new modules suitable for full-scale implementation. To date, most applications have used flat-sheet, plate, and frame elements.

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<sup>1</sup> Hydrophilic membranes, or ‘water loving’ membranes have a high affinity for water and possess an air/water/membrane contact angle of 60 degrees or less (there is much debate in the water treatment industry regarding specific definitions of what contact angle corresponds to hydrophilic versus hydrophobic). Many membranes currently used in water treatment are more hydrophobic, and preferentially adsorb organic compounds from the raw water source. To minimize organic fouling, hydrophilic membranes are typically preferential in FO or RO applications.

- The extraction of water from the draw solution and the subsequent reconstitution of the draw solution, where applicable. Some FO processes are being developed that do not require reconstitution (for example, fertilizers used as a draw solution).

Relative energy consumption of desalination processes using FO-LT, multi-stage flash (MSF), multiple effect distillation (MED), and RO have been estimated (McGinnis and Elimelech, 2007). Figure 2.3 compares the energy consumption between the processes. It was estimated by the authors that FO with a thermally decomposing draw solution (such as in the FO-LT process) would use less than one-third the work energy of RO for desalination.



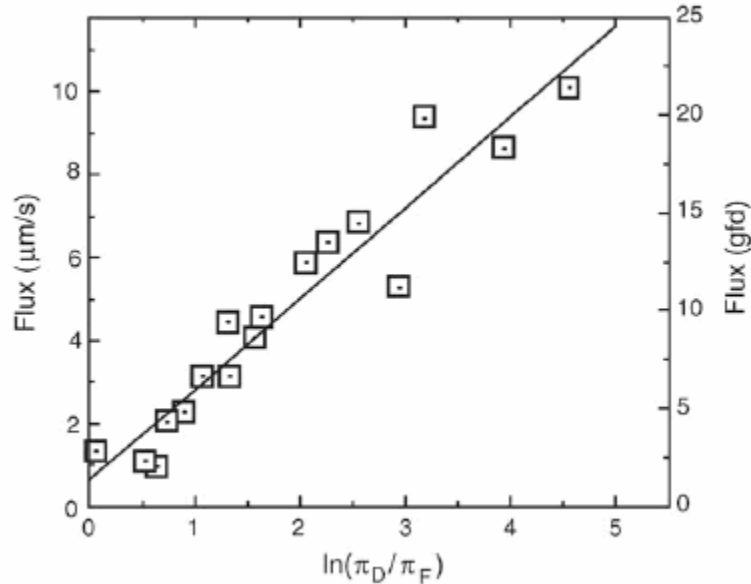
**Figure 2.3 Estimated Energy Consumption for Desalination Processes (McGinnis and Elimelech, 2007). The FO-LT process incorporates the use of low-quality heat for thermal decomposition of the draw solution and recovery using distillation columns.**

Given that the theoretical thermodynamic minimum energy for desalination<sup>2</sup>, at 50 percent recovery, has been reported to be 1 kWh/m<sup>3</sup> (Spiegler and El-Sayed, 2001), the authors' claims are interesting. It is assumed that the thermal energy required for decomposition of the ammonium bicarbonate draw solution is not included.

For existing FO membranes, flux is highly dependent upon the osmotic driving force. Figure 2.4 illustrates the impact of the logarithmic ratio of draw solution osmotic pressure ( $\pi_D$ ) and impaired feed stream osmotic pressure ( $\pi_f$ ) on flux. Ultimately, the greatest potential application of FO may be in integrated wastewater recovery and desalination processes. In such a process,

<sup>2</sup> Numerous authors have calculated the thermodynamic minimum energy required for desalination, including Spiegler and El-Sayed (2001). Two different approaches are typically used. In the more rigorous approach, the work required is calculated using thermodynamic theory. In the less rigorous approach, researchers have indicated that the thermodynamic minimum energy for desalination is equal to the energy required to produce 1 drop of water from a RO membrane operating at very low recovery (<1%), that is, the osmotic pressure of seawater. In practice, RO systems operate at recoveries between 30% and 50%, so that the thermodynamic minimum energy at that point is equal to the log mean feed-concentrate osmotic pressure. In practice, the energy required for desalination using RO is in the range of 3 to 4 kWh/m<sup>3</sup>.

seawater is used as the draw solution to recover water from a treated wastewater stream. Given typical seawater and wastewater total dissolved solids (TDS) concentrations of 35,000 milligrams per liter (mg/L) and 1,000 mg/L, respectively, fluxes of greater than 15 gallons per square foot per day (gfd) may be possible in the lead element. This is comparable to existing sea water reverse osmosis (SWRO) fluxes on the order of 8 to 10 gfd, depending on membrane characteristics.



**Figure 2.4 Impact of Osmotic Pressure on FO Membrane Flux (McCutcheon and others, 2006)**

Potential advantages of FO-RO using seawater as the draw solution include the reduction in the volume of seawater extracted and dilution of the SWRO feed-stream resulting in lower pressure requirements. While the energy consumption associated with FO is quite low, energy consumption would be greatly increased if reconstitution of the draw solution is required.

Current commercialization of industrial FO processes has been limited. While Statkraft has constructed a demonstration power plant based upon FO principles, and Modern Water PLC (a United Kingdom-based firm) has constructed several small demonstration plants for seawater desalination, no sizable desalination plant exists that depends upon FO as the main desalination process. Commercialization efforts have largely focused on processes that do not require a dedicated draw solution re-constitution process. The ‘Green Machine’ commercialized by HTI in conjunction with Bear Creek Services is used for volume reduction of wastes produced in the hydraulic fracturing process (fracking) used for increased natural gas production. The concentrated salt solution injected in the process is used as the draw solution to extract water from the waste stream recovered from the wells (HTI, 2010). Modern Water PLC, in a demonstration project located in Oman, has retrofitted a cooling tower to operate with very high concentrations of a proprietary salt. As water evaporates in the cooling tower, the concentration increases, so that the resulting stream in the cooling loop can be used as draw solution to extract water from seawater. While seawater cooling towers are typically limited to 1.3 to 1.5 cycles of concentration due to scaling potential, the process described by Modern Water PLC can operate

at much higher cycles, requiring less blowdown from the cooling tower. This consequentially reduces the consumption of biocides and other chemicals used in the cooling tower (Modern Water, 2010).

### 2.1.1. General Theory of Forward Osmosis

Osmosis is the transport of water across a selectively permeable membrane from a region of higher chemical potential to a region of lower chemical potential. This chemical potential gradient is a result of the difference in solute concentrations across the membrane, which allows passage of water, but rejects most solute molecules or ions.

Osmotic pressure ( $\pi$ ) is the pressure which, if applied to the more concentrated solution, would prevent net transport of water across the membrane. Increasing the applied pressure on the concentrated solution in excess of the osmotic pressure would reverse net flow across the membrane, as in RO systems. FO, on the other hand, uses the osmotic pressure differential ( $\Delta\pi$ ) across the membrane, rather than the hydraulic pressure differential (as in RO), as the driving force for transport of water through the membrane. The FO process results in concentration of a feed stream and dilution of a highly concentrated stream (referred to as the draw solution). The general equation describing water transport in FO is:

$$J_w = K_w(\sigma\Delta\pi - \Delta P) \quad (2-2)$$

where:

- $J_w$  = water flux
- $\Delta\pi$  = differential osmotic pressure across the membrane
- $K_w$  = water permeability coefficient of the membrane
- $\sigma$  = reflection coefficient (a measure of the relative permeability of a particular membrane to a particular solute)
- $\Delta P$  = differential applied pressure across the membrane

### 2.1.2. Impacts of Concentration Polarization

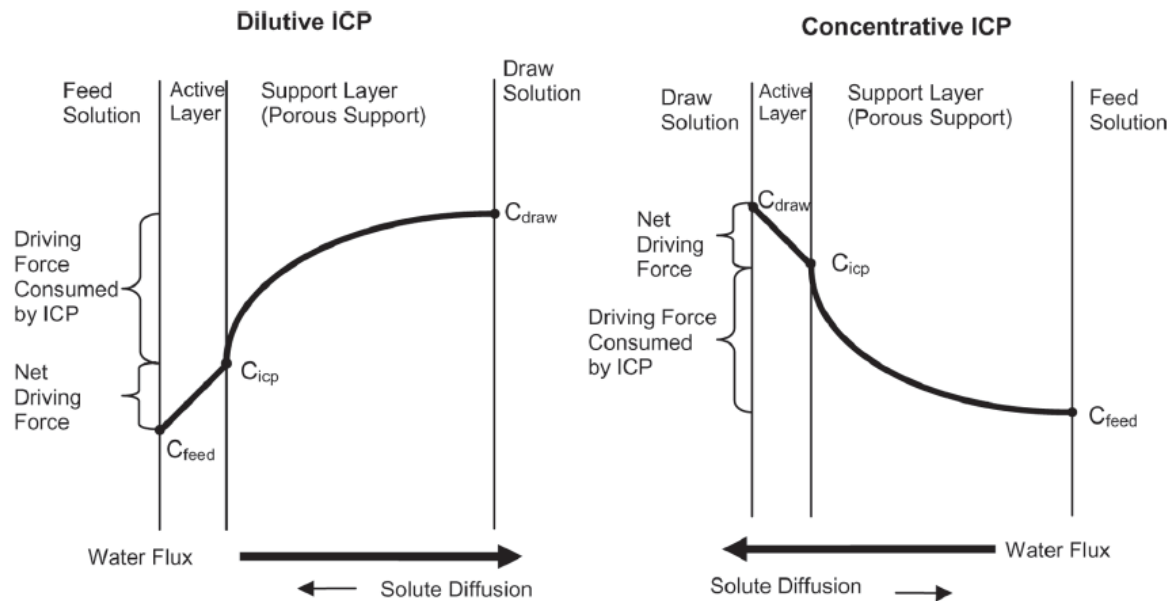
In pressure-driven membrane processes, convective permeate flow causes a buildup of solute at the surface of the membrane active layer due to relatively lower diffusive transport of solute back into solution. Referred to as concentration polarization (CP), this phenomenon reduces permeate water flux by increasing osmotic pressure of the solution at the membrane surface, which must be overcome with hydraulic pressure. CP is not limited to pressure-driven membrane processes, but also occurs during osmotic-driven membrane processes, on both the feed and permeate sides of the membrane (Cath and others, 2006). Researchers have shown that at higher osmotic potentials, the water flux predicted by Equation 2-2 is frequently lower than the anticipated flux. This lower-than-expected water flux is often attributed to concentration polarization.

When the feed solution flows next to the active layer of the membrane (such as in RO), solutes build up near the active layer, resulting in concentrative external concentration polarization (ECP). Simultaneously, the draw solution in contact with the permeate side of the membrane is being diluted at the membrane interface by the permeating water. This is called dilutive ECP. Both concentrative and dilutive ECP phenomena reduce the effective osmotic driving force. The adverse effect of ECP on osmotic-driven membrane processes can be minimized by increasing

flow velocity and turbulence at the membrane surface or by manipulating the water flux. However, because water flux in FO is already low, the ability to diminish ECP by reducing flux is limited. The impact of concentrative ECP is generally much smaller than dilutive ECP for FO, and is often neglected (Gray and others, 2006).

In FO, depending upon the membrane side that faces the feed solution (active or support layer), dilutive or concentrative internal concentration polarization (ICP) can also occur. When the FO membrane used is a composite or asymmetric membrane, with a dense separating layer and a porous support layer, ICP can take place within the support layer. This type of CP cannot be minimized by increased cross-flow. In FO applications for desalination and water treatment, the active layer of the membrane faces the feed solution and the porous support layer faces the draw solution. As water permeates the active layer, the draw solution within the porous substructure becomes diluted, resulting in dilutive ICP. Conversely, concentrative ICP occurs when the active layer faces the draw solution, and convective flux through the membrane causes a buildup of solute within the support layer. Both types of ICP cause a decrease in the osmotic pressure gradient, which decreases permeate water flux.

Figure 2.5 schematically represents the differences between dilutive and concentrative ICP in a FO application. Most FO studies are being conducted with feed solution contacting the active dense layer of the membrane to minimize blockage of the support layers by contaminants in the feed solution. As a result, dilutive ICP is the most relevant concentration polarization encountered in FO applications.



**Figure 2.5 Comparison of Dilutive Internal Concentration Polarization (left) and Concentrative Internal Concentration Polarization (right) in Forward Osmosis processes (Gray and others, 2006)**

In pressure driven membrane processes, such as RO and nanofiltration (NF), some salts (or feed solutes) diffuse through the membrane into the permeate (product water). The salt diffusion in



these processes is driven predominantly by the concentration gradient across the membrane (refer to Equation 2-2), with a significantly smaller contribution due to pressure gradient.<sup>3</sup> The same phenomenon occurs in FO. Salt diffuses from a high concentration to a low concentration, governed by Fick's Law. In FO, there is a very high concentration gradient from the draw solution to the feed. This gradient forces diffusion of solutes from the draw solution into the feed stream. This diffusion from the draw solution is termed "reverse solute flux" because the solutes are diffusing against the flow (flux) of water into the draw solution.

The reverse solute flux can be estimated by Equation 2-3:

$$J_s = K_s(\Delta C) \quad (2-3)$$

where:

- $J_s$  = solute flux
- $K_s$  = solute permeability coefficient of the membrane
- $\Delta C$  = differential concentration of solute across the membrane

Internal concentration polarization was originally described by Loeb and others (1997). Equation 2-4 describes the impact of dilutive ICP on FO flux.

$$K_w = \left( \frac{1}{J_w} \right) \ln \left( \frac{B + A\pi_D}{B + J_w + A\pi_F} \right) \quad (2-4)$$

where:

- $J_w$  = water flux
- $K_w$  = water permeability coefficient of the membrane
- $A$  = mass transfer coefficient for water
- $B$  = mass transfer coefficient for solute
- $\pi_D$  = osmotic pressure of the draw solution
- $\pi_f$  = osmotic pressure of the feed solution

By examination of Equation 2-4, the flux is nonlinear with respect to the osmotic pressure differential between draw solution and feed solution as a result of ICP. As the osmotic pressure differential increases, the magnitude of the deviation in flux from a linear model (Equation 2-2) increases.

## 2.2. Fundamentals of Reverse Osmosis

RO is a membrane process that utilizes hydraulic pressure to offset osmotic pressure and induce mass transport of water across a semi-permeable membrane. Membrane desalination processes using RO or NF are diffusion-controlled membrane processes. Mass transfer of solvent, typically

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<sup>3</sup> From a theoretical perspective, FO is typically defined as having a feed pressure of zero. In reality, minor losses in feed spacers, fittings, etc., requires some pressure to permit the desired cross-flows through the system. If the hydraulic pressures are not equal, a hydraulic pressure gradient may be formed. For most FO applications, the hydraulic pressure gradient is significantly smaller than the osmotic pressure gradient, and may be ignored.

water, through the membrane occurs as a result of a pressure differential, while solute passes through the membrane based upon concentration differential (diffusion). The performance of RO and NF processes is typically described using the homogenous solution diffusion (HSD) model, which describes the effect of feed water quality concentration, membrane characteristics, hydraulic recovery, and operating pressure on the permeate concentration. Other theories describing RO and NF processes exist, such as the solution-diffusion model. A typical RO system is illustrated in Figure 2.6.



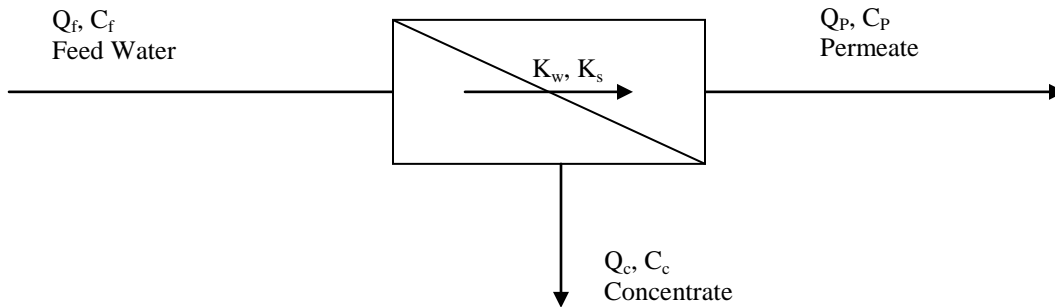
**Figure 2.6 Kay Bailey Hutchison Desalination Plant, El Paso (Photo credit – El Paso Water Utilities)**

Figure 2.7 shows a cut-away view of an individual spiral wound membrane element.



**Figure 2.7 Membrane Element (photo credit – R. Huehmer)**

A basic schematic of a reverse osmosis element showing mass balance parameters is illustrated in Figure 2.8.



**Figure 2.8 Schematic diagram of a reverse osmosis membrane element**

The following mass balance equations are commonly used to describe RO and NF membrane process performance. Equations 2-5 through 2-15 are adapted from AWWA Manual of Water Supply Practices, Second Edition, M46 (2007).

Equation 2-5 indicates mass balance for water flow:

$$Q_f \rho_f = Q_p \rho_p + Q_c \rho_c \quad (2-5)$$

where:

- $Q_f$  = feedwater flow rate, e.g., gpd [ $m^3/d$ ]
- $\rho_f$  = density of feedwater
- $Q_p$  = permeate flow rate, e.g., gpd [ $m^3/d$ ]
- $\rho_p$  = density of permeate
- $Q_c$  = concentrate flow rate, e.g., gpd [ $m^3/d$ ]
- $\rho_c$  = density of concentrate

Equation 2-6 describes mass balance for solute flux:

$$Q_f C_f = Q_p C_p + Q_c C_c \quad (2-6)$$

where:

- $C_f$  = feedwater solute concentration, units of mass per volume, e.g., mg/L
- $C_p$  = permeate solute concentration, units of mass per volume, e.g., mg/L
- $C_c$  = concentrate solute concentration, units of mass per volume, e.g., mg/L

Equation 2-7 indicates product recovery rate:

$$R = \frac{Q_p}{Q_f} \quad (2-7)$$

where:

- $R$  = decimal fraction of product water recovered from feedwater
- $Q_p$  = permeate flow rate, e.g., gpd [ $m^3/d$ ]
- $Q_f$  = feedwater flow rate, e.g., gpd [ $m^3/d$ ]

Equation 2-8 indicates water flux:

$$J_w = K_w (\Delta P - \Delta \pi) = \frac{Q_p}{A_{mem}} \quad (2-8)$$

where:

- $J_w$  = water flux, units of flow rate per unit area, e.g., gpd/ft<sup>2</sup> or gfd [L/hr-m<sup>2</sup> or LMH]
- $K_w$  = water mass transfer coefficient (see Equation 2-9), units of flux per pressure, e.g., gpd/ft<sup>2</sup>/psi or gfd/psi [L/m<sup>2</sup>-hr-kPa or L/m<sup>2</sup>-hr-bar]
- $\Delta P$  = transmembrane pressure differential, units of pressure, e.g., psi [kPa or bar]
- $\Delta \pi$  = transmembrane osmotic pressure difference, units of pressure, e.g., psi [kPa or bar]
- $Q_p$  = permeate flow rate, e.g., gpd [ $m^3/d$ ]
- $A_{mem}$  = effective membrane area, e.g., ft<sup>2</sup> [m<sup>2</sup>]

Equation 2-9 defines dissolved solute flux:

$$\begin{aligned}
 J_s &= K_s (C_m - C_p) \\
 &= K_s \left[ \left( \frac{C_f + C_c}{2} \right) - C_p \right] = \frac{Q_p C_p}{A_{mem}}
 \end{aligned}
 \tag{2-9}$$

where:

- $J_s$  = solute flux, units of mass per time per area, e.g., lb/ft<sup>2</sup>/d [g/m<sup>2</sup>/s]
- $K_s$  = solute mass transfer coefficient (see Equation 2-7), units of length per time, e.g., ft/d [m/s]
- $C_m$  = concentration at the membrane surface
- $A_{mem}$  = effective membrane area, e.g., ft<sup>2</sup> [m<sup>2</sup>]

Equation 2-10 indicates water mass transfer coefficient:

$$K_w = \frac{Q_p}{A_{mem}(\Delta P - \Delta \pi)} = \frac{Q_p}{A_{mem}(NAP)}
 \tag{2-10}$$

where:

- $K_w$  = water mass transfer coefficient (see Equation 2-9), units of flux per pressure, e.g., gpd/ft<sup>2</sup>/psi or gfd/psi [L/m<sup>2</sup>-hr-kPa or L/m<sup>2</sup>-hr-bar]
- $Q_p$  = permeate flow rate, e.g., gpd [m<sup>3</sup>/d]
- $\Delta P$  = transmembrane pressure differential, units of pressure, e.g., psi [kPa or bar]
- $\Delta \pi$  = transmembrane osmotic pressure difference, units of pressure, e.g., psi [kPa or bar]
- $NAP$  = Net Applied Pressure ( $\Delta P - \Delta \pi$ ), e.g., psi [kPa or bar]
- $A_{mem}$  = effective membrane area, e.g., ft<sup>2</sup> [m<sup>2</sup>]

Equation 2-11 describes solute mass transfer coefficient:

$$K_s = \frac{Q_p C_p}{A_{mem} \Delta C}
 \tag{2-11}$$

where:

- $K_s$  = solute mass transfer coefficient, units of length per time, e.g., ft/d [m/s]
- $Q_p$  = permeate flow rate, e.g., gpd [m<sup>3</sup>/d]
- $C_p$  = permeate solute concentration, units of mass per volume, e.g., mg/L
- $A_{mem}$  = effective membrane area, e.g., ft<sup>2</sup> [m<sup>2</sup>]
- $\Delta C$  = concentration differential (see Equation 2-12)

Equation 2-12 indicates concentration differential:

$$\Delta C = (C_m - C_p) = \left( \frac{C_f + C_c}{2} \right) - C_p \quad (2-12)$$

where:

- $\Delta C$  = concentration differential
- $C_m$  = concentration at the membrane surface
- $C_p$  = permeate solute concentration, units of mass per volume, e.g., mg/L
- $C_f$  = feedwater solute concentration, units of mass per volume, e.g., mg/L
- $C_c$  = concentrate solute concentration, units of mass per volume, e.g., mg/L

Equation 2-13 describes the effects of each of the five independent variables ( $K_s$ ,  $C_r$ ,  $K_w$ ,  $\Delta P$ , and  $R$ ) on the permeate concentration:

$$C_p = \frac{K_s C_f}{K_w \Delta P + \Delta \pi \left( \frac{2 - 2R}{2R} \right) + K_s} \quad (2-13)$$

where:

- $C_p$  = permeate solute concentration, units of mass per volume, e.g., mg/L
- $K_s$  = solute mass transfer coefficient, units of length per time, e.g., ft/d [m/s]
- $C_f$  = feedwater solute concentration, units of mass per volume, e.g., mg/L
- $K_w$  = water mass transfer coefficient, units of flux per pressure, e.g., gpd/ft<sup>2</sup>/psi or gfd/psi [L/m<sup>2</sup>-hr-kPa or L/m<sup>2</sup>-hr-bar]
- $\Delta P$  = transmembrane pressure differential, units of pressure, e.g., psi [kPa or bar]
- $\Delta \pi$  = transmembrane osmotic pressure difference, units of pressure, e.g., psi [kPa or bar]
- $R$  = decimal fraction of product water recovered from feedwater

### 2.3. Residuals Production and Management in a Desalination Plant

When examining the use of RO system concentrate as a draw solution, a basic examination of the production of concentrate is warranted. In general, as the TDS in the permeate stream is very low relative to the concentrate stream, it is useful to assume that membrane processes have complete rejection of dissolved salts. Using this assumption, the flow and concentration of the concentrate stream may be estimated using Equations 2-14 and 2-15, respectively.

$$Q_c = Q_f (1 - R) \quad (2-14)$$

where:

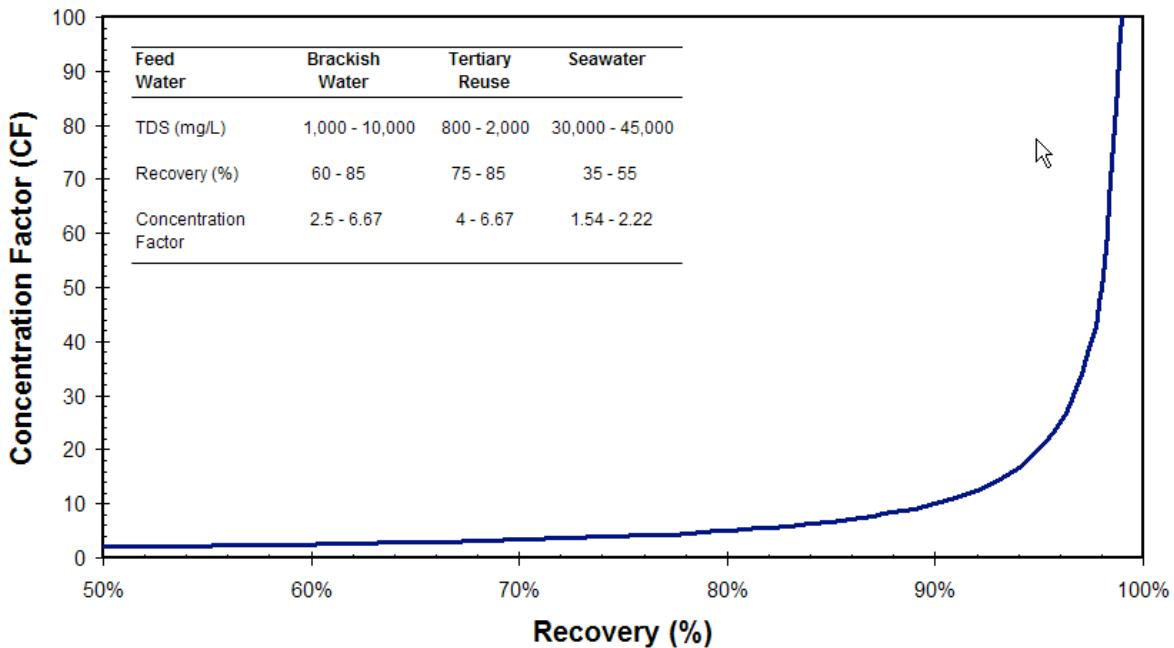
- $Q_c$  = concentrate flow rate, e.g., gpd [m<sup>3</sup>/d]
- $Q_f$  = feedwater flow rate, e.g., gpd [m<sup>3</sup>/d]
- $R$  = decimal fraction of product water recovered from feedwater

$$C_c = CF \times C_f = 1/(1-R) \times C_f \quad (2-15)$$

where:

- $C_c$  = concentrate solute concentration, units of mass per volume, e.g., mg/L
- $CF$  = concentration factor =  $1/(1-R)$
- $C_f$  = feedwater solute concentration, units of mass per volume, e.g., mg/L

Figure 2.9 illustrates the impact of the hydraulic recovery on the concentration factor over a range of hydraulic recovery values. The practical limits of hydraulic recovery are governed predominantly by the water chemistry in brackish water desalination. As the feed water is concentrated, the concentration of sparingly soluble salts can approach the solubility limit, resulting in scale formation. While scale-inhibiting chemicals can be applied to limit scale formation, they are effective over a limited range of super-saturation. In seawater desalination applications, the maximum hydraulic recovery is typically limited by the operating pressures required to generate a positive net driving pressure. Typically, in a seawater desalination environment, recovery ranges from 35 to 55 percent.



**Figure 2.9 Nominal Concentration Factor of Reverse Osmosis Concentrate**

Typically, concentrate from desalination processes is managed using one or more of the following approaches.

*Discharge to a receiving waterbody.* In discharge to a receiving body, concentrate is typically discharged through an underwater diffuser system either directly or after treatment (for example, discharge in a sanitary sewer system and discharged as part of the wastewater treatment plant effluent). The diffuser system is designed such that rapid dilution of the concentrate into the receiving water occurs in the “near field,” minimizing impacts on benthic organisms. While this technology is applicable to most receiving bodies, it is most effective where there is a natural current to induce additional mixing in the “far field” and to transport the diluted concentrate away from the diffuser. As a result, diffusers are highly effective in rivers and oceans, where currents are present at reasonable velocities. With high levels of TDS in the concentrate, disposal into a river may not be acceptable due to potential toxicity impacts on freshwater flora and fauna. For seawater desalination plants, ocean disposal is typically practiced, with diffuser lines used in most of the western world.

*Deep well injection.* In deep well injection, concentrate is pumped into deep wells below confining aquifers. While deep well injections have been used with great success in brackish water systems, increasingly regulations are limiting the installation of new wells. Few large seawater systems use deep well injection, as the cost of ocean disposal is usually more economically feasible. There are a number of examples of shallow well injection for seawater disposal, including the desalination plants in Sand City, CA and on San Nicholas Island, CA.

*Evaporation ponds.* The use of evaporation ponds for desalination concentrate disposal can be very effective in arid climates and have been successfully applied. They are typically very land intensive, and concerns exist regarding the potential for contamination of groundwater as a result of leaks in pond liners. While some regions of Texas possess pan evaporation rates that are positive, their use in coastal regions of Texas is typically not feasible, as a result of large rainfalls associated with hurricane landfall. Evaporation ponds are typically not used for seawater desalination unless salt recovery is desired. It is typically more economical to dispose of the concentrate to the ocean.

*Mechanical Zero-Liquid Discharge (ZLD).* For inland desalination plants, the use of evaporation ponds, receiving waterbody disposal, or deep water injection may not be feasible. In those cases, ZLD processes may be employed. ZLD is extremely expensive, and, as a result, almost no municipal plants use ZLD for concentrate disposal worldwide. In most instances, alternative water supplies are currently implemented. As mentioned above, ZLD processes are typically not used for seawater desalination unless salt recovery is desired.

For additional information on concentrate management techniques, refer to Mickley (2001).

In this study, the potential for using previously disposed concentrate from desalination plants as a draw solution to recover product water from a wastewater stream is evaluated. The coupling of FO and RO possesses the potential to significantly minimize requirements for additional concentrate management depending on the configuration utilized. Alternatively, and not covered in the scope of this study, is the use of concentrate management techniques to increase the concentration of concentrate streams from inland desalination plants, improving their potential for FO applications.



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### **3. Survey of Water Categories and Quality**

For operation of a FO process, two aqueous streams are needed: first, an impaired water feed stream (i.e., source water), and secondly, a draw solution (i.e., brine stream), which has an elevated osmotic pressure to provide the driving force for water flow from, and concentration of, the feed stream.

The quality and presence of specific constituents in various source waters will dictate the long-term performance of the FO process and the auxiliary RO system in a FO-RO process. Source water quality data will also assist engineering practitioners in the selection of pretreatment and post-treatment technologies suitable for achieving water quality goals and desirable water recovery.

In the TWDB assessment of FO, the specific pairing of desalination plant concentrate and treated wastewater treatment streams is of interest. This chapter reports on Task 1 and Task 2 of the project scope. The purpose of Task 1 is to collect water quality and quantity data from wastewater treatment facilities and desalination plants based in Texas. The following sub-tasks have been completed by the research team to achieve this objective:

- Locating existing desalination plants within Texas
- Identifying wastewater treatment plants in close proximity to the desalination facilities
- Short-listing of facilities and utilities to contact for water quality data
- Summarizing of water quality data

Task 2 consists of estimating the osmotic pressure of the various waters, and selecting a combination of the different water qualities that represent favorable differences in osmotic pressure, the key driving force in FO applications.

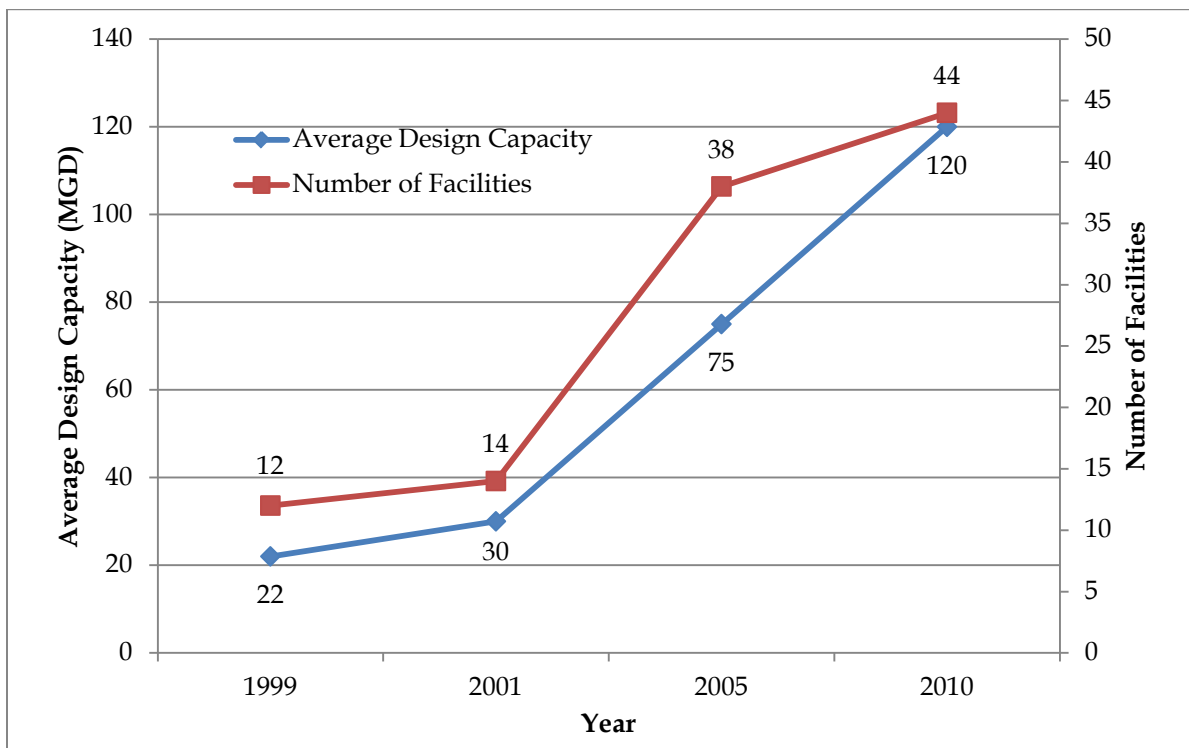
The team characterized and classified various impaired water sources, including secondary and tertiary treated reclaimed water, and various concentrate sources, from seawater and brackish water desalination facilities from participating utilities. The researchers were then able to develop correlations between water quality and the expected performance of the FO process under various scenarios, and perform a screening of potential applications.

#### **3.1. Desalination in Texas**

Several reports have been written in the last decade on desalination facilities in Texas. In a report funded by the U.S. Bureau of Reclamation, Mickley (2001) surveyed 12 municipal desalination facilities in operation in Texas in 1999, with a total capacity of approximately 22 million gallons per day (mgd). These plants employed a combination of RO, electrodialysis (ED), and electrodialysis reversal (EDR) technologies to treat predominantly brackish surface water and ground water. One facility, the Harlington Water Works System, utilized RO to treat secondary wastewater for non-potable purposes.

In 2002, the Wagnick report indicated that at least 14 municipal and 104 industrial desalination facilities existed in Texas in 2001, totaling about 30 mgd and 71 mgd, respectively (Wagnick, 2002).

In 2005, the TWDB funded a study to identify existing municipal desalination plants within Texas (Nicot and others, 2005). Of 105 public water system facilities contacted by the researchers, 38 facilities were identified as having desalination capacity greater than 0.025 mgd. An additional 47 facilities were identified that had desalination capacity below 0.025 mgd. A database operated by Global Water Intelligence (GWI) called DesalData.com (formerly the Wagnick reports) aims to compile the global installation base of desalination facilities. In 2010, Desaldata.com identified 32 municipal desalination plants in Texas; however, several of the entries are considered duplicates. In 2010, the American Membrane Technology Association (AMTA) conducted a review of membrane-based treatment facilities, including RO/NF and ED/EDR facilities. The unpublished database builds upon the previous work of Wagnick, GWI and Nicot and others to establish the most up-to-date inventory of municipal desalination plants in the United States (Movahed, 2010). Additional plants included in this work include the Kay Hutchison Desalination Plant in El Paso, Texas. The database created by Nicot and others was updated by TWDB (2011), which reports 44 plants with a capacity of greater than 0.025 mgd. Figure 3.1 presents the cumulative design capacity and average capacity of Texas municipal desalination facilities through 2010.



**Figure 3.1 Cumulative Design Capacity of Desalination Facilities in Texas with capacity  $\geq 0.025$  mgd (TWDB, 2011)**

Table 3.1 presents the 44 facilities that were identified by TWDB in 2011. The processes utilized consist of a combination of RO and EDR. The source water is varied, consisting of either brackish surface water or brackish groundwater.

**Table 3.1 Summary of Texas Desalination Facilities with Capacity  $\geq 0.025$  mgd (TWDB, 2011)**

Desalination Facility Name	Status	Start Up Year	Source Water	Process	Design Capacity (mgd)*
Big Bend Motor Inn	Operating	1989	GW	RO	0.057
City of Abilene (Hargesheimer Treatment Plant)	Operating	2003	SurW	RO	7.95
City of Bardwell	Operating	1990	GW	RO	0.252
City of Bayside	Operating	Original plant in 1990; replaced in 2010	GW	RO	0.045
City of Beckville	Operating	2004	GW	RO	0.216
City of Brady	Operating	2005	SurW	RO	3.0
City of Clarksville City	Operating	2006	GW	RO	0.288
City of Evant	Operating	2010	GW	RO	0.1
City of Fort Stockton	Operating	1996	GW	RO	6.5
City of Granbury	Idle	Original EDR Plant built in 1984; in 2007 RO Plant was mounted in trailer	SurW	RO	0.462
City of Hubbard	Operating	2002	GW	RO	0.648
City of Kenedy	Operating	1995	GW	RO	2.858
City of Laredo Santa Isabel	Operating	1996	GW	RO	0.1
City of Los Ybanez	Idle	1991	GW	RO	0.025
City of Robinson	Operating	1994	SurW	RO	2.3
City of Seadrift	Operating	1998	GW	RO	0.61
City of Seymour	Operating	2000	GW	RO	3.0
City of Sherman	Operating	1993	SurW	EDR	11.0
City of Tatum	Operating	1999	GW	RO	0.324
Cypress Water Treatment Plant	Operating	2008	SurW	RO	10
Dell City	Operating	1997	GW	EDR	0.1
DS Waters of America, LP	Operating	1997	GW	RO	0.09
Esperanza Fresh Water Supply	Operating	1990	GW	RO	0.023
Holiday Beach WSC	Operating	2002	GW	RO	0.15
Horizon Regional MUD	Operating	2001	GW	RO	6.0
Kay Bailey Hutchison Desalination Plant	Operating	2007	GW	RO	27.5
Lake Granbury Surface Water Advanced Treatment System	Operating	2003	SurW	RO	12.5
Longhorn Ranch Motel	Operating	1990	GW	RO	0.023
Midland Country Club - fairways & greens	Operating	2004	GW	RO	0.023
North Alamo WSC (Lasara)	Operating	2005	GW	RO	1.2
North Alamo WSC (Owassa)	Operating	2008	GW	RO	1.5
North Alamo WSC (Doolittle)	Operating	2008	GW	RO	3.75
North Cameron Regional WSC	Operating	2006	GW	RO	2.5
Oak Trail Shores	Operating	EDR installed in 1998; RO replaced EDR in 2007	SurW	RO	0.792
Possum Kingdom WSC	Operating	2003	SurW	RO	1.0
River Oaks Ranch	Operating	1987	GW	RO	0.14

<b>Desalination Facility Name</b>	<b>Status</b>	<b>Start Up Year</b>	<b>Source Water</b>	<b>Process</b>	<b>Design Capacity (mgd)*</b>
Southmost Regional Water Authority	Operating	2004	GW	RO	7.5
Sportsmans World MUD	Operating	1984	SurW	RO	0.083
Study Butte Terlingua Water System	Operating	2000	GW	RO	0.14
The Cliffs (Double Diamond Utilities)	Operating	1991	SurW	RO	0.25
Valley MUD #2	Operating	2000	GW	RO	1.0
Veolia Water Treatment Plant	Idle	1992	SurW	RO	0.245
Water Runner, Inc.	Operating	2001	GW	RO	0.028
Windermere Water System	Idle	2003	GW	RO	2.88

*Notes:*

*mgd: million gallons per day*

*GW: Groundwater*

*RO: Reverse osmosis*

*SurW: Surface water*

*EDR: Electrodialysis reversal*

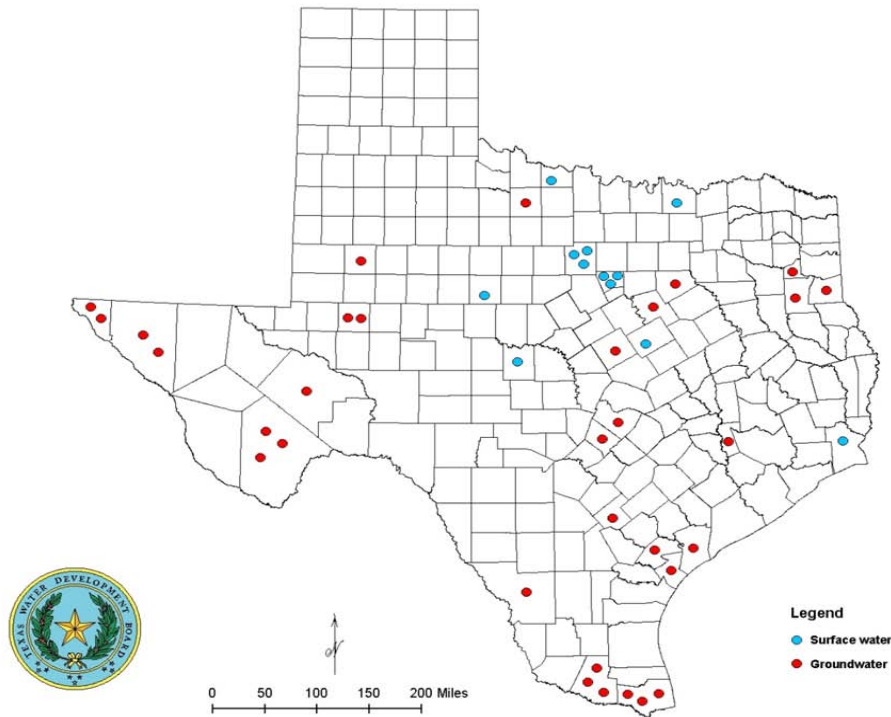
*MUD: Municipal Utility District*

*WSC: Water Supply Corporation*

*\*Design capacity in Table includes blending.*

*Source: Desalination Database of Texas; TWDB, 2011*

To illustrate the geographic distribution of desalination plants in Texas, TWDB plotted desalination plant locations on a map of Texas (Figure 3.2). Maps showing plant location by capacity and by source water type are shown in the figure.



**Figure 3.2** Locations of desalination facilities in Texas (TWDB, 2011)

### 3.2. Characterization of Desalination Plant Concentrate

Based upon the literature review of the existing Texas desalination facilities, and researchers’ knowledge, 20 facilities were identified that had wastewater treatment plants in close proximity to the desalination plants. Letters were sent to these facilities requesting water quality data for the desalination concentrate and wastewater streams. Follow-up to the letters was conducted via email and telephone for each facility.

Early commitments to provide data were made by seven facilities, including El Paso Water Utilities (EPWU), Brownsville Public Utilities Board, City of Laredo, and the City of Abilene. The concentrate quality data obtained is from brackish water desalination facilities (there are no full-scale municipal seawater desalination plants in Texas). In order to obtain characteristic water quality data for Texas seawater and seawater concentrate, information was extracted from documents prepared by Henthorne (2007) and NRS (2009).

The data provided by the utilities are summarized in the following sections.

#### 3.2.1. *Kay Bailey Hutchison Desalination Plant, El Paso, Texas*

The Kay Bailey Hutchison Desalination Plant was constructed as a joint project of EPWU and Ft. Bliss. The plant is designed with a permeate capacity of 15.5 mgd and a maximum finished water blended capacity of 27.5 mgd. Source water for the plant is brackish water from the Hueco Bolson. The \$87 million facility uses a two-stage RO system, and disposes of concentrate via deep well injection. The geologic formation confines the concentrate and prevents its migration to fresh water.

As part of this study, EPWU provided water quality monitoring data for the concentrate stream for 2007 to 2008. The TDS of the concentrate stream averaged 3,600 mg/L, with maximum values of 5,500 mg/L measured. These data were used to calculate osmotic pressure estimates prior to pilot testing. Data from 2008 to 2010 were provided subsequent to this analysis, and are presented in Tables 3.2, B.6, and B.7. Note that data in Tables 3.12 and 3.13 reflect 2007-2008 data. Table 3.2 summarizes the minimum, maximum, and average values for each water quality constituent analyzed.

**Table 3.2 Concentrate Analysis from Kay Bailey Hutchison Desalination Plant**

Parameter <sup>a</sup>	2008			2009			2010			2008-10		
	Min	Avg.	Max	Min	Avg.	Max	Min	Avg.	Max	Min	Avg.	Max
Cl <sup>-</sup>	265	4,699	9,710	2,590	5,439	6,740	4,040	6,099	10,100	265.0	5,520.2	10,100.0
SO <sub>4</sub> <sup>-2</sup>	127	1,039	2,110	1,020	1,410	2,260	1,020	1,410	2,260	127.0	1,406.2	2,260.0
ALK-T <sup>a</sup>	418.8	417	498	400	427	445	130	397	477	130.0	401.1	498.0
EC	10,400	16,267	22,100	8,740	18,712	21,500	10,700	20,565	28,800	8,740.0	17,531.6	28,800.0
Fe-T <sup>b</sup>	0.03	0.13	0.57	0.032	0.121	0.42	0.06	0.16	0.41	0.03	0.21	0.57
Hard-T <sup>c</sup>	528	2,089	3,030	1,200	2328	3,050	540	2,490	4,330	528.0	2,176.1	4,330.0
Mn <sup>+2</sup>	0.1	0.16	0.21	0.08	0.18	0.22	0.14	0.21	0.34	0.08	0.18	0.34
PO <sub>4</sub> <sup>-3</sup>	0.1	0.16	1.26	0.1	0.16	0.49	0.1	0.17	0.96	0.10	0.39	1.26
pH	7.1	8	8.3	7.5	7.8	8	7.6	8	8	7.1	7.8	8.3
Ca <sup>+2</sup>	376	589	793	281	640	937	176	722	1,220	176.0	637.1	1,220.0
K <sup>+</sup>	4.9	76	99.7	38.9	77	114	61	78	89	4.9	70.9	114.0
Mg <sup>+2</sup>	0.9	153	208	85.7	171	250	146	191	260	0.9	162.8	260.0
Na <sup>+</sup>	172	2,674	4,200	1,490	3,097	4,140	2,710	3,412	4,440	172.0	2,926.1	4,440.0
SiO <sub>2</sub>	28.7	148	228	23.4	128	173	36.7	145	187	23.4	122.0	228.0
TDS	6,740	10,412	13,200	5,730	11,520	13,600	9,750	12,763	17,900	5,730.0	11,290.6	17,900.0
Temp <sup>d</sup>	18.9	22	26	22.6	25	26.3	22.1	26	26.6	18.9	23.9	26.6

Notes:

\*All values are in mg/L except EC (µS/cm), pH (pH units), and Temp (°C).

% change is based on average values.

<sup>a</sup> ALK-T refers to Total Alkalinity.

<sup>b</sup> Fe-T refers to Total Iron Concentration.

<sup>c</sup> Hard-T refers to Total Water Hardness.

<sup>d</sup> Temp refers to Temperature in °C.

### 3.2.2. Abilene, Texas

In 2000, CH2M HILL conducted conceptual design for the City of Abilene Southside Water Treatment Plant. The system uses RO to desalinate brackish surface water, containing approximately 1,200 mg/L TDS, producing 3 mgd of RO permeate (CH2M HILL, 2001). During the initial phase of operation, it was anticipated that the plant would operate at recoveries ranging from 68 percent to 80 percent. Table 3.3 presents the design water quality data for the facility.

The source water is from the O.H. Ivie Reservoir located approximately 50 miles from the plant. The plant capacity is 7.5 mgd with all water filtered through a microfiltration (MF) system with pre-coagulation to improve organics removal. A portion of the MF filtrate bypasses the downstream RO system and blends with the RO permeate (e.g., split-treatment scheme).

The plant has concentrate evaporation ponds that reduce the volume of waste RO concentrate to be discharged to the sanitary sewer system.

**Table 3.3 Design Concentrate Water Quality for Abilene, Texas RO Desalination Facility**

Description	Units	Initial Design Basis (at 68% Recovery)	Potential High-Recovery Design Basis (Highest expected case TDS and 80% Recovery)
<b><u>Design Criteria Assumptions:</u></b>			
RO Recovery	%	68%	80%
RO Permeate Flow	mgd	3.00	3.00
RO Feed Flow	mgd	4.41	3.75
Temperature	°C	20.0	10.0
Temperature Range	°C	10 – 30	10 – 30
Concentrate flow	mgd	1.41	0.75
RO Permeate/Finished Flow	%	50%	50%
Finished water flow	mgd	6.00	6.00
Overall plant recovery		80.5%	88.4%
Raw water flow	mgd	7.45	6.79
<b><u>Water Quality Data</u></b>			
Ammonium (NH <sub>4</sub> <sup>+</sup> )	mg/L	1.2	1.9
Barium (Ba <sup>2+</sup> )	mg/L	0.8	1.2
Bicarbonate (HCO <sub>3</sub> <sup>-</sup> )	mg/L	332	531
Bromide (Br <sup>-</sup> )	mg/L	5.3	8.5
Calcium (Ca <sup>2+</sup> )	mg/L	342	547
Carbonate (CO <sub>3</sub> <sup>2-</sup> )	mg/L	0.0	0.2
Chloride (Cl <sup>-</sup> )	mg/L	1,235	1,975
CO <sub>2</sub> (aq)	mg/L	23.1	27.6
Color (RO feed color = 8 cu)	CU	25	40
Fluoride (F <sup>-</sup> )	mg/L	1.2	1.9
Magnesium (Mg <sup>2+</sup> )	mg/L	187	298
Iron (Fe <sup>2+/3+</sup> )	mg/L	1.2	1.9
Manganese (Mn <sup>2+/4+</sup> )	mg/L	0.1	0.2
Nitrate (NO <sub>3</sub> <sup>-</sup> )	mg/L	1.4	2.2
pH (Calc)	SU	7.37	7.50
Phosphate (PO <sub>4</sub> <sup>3-</sup> )	mg/L	0.3	0.5
Potassium (K <sup>+</sup> )	mg/L	27	43
SiO <sub>2</sub> (Reactive)	mg/L	36.0	57.6
Sodium (Na <sup>+</sup> )	mg/L	575	920
Strontium (Sr <sup>2+</sup> )	mg/L	10.9	17.4
Sulfate (SO <sub>4</sub> <sup>2-</sup> )	mg/L	871	1,393
TDS (total ions + silica)	mg/L	3,627	5,801
TOC (RO feed TOC = 3.6 mg/L)	mg/L	11.3	18.0

Source: CH2M HILL (2001)

### 3.2.3. Regional Desalination Plant, Brownsville, Texas

The Southmost Regional Water Authority (SRWA) regional desalination plant, located near Brownsville, Texas, uses RO to treat brackish groundwater as an alternative water supply. Commissioned in the spring of 2004, the \$22 million facility has been producing 7.5 mgd of drinking water, equivalent to more than 40 percent of the City's annual potable water needs. Table 3.4 summarizes the waste stream discharges from the plant.



**Table 3.4 SRWA Regional Desalination Plant – Monthly Product Flow, Feed and Concentrate Quality**

		<b>Jan</b>	<b>Feb</b>	<b>Mar</b>	<b>April</b>	<b>May</b>	<b>June</b>	<b>July</b>
Product Flow (mgd)	Min	2.42	1.92	1.67	2.63	2.13	2.17	0.03
	Avg	2.73	2.51	2.56	2.80	2.66	2.62	2.28
	Max	3.07	3.10	2.95	2.97	3.30	2.73	3.01
Feed TDS (mg/L)	Min	3,050	2,960	3,040	3,160	3,110	2,960	3,000
	Avg	3,307	3,428	3,257	3,293	3,398	3,363	3,341
	Max	3,700	4,000	3,480	3,410	4,090	3,610	3,520
Concentrate TDS (mg/L)	Min	8,170	7,450	3,300	9,171	9,200	9,270	3,400
	Avg	9,480	10,483	10,305	10,219	10,410	10,776	9,708
	Max	10,590	12,610	11,880	10,710	11,730	11,570	12,040

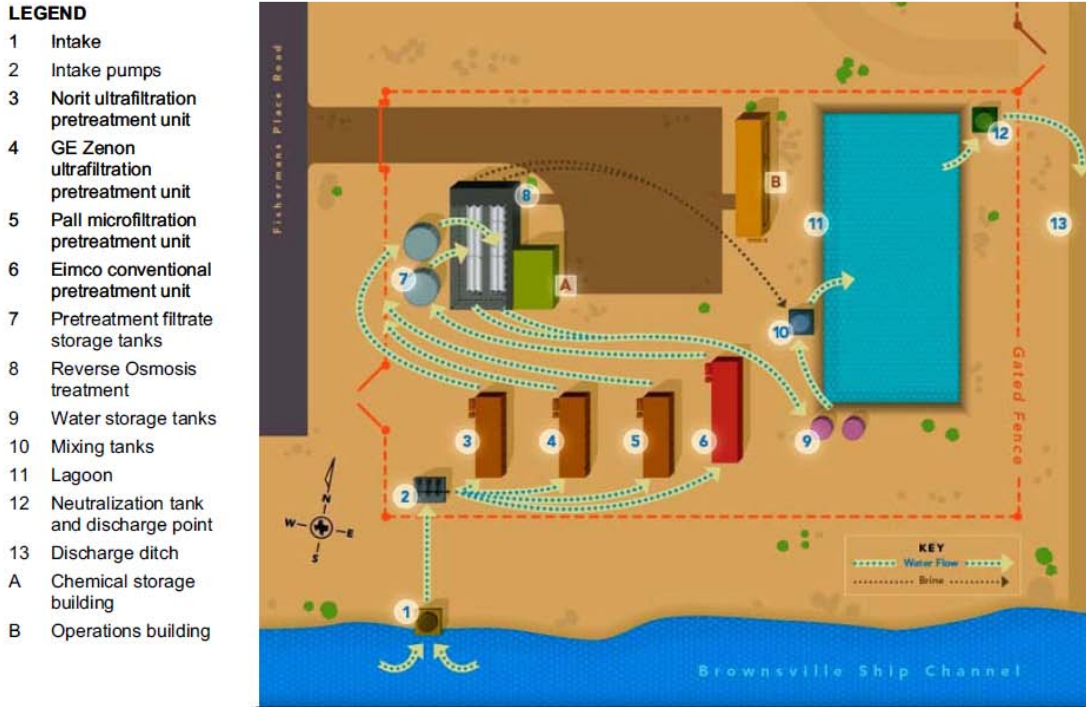
  

		<b>August</b>	<b>Sept</b>	<b>Oct</b>	<b>Nov</b>	<b>Dec</b>	<b>Annual</b>
Product Flow (mgd)	Min	0.47	0.90	0.82	1.10	1.61	1.49
	Avg	1.51	1.20	1.23	1.73	2.08	2.16
	Max	2.63	1.92	2.64	2.77	2.45	2.79
FeedTDS (mg/L)	Min	2,690	2,620	2,600	2,880	2,930	2,917
	Avg	3,387	3,324	3,267	3,302	3,353	3,335
	Max	5,490	3,940	3,860	3,750	3,740	3,883
Concentrate TDS (mg/L)	Min	4,470	4,610	5,860	5,960	9,200	6,672
	Avg	8,450	7,905	8,136	9,206	11,174	9,688
	Max	10,880	10,710	10,000	12,320	12,930	11,498

Source: NRS (2009)

**3.2.4. Brownsville, Texas Seawater Desalination Pilot Study**

In 2007, Brownsville Public Utilities Board (BPUB) and TWDB partnered to conduct a seawater desalination pilot study. The pilot facility was located on the north shore of the Brownsville Ship Channel, on land made available by the Port of Brownsville. The primary purpose of the pilot was to evaluate actual performance of the proposed desalination process under site-specific conditions. Piloting results were used to refine the design and cost estimate for a full-scale (25 mgd) seawater desalination facility. The Brownsville Seawater Desalination Pilot Project operated from February 2007 to July 2008. The results of the study were published in a report prepared in 2009 (NRS, 2009). An equipment layout of the pilot site is shown in Figure 3.3.



**Figure 3.3 Schematic of the Brownsville Seawater Desalination Pilot Study (NRS, 2009)**

The seawater RO unit was operated at a recovery of 48.8 percent. During the study, the average feed TDS was 30,932 mg/L. Feed water quality is summarized in Table 3.5.

**Table 3.5 Feed Water Quality for the Brownsville, Texas Seawater Desalination Pilot Study**

Parameter	Units	Minimum	Average	95th Percentile	Maximum
Alkalinity	mg/L as CaCO <sub>3</sub>	109.40	140.96	155.20	318.50
Barium (Ba <sup>2+</sup> )	mg/L	0	0.086	0.242	0.318
Bicarbonate (HCO <sub>3</sub> <sup>-</sup> )	mg/L as CaCO <sub>3</sub>	144	171	313	433
Boron (B)	mg/L	3.02	7.75	17.80	19.30
Calcium (Ca <sup>2+</sup> )	mg/L	357	386	418	434
Carbonate (CO <sub>3</sub> <sup>2-</sup> )	mg/L as CaCO <sub>3</sub>	2.49	3.00	5.99	6.46
Chloride (Cl <sup>-</sup> )	mg/L	13,900	17,083	24,360	25,500
Color, Apparent	PCU	0	12	25	25
Color, True	PCU	0	8	10	10
Conductivity	µS/cm	28,400	48,100	53,800	55,500
Fluoride (F <sup>-</sup> )	mg/L	0	0	0	0
Magnesium (Mg <sup>2+</sup> )	mg/L	911	1,135	1,310	1,330
Nitrate (NO <sub>3</sub> -N)	mg/L	0	2.62	1.05	2.62
pH (Calc)	SU	7.12	8.01	8.27	8.66
Potassium (K <sup>+</sup> )	mg/L	417	487	661	684
Sodium (Na <sup>+</sup> )	mg/L	6,390	8,468	10,175	10,500
Silica (SiO <sub>2</sub> )	mg/L	0	24.0	29.5	116.
Strontium (Sr <sup>2+</sup> )	mg/L	2.23	5.69	7.73	7.98

Parameter	Units	Minimum	Average	95th Percentile	Maximum
Sulfate (SO <sub>4</sub> <sup>2-</sup> )	mg/L	1,850	2,642	4,365	6,380
Temperature	°C	14.5	25.0	30.0	31.8
TDS	mg/L	17,600	29,800	33,300	34,400
TOC	mg/L	2.03	3.53	4.52	7.77
DOC	NTU	1.66	3.25	4.12	6.35
Turbidity	cm <sup>-1</sup>	0.31	44.7	121.8	2,745
UV254		0.019	0.047	0.070	0.130

Notes:

CaCO<sub>3</sub> = calcium carbonate

CU = color unit

µS/cm = micro Siemens per centimeter

SU = standard unit

Source: NRS (2009)

The TDS of the RO concentrate ranged from 55,800 mg/L to 68,700 mg/L; however, individual chemical constituent parameters were not reported in the pilot study report. CH2M HILL estimated concentrate quality using Dow Filmtec's ROSA projection software. The results are presented in Table 3.6.

**Table 3.6 Brownsville, Texas Seawater Desalination Pilot Study Estimated Concentrate Quality**

Parameter	Units	Feed <sup>1</sup>	Adjusted Feed <sup>2</sup>	Concentrate Stage 1	Permeate Stage 1	Permeate Total
Ammonium (NH <sub>4</sub> <sup>+</sup> )	mg/L	0	0	0	0	0
Barium (Ba <sup>2+</sup> )	mg/L	0.24	0.24	0.47	0	0
Bicarbonate (HCO <sub>3</sub> <sup>-</sup> )	mg/L	313.	364.	683.	3.7	3.7
Boron	mg/L	17.86	17.80	30.3	4.3	4.3
Calcium (Ca <sup>2+</sup> )	mg/L	434.	434.	839.	0.78	0.78
Carbonate (CO <sub>3</sub> <sup>2-</sup> )	mg/L	79.73	4.68	16.82	0	0
Chloride (Cl <sup>-</sup> )	mg/L	19,000	19,012.	36,665.	135.8	135.8
CO <sub>2</sub>	mg/L	0.79	18.35	24.5	20.35	20.35
Fluoride (F <sup>-</sup> )	mg/L	0	0	0	0	0
Magnesium (Mg <sup>2+</sup> )	mg/L	1,330	1,330	2,571.5	2.43	2.43
Nitrate (NO <sub>3</sub> <sup>-</sup> )	mg/L	1.05	1.05	1.98	0.06	0.06
pH	SU	8.30	7.00	7.20	5.41	5.41
Potassium (K <sup>+</sup> )	mg/L	684.00	684.	1,317.	6.16	6.16
Silica (SiO <sub>2</sub> )	mg/L	29.50	29.5	56.7	0.37	0.37
Sodium (Na <sup>+</sup> )	mg/L	10,175.	10,354.	19,961.	81.2	81.2
Strontium (Sr <sup>2+</sup> )	mg/L	7.73	7.73	14.9	0.01	0.01
Sulfate (SO <sub>4</sub> )	mg/L	2,642.	2,721.	5,264.	1.95	1.95
TDS	mg/L	34,798.	35,045.	67,568.	257.	257.

<sup>1</sup>Based on maximum feedwater quality from Table 3.5.

<sup>2</sup>Following acid addition.

Note: Projected Concentrate using Dow Filmtec Projection Software.

Source: CH2M HILL

### 3.2.5. Corpus Christi, Texas Seawater Desalination Pilot Study

Under funding provided by the United States Bureau of Reclamation, Aqua Resources International conducted a pilot study in Corpus Christi to evaluate membrane pretreatment requirements for SWRO. Pilot testing was conducted in 2003-2004. The principal objectives of the project were to evaluate and compare membrane and conventional pretreatment, as measured by pretreated (RO feed) water quality and RO performance, and to estimate life-cycle costs for both alternatives.

Several different membrane filtration and conventional media filtration technologies were evaluated. The project was implemented at the San Patricio Municipal Water District (SPMWD) water treatment facility near Corpus Christi, Texas. The feed water to the pilot plant was seawater from nearby Corpus Christi Bay, which is located in the northwestern area of the Gulf of Mexico. Feed water quality is presented in Table 3.7.

**Table 3.7 Corpus Christi, Texas Desalination Pilot Study Raw Water Quality**

Constituent	Units	Value
Aluminum (Al <sup>3+</sup> )	mg/L	0.1
Ammonia (NH <sub>3</sub> )	mg/L	0.11
Total Alkalinity	mg/L	143
Barium (Ba <sup>2+</sup> )	mg/L	0.03
Bicarbonate (HCO <sub>3</sub> <sup>-</sup> )	mg/L	153
Bromide (Br <sup>-</sup> )	mg/L	90
Calcium (Ca <sup>2+</sup> )	mg/L	395
Carbonate (CO <sub>3</sub> <sup>2-</sup> )	mg/L	110
Calcium Hardness (as CaCO <sub>3</sub> )	mg/L	986
Chloride (Cl <sup>-</sup> )	mg/L	15,700
Copper (Cu <sup>+2+</sup> )	mg/L	5.5
Conductivity	uS/cm	40.9 – 43.9
Fluoride (F <sup>-</sup> )	mg/L	0.86
Heterotrophic Plate Count (HPC)	CFU	94
Total Iron (Fe <sup>2+3+</sup> )	mg/L	0.03
Dissolved Iron (Fe <sup>2+</sup> )	mg/L	<0.01
Magnesium (Mg <sup>2+</sup> )	mg/L	990
Nitrate (NO <sub>3</sub> <sup>-</sup> )	mg/L	<0.01
pH	SU	8.4
Total Phosphorus (as phosphate PO <sub>4</sub> <sup>3+</sup> )	mg/L	0.05
Potassium (K <sup>+</sup> )	mg/L	355
Silica, Reactive (SiO <sub>2</sub> )	mg/L	<1
Sodium (Na <sup>+</sup> )	mg/L	9,100
Sulfate (SO <sub>4</sub> <sup>2-</sup> )	mg/L	2,230
Temperature	°C	17.0
Total Dissolved Solids (TDS - calculated)	mg/L	30,200
Total Hardness (as CaCO <sub>3</sub> )	mg/L	5,060
Total Organic Carbon (TOC)	mg/L	4
Total Solids (TS)	mg/L	32,000 – 36,000
Total Suspended Solids (TSS)	mg/L	10 – 70
Turbidity	NTU	1.7 – 1.93
UV-254	cm <sup>-1</sup>	0.051 – 0.058

Source: Henthorne (2007)

Although concentrate sampling was conducted during the study, concentrate quality results were not included in the report. Based upon the raw water quality, RO system flux (8 gfd, and recovery of 50 percent, CH2M HILL estimated concentrate quality using Dow Filmtec’s ROSA projection software. Table 3.8 presents estimated quality.

**Table 3.8 Corpus Christi, Texas Desalination Pilot Study Estimated Concentrate Quality**

Constituent	Units	Value
Barium (Ba <sup>2+</sup> )	mg/L	0.06
Bicarbonate (HCO <sub>3</sub> <sup>-</sup> )	mg/L	286.17
Bromide (Br <sup>-</sup> )	mg/L	179
Calcium (Ca <sup>2+</sup> )	mg/L	789
Carbonate (CO <sub>3</sub> <sup>2-</sup> )	mg/L	13.3
Chloride (Cl <sup>-</sup> )	mg/L	31279
Fluoride (F <sup>-</sup> )	mg/L	1.71
Iron (Fe <sup>2+/3+</sup> )	mg/L	<0.01
Magnesium (Mg <sup>2+</sup> )	mg/L	1,977
pH	SU	8.42
Total Phosphorous (as Phosphate PO <sub>4</sub> <sup>3+</sup> )	mg/L	0.1
Potassium (K <sup>+</sup> )	mg/L	705
Silica, Reactive (SiO <sub>2</sub> )	mg/L	<1
Sodium (Na <sup>+</sup> )	mg/L	18125
Sulfate (SO <sub>4</sub> <sup>2-</sup> )	mg/L	4454
Temperature	°C	19.0
Total Dissolved Solids (TDS)	mg/L	57,816

### 3.3. Characterization of Wastewater Treatment Plant Effluent

In response to CH2M HILL’s request for data, only two utilities responded with typical wastewater quality data: BPUB and EPWU. In order to provide a broader examination of available wastewater quality, data were also extracted from the literature.

#### 3.3.1. Typical Water Quality of Raw Domestic Wastewater

Domestic wastewater consists of raw potable water with increased total dissolved solids as a result of contamination from human wastes, soaps, etc. The increase in concentration typical in wastewater for specific constituents is listed in Table 3.9 (Metcalf and Eddy, 2003). The most significant additions in salinity occur due to increases in alkalinity, sodium, and chloride. With many Texas water supplies possessing TDS concentrations of up to 1,000 mg/L, it is anticipated that wastewater can contain TDS concentrations of up to 1,380 mg/L, where there are no industrial discharges.

**Table 3.9 Typical Chemical Constituent Increase in Potable Water from Domestic Use**

Constituent	Units	Increase in concentration
Total Alkalinity (as CaCO <sub>3</sub> )	mg/L as CaCO <sub>3</sub>	60 – 120
Aluminum (Al <sup>3+</sup> )	mg/L	0.1 – 0.2
Bicarbonate (HCO <sub>3</sub> <sup>-</sup> )	mg/L	50 – 100
Boron	mg/L	0.1 – 0.4
Calcium (Ca <sup>2+</sup> )	mg/L	6 – 16
Carbonate (CO <sub>3</sub> <sup>2-</sup> )	mg/L	0 – 10

Constituent	Units	Increase in concentration
Chloride (Cl <sup>-</sup> )	mg/L	20 – 50
Fluoride (F <sup>-</sup> )	mg/L	0.2 – 0.4
Magnesium (Mg <sup>2+</sup> )	mg/L	4 – 10
Manganese (Mn <sup>2+/3+</sup> )	mg/L	0.2 – 0.4
Nitrate (NO <sub>3</sub> <sup>-</sup> )	mg/L	20 – 40
Phosphate (PO <sub>4</sub> <sup>3-</sup> )	mg/L	5 – 15
Potassium (K <sup>+</sup> )	mg/L	7 – 15
Sodium (Na <sup>+</sup> )	mg/L	40 – 70
Silica, Total (SiO <sub>2</sub> )	mg/L	2 – 10
Sulfate (SO <sub>4</sub> <sup>2-</sup> )	mg/L	15 – 30
Total Dissolved Solids (TDS)	mg/L	150 – 380

Source: Metcalf and Eddy (2003)

During wastewater treatment, the concentration of various constituents change as the wastewater is treated through primary, secondary, and tertiary treatment stages. Table 3.10 displays typical wastewater concentrations for raw wastewater, primary treated wastewater, and secondary treated wastewater.

**Table 3.10 Typical Water Quality for Various Wastewater Streams**

Constituent	Units	Raw wastewater	Primary treated wastewater	Secondary treated wastewater
Total Alkalinity (as CaCO <sub>3</sub> )	mg/L as CaCO <sub>3</sub>	NR	NR	NR
Aluminum (Al <sup>3+</sup> )	mg/L	NR	NR	NR
Bicarbonate (HCO <sub>3</sub> <sup>-</sup> )	mg/L	NR	NR	NR
Boron	mg/L	0.35	0.38	0.42
Calcium (Ca <sup>2+</sup> )	mg/L	74.4	72.2	66.7
Carbonate (CO <sub>3</sub> <sup>2-</sup> )	mg/L	NR	NR	NR
Chloride (Cl <sup>-</sup> )	mg/L	240	232	238
Fluoride (F <sup>-</sup> )	mg/L	NR	NR	NR
Magnesium (Mg <sup>2+</sup> )	mg/L	38.5	38.1	39.3
Manganese (Mn <sup>2+/4+</sup> )	mg/L	0.065	0.062	0.039
Nitrate (NO <sub>3</sub> <sup>-</sup> )	mg/L	0.1	0.1	1.4
Phosphate (PO <sub>4</sub> <sup>3-</sup> )	mg/L	6.1	5.1	3.4
Potassium (K <sup>+</sup> )	mg/L	NR	NR	NR
Sodium (Na <sup>+</sup> )	mg/L	198	192	198
Sulfate (SO <sub>4</sub> <sup>2-</sup> )	mg/L	312	283	309
Silica, Total (SiO <sub>2</sub> )	mg/L	NR	NR	NR
Total Dissolved Solids (TDS)	mg/L	270 – 860	270 – 860	500 – 700

Note:

Typical wastewater quality is comprised of potable water that possesses increasing TDS as a result of wastewater constituents.  
NR = Not Reported

Source: Metcalf and Eddy (2003)

### 3.3.2. El Paso, Texas

The El Paso Water Utility provided comprehensive water quality data for all of their wastewater treatment facilities, Northwest Wastewater Treatment Plant (WWTP), Roberto Bustamante (RB) WWTP, Haskell Street (HS) WWTP and Fred Harvey (FH) WWTP. Table 3.11 summarizes data provided by EPWU.

**Table 3.11 Treated Wastewater Quality for the WWTPs in El Paso, Texas**

Parameter	Units	Northwest			RB	HS			FH		
		Min	Avg	Max	Avg	Min	Avg	Max	Min	Avg	Max
T-ALK	mg/L	116.2	152.6	199.3	125.6	104.2	133.0	165.2	53.5	149.5	181.9
Ammonia (NH <sub>3</sub> )	mg/L	0.025	0.47	3.3	3.48	<0.020	0.68	5.8	<0.020	<0.020	1.6
Bromide (Br <sup>-</sup> )	mg/L	0.05	0.28	0.55	<0.25	<0.25	<0.25	<0.25	<0.05	<0.05	0.14
Calcium (Ca <sup>2+</sup> )	mg/L	44.5	65.38	78.3	75.9	38.5	61.0	84.2	33.4	58.11	79.5
CBOD5	mg/L	2	3.14	4.7	4.325	<2.0	3.09	5.1	NR	NR	NR
COD	mg/L	8	20.23	39	39	13	24.2	36	NR	NR	NR
Conductivity	mS/cm	1650	1923.8	2450	1732.5	1260	1468.1	1790	898	1185.7	1440
Fluoride (F)	mg/L	0.17	0.81	1.09	0.8	0.62	0.86	1.29	0.44	0.66	0.88
Total Hardness	mg/L	176	213.5	257	268.3	146	203.3	280	116	168.9	216
Magnesium (Mg <sup>2+</sup> )	mg/L	9.2	13.02	16.8	17.2	10.8	13.4	17.8	2.8	6.74	12.3
Nitrate (NO <sub>3</sub> <sup>-</sup> )	mg/L	NR	NR	NR	NR	5	12.08	18.1	NR	NR	NR
Nitrite (NO <sub>2</sub> <sup>-</sup> )	mg/L	0.06	0.49	0.87	0.85	<0.25	<0.25	0.33	<0.05	<0.05	<0.05
Total TKN	mg/L	2.1	3.367	7	5.68	<2.0	3.99	17.4	NR	NR	NR
Ortho Phosphate (o-PO <sub>4</sub> )	mg/L	0.11	1.76	5.49	4.13	0.28	1.76	6.35	0.32	0.61	0.98
pH	-	7.6	7.93	8.1	NR	7.1	7.4	7.6	6.3	6.99	162
Total P	mg/L	0.3	2.04	6.3	3.95	0.4	2.19	7.1	0.3	0.73	NR
Potassium (K <sup>+</sup> )	mg/L	10.6	12.79	16.2	19.8	13.4	17.3	20.5	12.2	17.8	NR
Silica (SiO <sub>2</sub> )	mg/L	22	35.72	41.5	26.1	5.4	31.3	40.8	18.1	27.05	37.1
Sodium (Na <sup>+</sup> )	mg/L	261	329.98	386	248.3	167	220.9	272	117	172.4	NR
Sulfate (SO <sub>4</sub> <sup>2-</sup> )	mg/L	139	310.2	363	294	129	180.3	254	38.3	96.13	NR
TDS	mg/L	1,070	1,236.4	1,590	1,087.5	812	922.7	1,160	580	709.6	1,110
Temp	°C	13.5	17.73	20.1	NR	15.7	18.4	22.4	11.6	22.04	39.2
TSS	mg/L	1	1.85	4	5.75	1	4.15	10	NR	NR	NR
Turbidity	NTU	0.08	1.18	2.46	3.55	0.9	2.51	5.08	0.12	0.31	NR
VSS	mg/L	1	1.55	3	1155	<1	3.42	8	NR	NR	NR

NR = Not Reported

### 3.4. Characterization of Osmotic Pressures of Various Waters in Texas

When examining the use of osmotic potentials for wastewater/desalination coupling, the following options were identified for initial screening:

- (1) Brackish water concentrate – primary effluent of domestic wastewater
- (2) Brackish water concentrate – secondary effluent of domestic wastewater
- (3) Seawater desalination concentrate – primary effluent of domestic wastewater
- (4) Seawater desalination concentrate – secondary effluent of domestic wastewater

Based upon discussion with Hydration Technologies, Inc. (HTI), the membranes provided for characterization and testing were cellulose triacetate (CTA) spiral-wound membranes, possessing poor solids tolerance and increased risk of fouling due to biological growth. As a result of these product limitations, coupling of concentrate with primary treated wastewater was not evaluated as part of this study. Additionally, given the variability of brackish waters in Texas, four different brackish waters were utilized in the analysis, as well as seawater and seawater concentrate (Table 3.12).

**Table 3.12 Calculated Total Dissolved Solids and Osmotic Pressure for Various Waters in Texas**

Item Description	Total Dissolved Solids (mg/L)	Osmotic Pressure (psi)
Raw Wastewater	1,106	9.1
Primary Effluent of Wastewater	1,079	8.9
Secondary Effluent of Wastewater	989	8.1
El Paso WWTP Effluent (avg)	1,236	10.9
El Paso Brackish Water Concentrate (avg)	3,561	36.8
El Paso Brackish Water Concentrate (max)	5,887	59.5
High Salinity Brackish Water	10,000	100
Seawater	34,630	351
Seawater Concentrate	67,423	672

Note: Values estimated based upon water quality data collected as part of this study. Data for El Paso are from 2007 and 2008 and differ from the 2008-2010 data presented in Tables 3.2, B.6, and B.7.

Prior to determining the experimental protocol, the anticipated flux for each stream was determined using data from the literature for the HTI membrane. Using RO concentrates from Section 3.2, the osmotic pressure for each stream was calculated utilizing a spreadsheet model. The model determines osmotic pressure using the method outlined in Stumm and Morgan (1996). The differential osmotic pressure (between the wastewater and the RO concentrate) represents the potential driving force available for mass transfer in a FO application. Represented by  $\Delta\pi_{\text{potent}}$  in Equation 3-1, it is determined by subtracting the osmotic pressure of the wastewater stream ( $\pi_{\text{wastewater}}$ ) from the osmotic pressure of the concentrate stream ( $\pi_{\text{Concentrate}}$ ) using Equation 3-1, below:

$$\Delta\pi_{\text{Potent}} = \pi_{\text{Concentrate}} - \pi_{\text{wastewater}} \quad (3-1)$$

The osmotic potential for each stream described in Table 3.12 is shown in Table 3.13.

**Table 3.13 Calculated Differential Osmotic Pressure for Various Combinations Coupling RO Concentrate with Wastewater**

Draw Solution	Wastewater	$\Delta$ Total Dissolved Solids (mg/L)	Differential Osmotic Pressure (psi)
A El Paso concentrate (avg)	El Paso WWTP Effluent (avg)	2,572	25.9
B El Paso concentrate (max)	El Paso WWTP Effluent (avg)	4,898	48.5
C Brackish water concentrate (high)	El Paso WWTP Effluent (avg)	9,011	89.9
D Seawater	El Paso WWTP Effluent (avg)	33,641	340
E Seawater concentrate	El Paso WWTP Effluent (avg)	66,441	662

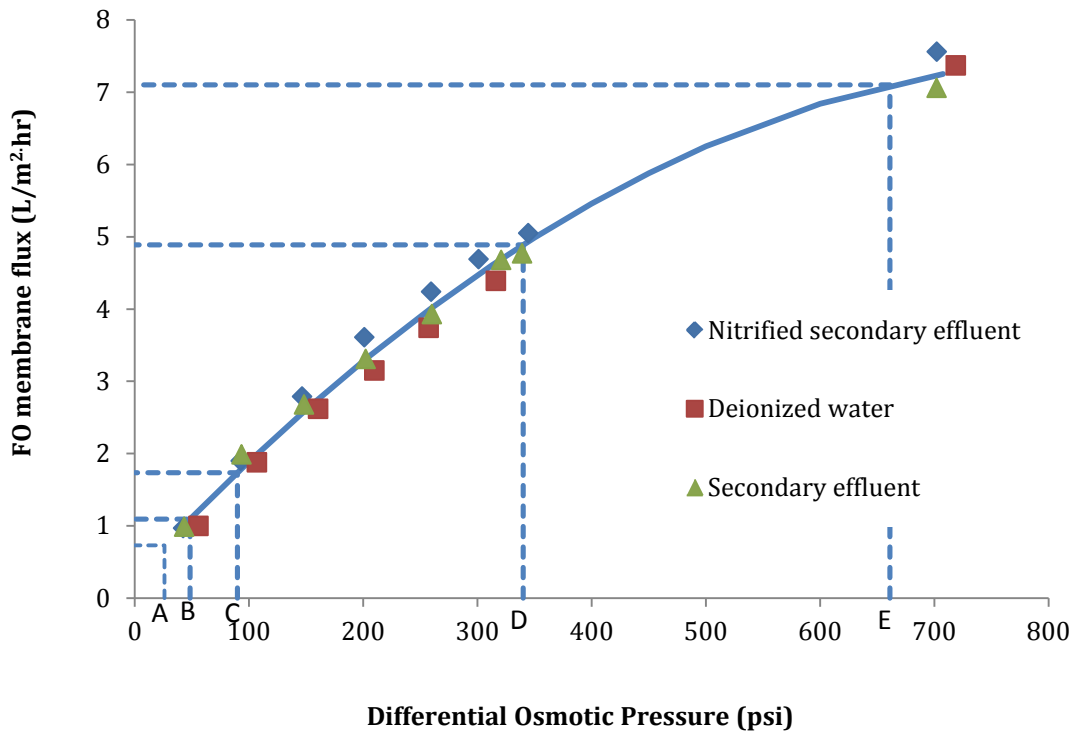
Note: Data for El Paso are from 2007 and 2008 and differ from the 2008-2010 data presented in Tables 3.2, B.6, and B.7.

In order to estimate the membrane flux that can be anticipated with each pairing, previous differential osmotic pressure versus flux data published using HTI membranes was examined (Cath and others, 2006). The differential osmotic pressure data contained in Table 3.3 was plotted versus membrane flux to provide an initial estimate of the flux anticipated using the HTI



cellulose acetate membrane. Figure 3.4 presents the resultant estimates of flux for each of the pairings in Table 3.13 using the data developed by Cath and others. The non-linear data developed by the researchers are a result of internal concentration polarization within the membrane.

Note that different, new membrane coupons were utilized for each measurement. The permeability of the membrane is subject to some variation. The data indicate that membrane flux for lower differential osmotic pressure is very low, which would likely result in very high capital equipment costs. As a result, the use of draw solutions with very high TDS, such as brackish water concentrate, seawater, or seawater concentrate streams are most relevant for schemes using desalination plant concentrate to recover wastewater using FO.



**Figure 3.4 Experimental Flux Versus Draw Solution Concentration Of Various Waters (adapted from Cath and others, 2009 to illustrate anticipated flux in the experimental portion of the study)**

Based upon the results of this initial screening, the primary focus of subsequent experimental testing will focus on higher concentration draw solutions. Currently, there are no municipal full-scale desalination facilities in Texas capable of providing highly concentrated brine/concentrate for FO applications.

## 4. Performance of Forward Osmosis Membranes

In order to examine the mechanics of forward and reverse osmosis water treatment and assess the feasibility of using high salinity streams to extract water out of wastewater streams, performance data for a system using RO concentrate to recover wastewater are required. To obtain this data, Colorado School of Mines (CSM) conducted laboratory and pilot testing of a FO system coupling the use of simulated RO concentrate with secondary wastewater as an impaired water feed stream. The testing conducted fulfills the requirements of Task 3 of the agreed upon scope of work for the project. The testing was conducted in two distinct phases. During the first phase of testing, FO was tested on the bench-scale using synthetic brines as the draw solution. The second phase of the tests included a pilot-scale FO-RO hybrid system. The pilot system used a novel spiral-wound-packed FO membrane provided by HTI using brine generated from synthetic saline water.

This chapter is structured as follows:

- Testing methodology – describes the apparatus used in the testing, including a description of the systems, membranes and draw solutions utilized in the study
- Performance of the bench-scale system – describes the results of the bench-scale testing on a flat-sheet membrane
- Performance of the pilot-scale system – describes the results of the pilot-scale testing of a spiral-wound membrane element

### 4.1. Testing Methodology

In order to evaluate the combined FO-RO process, the performance (i.e., water production rate and reverse salt diffusion) of both a small bench-scale system and on a novel pilot scale system were evaluated. This section describes the apparatus used, including the membranes and draw solutions utilized.

#### 4.1.1. Bench-scale FO System

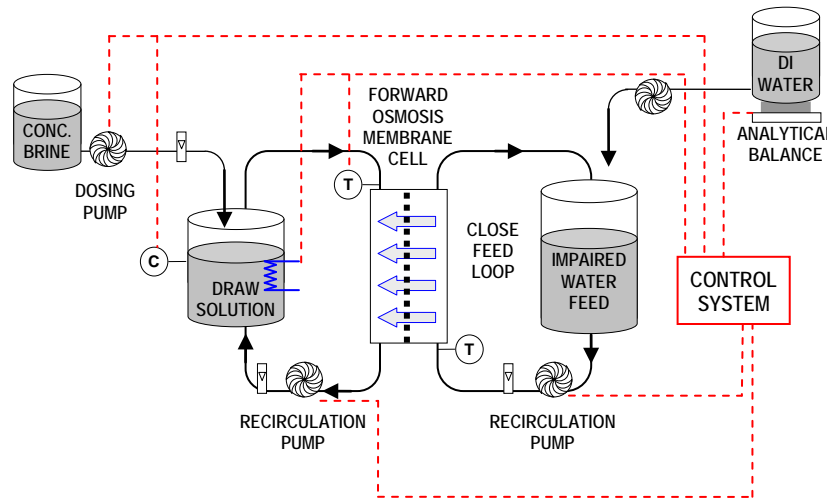
The bench-scale system utilized in the testing consists of a plate-and-frame membrane cell having 500 cm<sup>2</sup> of membrane surface area; a 3 liter plastic feed tank; and a 3 liter draw solution tank (Figure 4.1). The bench-scale system was fully controlled by a supervisory control and data acquisition (SCADA) system. The membrane used in the experiment was a commercially available, proprietary, dense CTA membrane that was specifically developed for FO applications by HTI. The membrane utilized in the study was harvested from a new casting batch that was also used for manufacture of the spiral-wound membrane element that was tested with the pilot scale system.

The FO apparatus was constructed such that two FO flow cells were constructed with symmetric flow chambers on both sides of the membrane that facilitated parallel, co-current or countercurrent flow along the membrane. The membrane surface area in cell #1 and cell #2 were 509 and 632 cm<sup>2</sup>, respectively. Each FO cell contains four narrow channels to minimize

fluctuations in hydraulic pressure that could negatively affect the operation and performance of the process. Turbulence-enhancement spacers were not installed in the flow channels in order to simplify hydrodynamic conditions and facilitate model validation with the experimental results. Cell #1 was utilized in most experiments conducted.

During the experiments, a constant cross-flow velocity and draw solution concentration was maintained on the draw solution side of the membrane. This was achieved by replenishing water that crossed the membrane with deionized (DI) water (controlled by the SCADA system). The amount of DI water that was added into the feed tank during the experimental run was used in calculation of the water flux across the membrane during the experimental run.

Additionally, reverse solute flux through the membrane can result in loss of salt into the impaired water feed solution. To ensure constant driving force, a make-up stream of concentrated draw solution was provided to the draw solution tank. The amount of solution added was determined by the SCADA system to maintain a constant conductivity in the draw solution. This maintained a constant osmotic pressure difference between the draw solution and the feed solution. A small peristaltic dosing pump was provided to automatically add the required draw solution concentrate. The reverse flux across the FO membrane was calculated based upon the volume and concentration of the added solution. Reconstitution of the draw solution was not practiced as part of the bench-top process.



**Figure 4.1 Schematic Drawing of the Forward Osmosis Bench Scale System**

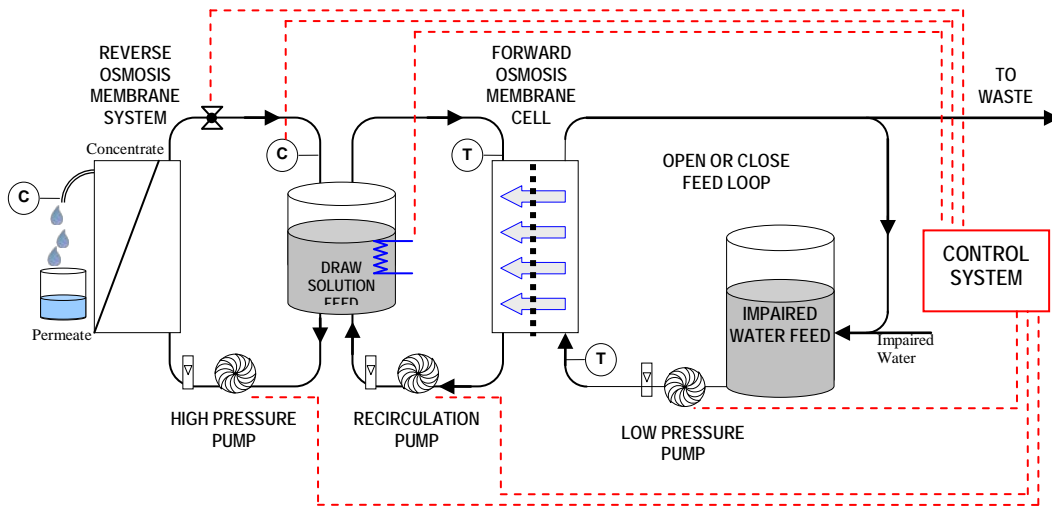
#### **4.1.2. Pilot-scale Forward Osmosis/Reverse Osmosis Hybrid System**

In the FO-RO hybrid system, the RO process produces purified water and concentrated brine that serves as the draw solution for the FO process. The concentrated draw solution is diluted during the FO process. The diluted draw solution is treated in a RO system to recover product water and to reconstitute the draw solution concentration. A schematic drawing of FO membrane configuration is shown in Figure 4.2. In the configuration studied, the system included the FO system, which treated an impaired feed stream consisting of secondary wastewater effluent from Golden, Colorado, and the RO reconcentration process. The FO system collected the impaired water stream into a tank and pumped the impaired water stream through the FO spiral-wound membrane. The FO subsystem consisted of a single spiral-wound membrane element. The FO

membrane was fed at one end with water from the FO impaired water feed tank and brine (draw solution) from the draw solution tank. A portion of the water was extracted from the impaired water stream across the membrane. The diluted draw solution stream and the concentrated feed solution came out through the opposite end of the FO membrane element, with the diluted draw solution stream flowing into the draw solution tank for reconstitution by the RO. During field testing, the FO impaired water feed was operated in an open loop, once through configuration. The impaired water feed tank was maintained at a constant level during the operation of the pilot study.

The RO reconcentration system consisted of a high-recovery, two-stage RO skid. The first stage consisted of three membrane elements, while the second stage was constructed with a single membrane element. Concentrate from the second stage was recycled to the draw solution tank. Draw solution concentration was maintained by adjusting the RO feed pressure to control the flux of permeate out of the system. Process parameters (e.g., conductivity, temperature, and pH) of the three streams were measured and reported by an integrated data acquisition system. Additional concentrated draw solution was occasionally added to the draw solution tank to maintain a constant draw solution concentration and osmotic pressure gradient across the FO membrane.

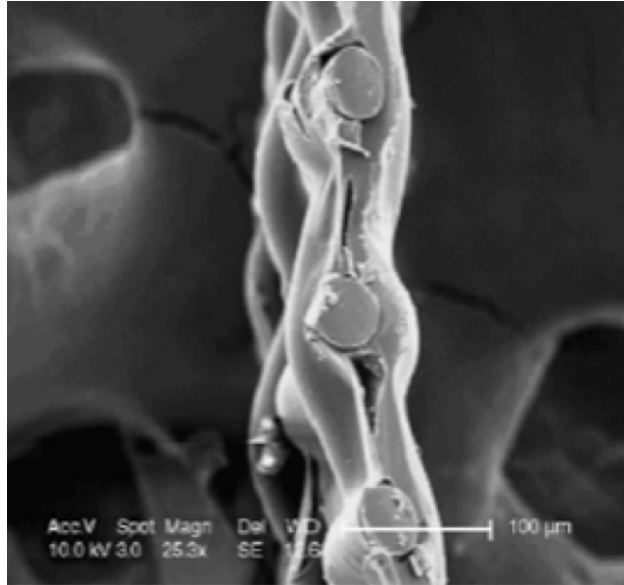
Additional details on the pilot system are provided in Section 4.5.



**Figure 4.2 Schematic Drawing of the Forward Osmosis Pilot Scale System**

#### 4.1.3. FO Membranes Used in the Study

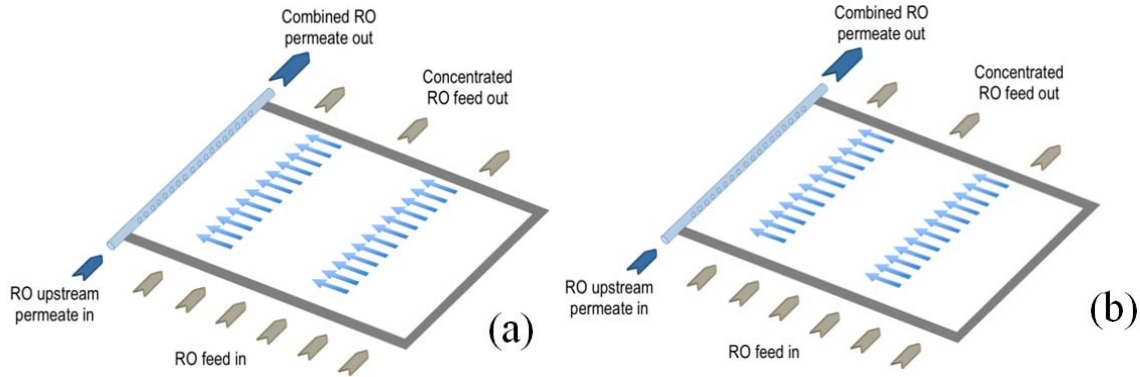
A proprietary FO membrane made from a hydrophilic cellulose-based polymer was used. The FO membrane is manufactured by HTI (Albany, Oregon) and is optimized for osmotically driven membrane processes. The membrane is thought to incorporate a CTA active layer (Lampi and others, 2005) cast directly onto a woven polyester mesh to create a flat sheet (Cath and others, 2006). An SEM micrograph of the membrane's cross-section is seen in Figure 4.3.



**Figure 4.3 SEM Micrographs of a Cross Section of the HTI CTA FO Membrane (McCutcheon and others, 2006)**

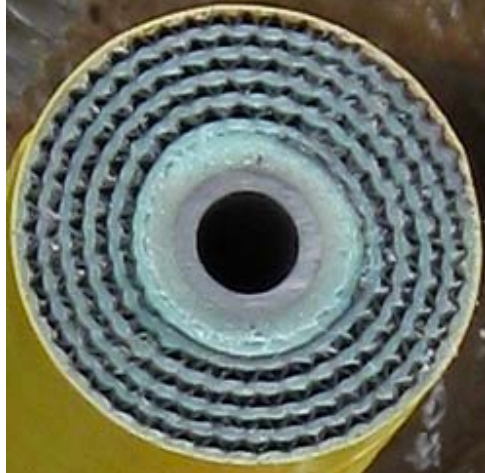
Previous studies (Elimelech and others, 1997; Hoek and others, 2002; Mi and Elimelech, 2010; Lee and others, 2010; Holloway and others, 2007; and Mi and Elimelech, 2008) have shown that the relatively smooth active layer surface of this membrane combined with the unique operating conditions of osmotically driven membrane processes (ODMP) substantially reduce the severity of irreversible membrane fouling during treatment of water with high concentrations of organic matter. These benefits may allow the osmotic dilution process with the CTA membrane to provide pretreatment of highly impaired wastewater prior to treatment with RO to facilitate potable reuse of impaired water.

For this study, the proprietary and commercially available CTA membrane was employed in a spiral-wound packaging configuration that improves membrane-packing density (i.e., membrane active surface area per module volume) for commercial applications. However, conventional spiral-wound module designs (e.g., spiral-wound RO modules) are inadequate for the unique hydraulic flows required for FO and osmotic dilution. In RO spiral-wound modules the product water is forced out of the membrane envelope by permeating water (Figure 4.4a); however, in a FO spiral-wound module, the draw solution (DS) has to flow through the membrane envelope (Figure 4.4b) to facilitate mass transfer across the membrane. As illustrated in Figure 4.4b, an additional glue line is added in the center of the membrane envelope that partially extends toward the outer edge of the membrane envelope, and a plug is placed at the center of the core tube to force the DS to flow inside the membrane envelope.



**Figure 4.4** Flow patterns in a spiral-wound module used for (a) RO and (b) modified for ODMP. The feed stream flows tangentially across the exterior of the rolled membrane envelope in both cases; however, for ODMP DS is forced to flow into the core tube and then through the interior of the membrane envelope before exiting the module.

The novel 4" diameter by 40" long spiral-wound FO module that was investigated in the pilot-study stage is made of a single membrane envelope with an active area of 1.58 m<sup>2</sup> and fits inside a conventional 4" diameter membrane pressure vessel. This spiral-wound membrane uses a larger feed spacer than those commonly employed in spiral-wound RO membranes. This feed spacer provides larger and more continuous channels for fluid flow to accommodate suspended solids and other debris that might be present in impaired water. Two standard RO permeate spacers are installed in the interior of the membrane envelope to reduce pressure drop in the DS channels and to enhance the flow of the DS. A cross-section view of the novel spiral-wound membrane is shown in Figure 4.5. Water permeance and NaCl rejection for the virgin CTA membrane were measured during experiments conducted in RO mode with a SEPA-CF membrane test cell (GE Osmonics, Minnetonka, MN).



**Figure 4.5** Photograph of the cross section of a novel spiral-wound membrane module for ODMP cast and wound by HTI. Distinct features include the draw solution core tube (schedule 80 PVC pipe), enhanced feed spacer (grey corrugation circling the central tube), and the low-pressure housing (yellow casing on the outer radius of membrane).

Feed at a constant temperature of 25 °C was either deionized water for permeance tests or a 2 g/L NaCl solution for rejection tests. Tests were conducted with feed pressures of 100 and 150 pounds per square inch gauge (psig). Water permeance and NaCl rejection for the SW30 thin-film polyamide (TFPA) (produced by Dow Filmtec) membrane were obtained from the manufacturer’s technical specifications data sheet. These values, along with the values for the CTA membrane, are summarized in Table 4.1. The TFPA RO membrane has higher water permeance and NaCl rejection compared to the CTA membrane; however, the thick porous support structure of commercial RO membranes substantially increases ICP effects during FO operation, thus limiting their application for ODMP.

**Table 4.1** Summary of Water Permeance and NaCl Rejection Values for the CTA (FO) and TFPA (RO) Membranes Used in this Study

Membrane	Water Permeance, L/m <sup>2</sup> ·hr·bar	NaCl Rejection, %
CTA	0.78 ± 0.02	93.2 ± 0.9
TFPA	1.31	99.4

## 4.2. Solution Chemistry

In the laboratory tests conducted, three draw solutions were used. These included: sodium chloride (NaCl), synthetic seawater, and synthetic brackish water. The three draw solutions were tested on the bench-scale system. The concentrate was recycled to the reconstitution tank. DI water was used in preparing all solutions for the bench-scale system. Three impaired feed solutions were tested; these included DI water, secondary effluent (impaired water) from Denver Water Recycling Plant (Denver, CO), and river water from the South Platte River (impaired water) upstream of downtown Denver. The secondary effluent had an average turbidity of 0.6 NTU, while the turbidity of the South Platte River water was 1.1 NTU. The compositions of the synthetic seawater and brackish water at 10 g/L TDS concentration are shown in Table 4.2. ACS

grade sodium chloride, acquired from Sigma-Aldrich/Fluka (Buchs, Switzerland), was used for preparation of sodium chloride draw solutions.

**Table 4.2** Composition of Synthetic Draw Solutions Utilized in the Study for the 10 g/L Solution Concentration. Note that various concentrations of the solutions were utilized, however the concentration ratios were maintained constant throughout the study

Constituent Ion	Units	SW Concentration at ~10 g/L TDS <sup>1</sup>	BW Concentration at ~10 g/L TDS
Sodium (Na <sup>+</sup> )	mg/L	3,080.00	2,387.4
Calcium (Ca <sup>2+</sup> )	mg/L	114.29	690.6
Magnesium (Mg <sup>2+</sup> )	mg/L	377.14	172.033
Potassium (K <sup>+</sup> )	mg/L	120.00	88.54
Carbonate / Bicarbonate (CO <sub>3</sub> <sup>2-</sup> / HCO <sub>3</sub> <sup>-</sup> )	mg/L	57.14	667.0
Chloride (Cl <sup>-</sup> )	mg/L	5,511.43	4,782.1
Sulfate (SO <sub>4</sub> <sup>2-</sup> )	mg/L	760.00	1,015.6
Bromide (Br <sup>-</sup> )	mg/L	16.00	1.45
Strontium (Sr <sup>+</sup> )	mg/L	2.51	
Barium (Ba <sup>2+</sup> )	mg/L	-	0.57
Boron (B)	mg/L	1.60	0.633
Fluoride (F <sup>-</sup> )	mg/L	0.29	4.1
Nitrate (NO <sub>3</sub> <sup>-</sup> )	mg/L	-	10.2
Total Phosphorous (P)	mg/L	-	1.26
Silica (SiO <sub>2</sub> )	mg/L	-	194.1

<sup>1</sup>Instant Ocean

### 4.3. Experimental Procedure

#### 4.3.1. Bench-scale Experimental Procedure

##### *DI water feed – NaCl draw solution experiments*

Baseline experiments were conducted with DI water as the feed solution. The draw solutions used in the experiments were 10, 20, 30, 50 g/L NaCl solutions. Before this set of experiments was conducted, a baseline performance test was performed with DI water as the feed and a 50 g/L NaCl draw solution. The last experiment in the series (with 50 g/L NaCl draw solution) was used as a comparison against the baseline. The similarity in results obtained for the first and second integrity tests serves as an assurance that the membrane and its performance did not degrade in the course of the experiments. At the start of each experiment, 1 L of draw solution was poured into the draw solution tank. Each experiment was conducted for 1.5-2 hours, and was terminated when a linear increase in the conductivity of the feed solution was observed; this linear increase serves as an indication of a constant rate of reverse solute diffusion, a phenomenon that refers to the diffusion of solute from the draw solution into the feed solution (forward diffusion refers to the diffusion of solute with water from the feed to the draw solution). At the termination of each experiment the system was thoroughly rinsed with DI water and drained before the start of a new experiment. Only one cell was used for these experiments. The membrane surface area was approximately 500 cm<sup>2</sup>.



**Experiments with DI alternate feeds and draw solutions**

For each of the four feed solution-draw solution (DS) combinations (i.e., DI-synthetic seawater, DI-synthetic brackish water, secondary effluent-synthetic seawater, and secondary wastewater effluent-synthetic brackish water), three different concentrations of the draw solution were tested (i.e., 10 g/L, 20 g/L, and 30 g/L). A membrane integrity test was conducted before and after each experiment. The integrity test was conducted with a DI feed solution and 50 g/L NaCl draw solution. The setup for these experiments was a 2-cell setup. Two membrane cells were connected in series with a total membrane surface area of approximately 1,150 cm<sup>2</sup>. Water samples from the feed and draw solutions were taken before and after each experiment, and later analyzed. Analyses included dissolved organic matter, salt composition (i.e., TDS, Ca<sup>2+</sup>, SO<sub>4</sub><sup>2-</sup>, Br<sup>-</sup>, PO<sub>4</sub><sup>3-</sup>, NO<sub>3</sub><sup>-</sup>, NH<sub>4</sub><sup>+</sup>, boron, and other constituents of interest), and general water quality parameters (i.e., pH, turbidity, TOC, and UVA). Rejection of different groups of contaminants, including organic and inorganic compounds, was calculated. Percent rejection of constituents in a given experimental sample was determined by conducting mass balances on the system for each pertinent experiment.

Calculation of the membrane flux during the experiment was conducted by dividing the total mass of DI water added to the impaired water feed tank by the FO membrane area and the duration of the experiment, as illustrated in Equation 4-1.

$$J_w = (M_{DI} \cdot sg) / (t \cdot A_{mem}) \tag{4-1}$$

where:

- $J_w$  = water flux
- $M_{DI}$  = mass of deionized water fed to the impaired water feed tank
- sg = specific gravity
- $t$  = duration of the experimental run
- $A_{mem}$  = area of the FO membrane in the test cell

The salt rejection of the membrane for each water quality constituent (i) is determined by measuring impaired water concentration before and after passing through the membrane cell. The calculation is shown in Equation 4-2.

$$R_i (\%) = 1 - \frac{C_{i,in}}{C_{i,out}} \tag{4-2}$$

where:

- $R_i$  = rejection of water quality constituent i
- $C_{i,in}$  = concentration of impaired water stream pumped into the FO cell
- $C_{i,out}$  = concentration of impaired water stream coming out of the FO cell

The salt reverse diffusion rate is calculated by determining the mass of concentrated draw solution added to the draw solution feed tank and dividing it by the duration of the experiment and the membrane area (Equation 4-3). The specific reverse diffusion rate is calculated by dividing the reverse diffusion rate by the membrane flux during the experiment.

$$J_s = (M_{DS \text{ Added}})/(t \cdot A_{mem}) \quad (4-3)$$

where:

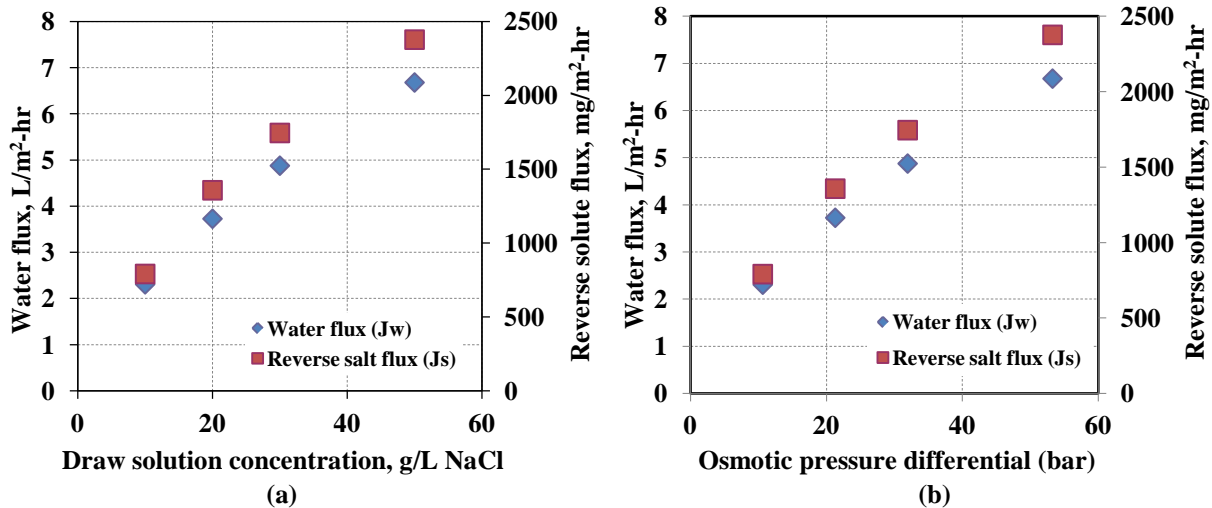
- $J_s$  = salt (solute) flux  
 $M_{DS}$  = solute mass of concentrated draw solution water fed to the draw solution tank over the duration of the experiment  
 $t$  = duration of the experimental run  
 $A_{mem}$  = area of the FO membrane in the test cell

## 4.4. Results

### 4.4.1. Bench-scale System

#### Performance tests with DI water feed and NaCl draw solution

Water flux and reverse diffusion of salt as a function of draw solution concentration are shown in Figure 4.6 for the four tests conducted with the DI water as the feed stream and different draw solution concentrations. From the data in Figure 4.6, it can be seen that water flux and the reverse diffusion of NaCl increased with increasing draw solution concentration. The reverse salt diffusion was determined from the rate of change of feed solution conductivity recorded by the SCADA system in conjunction with an experimentally determined conversion factor between conductivity and NaCl concentration.



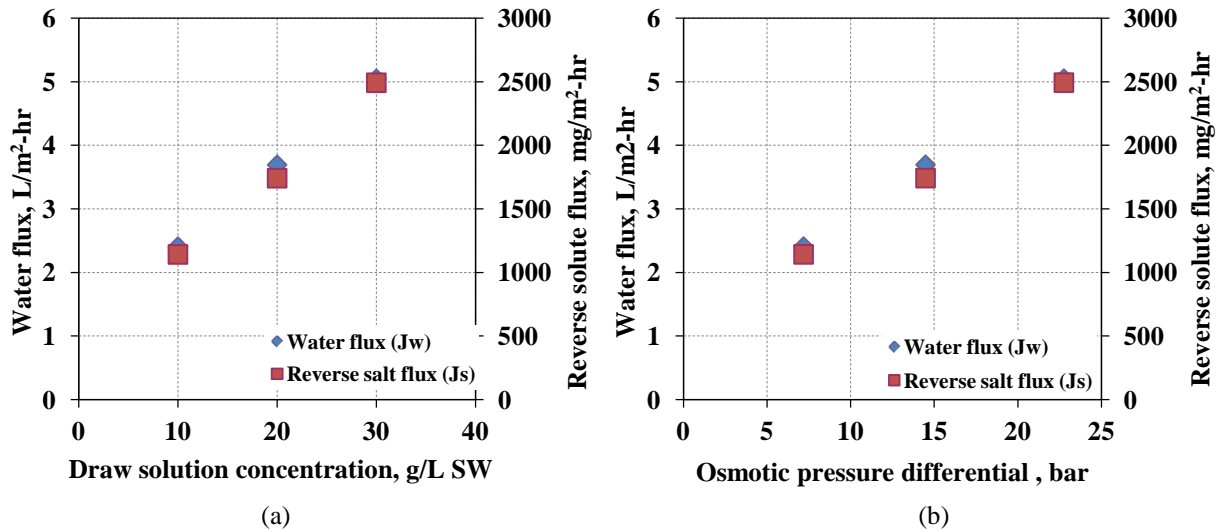
**Figure 4.6** Water flux and reverse solute (NaCl) flux as functions of draw solution concentration (a), and osmotic pressure differential (b).  $J_w$  represents water flux, and  $J_s$  represents reverse solute flux.

The observed increase in water flux with increasing draw solution concentration is due to an increase in osmotic pressure differential ( $\Delta\pi$ ) across the membrane, which is the driving force for mass transport. The increase in water flux is non-linear; this effect is likely due to increasing dilutive internal concentration polarization with increasing draw solution concentrations, which decreases slightly the osmotic pressure at the membrane interface. Similarly, the increase in reverse salt flux was expected because of the chemical potential gradient that exists between the

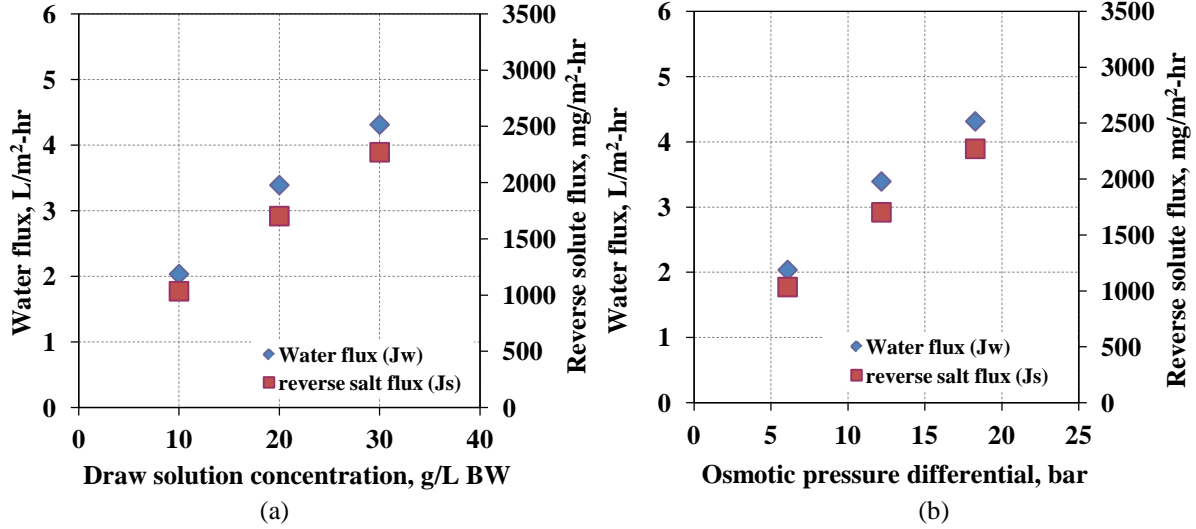
high NaCl concentration in the draw solution and the relatively low NaCl concentration in the feed stream. Membrane integrity tests were conducted to assure that the membrane was not damaged during the experiments. Similar results were obtained from the first integrity test and the second integrity test (end of experiment); this is an indication that the integrity of the membrane was not compromised during the course of the experiments.

**Performance testing of FO elements**

Water fluxes and reverse solute flux for the experiments involving DI feed water and synthetic seawater and brackish water draw solutions are illustrated in Figure 4.7 and Figure 4.8, respectively. In general, water fluxes during the DI-SW experiment were higher than those observed in the DI-BW experiment. This behavior is attributed to the higher concentration of NaCl in seawater compared to brackish water, and a corresponding increase in the osmotic pressure available. Reverse fluxes of solutes from the draw solution into the feed stream were generally higher during the DI-SW experiments.

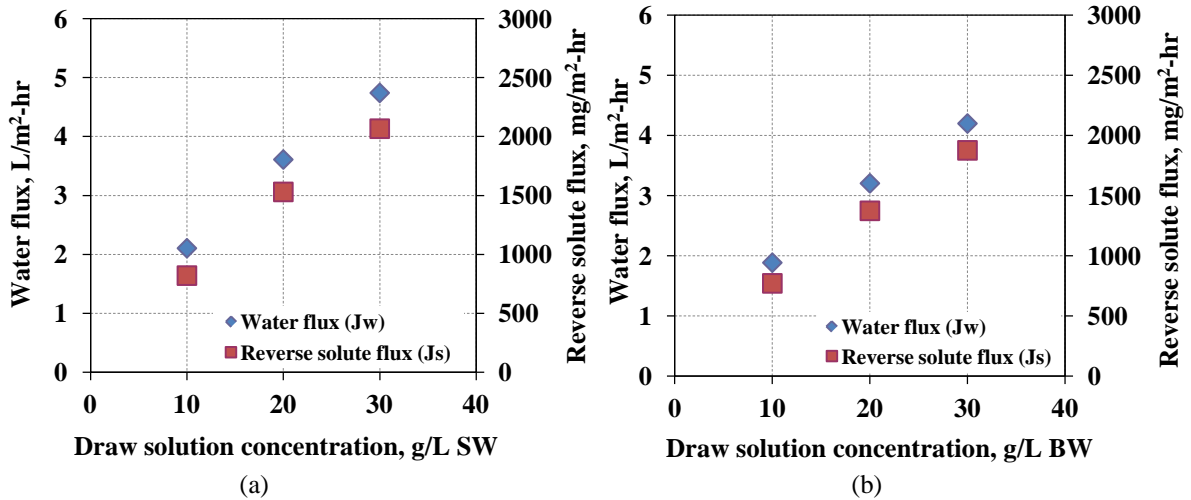


**Figure 4.7** Water flux and reverse solute flux as a function of synthetic seawater (SW) draw solution concentration (a), and osmotic pressure differential (b). Feed water was DI water.  $J_w$  represents water flux, and  $J_s$  represents reverse solute flux.



**Figure 4.8** Water flux and reverse solute flux as a function of synthetic brackish water (BW) draw solution concentration (a), and osmotic pressure differential (b). Feed water was DI water.  $J_w$  represents water flux, and  $J_s$  represents reverse solute flux.

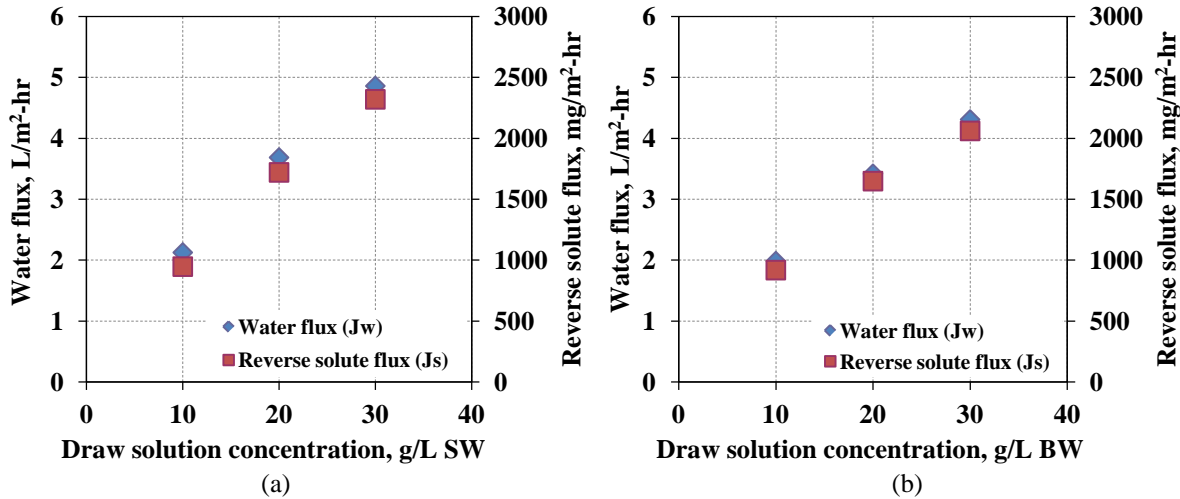
Water fluxes and reverse solute fluxes for the tests involving secondary effluent (SecEff) feed and synthetic seawater and brackish water draw solutions are illustrated in Figure 4.9. The terms SW (seawater) and BW (brackish water) in the following figures represent draw solutions with ionic concentrations in similar ratios as described in Table 4.2 which are characteristic of seawater and brackish water. The total dissolved solids of the solutions are varied to permit the impacts of each solution on FO performance to be evaluated.



**Figure 4.9** Water flux and reverse solute flux as functions of synthetic seawater draw solution concentration (a) and brackish water draw solution concentration (b). The impaired feed water was secondary effluent from a Golden (Colorado) wastewater treatment plant.

Similar to the results of the DI-SW and DI-BW tests, water fluxes observed for the SecEff-SW experiments were also higher (at all draw solution concentrations) than the SecEff-BW tests. The reverse solute fluxes at all concentrations were also higher for the SecEff-SW tests.

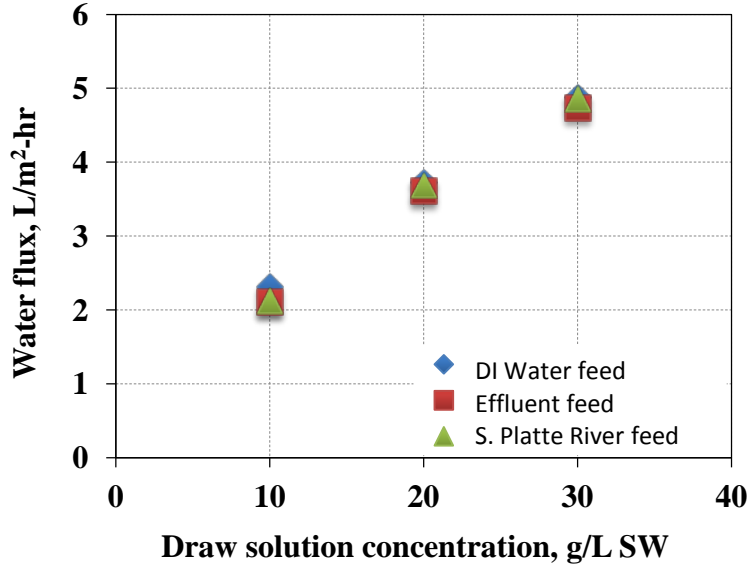
Water fluxes and reverse solute fluxes for the tests involving South Platte River water (SPLT) feed and synthetic seawater and brackish water draw solutions are illustrated in Figure 4.10.



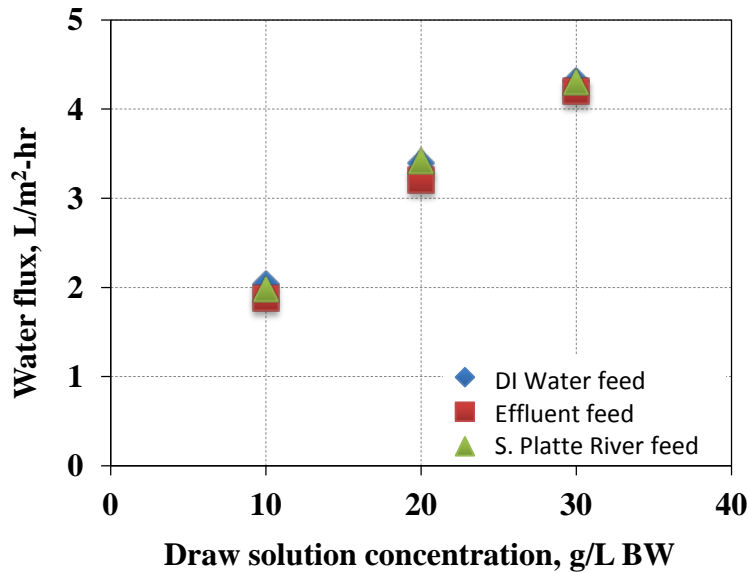
**Figure 4.10** Water flux and reverse solute flux as functions of seawater draw solution concentration (a) and brackish water draw solution concentration (b). Feed water was South Platte River collected close to downtown Denver.

Similar to the trends observed for the experiments involving DI water and secondary effluent feeds, at equal draw solution concentration, water flux and reverse solute flux values were higher for the SPLT-SW experiments than the SPLT-BW experiments. Based on these results, seawater appears to be a better draw solution than brackish water at the same concentration.

Figures 4.11 and 4.12 provide a summary of water fluxes for all feed solutions tested on the bench-scale with seawater and concentrated brackish water draw solutions. Results reveal that feed water quality has minimal effect on the short-term performance of FO. Results in Figures 4.13 and 4.14 revealed that bulk reverse salt diffusion into the feed is almost independent of the type of draw solution used (seawater or brackish water ionic constituents). The concentration gradient is the driving factor in the reverse solute flux.



**Figure 4.11** Water flux as function of seawater draw solution concentration. Feed water samples were DI water, secondary effluent from Golden wastewater treatment plant, and South Platte River water collected close to downtown Denver.



**Figure 4.12** Water flux as function of brackish water draw solution concentration. Feed water samples were DI water, secondary effluent from Golden wastewater treatment plant, and South Platte River water collected close to downtown Denver.

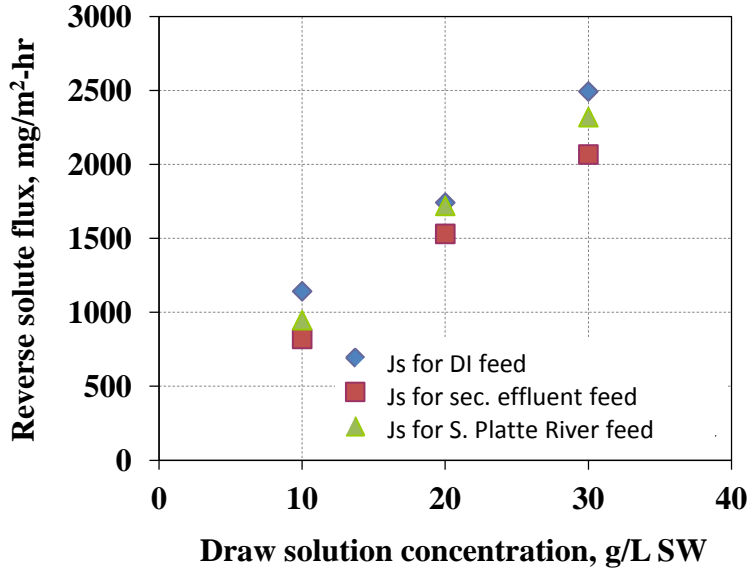


Figure 4.13 Reverse solute flux as function of seawater draw solution concentration. Feed water samples were DI water, secondary effluent from Golden wastewater treatment plant, and South Platte River water.

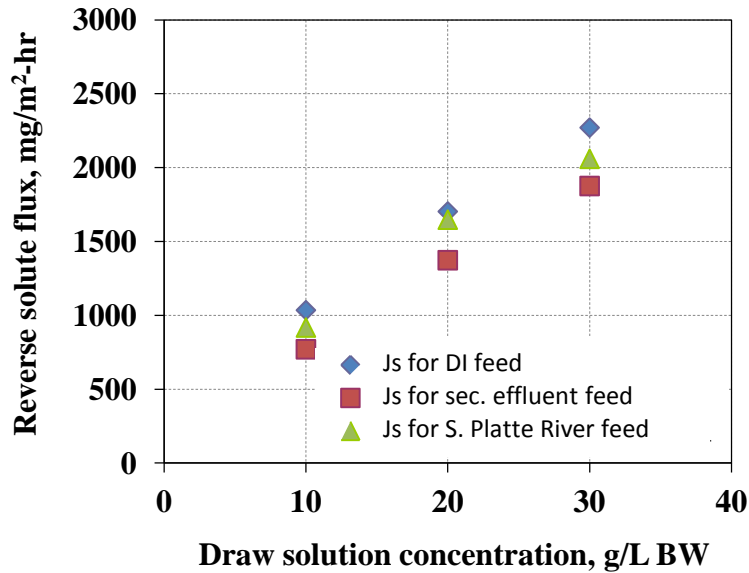


Figure 4.14 Reverse solute flux as function of brackish water draw solution concentration. Feed water samples were DI water, secondary effluent from a wastewater treatment plant, and South Platte River water.

***Ion Analysis for the DI-SW experiments***

***Rejection of dissolved organic matter by FO***

Aromatic organic compounds absorb from the UV region of the electromagnetic spectrum and therefore, the presence of dissolved organic carbon (DOC) in a given sample can be verified by measuring the UV absorbance of the sample. TDS interferences limit the ability to evaluate the rejection for DOC using TOC analyzers. As a surrogate for DOC, UV absorbance was measured to characterize the DOC. Results in Table 4.3 for the secondary effluent feed and brackish water draw solution indicate that the rejection of DOC (represented by UV absorbance) was very high. Similarly with seawater draw solution (Table 4.4), high rejection of DOC was observed. The UVA observed for draw solution samples obtained at the end of the experiments were very close (equal or lower in some cases – in any event within the accuracy of the UVA measurement method) to what was observed in samples taken at the beginning of the experiments, indicating that the water that diffused through the membrane from the feed to the draw solution had very little or no DOC. The accuracy for the UVA measurement was  $\pm 0.004$ .

**Table 4.3 UV Absorbance at 254 nm for Samples Obtained during the Secondary Effluent-BW Experiments**

DS Conc.	Initial Draw sample	Initial Feed sample	Final Draw sample	Final Feed sample	Rejection
(g/L)	UVA	UVA	UVA	UVA	%
10	0.009	0.158	0.012	0.155	98.1
20	0.011	0.156	0.008	0.152	>99
30	0.01	0.167	0.008	0.162	>99

**Table 4.4 UV Absorbance at 254 nm for Samples Obtained during the Secondary Effluent-SW Experiments**

DS Conc.	Initial Draw sample	Initial Feed sample	Final Draw sample	Final Feed sample	Rejection
(g/L)	UVA	UVA	UVA	UVA	%
10	0.007	0.172	0.011	0.171	97.7
20	0.006	0.17	0.008	0.17	98.8
30	0.012	0.171	0.01	0.167	>99

***Rejection of ammonia using secondary effluent as feed***

Ammonia concentration in each sample obtained during the experiments involving secondary effluent was also measured. Results in Table 4.5 for the secondary effluent feed and brackish water draw solution indicate that the rejection of ammonia was lower at 10 g/L BW concentration (approximately 81%) and higher at 20 g/L (approximately 93%) and 30 g/L BW (approximately 91%). Similarly with seawater draw solution (Table 4.6), approximately 62% rejection of ammonia was observed at 10 g/L SW; 73% and 74% ammonia rejection were calculated at 20 g/L and 30 g/L SW, respectively. One plausible explanation for the low rejection of ammonia observed at 10 g/L draw solution (BW and SW) concentration is the fact that the



higher water fluxes observed at 20 g/L and 30 g/L draw solution would result in a corresponding higher dilution of ammonia concentration in the permeate stream.

**Table 4.5 Ammonia Concentration in Samples Obtained during the Secondary Effluent-BW Experiments**

DS Conc.	Initial Draw sample	Initial Feed sample	Final Draw sample	Final Feed sample	Rejection
(g/L)	NH <sub>3</sub> -N (mg/L)	NH <sub>3</sub> -N (mg/L)	NH <sub>3</sub> -N (mg/L)	NH <sub>3</sub> -N (mg/L)	%
10	0	6.65	1.25	6.3	80.7
20	0	3.75	0.35	5.8	92.7
30	0.05	6.75	0.7	8.4	91.4

**Table 4.6 Ammonia Concentration in Samples Obtained during the Secondary Effluent-SW Experiments**

DS Conc.	Initial Draw sample	Initial Feed sample	Final Draw sample	Final Feed sample	Rejection
(g/L)	NH <sub>3</sub> -N (mg/L)	NH <sub>3</sub> -N (mg/L)	NH <sub>3</sub> -N (mg/L)	NH <sub>3</sub> -N (mg/L)	%
10	0	7.45	2.9	7.9	62.2
20	0	8.1	2.25	8.3	72.6
30	0	8.7	2.2	8.4	74.3

***Rejection of ammonia using South Platte River water as feed***

Ammonia concentration in each sample obtained during the experiments involving South Platte River water was also measured. Results in Table 4.7 for the South Platte River feed and brackish water draw solution indicate that there was no rejection of ammonia at 10 g/L and 20 g/L BW concentration. Removal of ammonia was approximately 90% at 30 g/L BW concentration. Similarly with seawater draw solution (Table 4.8), there was no rejection of ammonia at 20 g/L SW concentration. However, 100% removal of ammonia was observed at 10 and 30 g/L SW concentration.

**Table 4.7 Ammonia Concentration in Samples Obtained during the South Platte River-BW Experiments**

DS Conc.	Initial Draw sample	Initial Feed sample	Final Draw sample	Final Feed sample	Rejection
(g/L)	NH <sub>3</sub> -N (mg/L)	NH <sub>3</sub> -N (mg/L)	NH <sub>3</sub> -N (mg/L)	NH <sub>3</sub> -N (mg/L)	%
10	0.25	0.65	ND	0.3	-
20	0.15	ND	0	0.35	-
30	0	0.1	0.05	0.95	90.5

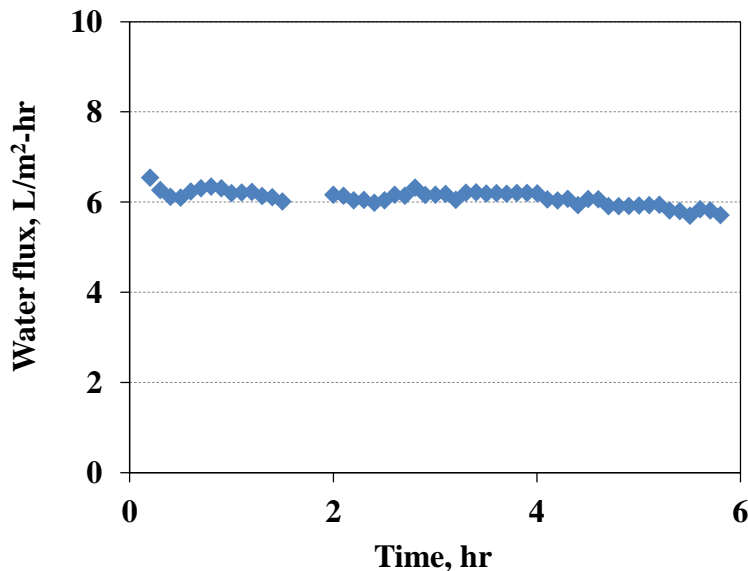
**Table 4.8 Ammonia Concentration in Samples Obtained during the South Platte River-SW Experiments**

DS Conc.	Initial Draw sample	Initial Feed sample	Final Draw sample	Final Feed sample	Rejection
(g/L)	NH <sub>3</sub> -N (mg/L)	NH <sub>3</sub> -N (mg/L)	NH <sub>3</sub> -N (mg/L)	NH <sub>3</sub> -N (mg/L)	%
10	0.1	0.3	0.1	0.05	100.0
20	0.05	0.2	0.3	0.05	-
30	0.25	0.3	0.25	0.1	100.0

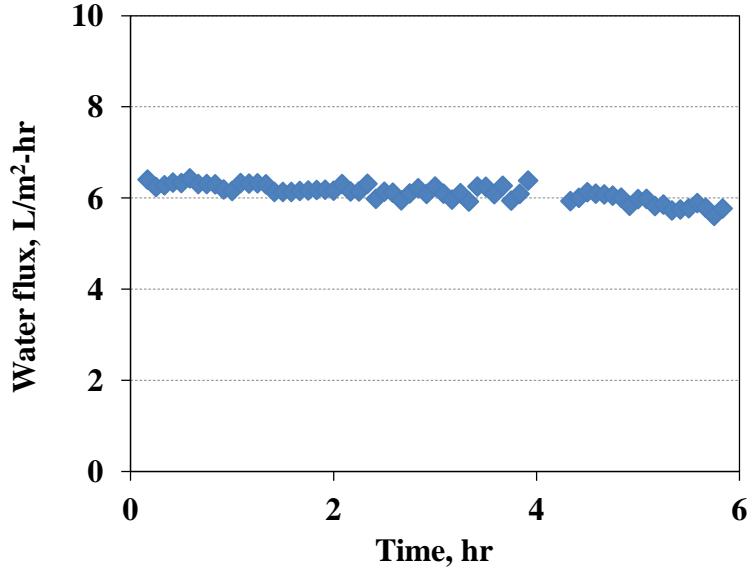
***FO rejection of turbidity and flux effects***

Bench-scale experiments were conducted with synthetic seawater draw solution and a synthetic high turbidity impaired feed solution. The impaired feed solution was prepared using deionized water mixed with different concentrations of bentonite to produce turbidity measurements of 5, 50, and 100 NTU. The main objective of the tests was to evaluate colloidal fouling during FO of impaired water. Previous bench- and pilot-scale studies have demonstrated that colloidal fouling is minimal in FO and mild membrane cleaning restored the performance of the FO membrane (Cath and others, 2006).

Water flux as a function of time for different feed water turbidity is illustrated in Figure 4.15 (50 NTU) and Figure 4.16 (100 NTU). Results indicate that, through 6 hours of testing, very minimal flux decline from colloidal fouling resulted from even very high feed water turbidity.



**Figure 4.15 FO water flux vs. time for the bench-scale testing with 50 NTU feed water turbidity. Water flux declines slowly towards the end of the experiment due to membrane fouling.**



**Figure 4.16** FO water flux vs. time for the bench-scale testing with 100 NTU feed water turbidity. Water flux declines slowly towards the end of the experiment due to membrane fouling.

#### **4.5. Field Pilot Testing of an Integrated Forward Osmosis/Reverse Osmosis Process On Secondary Wastewater**

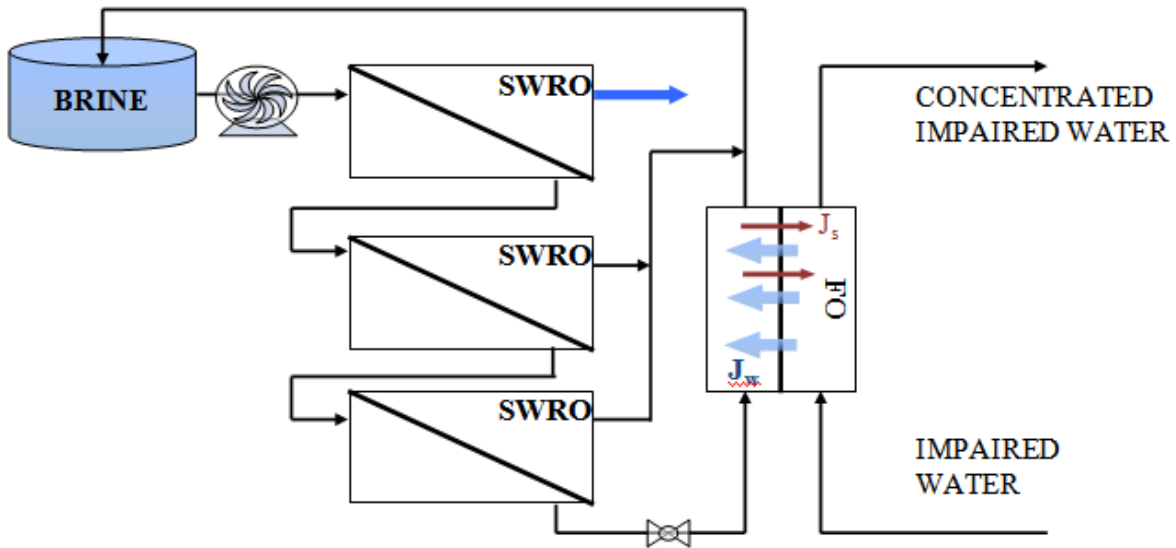
Pilot-scale experiments were conducted at the wastewater treatment research facility at the CSM (<http://aqwatec.mines.edu/research/SBMBR/>). The site supports a 7,000 gal/day hybrid wastewater treatment system combining sequencing batch reactors (SBR) with submerged ultrafiltration membrane bioreactors (MBR) (SBMBR). The Aqua Aerobic<sup>®</sup> MBR demonstration system is designed for small communities or cluster homes, providing an effluent quality that is suitable for onsite reuse. The effluent/filtrate from the SBMBR system was used as the main source of feed water to the FO pilot-system. A photograph of the SBMBR system at CSM is shown in Figure 4.17.



**Figure 4.17** The SBMBR demonstration system at CSM (Golden, CO)

#### **4.5.1. FO-RO Test System**

The pilot-scale FO-RO system consists of a RO sub-system and an FO sub-system. The RO sub-system consists of a three-stage array of 2.5” diameter/40” long seawater RO membrane elements (SW30-2540, Dow Filmtec, Edina, MN) connected in series. These membranes were chosen for their high salt rejection and high operating pressure that enable production of high draw solution concentrations. The three membranes are staged in 1:1:1 array. Depending on the operational settings, the system is capable of producing up to 2 L/min permeate from stage 1 and up to 3 L/min of final concentrate (i.e., draw solution). For the pilot study, permeate from the first stage only was collected as the final permeate of the pilot system. For the pilot study, permeate from the first RO stage only was collected as the final permeate of the pilot system. RO stages 2 and 3 were used to further concentrate the brine for use as the draw solution in the FO process. The permeate from stages 2 and 3 was blended with diluted draw solution from the FO subsystem, and returned to the brine tank (as feed for RO stage 1). This permeate was recirculated, creating a semi-batch process, in order to facilitate measurement and calculation of water flux and solute rejection in the FO subsystem. A schematic drawing of the RO sub-system is illustrated in Figure 4.18.



**Figure 4.18** Schematic drawing of the forward osmosis/reverse osmosis pilot test system

A high-pressure positive displacement pump (HydraCell M03, Wanner Engineering, Inc., Minneapolis, MN) is used in the RO sub-system and is capable of producing high pressure at low and variable flow rates. All high-pressure pipes and fittings are stainless steel while low-pressure pipes/tubing and fittings are polyethylene, PVC, or nylon. All tanks are made of PVC. Photos of the pilot are shown in Figure 4.19.



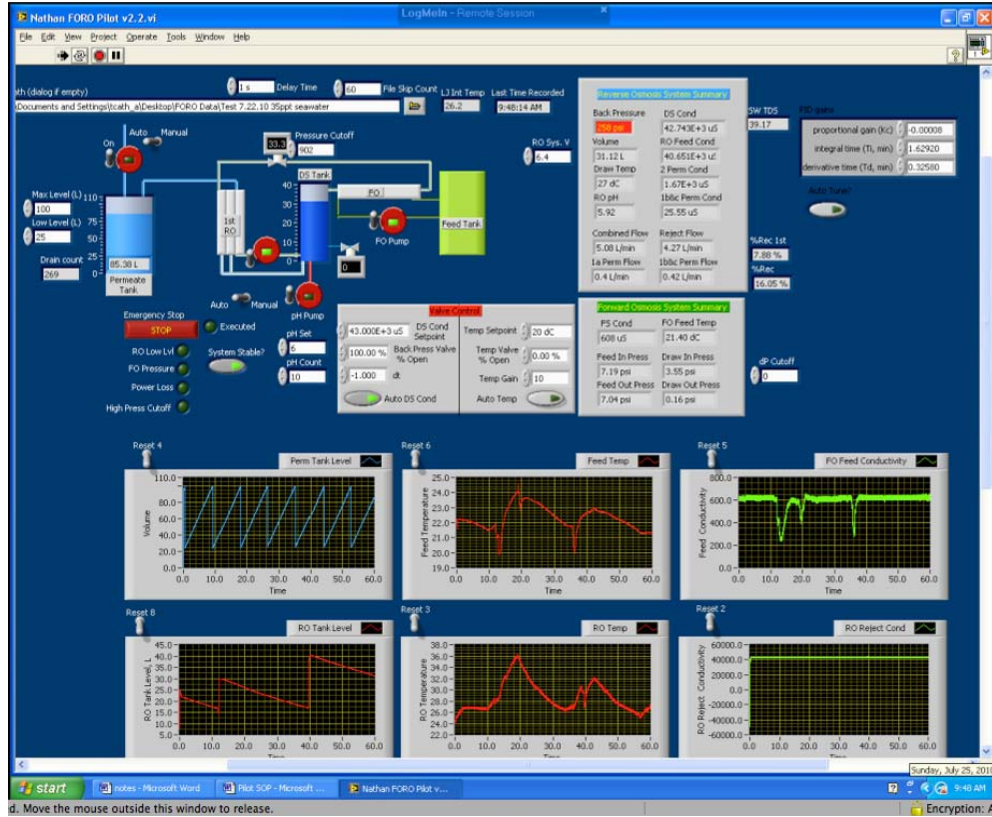
**Figure 4.19** The pilot forward osmosis/reverse osmosis system deployed at CSM Wastewater Research Facility

#### 4.5.2. Supervisory Control and Data Acquisition (SCADA) System

Various sensors are installed in the RO and FO sub-systems that measure stage 1 permeate and final concentrate conductivities, flow rates of all streams, pH, temperatures, and operating pressures. Information from the sensors is continuously recorded by a SCADA system. The operation of the RO sub-system is fully controlled by the SCADA system to produce a stream of draw solution at constant, predetermined TDS concentration. Operating parameters for the pilot-scale RO system are summarized in Table 4.9. A screenshot of the SCADA system during a typical experiment is illustrated in Figure 4.20.

**Table 4.9 Pilot-scale RO Sub-system Operating Parameters**

Parameter	Unit	Value
Feed flow rate	L/min	4.4
First permeate flow rate	L/min	1.0
Concentrate flow rate	L/min	1.0 – 2.4
Average water flux	LMH (gfd)	21.4 (13.1)



**Figure 4.20 Screenshot of the SCADA system during a typical experiment.**

**4.5.3. Pilot Forward Osmosis Membrane Elements**

For the pilot scale study, special spiral-wound membrane elements were acquired from Hydration Technology Innovations (Albany, OR). The total membrane surface area of the FO membrane elements was 1.58 m<sup>2</sup>. The membrane elements contained different feed spacers that provided different levels of fouling control in the feed channels. The elements were used in the laboratory pilot test and were described in detail in Section 4.1 and Table 4.1.

**4.5.4. Pilot System Operation**

The continuous supply of treated wastewater as FO feed at the CSM facility enabled continuous operation of the pilot system for over 1,300 hours (~55 days). The location of the pilot system allowed access to both tertiary effluent as well as lower quality effluent. Lower quality water was achieved by continuous dosing of activated sludge to achieve an FO feed stream with up to 50 mg/L TSS. Average water properties for the tertiary effluent are summarized in Table 4.10.

**Table 4.10 Average Properties of Reclaimed Water Streams used as Feed to the Forward Osmosis Sub-system**

Alkalinity (mg/L-CaCO <sub>3</sub> )	Total P (mg/L-P)	Total N (mg/L-N)	COD (mg/L)	Ortho-P (mg/L-P)	NH <sub>3</sub> (mg/L-N)	NO <sub>3</sub> (mg/L-N)	DOC (mg/L)
39.5	5.5	14.2	10.5	3.4	0.1	13.1	3.2

Effluent water from the SBMBR fed a 300 gallon buffer tank at a rate of approximately 5 gpm and a constant speed rotary vane pump (Procon, Murfreesboro, TN) was used to draw water from the buffer tank and feed the FO sub-system. The pump was operated with a bypass that directed approximately 3.4 gpm feed water to the FO membrane element and the rest returned back to the buffer tank. After passing through the membrane, the feed was wasted.

Synthetic sea salt was dissolved into collected RO permeate to make the brine for the RO sub-system. The brine was concentrated in the RO system and the concentrate became the draw solution for the FO process. A portion of the permeate stream from stage 1 of the RO subsystem was removed from the system; the flow rate of this stream was equivalent to the flowrate of water through the FO membrane. The remaining permeate stream was returned to the RO feed tank; therefore, stage 1 permeate flow rate had to be greater than the flow rate of water diffusing through the FO membrane and had to be maintained without over concentrating the brine. Therefore, the brine was mixed at a concentration of approximately 5 g/L below the desired draw solution concentration and then concentrated in the RO system to the desired draw solution concentration. Water flux through the FO membrane was calculated from the rate of accumulation of RO permeate in the permeate tank. The volume of liquid in the RO system was set to 50 L, of which 40 L was stored in the RO feed tank. The brine was pumped from the brine tank to the RO system, flowing through the RO membrane elements and concentrated to the desired draw solution concentration. The draw solution then flowed through the FO membrane element on the draw solution side of the membrane, was diluted by water diffusing from the feed stream, and returned to the RO feed tank to be concentrated again. Due to reverse diffusion of salt into the feed, the water level in the RO feed tank constantly declined and a batch of approximately 20 L brine was added to the tank every day to maintain osmotic pressure difference as explained in more detail in Section 4.1.1.

**4.5.5. Pilot Experiments**

Pilot-scale experiments were conducted with either tertiary treated feed water or the same stream dosed with activated sludge. All experiments with SBMBR effluent were conducted with a draw solution concentration of approximately 30 g/L or 60 g/L synthetic sea salt. Experiments with SBMBR effluent amended with activated sludge were conducted with draw solution concentrations of 30 g/L sea salt. The FO membrane was replaced once during the testing. Membranes subject to fouling were replaced with new elements containing improved spacer design. Table 4.11 through Table 4.12 present the composition of the SBMBR effluent used as feed and the synthetic SW used in these experiments. Table 4.13 presents the calculated osmotic pressure of the synthetic seawater draw solution.



**Table 4.11 Composition of SBMBR Effluent Used as a Feed for the Pilot Experiments**

Parameter	Value	Units
pH	6.85	-
Alkalinity	50	mg/L CaCO <sub>3</sub>
Ortho-P	4.3	mg-P/L
Total N		mg-N/L
NO <sub>3</sub>	8.5	mg-N/L
Ammonia	0	mg-N/L
UVA 254	0.1097	
COD	16	mg/L
DOC	4.45	mg/L

**Table 4.12 Instant Ocean Synthetic Sea Salt Composition Used for MBR Pilot**

Ion	Instant Ocean	Instant Ocean
	(@ 35 g/L TDS) (mg/L)	(@ 60 g/L TDS) (mg/L)
Chloride (Cl <sup>-</sup> )	19,290	33,069
Sodium (Na <sup>+</sup> )	10,780	18,480
Sulfate (SO <sub>4</sub> <sup>2+</sup> )	2,660	4,560
Magnesium (Mg <sup>2+</sup> )	1,320	2,263
Potassium (K <sup>+</sup> )	420	720
Calcium (Ca <sup>2+</sup> )	400	686
Carbonate/bicarbonate (CO <sub>3</sub> <sup>2-</sup> /HCO <sub>3</sub> <sup>-</sup> )	200	343
Bromide (Br <sup>-</sup> )	56	96
Strontium (Sr <sup>2+</sup> )	8.8	15
Boron (B)	5.6	10
Fluoride (F <sup>-</sup> )	1	2

**Table 4.13 Composition of Synthetic Seawater Used as a Draw Solution for the Process. Osmotic pressure was calculated using OLI Stream Analyzer chemical engineering software package (@35 mg/L).**

DS Concentration (g/L)	Osmotic Pressure (psi)
5	50.8
10	101.5
15	152.3
20	201.6
25	252.4
30	303.1
35	355.3
70	710.1



#### 4.5.6. Sampling

Samples were collected throughout the course of each experiment. For all experiments, 250-mL samples were collected for analysis from the feed, draw solution, and permeate on the first day and every other day thereafter until the experiment was terminated.

#### 4.5.7. Long Term Pilot FO-RO Performance Test

The pilot study generated a diverse and considerable amount of data during consecutive 1300 hours (55 days) of operation. Operating conditions, including draw solution (DS) concentration, DS flow rate, feed total suspended solids (TSS) concentration, and applied feed backpressure (measured at the feed inlet) for specific time intervals during the pilot study are summarized in Table 4.14.

**Table 4.14 Summary of Operating Conditions During the Pilot Study. Letter indices enumerated in the days of operation column are used to correlate operation intervals identified in Figure 4.21.**

Days of Operation	DS Concentration mg/L	DS Flowrate L/min	TSS Conc. mg/L	Applied Pressure on FO feed, psig
0-14 (A)	30,000 ± 600 (est.)	1.6 ± 0.3	0	0
14-21 (B) <sup>1</sup>	54,486 ± 1,341	2.0 ± 0.3	0	0
22-27 (C) <sup>1</sup>	27,550 ± 296	2.3 ± 0.2	0	0
27-31 (D)	29,517 ± 349	2.2 ± 0.1	5.3 ± 1.6	0
31-37 (D) <sup>2</sup>	28,661 ± 1,242	2.4 ± 0.3	15.9 ± 2.0	0
37-39 (E) <sup>2</sup>	29,585 ± 1,343	2.3 ± 0.1	50.2 ± 6.3	0
39-49 (F)	29,585 ± 1,343	2.3 ± 0.1	16.4 ± 2.8	0
50-52 (G)	29,585 ± 1,343	2.3 ± 0.1	16.4 ± 2.8	10
53-55 (G)	29,585 ± 1,343	1.9 ± 0.2	16.4 ± 2.8	20

<sup>1</sup>Chemical cleaning of the RO subsystem performed between intervals B and C.

<sup>2</sup>Feed flow reversal performed on FO spiral-wound membrane for one hour prior to increasing TSS concentration during interval E.

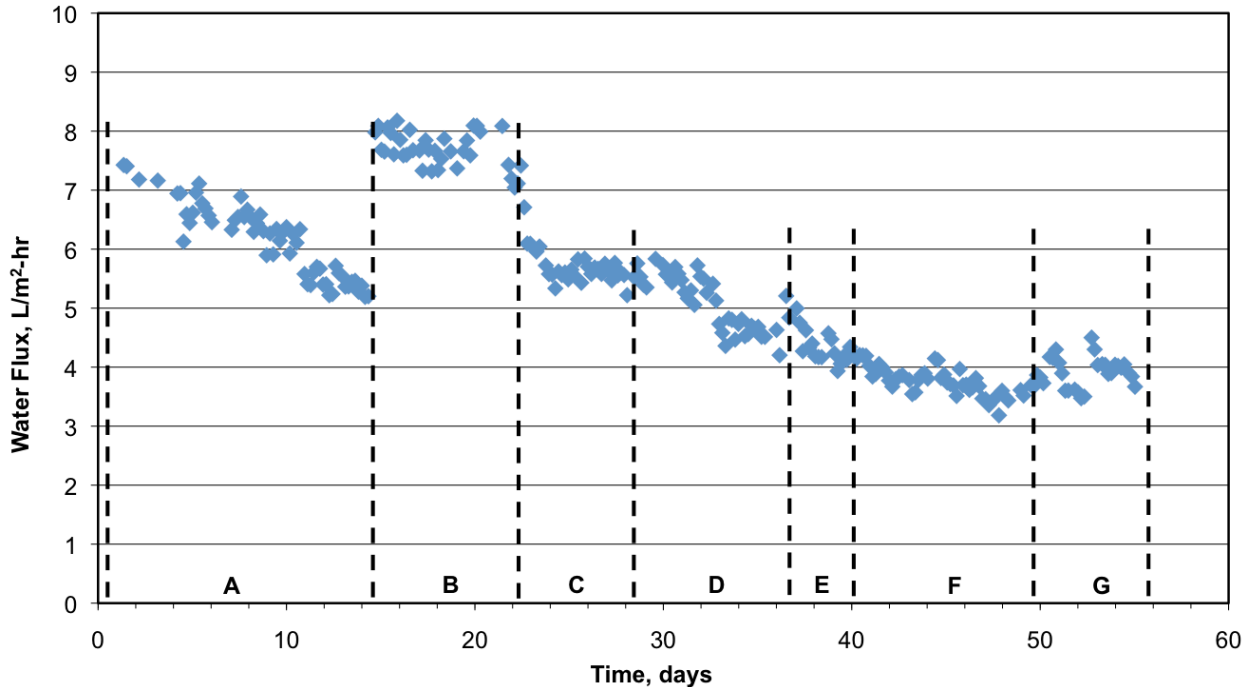
#### 4.5.8. Water Productivity

Water flux was continually monitored during the study. An overview of water flux performance as a function of operation time is presented in Figure 4.21. Letter indices represent operation intervals enumerated in Table 4.14. During 1300 hours (55 days) of continuous operation the hybrid process produced approximately 10,000 L of RO product water and processed more than 900,000 L of wastewater effluent (FO feed).

During interval A, the hybrid process operated approximately 250 hours with SBMBR permeate before reaching a steady state. This likely indicates that a film of organic foulants was accumulating on the membrane surface and reached its optimal thickness (equal rates of foulant deposition and sloughing off) during this time interval. During the last 100 hours of interval A water flux was relatively constant. During interval B the DS concentration was increased about 80% (30,000 to 55,000 mg/L) according to the operations schedule summarized in Table 4.14. Although the driving force for the process (i.e., the osmotic pressure of the DS) was nearly doubled during this interval, water flux did not increase proportionally. This phenomenon is attributed to internal concentration polarization phenomenon. Water flow rate during this interval

is almost constant, and additional flux decline from membrane fouling was not observed. Chemical cleaning of the DS hydraulic system was performed between interval *B* and *C* to remove mineral scale within the RO subsystem. During interval *C*, DS concentration was reduced to approximately 30,000 mg/L sea salt. An equivalent water flux to the end of interval *A* was observed during this time period. During interval *D* activated sludge was dosed into the FO feed stream to achieve 5 mg/L and later 16 mg/L TSS (Table 4.14), and a decrease in water flux was observed.

Immediately following interval *D*, the direction of feed flow within the FO membrane module was reversed for one hour in an attempt to remove settled activated sludge from the feed channels of the membrane module. During interval *E*, immediately following feed flow reversal, the dosing rate of activated sludge solids was increased to achieve approximately 50 mg/L total suspended solids (TSS) in the FO feed, and additional water flux decline was observed. During interval *F*, the dosing of activated sludge was reduced to approximately 15 mg/L TSS in the FO feed for 10 days of continuous operation. And lastly, FO feed pressure was increased during interval *G*. Data in Figure 4.21 indicate that additional pressure negligibly increased the flux.



**Figure 4.21** Compiled water flux data from upgraded pilot-scale system. Letters marked on graph indicate intervals of interest during the pilot study and are defined in Table 4.14.

## **5. System and Process Modeling For Forward Osmosis**

In the TWDB assessment of FO, the specific pairing of desalination plant concentrate and treated wastewater treatment streams is of interest. This Chapter reports on Task 4 of the project scope. The purpose of Task 4 was to develop process models with which to better understand the impact of variables on the performance of the FO membrane elements.

This chapter is structured as follows:

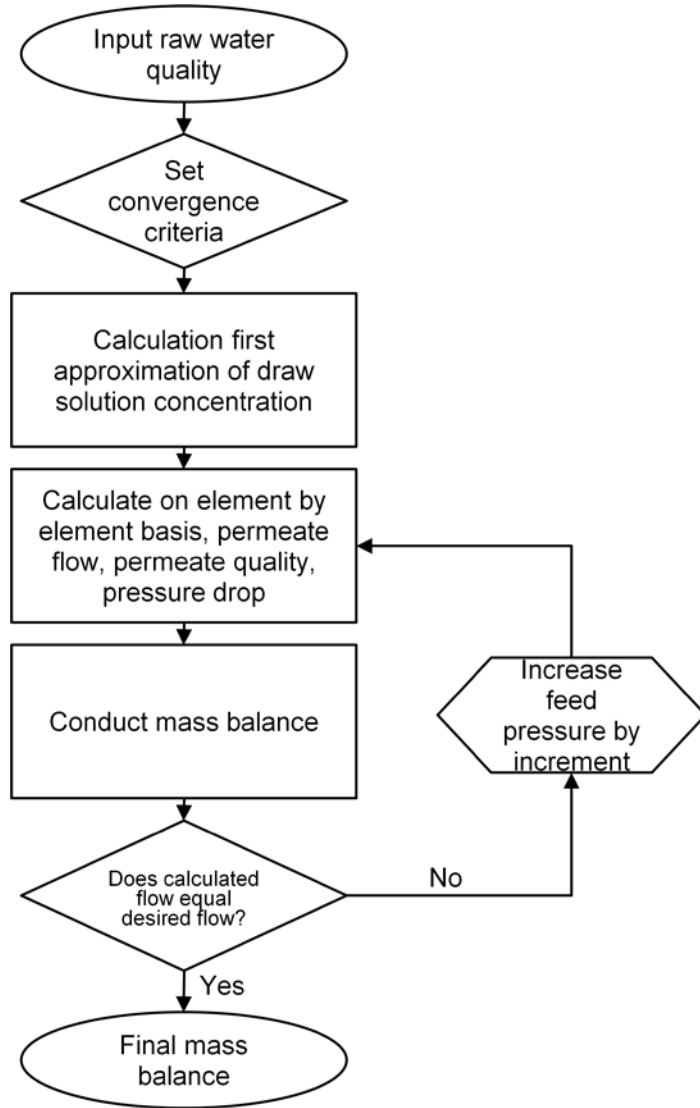
- Development of a FO model for a full-scale system
- Development of a RO model
- Validation of the RO model

### **5.1. Forward Osmosis Model Development**

Currently, most small FO applications utilize a semi-batch approach similar to the flow schematic for the bench-scale apparatus described in Chapter 3. While industrial applications may prefer this approach, as is the case for many industrial RO applications, implementation at a large municipal plant would likely require multiple elements installed into a single pressure vessel, with staging required to obtain very high recovery values without recirculation.

The development of FO module elements is not yet mature. Significant opportunities for the optimization of element design, hydraulics, etc., currently exist within the field. Currently, the high differential pressure within the elements limits the system to two elements in series. Future developments may alter the configurations envisioned within this study. Personal communication with HTI indicates that development work is ongoing and will result in the ability to stage more membranes. It was assumed that up to 8 elements can be installed in a stage for the purposes of the modeling effort. Additionally, complete characterization of the FO element performance was not achieved in the testing; as a result, a number of assumptions need to be made, and validation of the model against experimental data is not possible.

A simplified FO performance model was developed using an empirical flux versus draw solution osmotic pressure data developed during pilot testing. The model assumes counter-current flow through the membrane element array, and calculates the flux and recovery based upon the initial impaired feed solution flow, pressure, and concentration and the initial draw solution flow, pressure and concentration. Figure 5.1 shows the basic algorithm used for FO model calculations, which were implemented in a spreadsheet format. Brute force iterations are utilized within the model.



**Figure 5.1 Calculation Schematic for Forward Osmosis Model**

Data collected during the pilot study were used as the major input into the model, including the values contained in Table 5.1.

**Table 5.1 FO Element Membrane Performance Parameters**

Parameter	Value
Nominal Area <sup>a</sup> (ft <sup>2</sup> )	32.3
Pressure Drop <sup>b</sup> (psi)	45 – 60
Flux (gfd)	3.53
Recovery (%)	4

Notes:

<sup>a</sup>Nominal area is based upon estimates following autopsy of typical membrane elements.

<sup>b</sup>The feed-concentrate pressure drop was measured to be approximately 45 psi.

Figure 5.2 displays the model developed and its output.

		Element Number								
		1	2	3	4	5	6	7	8	
Number of Online Trains	#	1								
Number of Stand-by Trains	#	0								
Number of Vessels per Train	#	50								
Number of Membranes per Vessel	#	6								
Membrane Area per Train	m <sup>2</sup>	1200								
Total Membrane Area	m <sup>2</sup>	1200								
Flow per Train	lpm	125								
Flow per Vessel	lpm	2.5								
		Draw Solution Concentration				mg/L	60,000			
		Feed Stream Concentration				mg/L	500			
		Inlet Pressure, Draw Solution				psi	20			
		Inlet Pressure, Feed Solution				psi	15			
		Flow per Train				lpm	300			
		Flow per Vessel				lpm	6			
<b>Counter-Current Flow</b>										
<b>Per Vessel Balance</b>										
<b>Draw Solution Balances</b>		→								
Inlet Flow	lpm	2.5	3.31	3.91	4.42	4.88	5.29	5.66	6.02	
Inlet Concentration	mg/L	60,000	45,373	38,333	33,899	30,758	28,375	26,483	24,932	
Inlet Osmotic Pressure	psi	765.9	579.2	489.3	432.7	392.6	362.2	338.0	318.3	
Inlet Pressure	psi	20	19.95	19.88	19.81	19.72	19.63	19.53	19.42	
Outlet Flow	lpm	3.31	3.91	4.42	4.88	5.29	5.66	6.02	6.35	
Outlet Concentration	mg/L	45372.5	38333.1	33899.4	30757.9	28374.5	26482.8	24932.1	23629.9	
Outlet Osmotic Pressure	psi	579.2	489.3	432.7	392.6	362.2	338.0	318.3	301.6	
Outlet Pressure	psi	19.95	19.88	19.81	19.72	19.63	19.53	19.42	19.31	
Pressure Drop	psi	0.0527	0.0655	0.0757	0.0844	0.0922	0.0994	0.1060	0.1122	
<b>Feed Solution Balances</b>		←								
Inlet Flow	lpm	2.96	3.57	4.08	4.53	4.94	5.32	5.67	6	
Inlet Concentration	mg/L	1014.2	841.5	735.8	662.4	607.5	564.3	529.2	500	
Inlet Osmotic Pressure	psi	12.9	10.7	9.4	8.5	7.8	7.2	6.8	6.4	
Inlet Pressure	psi	14.41	14.47	14.54	14.62	14.70	14.79	14.89	15	
Outlet Flow	lpm	2.15	2.96	3.57	4.08	4.53	4.94	5.32	5.67	
Outlet Concentration	g/L	1394.0	1014.2	841.5	735.8	662.4	607.5	564.3	529.2	
Outlet Osmotic Pressure	psi	17.8	12.9	10.7	9.4	8.5	7.8	7.2	6.8	
Outlet Pressure	psi	14.36	14.41	14.47	14.54	14.62	14.70	14.79	14.89	
Pressure Drop	psi	0.0464	0.0592	0.0694	0.0781	0.0859	0.0931	0.0997	0.1059	
Concentration Polarization										
Driving Force	psi	747.3	562.9	474.5	419.0	379.8	350.1	326.6	307.3	
Solvent Transfer	lpm	0.8060	0.6071	0.5118	0.4519	0.4096	0.3776	0.3523	0.3316	
Flux	lmh	16.12	12.14	10.24	9.04	8.19	7.55	7.05	6.63	
Recovery	%	0.27	0.17	0.13	0.10	0.08	0.07	0.06	0.06	
Flow Recovered	lpm	0.81	1.41	1.92	2.38	2.79	3.16	3.52	3.85	
Water Recovered	lpm	3.84789								
Average Flux	lmh	3.61972								
Total Recovery	%	64%								

Figure 5.2 FO Model Output Parameters and Structure

5.1.1. Model Limitations

- The model was developed for application as a staged, continuous flow system. [It cannot model the experiments discussed previously because the model is not configured to model semi-batch conditions.]
- The model is configured to evaluate only NaCl rejection. Rejection of other salts or compounds is not incorporated into the model.

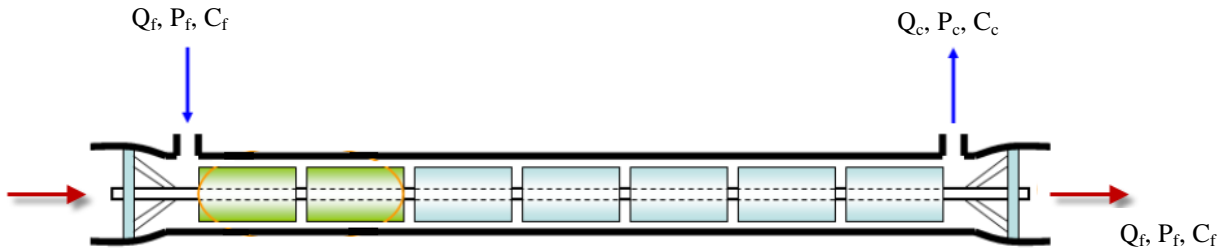
- The existing membrane is installed with only two elements in series due to high draw solution pressure drop, configured for counter-current flow. Currently, very high-pressure drops exist on the draw solution side of the membrane, impacting the hydraulic design of a system. Consequently, validation of the model isn't possible against the experimental data at this time.

## 5.2. Seawater Reverse Osmosis Model Development

In previous work conducted by CH2M HILL, a simplified RO projection tool was developed to permit evaluation of new technologies. In a simple spreadsheet, using minimal Visual Basic code, the model can be readily configured for any number of new processes, simply by altering the mass balance of the system.

The model utilizes element-by-element mass balance calculations for a single pressure vessel (Figure 5.3) to simulate performance of a system with a user configurable number of pressure vessels containing a user configurable number of elements. The basic calculation algorithm was shown previously in Figure 5.1. The model, as currently configured, is limited to use on sodium chloride solutions, which is satisfactory for the preliminary screening study conducted as part of this work, particularly given that sodium and chloride ions are the main constituents of seawater.

In the initial validation of the CH2M HILL model, projection data from a Toray TM820-400 membrane element were utilized. Temperature correction factors, mass transfer coefficients and pressure drop data were incorporated into the model from Toray's RO design and performance software program, TorayRO (Version 2.032).<sup>4</sup>



**Figure 5.3 Typical RO Pressure Vessel Schematic Using Hybrid Membrane Configuration<sup>5</sup>**

The model input parameters are similar to most projection programs produced by manufacturers. A graphic of the input is shown in Figure 5.4. To conduct a projection, the user presses the solve button to alter the feed pressure until the permeate flow produced equals the desired permeate flow.

<sup>4</sup> Proprietary RO projection software distributed by Toray.

<sup>5</sup> Increasingly, the use of different models of membrane in a single pressure vessel is being adopted. By installing high rejection membranes in the lead elements, and lower rejection, higher productivity elements in the later positions, it is possible to produce better flux balancing within the vessel, with a slight reduction in energy consumption. The green and blue elements of the figure (if viewed in color) represent the higher and lower rejection elements, respectively.

User Entered Data					
Permeate Flow (system)	qpm	33	Permeate Flow	qpd	47,520
Target Recovery	%	40%	Flux	qfd	9.0
Temperature	°C	18	Form Q (spec) qpm		33
Fouling Factor		0.85	Form Q (calc) qpm		32.1
Feed Water TDS	mg/L	35,000	Delta		0.9
Number of Vessels	-	2	Run #		
Number of Elements/FV	-	6	Project Name	TWDB Model	
Supply Press. [pS]	pri	15	Date	23-Aug-11	
Permeate Pressure	pri	0			
Discharge Press. [pD]	pri	0			

Concentr					
R	liter-atm/mal-Doqt	0.08205	TCF - Water	-	0.76
			TCF - Salt	-	0.71
			TCF - CP	-	0.80
			System Feed Flow	qpm	82.50
			Press Vessel Feed Flow	qpm	41.25
			First Element Feed	qpm	41.25
Sodium	q/mal	22.98977	Actual Recovery		38.87%
Chloride	Valence	35.453	Total Feed	qpm	123.75

First Guess				
Feed Pressure	priq	878.5	atm	59.8

		Element						
		1	2	3	4	5	6	
TDS (inlet)	mg/L	-	35,000	38,374	42,003	45,817	49,713	53,569
Sodium	mg/L	-	13,768	15,095	16,523	18,023	19,556	21,072
Chloride	mg/L	-	21,232	23,279	25,480	27,794	30,158	32,496
TDS (outlet)	mg/L	-	38,374	42,003	45,817	49,713	53,569	57,251
Sodium	mg/L	-	15,095	16,523	18,023	19,556	21,072	22,521
Chloride	mg/L	-	23,279	25,480	27,794	30,158	32,496	34,730
TDS (Permeate)	mg/L	-	76	92	114	143	181	233
Sodium	mg/L	-	30	36	45	56	71	92
Chloride	mg/L	-	46	56	69	87	110	141
Inlet Flow (per vessel bar)	qpm	-	41.3	37.6	34.4	31.5	29.0	27.0
Outlet Flow (per vessel bar)	qpm	-	37.6	34.4	31.5	29.0	27.0	25.2
Avg Flow (per vessel bar)	qpm	-	39.4	36.0	32.9	30.3	28.0	26.1
Permeate Flow (per vessel bar)	cm <sup>3</sup> /hr	-	228.8	205.1	180.5	155.8	131.9	109.4
	qpm	-	3.6	3.3	2.9	2.5	2.1	1.7
Feed Pressure	pri	-	879	876	874	872	870	869
Outlet Pressure	pri	-	876	874	872	870	869	867
Pressure Drop	pri	-	2.44	2.17	1.94	1.74	1.57	1.44
Permeate Pressure	pri	-	0	0	0	0	0	0
Not Driving Pressure	pri	-	428.2	383.7	337.7	291.6	246.7	204.6
	atm	-	29.1	26.1	23.0	19.8	16.8	13.9
Flux	qfd	-	11.9	10.6	9.4	8.1	6.8	5.7
Recovery	%	-	8.8%	8.6%	8.3%	7.8%	7.2%	6.4%
Avg Concentration (to calculate F	mg/L	-	36,661	40,161	43,882	47,739	51,617	55,389
Conc Polarization	-	-	1.10	1.09	1.09	1.08	1.08	1.07
Permeate Conc	mg/L	-	75.6	92.3	114.1	143.0	181.4	232.7
Osmotic Pressure, Conc - Food	pri	-	390.4	428.0	468.5	511.0	554.5	597.5
Osmotic Pressure, Conc - Bulk	pri	-	428.0	468.5	511.0	554.5	597.5	638.5
Osmotic Pressure, Form	pri	-	0.8	1.0	1.3	1.6	2.0	2.6
Avg Osmotic Pressure - Membrane	pri	-	448.2	490.2	533.9	577.9	620.7	660.7
Total Permeate Flow (system)	qpm	-	32.1					
System Recovery	%	-	38.9%					
Permeate Quality (est)	mg/L	-	127.0					

Figure 5.4 Example of RO Model Input

5.2.1. Validation of SWRO Model

Following adaptation of the model to utilize the TM820-400 SWRO element, CH2M HILL performed a series of runs (commonly called “projections”) utilizing both the TorayRO model and the CH2M HILL model. The purpose of these runs was to validate the CH2M HILL model relative to the performance of the Toray RO model, in terms of feed pressure, permeate TDS, and differential pressure (dP). Table 5.2 summarizes the twelve validation runs that were performed and the associated output from each. Validation runs for the model were performed at fluxes ranging from 6 gfd to 10 gfd, at two temperatures bounding the range of typical offshore seawater temperatures (8 and 14 °C) and two different recoveries, 40 and 50 percent. A feed solution concentration of 35,000 mg/L NaCl was used for the modeling effort.

**Table 5.2 Validation Cases for CH2M HILL Model Utilizing TORAY TM820-400 Membrane Element**

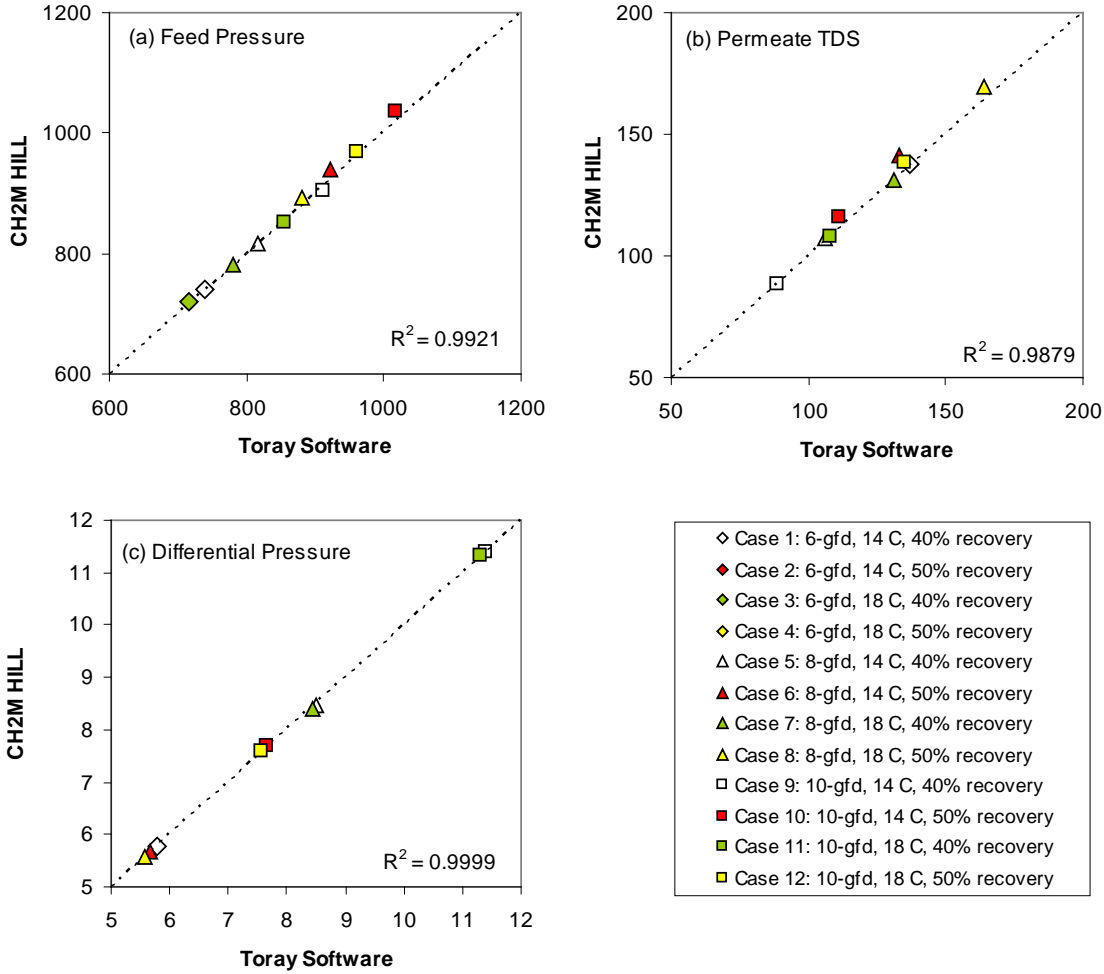
Case	Flux (gfd)	Temp (°C)	Recovery (%)	Toray RO Model			CH2M HILL Model		
				Feed Pressure (psig)	Permeate TDS (mg/L)	dP <sup>a</sup> (psid)	Feed Pressure (psig)	Permeate TDS (mg/L)	dP (psid)
1	6	14	40%	740	137	5.79	741	138	5.77
2	6	14	50%	Did not converge.			857	184	3.81
3	6	18	40%	715	170	5.74	720	170	5.71
4	6	18	50%	Did not converge.			831	223	3.74
5	8	14	40%	817	106	8.49	816	107	8.48
6	8	14	50%	921	133	5.65	939	141	5.67
7	8	18	40%	780	131	8.43	781	131	8.41
8	8	18	50%	880	164	5.57	894	169	5.58
9	10	14	40%	911	89	11.4	903	89	11.40
10	10	14	50%	1017	111	7.65	1037	116	7.69
11	10	18	40%	855	108	11.3	852	108	11.32
12	10	18	50%	960	135	7.56	970	138	7.59

<sup>a</sup> Vessel differential pressure

Figure 5.5 graphically illustrates the match between the results of the two models for the three performance parameters: (a) feed pressure, (b) permeate TDS, and (c) vessel feed-concentrate differential pressure. In all twelve runs, results from the CH2M HILL model correlated to a very high degree with those of the Toray model (i.e., a very high degree of statistical correlation or fit). Correlation coefficient (R<sup>2</sup>) values for each of the three parameters were greater than 0.9879, where a value of 1.0 signifies a perfect match or correlation.<sup>6</sup> Figure 5.5 illustrates that the CH2M HILL model is quite accurate when compared to the Toray model, with the CH2M HILL model slightly under-predicting concentration polarization. Based upon the results of the validation runs for a conventional seawater RO configuration, it was concluded that the CH2M HILL model provides accurate results for the TM820-400 within the range of variables outlined in Table 5.2.

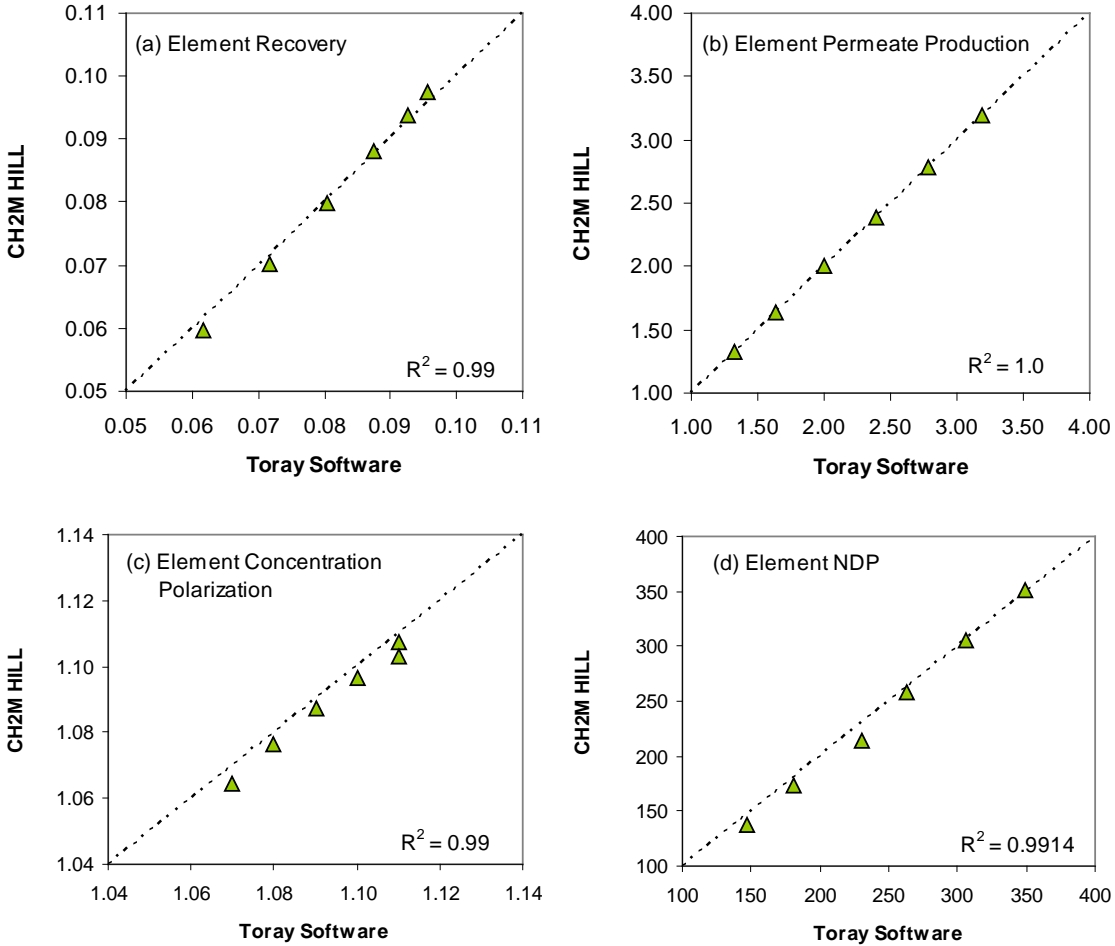
<sup>6</sup> With the graphical comparison method shown in Figure 5.5, a perfect match or correlation between the results of the two models would result in all values falling exactly on the 45-degree line. The farther a value falls to either side of the line indicates the less correlated the result of the CH2M HILL model is to the result of the Toray RO model.





**Figure 5.5 Validation of CH2M HILL model utilizing TORAY TM820-400 Membrane Element**

Additional comparisons between the two models were performed on an element-by-element basis for Case 7. For the case examined, utilizing a flux of 8 gfd, temperature of 18°C and recovery of 40 percent, the CH2M HILL model provides nearly identical results to the Toray RO model and a very high degree of correlation ( $R^2$  values ranging from 0.99 to 1.0). Graphical representation of the degree of correlation on an element-by-element basis for the parameters of (a) recovery, (b) permeate production, (c) concentration polarization and (d) net driving pressure (NDP) are shown in Figure 5.6.

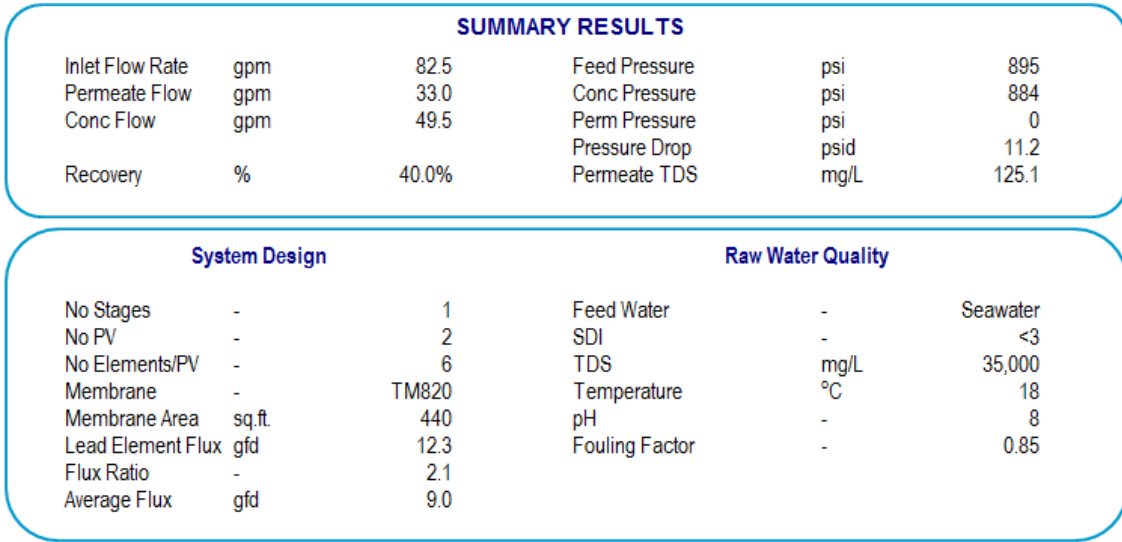


**Figure 5.6 Validation of CH2M HILL model utilizing TORAY TM820-400 Membrane Element**

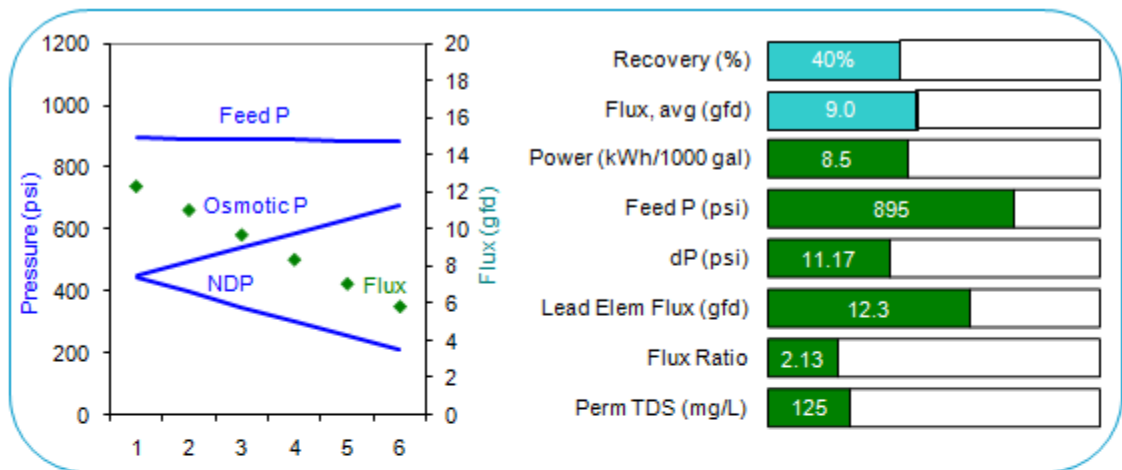
Based upon the validation efforts, the CH2M HILL model accurately reproduces the TorayRO results for the TM820-400 membrane element under the range of conditions tested. Based on these results, it can be concluded that the CH2M HILL model is an appropriate model to use in screening evaluations of the coupled FO-RO process.

**5.2.2. Model Output**

Typical model input data for the SWRO model are shown in Figure 5.7. Based upon this input scenario, the output is shown in Figure 5.8.



**Figure 5.7 Typical SWRO Model Input**



		Element					
		1	2	3	4	5	6
Feed TDS	mg/L	35,000	38,499	42,283	46,279	50,377	54,439
Feed P	psi	895	893	890	888	887	885
dP	psi	2.44	2.16	1.92	1.71	1.54	1.41
Osmotic P	psi	390.4	429.4	471.6	516.2	561.9	607.2
NDP	psi	442.6	396.2	348.1	299.6	252.4	208.2
Feed Q	gpm	41.3	37.5	34.1	31.2	28.7	26.5
Conc Q	gpm	37.5	34.1	31.2	28.7	26.5	24.8
Perm Q	gpm	3.7	3.4	2.9	2.5	2.1	1.8
Perm TDS	mg/L	73.5	90.1	112.0	141.3	180.5	233.3
Flux	gfd	12.3	11.0	9.6	8.3	7.0	5.8
Recovery	%	9.1%	8.9%	8.6%	8.1%	7.5%	6.6%
CP	-	1.10	1.10	1.09	1.09	1.08	1.07

**Figure 5.8 Typical SWRO Model Output (based on previous input data)**

### **5.2.3. Model Limitations**

The model is currently configured to only model rejection of NaCl. Although this is not accurate for actual applications, it is sufficient for determining process trends.

### **5.3. Summary**

Spreadsheet-based Excel project models were developed to permit the mass balance to be determined for a combined FO-RO system. Used in conjunction with the cost model developed in Chapter 6, the economic feasibility of the FO-RO process can be determined.

## **6. Development of a Forward Osmosis/Reverse Osmosis System Cost Model**

In order to examine the feasibility of a hybrid FO-RO process, an economic model is required. Based on the information presented in Sections 3 and 4 of this report, a cost model was developed to allow readers to calculate order of magnitude construction and operations and maintenance (O&M) costs for forward osmosis systems. The development of a cost model fulfills Task 5 of the scope of work. The cost models represent a Class 5 cost estimate using the guidelines established by the Association for the Advancement of Cost Engineering (AACE) and represent a +50%;-30% level of accuracy.

The cost models for the FO process and the RO reconcentration process were separately developed intentionally. The purpose is to allow users to provide an initial estimate for FO separately in waste volume reduction applications are specifically examined. To develop the cost models, and evaluate the economic feasibility of the FO-RO model, the following subtasks were completed:

- Development of a model process flow diagram for a full-scale FO system
- Development of a typical layout for a full-scale FO system
- Development of a construction cost model based upon the process flow diagram and layout
- Development of an operating cost model for the FO process
- Development of a model process flow diagram for a full-scale RO draw solution reconcentration system
- Development of a typical layout for a full-scale RO draw solution reconcentration system
- Development of a construction cost model based upon the process flow diagram and layout for a draw solution reconcentration
- Development of an operating cost model for the RO draw solution reconcentration process
- Comparison of costs versus a conventional advanced wastewater reuse facility

### **6.1. Forward Osmosis Cost Model Development**

The major steps taken in the development of the FO cost model include:

- Development of a model process flow diagram for a full-scale FO system
- Development of a typical lay-out for a full-scale FO system
- Development of a construction cost model based upon the process flow diagram and lay-out
- Development of an operating cost model for the FO process

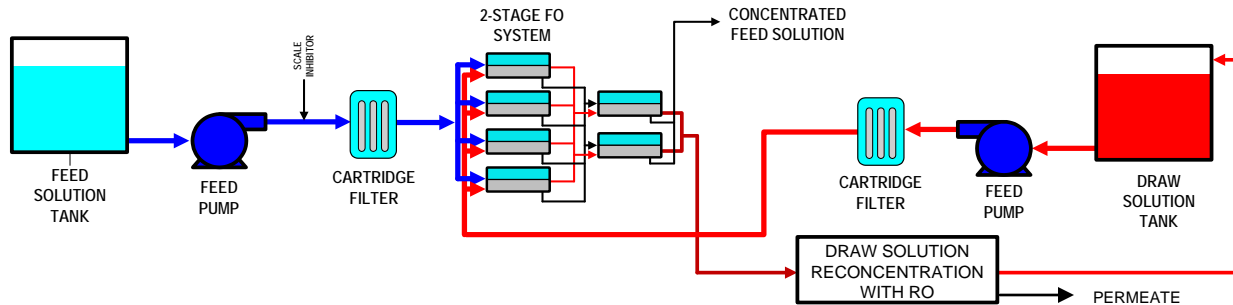
Each of these steps are described separately below.

#### ***6.1.1. Development of a FO Process Flow Schematic***

Prior to developing a detailed cost model of forward osmosis, the general process flow diagram, major equipment list must be prepared and major constraints and criteria identified. While most FO applications to date are utilizing a semi-batch recirculation process due to the early element development stage, it is assumed that a future commercially available system will utilize a staged

continuous flow approach similar to that used in RO applications – but may maintain the ability for recirculation currently in practice.

A schematic process flow diagram of a FO system is presented in Figure 6.1.



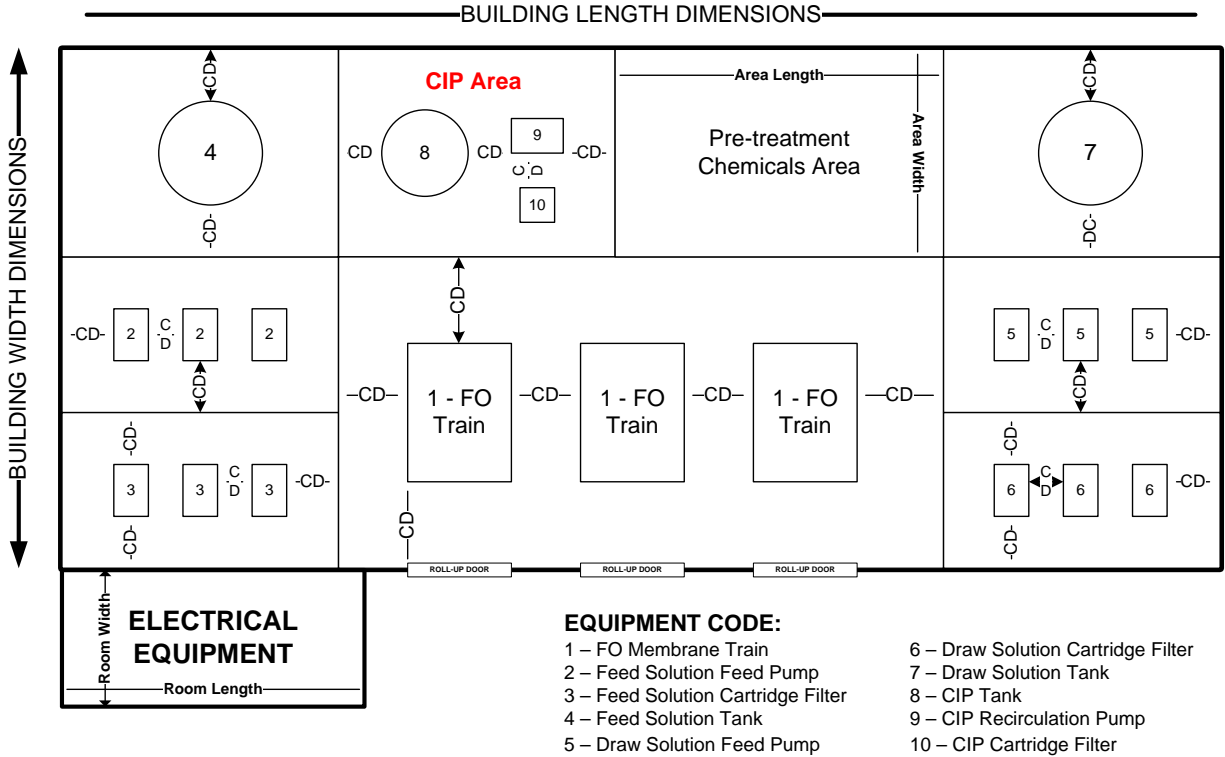
**Figure 6.1 Schematic Process Flow Diagram for Forward Osmosis**

The conceptualized FO system shown in Figure 6.1 is composed of the following subsystems:

- Feed solution tank and pumping system – for collecting and pumping the impaired feed water into the feed side of the FO elements
- Feed solution pretreatment chemicals – provision has been made for the dosing of acid or scale inhibitor in minimize impacts of scale and biological growth in the FO system
- Feed solution cartridge filters – cartridge filters are provided to protect the membrane elements from feed spacer damage due to particulate matter in the feedwater.
- FO skids – While current HTI systems such as the “Green Machine” utilize vertical module configurations, it is assumed that future developments will permit horizontal installation of multiple elements into a single pressure vessel.
- Draw solution tank and pumping system – for make-up, collection and circulation of draw solution into the FO membrane
- Draw solution cartridge filters – to protect the inside of the membrane from particle damage
- Draw solution re-concentration system (RO system including cartridge filters, pretreatment chemicals and RO skids) – to provide reconstitution of the draw solution and to restore the performance of the membrane elements

### 6.1.2. Development of FO Building Layout

CH2M HILL developed a conceptualized layout for a large-scale FO installation based upon the process flow diagram, equipment list and typical RO membrane installations. Figure 6.2 presents the proposed layout of the forward osmosis system. This layout is parametric in nature (can be scaled up and down) and was utilized to calculate building footprint as well as quantity take-offs for components of a forward osmosis system ranging in size from 0.5 to 15-mgd of recovered water from the feed solution.



**Figure 6.2 Proposed Building Layout for Forward Osmosis**

The layout presented in Figure 6.2 served as the basis to calculate quantity take-offs for site-work (excavation, imported structural backfill, native backfill and hauling excess), concrete, building footprint, metals, doors and windows, equipment, instrumentation and controls, conveying systems, mechanical (piping, valves, fittings) and electrical. A custom parametric cost estimating module, based on CH2M HILL’s proprietary cost estimating platform, was developed for the proposed FO process.

Table 6.1 presents the design criteria utilized for setting up the FO systems in the cost model.

**Table 6.1 FO System Design Criteria**

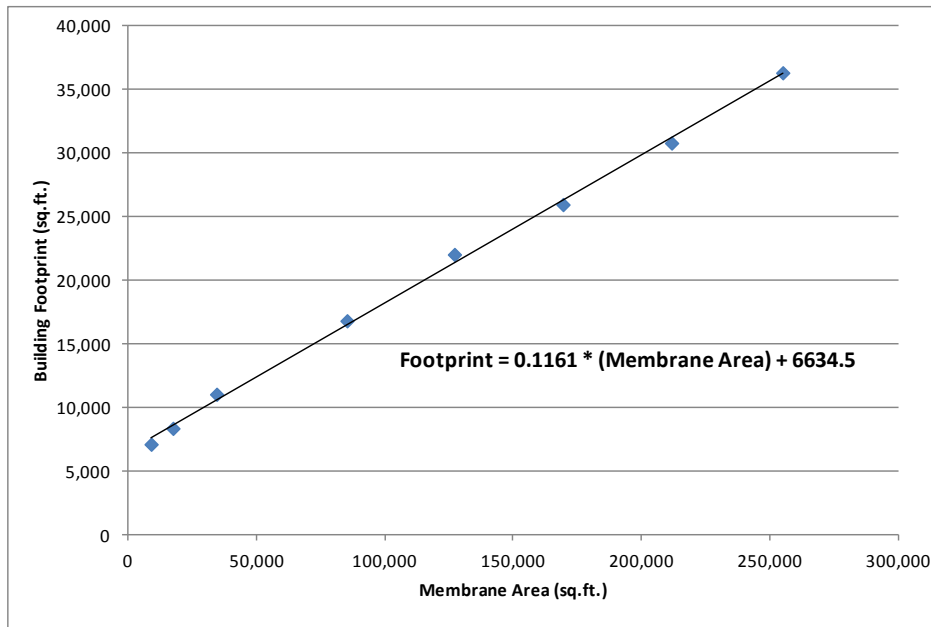
Parameter	Units	Value
FO Flux	gfd	5.5
FO Feed Solution Pressure	psi	58.0
FO Draw Solution Pressure	psi	38.0
Number of Stages	#	2
Diameter of Membrane Element	in	8
Length of Membrane Element	in	40
FO Membrane Area per Element	sf	172
Projected Water Recovery	%	55
Maximum FO Train Capacity	mgd	1.43
Feed Solution Flow to Membranes per Train	mgd	2.6
Draw Solution Flow to Membranes per Train	mgd	0.2
Number of Trains	#	Varies

Parameter	Units	Value
CIP System Included?	Y/N	Yes
Pretreatment Chemicals Included?	Y/N	Yes
Feed Solution Cartridge Filters?	Y/N	Yes
Draw Solution Cartridge Filters?	Y/N	Yes

**6.1.3. Development of the FO Cost Model**

Design criteria presented in Table 6.1 were used to develop cost estimates for FO systems with a capacity of 0.5, 1, 2, 5, 7.5, 10, 12.5 and 15-mgd of recovered water from the impaired water feed solution. Quantity take-offs were then calculated from each parametric layout and major equipment list. Finally, cost curves were generated based on the information calculated from the individual spreadsheet models for each system size. Unit cost data for the different components were obtained from CH2M HILL’s parametric cost estimating system (CPES) and equipment vendors. Empirical models were developed from the material take-off and unit costs for presentation in this work. Figure 6.4 represents that building area required for each amount of membrane area utilized in the plant. Historical work by CH2M HILL indicates that the construction costs in parametric models are better represented by membrane area than the installed flow-rate. All of the cost curves utilize total installed membrane area as the x-axis of empirical curves.

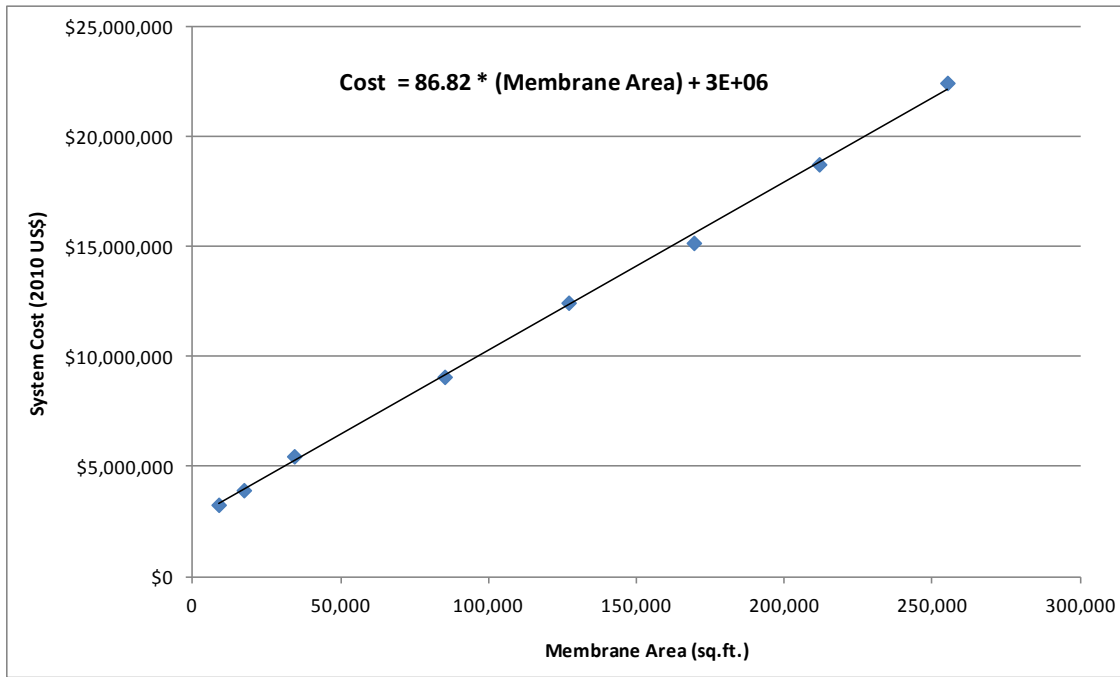
Figure 6.3 displays the building footprint required to contain the membrane area shown on the x-axis, including all major equipment, required clearance for maintenance and operations, ancillary equipment, motor control centers and all other equipment shown in Figure 6.2. The general layout shown in Figure 6.2 was used for all area estimates.



**Figure 6.3 Building Area Curve for a FO System**



Figure 6.4 presents the curve developed to estimate construction cost for FO systems (excluding the RO draw solution reconcentration portion of the project).



**Figure 6.4 Forward Osmosis System Construction Cost Curve**

Figure 6.5 presents the cost curve developed to estimate the equipment cost portion of a forward osmosis system, based upon the process flow diagram shown and typical equipment costs.

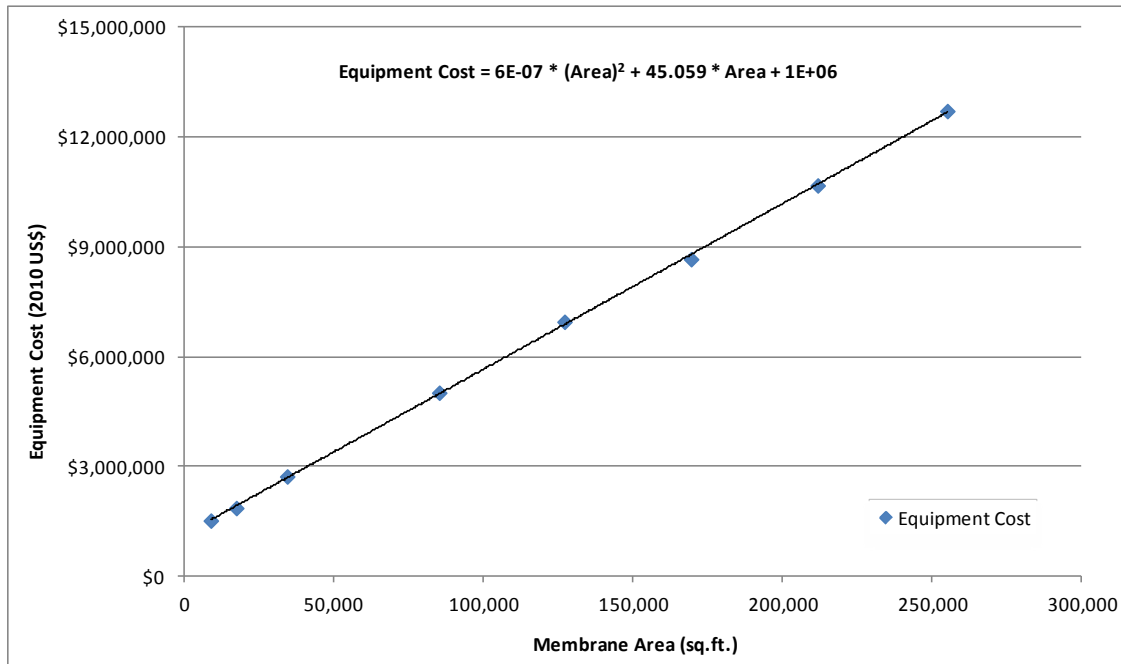
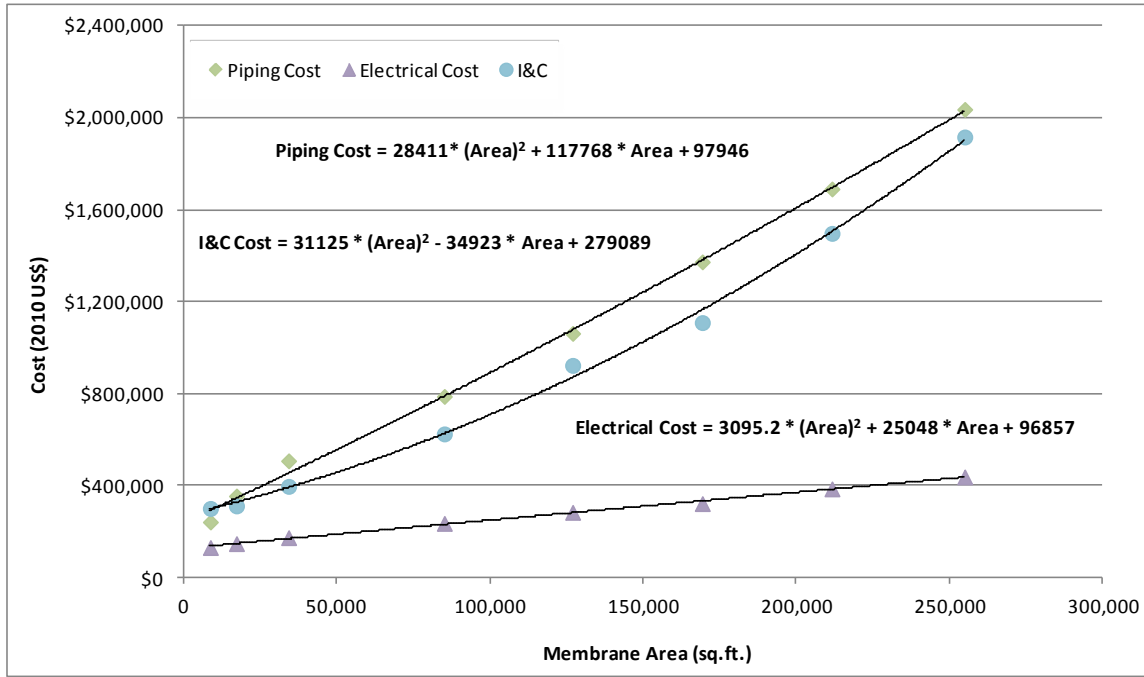


Figure 6.5 Process Equipment Cost Curve

Figure 6.6 presents the cost curves developed to estimate mechanical, Instrumentation and control (I&C) and electrical costs associated with the conceptual FO systems.



**Figure 6.6 Capital Cost Curves for Other Engineering Disciplines**

**6.1.4. Operations and Maintenance Costs**

In development of the operations and maintenance (O&M) costs, a model was prepared examining the system component consumables including energy, chemical, labor, and other items (e.g., membrane and cartridge filter replacements). Table 6.2 lists the assumptions utilized to develop annual O&M costs for FO systems. Based upon the model developed and cost assumptions contained in Table 6.2, operating cost estimates were developed for a range of flows. Using regression analysis, CH2M HILL determined that the most accurate method of cost estimating, over a range of flux, recovery and other data, is to use the total active membrane area.

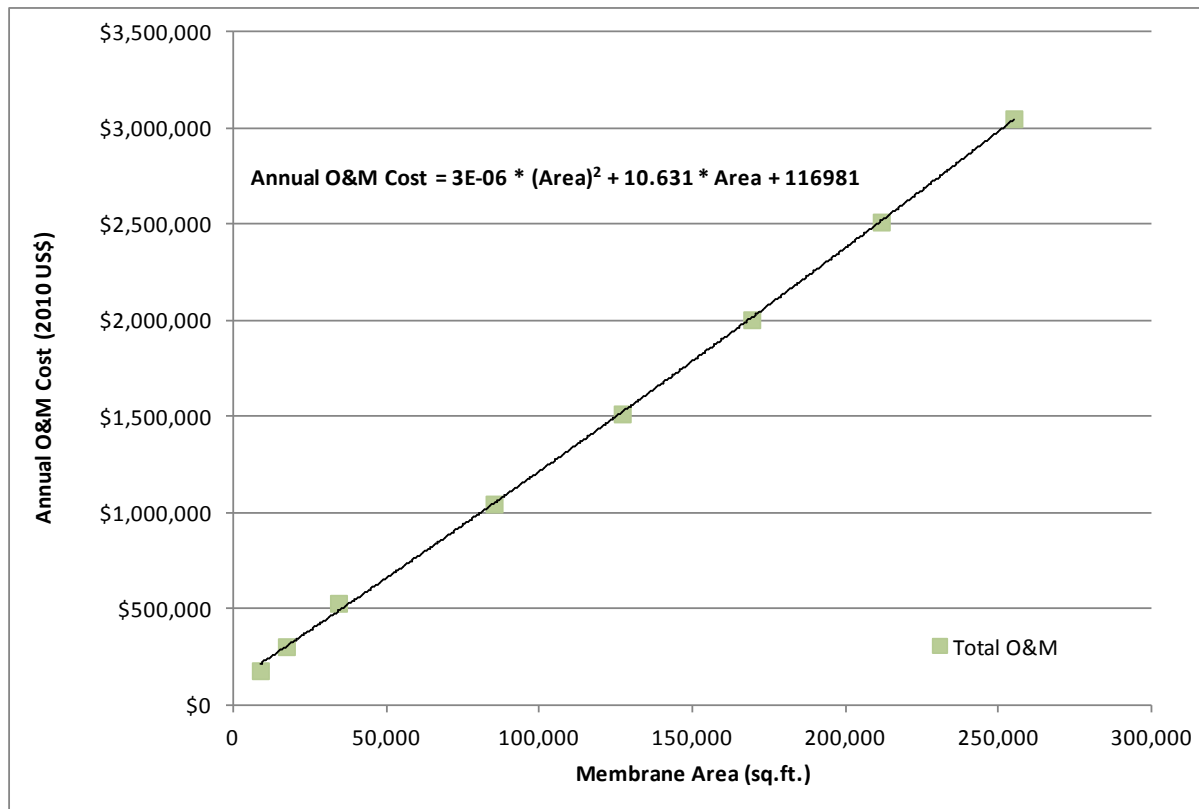
**Table 6.2 O&M Assumptions for FO Operating Cost**

Parameter	Units	Value
Maximum to Average Flow Factor	#	1.0
Number of Hours per Day the Plant Operates	Hr	24
Number of Days per Year the Plant Operates	Days	329
Power Cost	\$/kwh	\$0.10
Sulfuric Acid	\$/dry ton	\$140.00
Scale Inhibitor	\$/dry ton	\$4,400.00
Citric Acid	\$/dry ton	\$2,500.00
Sodium Hydroxide	\$/dry ton	\$825.00
Sodium EDTA	\$/dry ton	\$1,260.00
FO Membrane Replacement Frequency	Years	6
Number of Membrane Replacements in 20	#	3

Parameter	Units	Value
Years		
Annual Discount Rate	%	6%
FO Membrane Element Replacement Cost	\$/Element	\$600.00
Cartridge Filter Element Replacement Cost	\$/Element	\$11.00
Maintenance and Repair Allowance	% of Equipment Cost	3%
Contingency	%	10%

Figure 6.7 presents the cost curve developed to estimate annual O&M costs for FO systems (excluding the RO draw solution reconcentration portion of the project) as a function of total FO membrane installed.

**Figure 6.7 Annual O&M Cost Curve based on FO membrane area**



## 6.2. Reverse Osmosis Draw Solution Re-concentration Cost Model Development

In order to recover water from the draw solution, a system designed for very high total dissolved solids is required. It is assumed that the reverse osmosis system cost model be based upon a generic seawater reverse osmosis system. An existing proprietary cost model was utilized for the development of the SWRO costs.

The major steps taken in the development of the RO cost model include:

- Development of a typical lay-out for a full-scale RO system
- Development of a construction cost model based upon the process flow diagram and lay-out
- Development of an operating cost model for the RO process

Each of these steps are described separately below.

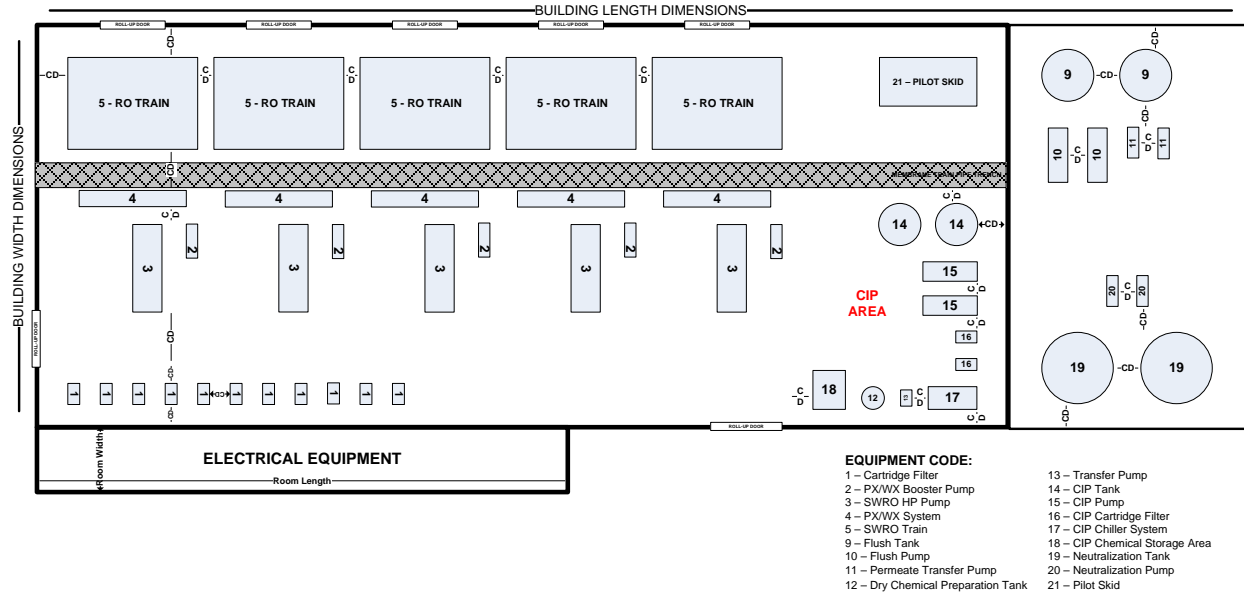
### ***6.2.1. Development of Conceptual Layout for SWRO***

CH2M HILL's proprietary cost model was utilized as the basis of the conceptual layout for the SWRO. This model has been previously utilized on cost development for numerous seawater desalination projects both domestically and abroad. The standard layout for the model is shown in Figure 6.8. The following major equipment is included in the concept layout and model:

- SWRO forwarding pumps
- Scale inhibitor dosing system
- Cartridge filters
- High-pressure feed pumps
- Isobaric energy recovery devices
- Energy recovery device booster pump
- SWRO membrane vessel racks
- Permeate flush tank and pumps
- Chemical cleaning tank, pump and cartridge filter
- CIP Heater and Chiller
- Bulk storage for chemical cleaning

The model assumes the use parallel of an individual high-pressure pump for each SWRO train, and individual isobaric energy recovery devices per train. Allowances for pipe-work, I&C, electrical and other miscellaneous components are included in the cost model.

**Figure 6.8 Layout for seawater desalination plant utilized for Cost Modeling**



**6.2.2. Development of the FO Cost Model**

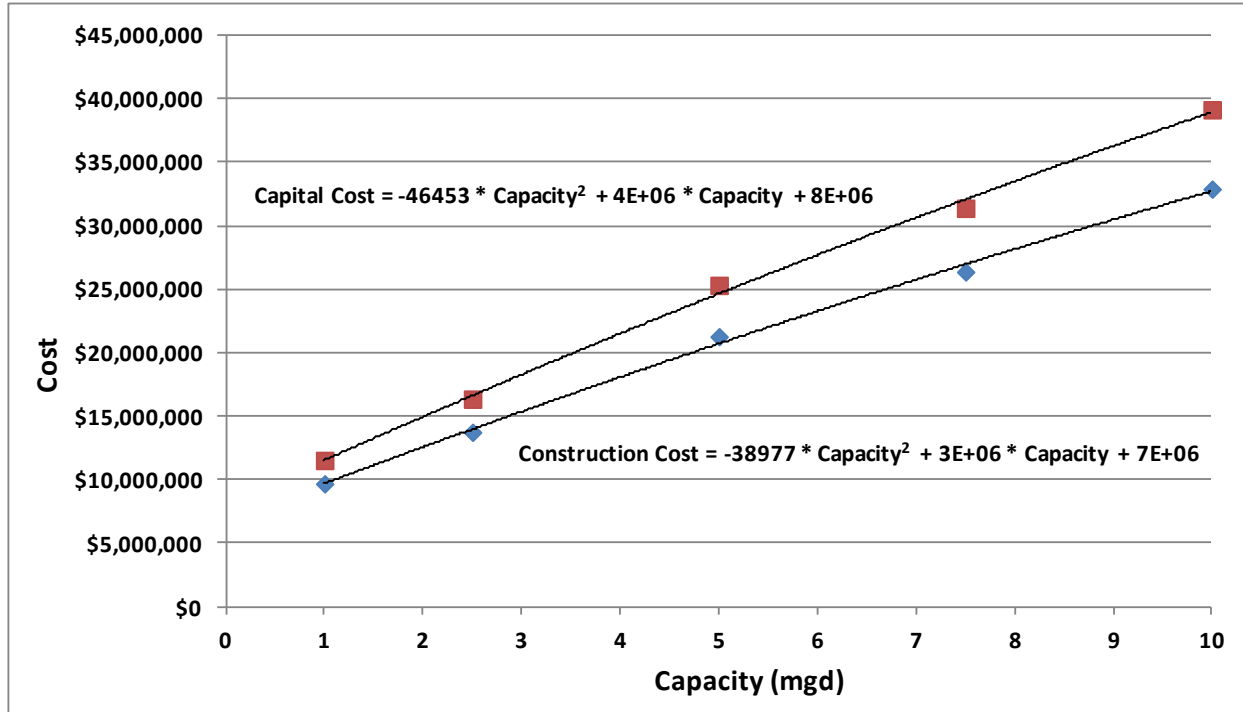
To develop construction cost curves for the conceptualized SWRO, material take-offs, unit costs, and construction costs were extracted from the CH2M HILL model for flow-rates of 1-, 2-, 2.5-, 5-, 7.5- and 10-mgd based upon the assumptions in Table 6.3 and layout illustrated in Figure 6.8. A regression analysis was conducted to develop an empirical cost model for inclusion in this report.

**Table 6.3 Reverse Osmosis System Design Criteria**

Parameter	Units	Value
Flux	gfd	8
Feed Pressure	psi	800
Number of Stages	#	1
Diameter of Membrane Element	in	8
Length of Membrane Element	in	40
Membrane Area per Element	sf	400
Projected Water Recovery	%	50
Number of Trains	#	Varies
CIP System Included?	Y/N	Yes
Pretreatment Chemicals Included?	Y/N	Yes
Feed Solution Cartridge Filters?	Y/N	Yes

Figure 6.9 presents unit construction costs for the seawater reverse osmosis system range from approximately \$9.00/gpd capacity to as low as \$3.16/gpd capacity for the RO desalination system alone. While this unit cost is lower than conventional seawater desalination installations, it is noted that in the FO-RO configuration, minimal pretreatment is required, and no allotments are made for for post-treatment, residual handling, intakes, or outfalls in this model.

**Figure 6.9 Construction Cost Curve for Seawater Reverse Osmosis System**



**6.2.3. Operations and Maintenance Costs**

Operations and maintenance costs associated with RO systems treating high salinity can be large, mostly due to the cost of electricity. Membrane replacement costs, cartridge filter replacement costs, labor and chemical costs also significant contributors to the O&M costs associated with the RO system. Table 6.4 contains the major assumptions used in development of the O&M costs.

**Table 6.4 O&M Assumptions for RO Operating Cost**

Parameter	Units	Value
Maximum to Average Flow Factor	#	1.0
Number of Hours per Day the Plant Operates	Hr	24
Number of Days per Year the Plant Operates	Days	329
Power Cost	\$/kwh	\$0.10
Sulfuric Acid	\$/dry ton	\$140.00
Scale Inhibitor	\$/dry ton	\$4,400.00
Citric Acid	\$/dry ton	\$2,500.00
Sodium Hydroxide	\$/dry ton	\$825.00
Sodium EDTA	\$/dry ton	\$1,260.00
FO Membrane Replacement Frequency	Years	6
Number of Membrane Replacements in 20 Years	#	3
Annual Discount Rate	%	6%

Parameter	Units	Value
FO Membrane Element Replacement Cost	\$/Element	\$600.00
Cartridge Filter Element Replacement Cost	\$/Element	\$11.00
Maintenance and Repair Allowance	% of Equipment Cost	3%
Contingency	%	10%

O&M costs were calculated for plant capacities ranging from 1 to 10 mgd. Figure 6.10 illustrates the anticipated operating costs associated with RO component of the FO-RO process.

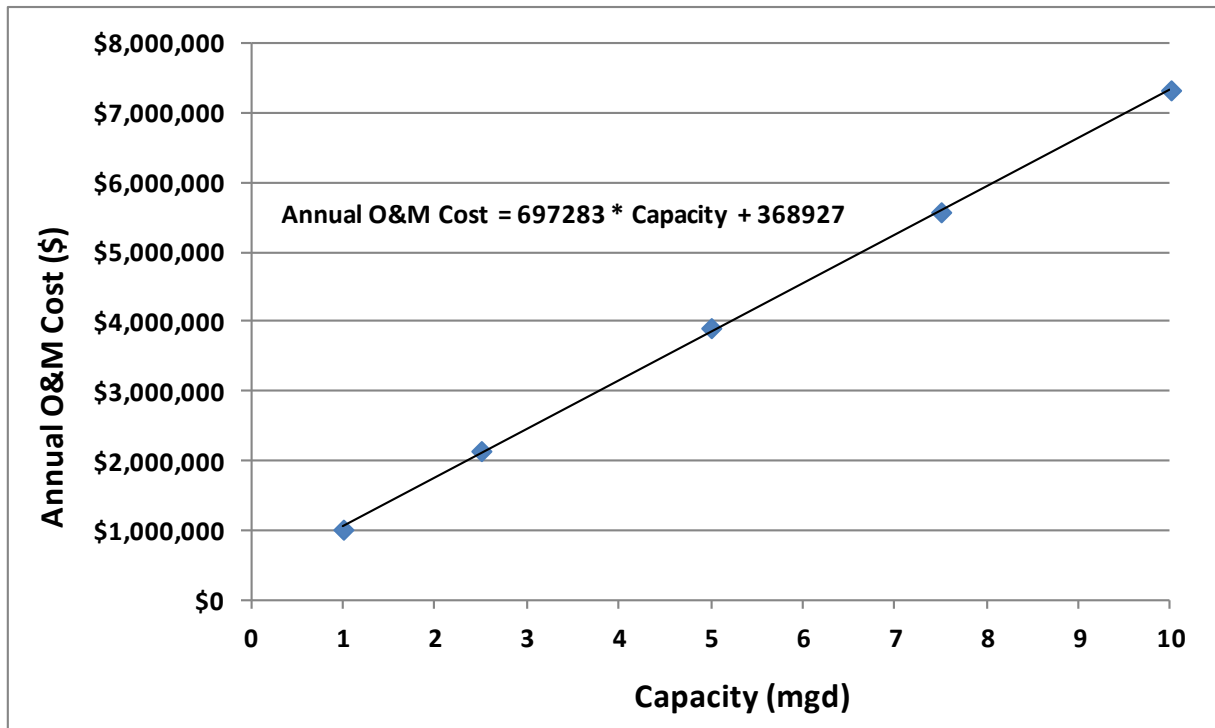


Figure 6.10 Annual O&M Cost Curves for Reverse Osmosis system

### 6.3. Comparison of FO-RO to Advanced Wastewater Treatment

Typical advanced wastewater treatment plants in the United States treat secondary treated wastewater using treatment processes consisting of microfiltration pretreatment, RO desalination, and advanced oxidation disinfection processes. These processes are able to provide a very good quality of treated water, which is currently being used for indirect potable reuse. In evaluating the relevancy of the FO-RO process, a likely application would be similar in nature to an advanced wastewater treatment plant. This section develops cost curves for advanced wastewater treatment systems that can be compared to the FO-RO process.

#### 6.3.1. Capital Cost for Advanced Wastewater Treatment Plant

To develop capital cost curves for an AWWTP, a number of assumptions were made regarding the process sizing prior to developing costs using CH2M HILL’s cost model. CH2M HILL analysts have determined using multivariate regression analysis that total membrane area and



recovery are the parameters most representative of the total capital costs. These process parameters are outlined in Table 6.5 and include flux and recovery. Additional allowances used in developing the capital costs are included in Table 6.6.

**Table 6.5 Process Parameters for an Advanced Wastewater Treatment Plant**

<b>Factor</b>	<b>Value</b>
<b>Additional Project Markups</b>	
MF	
Unit Flux	50 gfd
Unit Recovery	92%
BWRO	
Unit Flux	14 gfd
BWRO Unit Recovery	75%
Number of stages	2
Membrane Area per Element	400"
Membrane Element Diameter	8"

**Table 6.6 Assumptions in Developing an Advanced Wastewater Treatment Plant Cost**

<b>Factor</b>	<b>Value</b>
<b>Additional Project Markups</b>	
Sitework	3.0%
Yard Piping	5.5%
Yard Electrical	4.5%
Plant Computer System	1.5%
Contractor Markups	
Overhead	7%
Profit	10%
Mobilization / Bonds / Insurance	3%
Contingency	30%

Construction and capital cost estimates developed for a range of flows from 1-mgd to 20-mgd are presented in Figure 6.11. Capital costs were calculated based on 19.5% allowances over the construction costs. Allowances included costs for permitting (1%), engineering (8%), services during construction (7%), commissioning and startup (3%) and legal and administrative costs (0.5%).

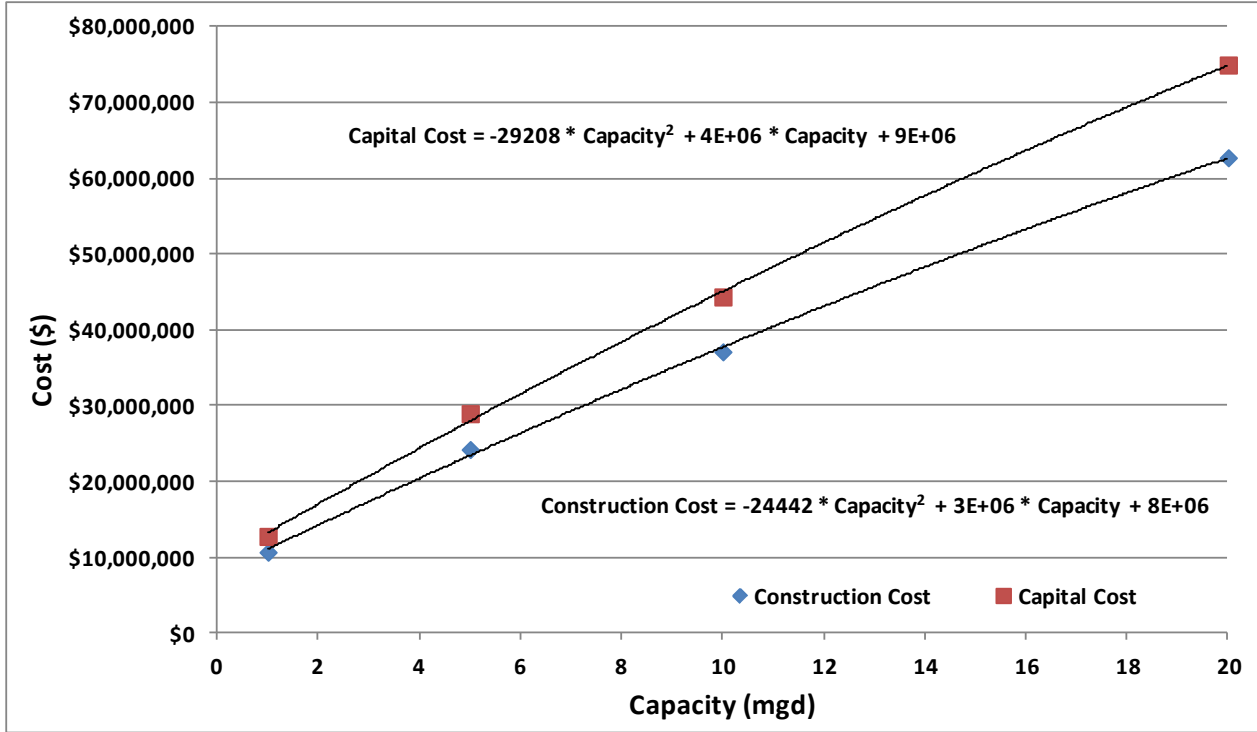


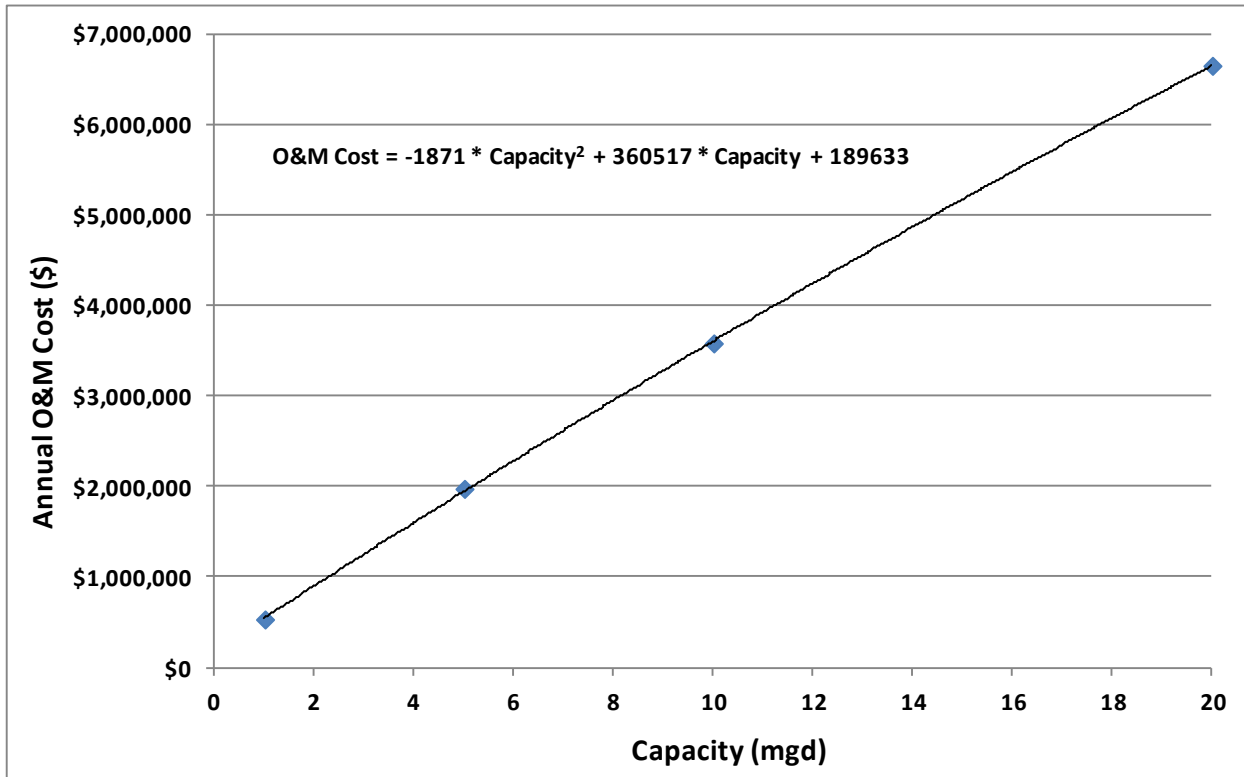
Figure 6.11 Capital Cost Curve for Advanced Wastewater Treatment Plant

6.3.2. Operating Costs for an Advanced Wastewater Treatment System

Operating costs for an advanced wastewater treatment plant were estimated using the assumptions contained in Table 6.7. An operations and maintenance cost curve for advanced wastewater treatment processes is shown in Figure 6.12.

Table 6.7 O&M Assumptions for an AWWTP Operating Cost

Parameter	Units	Value
Maximum to Average Flow Factor	#	1.0
Number of Hours per Day the Plant Operates	hr	24
Number of Days per Year the Plant Operates	Days	329
Power Cost	\$/kwh	\$0.10
Sulfuric Acid	\$/dry ton	\$140.00
Scale Inhibitor	\$/dry ton	\$4,400.00
Citric Acid	\$/dry ton	\$2,500.00
Sodium Hydroxide	\$/dry ton	\$825.00
Sodium EDTA	\$/dry ton	\$1,260.00
RO Membrane Replacement Frequency	Years	6
Number of Membrane Replacements in 20 Years	#	3
Annual Discount Rate	%	6%
RO Membrane Element Replacement Cost	\$/Element	\$600.00
Cartridge Filter Element Replacement Cost	\$/Element	\$11.00
Maintenance and Repair Allowance	% of Equipment Cost	3%
Contingency	%	10%



**Figure 6.12 O&M Cost Curve for Advanced Wastewater Treatment Processes**

**6.3.3. Comparison of FO-RO versus AWWTP**

To place the FO-RO process in context with other commercially available water and wastewater treatment technologies, the construction and annual O&M costs for the FO-RO process were compared against the costs for an AWWTP and a SWRO plant.

Figure 6.13 shows the comparison of construction costs. Based upon the estimates developed, it is anticipated that the cost for an FO-RO process will be about 30% higher than that of an AWWTP.

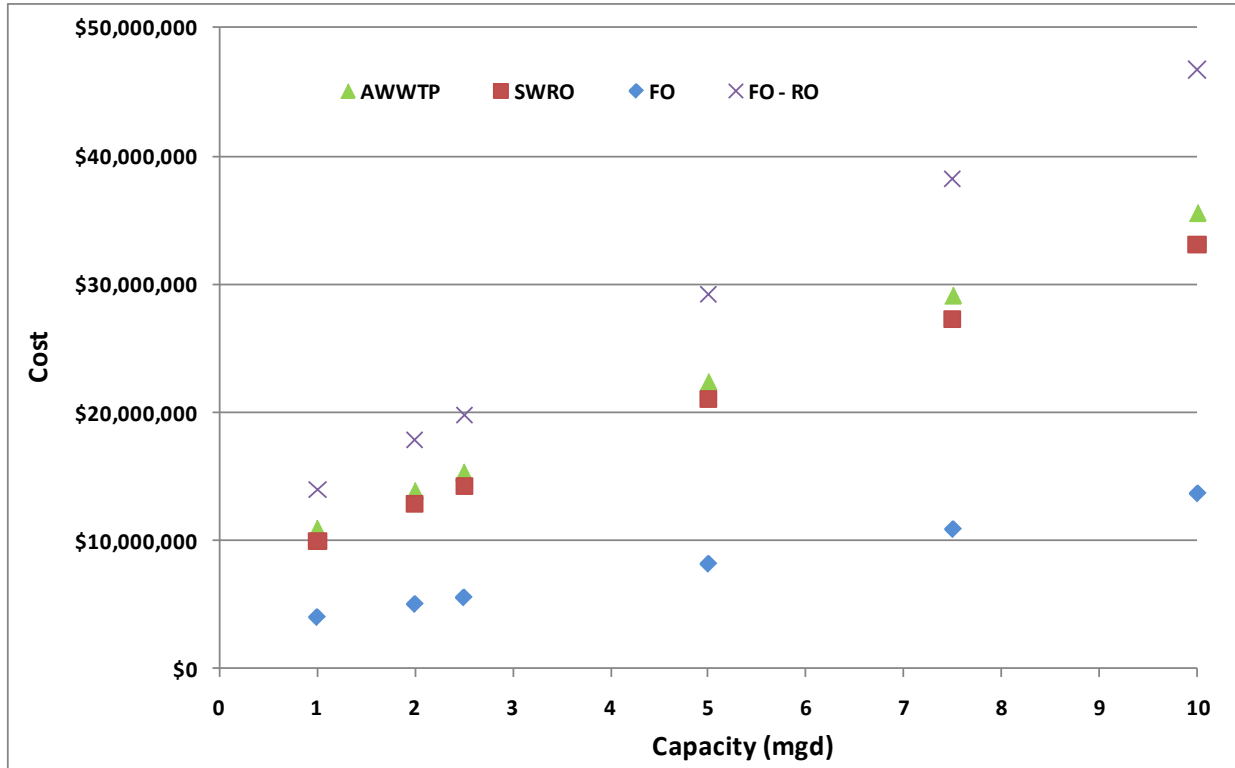


Figure 6.13 Comparison of FO-RO and AWWTP Process Construction Costs

Figure 6.14 provides an estimate of the operating costs. As expected, there is a premium (about 2.5 times) involved for an FO-RO process. Much of this premium is a result of the much higher energy consumption of an FO-RO process relative to an AWWTP. Higher pressures are required in the RO portion of the FO-RO process when compared to the driving pressures required for the RO portion of the AWWTP process due to the different salinity, with pressures of less than 300 psi anticipated for the AWWTP process and pressures exceeding 800 psi anticipated for the FO-RO process.

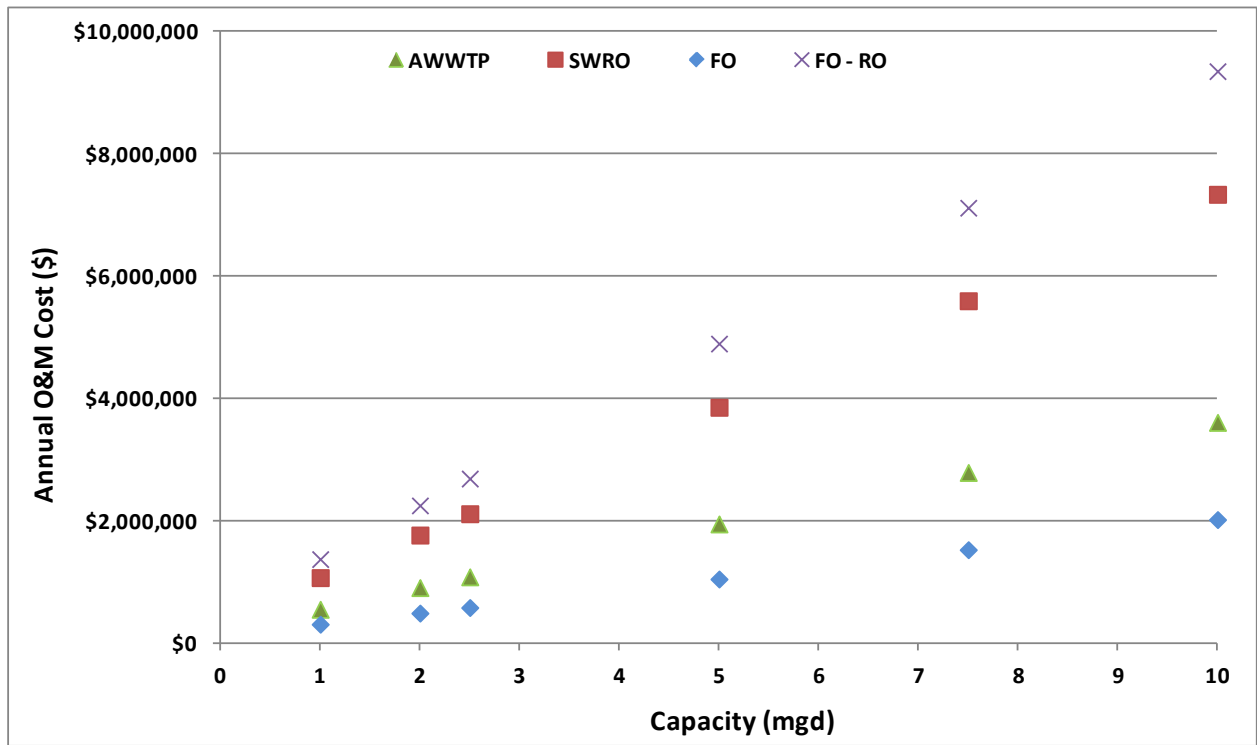


Figure 6.14 Comparison of FO-RO and AWWTP Process O&M Costs

#### 6.4. Economic Feasibility of a FO-RO Hybrid Process

While it appears that existing development paths in FO technology will result in a feasible FO-RO process, it does not compare favorably in terms of either capital costs or operations cost with the AWWTP for the recovery of wastewater at this time. While niche applications of this specific application are likely to be installed around the world (and have already been installed for treating leachate as a impaired water) the specific application of using reverse osmosis concentrate as a draw solution to recovery water from wastewater effluent is not likely to be utilized, in the configuration studied.

As shown in Figure 6.13 and Figure 6.14, the construction costs and operating costs associated with the FO-RO hybrid process are more expensive than existing AWWTP technology. There is little advantage to implementing an FO-RO process over an AWWTP plant – particularly given

that the AAWTP technology is a proven process used for indirect potable reuse in the United States, Australia and Singapore, and direct potable reuse in Namibia.

In addition, the proposed FO-RO process, if applied to potable water, is in effect a FO based AWWTP for either direct or indirect potable reuse. Due to the use of the reconcentration step to recover beneficial water, and the high capital and operating costs associated with the extraction of beneficial water, the process isn't particularly viable compared to existing best available technology. Additionally, due to the requirement of the reconcentration step for extraction of beneficial water, only a very small amount of concentrate is utilized. From a practical operations perspective, tailoring a custom, low scaling draw solution may be more viable than utilizing concentrate in this configuration. At the current level of development, we do not see a FO-RO process using RO concentrate from an existing desalination plant as the draw solution as a good investment. There are certainly other niche applications where an FO-RO process is likely viable.

As we specifically examine Texas, we also conclude that the specific application studied is not likely viable. First, given the low specific flux of existing FO membranes, very high osmotic pressure gradients are required. The concentrate osmotic pressures required for the application realistically requires that very high concentrations of TDS be required. No full-scale facility in Texas currently has concentrate streams with TDS high enough to obtain reasonable fluxes. Secondly, for most reuse applications, an AWWTP is better value for the utilities at this time.

## **6.5. Volume Minimization Utilizing FO**

While not specifically studied within the scope of this project, we do believe that RO concentrate could play a very beneficial role in future waste minimization applications. In such an application, seawater, SWRO concentrate, or other high osmotic pressure streams would be used to extract water from an impaired wastewater stream in a once through process. The high osmotic pressure stream would be diluted by FO to levels closer to (or below) ambient seawater concentrations, potentially reducing environmental impacts associated with TDS toxicity in concentrate effluent streams. The impaired waste stream would have its volume significantly reduced, concentrating the waste, and making subsequent non-open water disposal of the waste more economical. In one specific application envisioned by the authors, wastewater streams from industrial processes, with BOD too high for economical activated sludge treatment, but too low for effective anaerobic treatment, could be concentrated as much as 10 to 20 times, providing high enough BOD to effectively utilize anaerobic treatment, and recovery energy from the waste using micro-turbines.

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## 7. Conclusions

The purpose of this study was to investigate the coupling of reverse and forward osmosis, using concentrate as a draw solution to extract water from a wastewater feed. Specifically, the study sought to:

- Examine the mechanics of forward and reverse osmosis water treatment
- Assess the feasibility of using high salinity streams to extract water out of wastewater streams
- Determine characteristics required for cost-effective application of this hybrid process

### 7.1. Mechanics of Forward Osmosis-Reverse Osmosis Treatment

Data from four Texas desalination plants and several wastewater treatment plants were used to characterize concentrate streams and wastewaters. Analysis of the data was conducted to determine the osmotic potential for FO applications. The subsequent estimated FO module flux based upon previously published data. The FO membrane flux calculated indicates that the osmotic potential between brackish water concentrate and wastewater is very low, resulting in very low flux and subsequently very high capital equipment costs. As a result, simulated seawater or seawater concentrate was used as the draw solution for the remainder of the study.

Bench-scale and testing of water fluxes and reverse-direction solute fluxes for the tests involving secondary effluent (SecEff) feed and synthetic seawater and brackish water draw solutions were conducted. Similar to the results of the DI-SW and DI-BW tests, water fluxes observed for the SecEff-SW experiments were also higher (at all draw solution concentrations) than the SecEff-BW tests. The reverse solute fluxes at all concentrations were also higher for the SecEff-SW tests. However, the calculated specific solute flux values were higher for the DI-BW experiments at all draw solution concentrations. Similar to the trends observed for the experiments involving DI water and secondary effluent feeds, at equal draw solution concentration, water flux and reverse solute flux values were higher for the SPLT-SW experiments than the SPLT-BW experiments.

Additional testing on particulate fouling of the membrane, using seeded turbidity, indicated that the membranes were somewhat tolerant to the presence of particulate matter, and did not form a particulate cake that decreased performance of the FO system. Experiments were conducted with turbidities as high as 100 NTU with low rates of flux decline.

While not specifically a concern in this scope of work, the rejection of ammonia in FO membranes is of relevance. Other FO processes not studied in this work utilize an ammonium carbonate draw solution. Testing indicated significant diffusion of ammonia into the draw solution from the feed water, resulting in concentrations of concern if discharged into the environment as a waste stream. Additional work is required to fully characterize the impact of discharging this ammonia laden stream into the sea.



## **7.2. Cost-effective Application of Forward Osmosis to Wastewater Treatment with Desalination Concentrate as a Draw Solution**

Cost modeling conducted indicates that the use of FO-RO is not cost competitive when compared to tertiary treatment of wastewater using an advanced treatment process including membrane filtration following by RO and UV disinfection. FO-RO results in increased operating costs, however. Assuming that membranes can be commercially produced at a reasonable price point, it is anticipated that use of FO-RO may be viable at some point in the future. The process studied, however, shows little promise as a method of beneficially using concentrate, as the volumes utilized are very small once the reconcentration RO system is utilized to beneficially recover water. While niche applications may well occur using FO-RO, it is likely with an engineered draw solution specifically prepared for the application.

The study found no concentrate streams available from existing full-scale RO applications in Texas that are suitable for use as an economic draw solution in a FO application.

## **8. Recommendations**

While the use of FO is currently limited to mostly experimental applications, such as well drilling frac water treatment and emergency portable water supply devices; there is increasing interest in potential large scale municipal and industrial applications. In conducting this study, several recommendations were developed for continuing the development of the process. These include:

- Evaluation and development of new draw solutions to minimize energy input for water recovery and/or draw solution reconstitution.
- Under the current low liquid mass transfer rates, water flux in FO elements under osmotic potentials including seawater concentrate are very low. Focused research into reducing internal concentration polarization and increasing the liquid mass transfer coefficient are critical for reducing the probable cost for FO applications.
- Recent research has indicated that hollow-fiber membrane configurations possess greater potential for reduction in internal concentration polarization. Development of a commercially viable FO hollow-fiber membrane is desirable.
- Long-term fouling data of FO membranes in an FO-RO configuration are unknown. Long-term testing of the FO-RO configuration is recommended to determine fouling characteristics with different feedwaters.
- Long-term durability of the FO membrane is not known at this time. Extended testing in the FO-RO configuration is desirable to benchmark probable membrane life.
- Examine the use of an FO process for waste stream minimization utilizing seawater or seawater concentrate as a draw solution.

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## Appendix A Osmotic Pressure Model

In order to calculate the osmotic pressure, the method utilized by Thomas Wolfe (2006) in encoding the Visual Basic code contained in the Total Flux and Scalant Program (TFSP) for USEPA was adopted. This utilizes a modified version of the method utilized in Stumm and Morgan “Aquatic Chemistry” (1996).

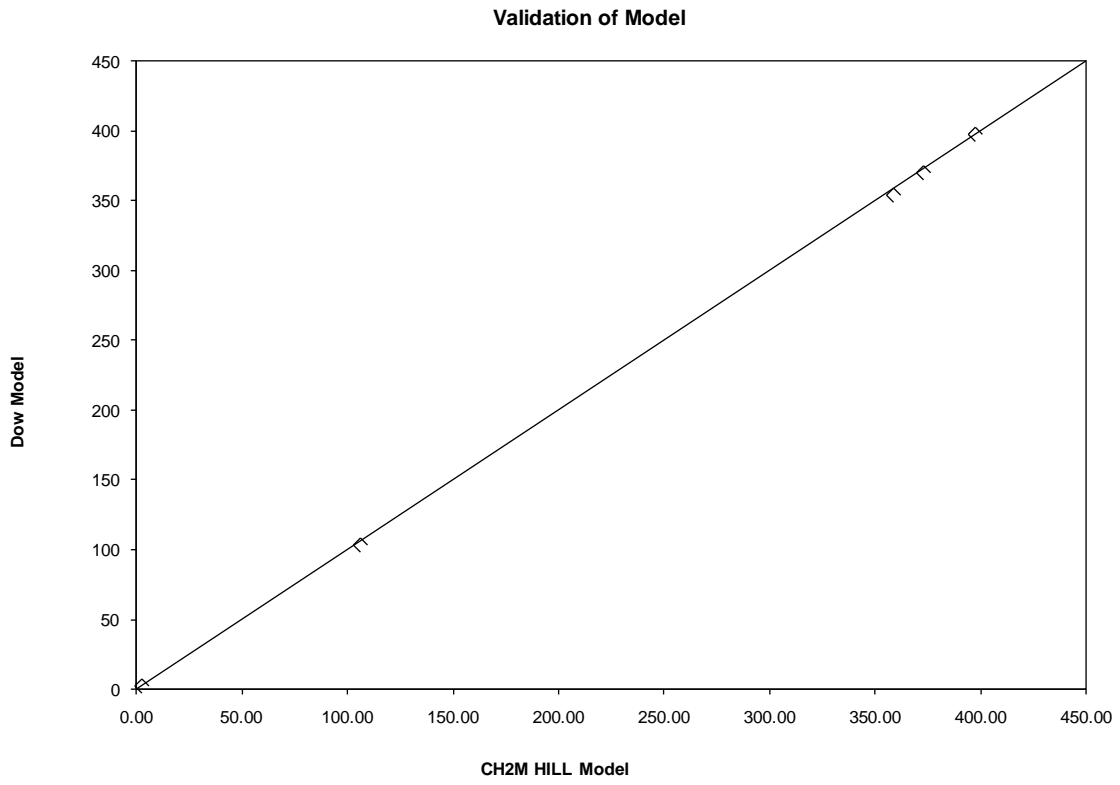
To validate the spreadsheet model used in this study, the osmotic pressure of various solutions were calculated using both this model and projections contained in the Dow Film Tec ROSA model. The different validation runs conducted are shown in Figure A.1.

**Figure A.1 Validation of Osmotic Pressure Model**

Osmotic Pressure Calculation							
Project Location	Tap Water	Brackish	Seawater @10C	Seawater @ 5C	Seawater @ 35C	Seawater @16C	
Temperature °C	18.50	18.50	16.00	5.00	35.00	16.00	
pH (valid from 2 to 12)	7.33	7.30	8.05	8.05	8.10	8.05	
Ionic Strength	0.0078	0.1776	0.7254	0.7254	0.7255	0.7254	
Osmotic Pressure psig	2.78	106.21	372.71	358.53	397.21	372.71	
<b>ION / Species</b>	<b>mg/L</b>	<b>mg/L</b>	<b>mg/L</b>	<b>mg/L</b>	<b>mg/L</b>	<b>mg/L</b>	<b>mg/L</b>
Sodium - Na	38.55	3220.00	11319.00	11319.00	11319.00	11319.00	
Potassium - K	2.77	400.00	457.00	457.00	457.00	457.00	
Calcium - Ca	36.99	200.00	391.00	391.00	391.00	391.00	
Magnesium - Mg	10.04	50.00	1316.00	1316.00	1316.00	1316.00	
Strontium - Sr	0.00	0.00	7.37	7.37	7.37	7.37	
Barium - Ba	0.04	0.04	6.10	6.10	6.10	6.10	
Ferric Iron - Fe+++ (as Fe)	0.00	0.00	0.00	0.00	0.00	0.00	
Aluminum - Al	0.08	0.08	0.00	0.00	0.00	0.00	
Ammonia - NH4 (as NH4)	10.00	10.00	0.01	0.01	0.01	0.01	
Hydrogen Ion - H	0.00	0.00	0.00	0.00	0.00	0.00	
<b>Sum all cations (incl metals):</b>	<b>98.48</b>	<b>3880.13</b>	<b>13496.48</b>	<b>13496.48</b>	<b>13496.48</b>	<b>13496.48</b>	
Bicarbonate - HCO3	155.88	300.00	149.00	149.00	149.00	149.00	
Carbon Dioxide - CO2 (calc'd)	12.17	20.28	1.00	1.00	1.00	1.00	
Carbonate - CO3 (calc'd)	0.18	0.75	3.51	2.72	6.13	3.51	
Bromide - Br	0.06	20.00	0.00	0.00	0.00	0.00	
Chloride - Cl	23.42	5326.00	20215.00	20215.00	20215.00	20215.00	
Fluoride - F	0.50	0.50	0.90	0.90	0.90	0.90	
Sulfate - SO4	77.88	445.00	2825.00	2825.00	2825.00	2825.00	
Nitrate - NO3	1.08	1.08	0.01	0.01	0.01	0.01	
Phosphate - PO4 (as PO4)	3.06	3.06	0.00	0.00	0.00	0.00	
Hydroxide - OH	0.00	0.00	0.02	0.02	0.02	0.02	
<b>Sum Total Anions:</b>	<b>274.24</b>	<b>6116.66</b>	<b>23194.44</b>	<b>23193.65</b>	<b>23197.06</b>	<b>23194.44</b>	
Silica (as SiO2)	4.76	4.76	4.76	4.76	4.76	4.76	
Difference meq/l	0.00	0.13	0.48	0.51	0.39	0.48	
<b>TDS mg/L</b>	<b>377.47</b>	<b>10001.54</b>	<b>36695.68</b>	<b>36694.89</b>	<b>36698.30</b>	<b>36695.68</b>	
Osmotic Pressure Calculated	2.78	106.21	372.71	358.53	397.21	372.71	
Dow	2.84528	103.45984	369.6044	353.58576	397.08592	369.6044	
	-4.97913445	-2.657516812	-0.840045	-1.39759512	-0.03153527	-0.84004455	
TDS Calculated	377.47		36695.68	36694.89	36698.30	36695.68	
Dow	367		36725	36715	36763	36725	

Some variation is indicated in osmotic pressure at lower salinities. It is hypothesized that this discrepancy is a different method of determining the carbonate equilibrium components. Given the magnitude of osmotic pressure differential required in FO applications, the model deviation for tap water (~5%) is of little significance in this study. Figure A.2 plots the validation data, showing good conformance over the range of results.

**Figure A.2 Validation Data for the Osmotic Pressure Model**



## Appendix B Osmotic Pressure Characterization

Figure B.1 Osmotic Pressure Calculation for Seawater

Osmotic Pressure Calculation			Osmotic Pressure calcs					
Project	Assessment of Osmotic Mechanisms		R_Gas	0.000082054				
Location	Seawater, Brownsville, TX		T_Kelvin	291.15				
Temperature	°C	18.0	RT	0.023890022				
pH (valid from 2 to 12)		8.3	Epsilon	0.5				
Ionic Strength		0.68739	sqrt(IS_Treated)	0.829088473				
Osmotic Pressure	psig	351.19						
ION / Species	mg/L	mEq/l	Valence	Molecular Weight	Ionic Strength	ActCoef	PI - atm Feed	
Sodium - Na	10175.00	442.59	1	22.990	442.588	0.93	9.83	
Potassium - K	684.00	17.49	1	39.098	17.495	0.93	0.39	
Calcium - Ca	434.00	21.66	2	40.080	43.313	0.75	0.19	
Magnesium - Mg	1330.00	109.44	2	24.305	218.885	0.75	0.98	
Strontium - Sr	7.73	0.18	2	87.620	0.353	0.75	0.00	
Barium - Ba	0.24	0.00	2	137.340	0.007	0.75	0.00	
Ferric Iron - Fe+++ (as Fe)	-	0.00	3	55.847	0.000	0.52	-	
Aluminum - Al	0.00	0.00	3	26.982	0.000	0.52	-	
Ammonia - NH4 (as NH4)	0.00	0.00	1	18.038	0.000	0.93	-	
Hydrogen Ion - H	5.01E-09	0.00E+00	1	1.000	0.000	0.93	0.00	
<b>Sum all cations (incl metals):</b>	<b>12,631.0</b>	<b>591.4</b>	Sum Cations		722.641		<b>11.39</b>	
Bicarbonate - HCO3	313.00	5.13	HC03_	-1	61.017	5.130	0.93	0.11
Carbon Dioxide - CO2 (calc'd)	1.86	0.08	CO2_	-2	44.010	0.169	0.75	0.00
Carbonate - CO3 (calc'd)	13.44		CO3_	-2	60.009	0.896	0.75	0.00
Bromide - Br		0.00	Br_	-1	79.904	0.000	0.93	-
Chloride - Cl	19000.0	535.92	Cl_	-1	35.453	535.921	0.93	11.90
Fluoride - F	-	0.00	F_	-1	18.998	0.000	0.93	-
Sulfate - SO4	2642.0	55.01	SO4_	-2	96.058	110.017	0.75	0.49
Nitrate - NO3	-	0.00	NO3_	-1	62.005	0.000	0.93	-
Phosphate - PO4 (as PO4)	-	0.00	PO4_	-3	94.971	0.000	0.52	-
Hydroxide - OH	3.39E-02	0.00E+00	OH_	-1	17.000	0.002	0.93	0.00
<b>Sum Total Anions:</b>	<b>21,970.33</b>	<b>-596.1</b>	Sum Anions			652.134		<b>12.51</b>
Silica (as SiO2)	29.50	0.98	SiO2_	-2	60.085	1.964	0.75	0.01
Difference meq/l	-4.70	-0.40%						
<b>TDS</b>	<b>mg/L</b>	<b>34630.8</b>	Ionic Strength			<b>0.68739</b>		
			Osmotic Pressure					<b>351.19</b>
Calculation By:	Name		Date					
Checked By:	R. Huehmer		1/15/2010					
	J Gomez		1/16/2010					



**Figure B.2 Osmotic Pressure Calculation for Seawater Concentrate**

Osmotic Pressure Calculation		
Project	Assessment of Osmotic Mechanisms	
Location	Seawater Concentrate, Brownsville, TX	
Temperature	°C	18.0
pH (valid from 2 to 12)		7.2
<b>Ionic Strength</b>		1.33932
<b>Osmotic Pressure</b>	psig	672.43

Osmotic Pressure calcs	
R_Gas	0.000082054
T_Kelvin	291.15
RT	0.023890022
Epsilon	0.5
sqrt(IS_Treated)	1.157290531


ION / Species	mg/L	mEq/l	Molecular		Ionic Strength	ActCoef	PI - atm Feed
			Valence	Weight			
Sodium - Na	19961.00	868.26	1	22.990	868.256	0.92	19.02
Potassium - K	1317.00	33.68	1	39.098	33.685	0.92	0.74
Calcium - Ca	839.00	41.87	2	40.080	83.733	0.71	0.35
Magnesium - Mg	2571.00	211.56	2	24.305	423.123	0.71	1.79
Strontium - Sr	14.95	0.34	2	87.620	0.682	0.71	0.00
Barium - Ba	0.47	0.01	2	137.340	0.014	0.71	0.00
Ferric Iron - Fe+++ (as Fe)	-	0.00	3	55.847	0.000	0.46	-
Aluminum - Al	0.00	0.00	3	26.982	0.000	0.46	-
Ammonia - NH4 (as NH4)	0.00	0.00	1	18.038	0.000	0.92	-
Hydrogen Ion - H	6.31E-08	0.00E+00	1	1.000	0.000	0.92	0.00
<b>Sum all cations (incl metals):</b>	<b>24,703.4</b>	<b>1,155.7</b>					
Bicarbonate - HCO3	683.00	11.19	-1	61.017	11.194	0.92	0.25
Carbon Dioxide - CO2 (calc'd)	47.76	2.17	-2	44.010	4.340	0.71	0.02
Carbonate - CO3 (calc'd)	3.04		-2	60.009	0.203	0.71	0.00
Bromide - Br		0.00	-1	79.904	0.000	0.92	-
Chloride - Cl	36665.9	1034.21	-1	35.453	1034.212	0.92	22.66
Fluoride - F	-	0.00	-1	18.998	0.000	0.92	-
Sulfate - SO4	5264.0	109.60	-2	96.058	219.202	0.71	0.93
Nitrate - NO3	-	0.00	-1	62.005	0.000	0.92	-
Phosphate - PO4 (as PO4)	-	0.00	-3	94.971	0.000	0.46	-
Hydroxide - OH	2.69E-03	0.00E+00	-1	17.000	0.000	0.92	0.00
<b>Sum Total Anions:</b>	<b>42,663.72</b>	<b>-1,155.0</b>					
Silica (as SiO2)	56.00	1.86	-2	60.085	3.728	0.71	0.02
Difference meq/l	0.71	0.03%					
<b>TDS</b>	<b>mg/L</b>	<b>67423.1</b>					

Calculation By:		Name	Date
Checked By:		R. Huehmer	1/15/2010
		J Gomez	1/16/2010

Osmotic Pressure	
Ionic Strength	1.33932
Osmotic Pressure	672.43

**Figure B.3 Osmotic Pressure Calculation for Raw Wastewater**

Osmotic Pressure Calculation			Osmotic Pressure calcs				
Project	Assessment of Osmotic Mechanisms		R_Gas	0.000082054			
Location	Typical Raw Wastewater		T_Kelvin	291.15			
Temperature	°C	18.0	RT	0.023890022			
pH (valid from 2 to 12)		7.8	Epsilon	0.5			
Ionic Strength		0.02367	sqrt(IS_Treated)	0.153846462			
Osmotic Pressure	psig	9.11					
ION / Species	mg/L	mEq/l	Valence	Molecular Weight	Ionic Strength	ActCoef	PI - atm Feed
Sodium - Na	198.00	8.61	1	22.990	8.613	0.98	0.20
Potassium - K	0.00	0.00	1	39.098	0.000	0.98	-
Calcium - Ca	74.40	3.71	2	40.080	7.425	0.92	0.04
Magnesium - Mg	38.50	3.17	2	24.305	6.336	0.92	0.03
Strontium - Sr	-	0.00	2	87.620	0.000	0.92	-
Barium - Ba	0.00	0.00	2	137.340	0.000	0.92	-
Ferric Iron - Fe+++ (as Fe)	-	0.00	3	55.847	0.000	0.82	-
Aluminum - Al	0.00	0.00	3	26.982	0.000	0.82	-
Ammonia - NH4 (as NH4)	22.00	1.22	1	18.038	1.220	0.98	0.03
Hydrogen Ion - H	1.58E-08	0.00E+00	1	1.000	0.000	0.98	0.00
<b>Sum all cations (incl metals):</b>	<b>332.9</b>	<b>16.7</b>	Sum Cations		23.593		<b>0.31</b>
Bicarbonate - HCO3	210.00	3.44	-1	61.017	3.442	0.98	0.08
Carbon Dioxide - CO2 (calc'd)	5.30	0.24	-2	44.010	0.482	0.92	0.00
Carbonate - CO3 (calc'd)	0.87		-2	60.009	0.058	0.92	0.00
Bromide - Br	-	0.00	-1	79.904	0.000	0.98	-
Chloride - Cl	240.0	6.77	-1	35.453	6.770	0.98	0.16
Fluoride - F	-	0.00	-1	18.998	0.000	0.98	-
Sulfate - SO4	312.0	6.50	-2	96.058	12.992	0.92	0.07
Nitrate - NO3	-	0.00	-1	62.005	0.000	0.98	-
Phosphate - PO4 (as PO4)	-	0.00	-3	94.971	0.000	0.82	-
Hydroxide - OH	1.07E-02	0.00E+00	-1	17.000	0.001	0.98	0.00
<b>Sum Total Anions:</b>	<b>768.18</b>	<b>-16.7</b>	Sum Anions		23.744		<b>0.31</b>
Silica (as SiO2)	5.00	0.17	-2	60.085	0.333	0.92	0.00
Difference meq/l	0.01	0.02%					
<b>TDS</b>	<b>mg/L</b>	<b>1106.1</b>	Ionic Strength		<b>0.02367</b>		
			Osmotic Pressure			<b>9.11</b>	
Calculation By:	Name	Date					
Checked By:	R. Huehmer	1/15/2010					
	J Gomez	1/16/2010					



**Figure B.4 Osmotic Pressure Calculation for Primary Treated Wastewater**

Osmotic Pressure Calculation		
Project	Assessment of Osmotic Mechanisms	
Location	Typical Primary Wastewater	
Temperature	°C	18.0
pH (valid from 2 to 12)		7.8
Ionic Strength		0.02281
Osmotic Pressure	psig	8.91


Osmotic Pressure calcs	
R_Gas	0.000082054
T_Kelvin	291.15
RT	0.023890022
Epsilon	0.5
sqrt(IS_Treated)	0.151040988

ION / Species	mg/L	mEq/l	Molecular		Ionic Strength	ActCoef	PI - atm Feed
			Valence	Weight			
Sodium - Na	192.00	8.35	1	22.990	8.352	0.98	0.20
Potassium - K	0.00	0.00	1	39.098	0.000	0.98	-
Calcium - Ca	72.20	3.60	2	40.080	7.206	0.92	0.04
Magnesium - Mg	38.10	3.14	2	24.305	6.270	0.92	0.03
Strontium - Sr	-	0.00	2	87.620	0.000	0.92	-
Barium - Ba	0.00	0.00	2	137.340	0.000	0.92	-
Ferric Iron - Fe+++ (as Fe)	-	0.00	3	55.847	0.000	0.83	-
Aluminum - Al	0.00	0.00	3	26.982	0.000	0.83	-
Ammonia - NH4 (as NH4)	20.00	1.11	1	18.038	1.109	0.98	0.03
Hydrogen Ion - H	1.58E-08	0.00E+00	1	1.000	0.000	0.98	0.00
<b>Sum all cations (incl metals):</b>	<b>322.3</b>	<b>16.2</b>	Sum Cations		22.936		<b>0.30</b>
Bicarbonate - HCO3	230.00	3.77	-1	61.017	3.769	0.98	0.09
Carbon Dioxide - CO2 (calc'd)	5.82	0.26	-2	44.010	0.529	0.92	0.00
Carbonate - CO3 (calc'd)	0.95		-2	60.009	0.063	0.92	0.00
Bromide - Br	-	0.00	-1	79.904	0.000	0.98	-
Chloride - Cl	232.0	6.54	-1	35.453	6.544	0.98	0.15
Fluoride - F	-	0.00	-1	18.998	0.000	0.98	-
Sulfate - SO4	283.0	5.89	-2	96.058	11.785	0.92	0.06
Nitrate - NO3	-	0.00	-1	62.005	0.000	0.98	-
Phosphate - PO4 (as PO4)	-	0.00	-3	94.971	0.000	0.83	-
Hydroxide - OH	1.07E-02	0.00E+00	-1	17.000	0.001	0.98	0.00
<b>Sum Total Anions:</b>	<b>751.78</b>	<b>-16.2</b>	Sum Anions		22.691		<b>0.31</b>
Silica (as SiO2)	5.00	0.17	-2	60.085	0.333	0.92	0.00
Difference meq/l	-0.01	-0.02%	Ionic Strength		<b>0.02281</b>		
<b>TDS</b>	<b>mg/L</b>	<b>1079.1</b>	Osmotic Pressure				<b>8.91</b>

Calculation By:	Name	Date
Checked By:	R. Huehmer	1/15/2010
	J Gomez	1/16/2010

**Figure B.5 Osmotic Pressure Calculation for Secondary Treated Wastewater**

Osmotic Pressure Calculation		
Project	Assessment of Osmotic Mechanisms	
Location	Typical Secondary Wastewater	
Temperature	°C	18.0
pH (valid from 2 to 12)		7.4
<b>Ionic Strength</b>		0.02217
<b>Osmotic Pressure</b>	psig	8.14

ION / Species	mg/L	mEq/l
Sodium - Na	198.00	8.61
Potassium - K	0.00	0.00
Calcium - Ca	66.70	3.33
Magnesium - Mg	39.30	3.23
Strontium - Sr	-	0.00
Barium - Ba	0.00	0.00
Ferric Iron - Fe+++ (as Fe)	-	0.00
Aluminum - Al	0.00	0.00
Ammonia - NH4 (as NH4)	0.00	0.00
Hydrogen Ion - H	3.98E-08	0.00E+00
<b>Sum all cations (incl metals):</b>	<b>304.0</b>	<b>15.2</b>
Bicarbonate - HCO3	120.00	1.97
Carbon Dioxide - CO2 (calc'd)	7.64	0.35
Carbonate - CO3 (calc'd)	0.20	
Bromide - Br		0.00
Chloride - Cl	238.0	6.71
Fluoride - F	-	0.00
Sulfate - SO4	309.0	6.43
Nitrate - NO3	1.40	0.02
Phosphate - PO4 (as PO4)	3.40	0.11
Hydroxide - OH	4.27E-03	0.00E+00
<b>Sum Total Anions:</b>	<b>679.64</b>	<b>-15.2</b>
Silica (as SiO2)	5.00	0.17
Difference meq/l	-0.07	-0.23%
<b>TDS</b>	<b>mg/L</b>	<b>988.6</b>

Osmotic Pressure calcs	
R_Gas	0.000082054
T Kelvin	291.15
RT	0.023890022
Epsilon	0.5
sqrt(IS_Treated)	0.148889914

	Valence	Molecular Weight	Ionic Strength	ActCoef	PI - atm Feed
Na_	1	22.990	8.613	0.98	0.20
K_	1	39.098	0.000	0.98	-
Ca_	2	40.080	6.657	0.92	0.04
Mg_	2	24.305	6.468	0.92	0.04
Sr_	2	87.620	0.000	0.92	-
Ba_	2	137.340	0.000	0.92	-
Fe3_	3	55.847	0.000	0.83	-
Al_	3	26.982	0.000	0.83	-
NH4_	1	18.038	0.000	0.98	-
H_	1	1.000	0.000	0.98	0.00
<b>Sum Cations</b>			<b>21.737</b>		<b>0.27</b>
HC03_	-1	61.017	1.967	0.98	0.05
CO2_	-2	44.010	0.694	0.92	0.00
CO3_	-2	60.009	0.013	0.92	0.00
Br_	-1	79.904	0.000	0.98	-
Cl_	-1	35.453	6.713	0.98	0.16
F_	-1	18.998	0.000	0.98	-
SO4_	-2	96.058	12.867	0.92	0.07
NO3_	-1	62.005	0.023	0.98	0.00
PO4_	-3	94.971	0.322	0.83	0.00
OH_	-1	17.000	0.000	0.98	0.00
<b>Sum Anions</b>			<b>22.599</b>		<b>0.28</b>
SiO2_	-2	60.085	0.333	0.92	0.00
<b>Ionic Strength</b>			<b>0.02217</b>		
<b>Osmotic Pressure</b>					<b>8.14</b>

Calculation By: R. Huehmer  
 Checked By: J Gomez

Date  
1/15/2010  
1/16/2010





**Figure B.6 Osmotic Pressure Calculation for Kay Bailey Hutchison Desalination Plant (average concentrations)**

Osmotic Pressure Calculation		
Project	Assessment of Osmotic Mechanisms	
Location	Avg Brackish Concentrate, El Paso	
Temperature	°C	23.9
pH (valid from 2 to 12)		7.8
Ionic Strength		0.22067
Osmotic Pressure	psig	110.60

ION / Species	mg/L	mEq/l
Sodium - Na	2926.10	127.28
Potassium - K	70.90	1.81
Calcium - Ca	637.10	31.79
Magnesium - Mg	162.80	13.40
Strontium - Sr	-	0.00
Barium - Ba	0.00	0.00
Ferric Iron - Fe+++ (as Fe)	0.21	0.01
Aluminum - Al	0.00	0.00
Ammonia - NH4 (as NH4)	0.00	0.00
Hydrogen Ion - H	1.58E-08	0.00E+00
<b>Sum all cations (incl metals):</b>	<b>3,797.1</b>	<b>174.3</b>
Bicarbonate - HCO3	401.10	6.57
Carbon Dioxide - CO2 (calc'd)	7.69	0.35
Carbonate - CO3 (calc'd)	3.94	-
Bromide - Br	-	0.00
Chloride - Cl	5520.2	155.70
Fluoride - F	-	0.00
Sulfate - SO4	1406.2	29.28
Nitrate - NO3	-	0.00
Phosphate - PO4 (as PO4)	0.39	0.01
Hydroxide - OH	1.07E-02	0.00E+00
<b>Sum Total Anions:</b>	<b>7,339.52</b>	<b>-191.6</b>
Silica (as SiO2)	122.00	4.06
Difference meq/l	-17.28	-4.72%
<b>TDS</b>	<b>mg/L</b>	<b>11258.6</b>


	Valence	Molecular Weight	Ionic Strength	ActCoef	PI - atm Feed
Na_	1	22.990	127.278	0.95	2.95
K_	1	39.098	1.813	0.95	0.04
Ca_	2	40.080	63.583	0.81	0.32
Mg_	2	24.305	26.793	0.81	0.13
Sr_	2	87.620	0.000	0.81	-
Ba_	2	137.340	0.000	0.81	-
Fe3_	3	55.847	0.034	0.63	0.00
Al_	3	26.982	0.000	0.63	-
NH4_	1	18.038	0.000	0.95	-
H	1	1.000	0.000	0.95	0.00
<b>Sum Cations</b>			<b>219.501</b>		<b>3.44</b>
HC03_	-1	61.017	6.574	0.95	0.15
CO2_	-2	44.010	0.699	0.81	0.00
CO3_	-2	60.009	0.262	0.81	0.00
Br_	-1	79.904	0.000	0.95	-
Cl_	-1	35.453	155.705	0.95	3.60
F_	-1	18.998	0.000	0.95	-
SO4_	-2	96.058	58.557	0.81	0.29
NO3_	-1	62.005	0.000	0.95	-
PO4_	-3	94.971	0.037	0.63	0.00
OH_	-1	17.000	0.001	0.95	0.00
<b>Sum Anions</b>			<b>221.833</b>		<b>4.05</b>
SiO2_	-2	60.085	8.122	0.81	0.04
<b>Ionic Strength</b>			<b>0.22067</b>		
<b>Osmotic Pressure</b>					<b>110.60</b>

Osmotic Pressure calcs	
R_Gas	0.000082054
T_Kelvin	297.05
RT	0.024374141
Epsilon	0.5
sqrt(IS_Treated)	0.469752377

Calculation By:	Name	Date
Checked By:	A Kupp	11/21/11
	R Huehmer	11/21/11

**Figure B.7 Osmotic Pressure Calculation for Kay Bailey Hutchison Desalination Plant (maximum concentrations)**

Osmotic Pressure Calculation		
Project	Assessment of Osmotic Mechanisms	
Location	Max Brackish Concentrate, El Paso	
Temperature	°C	26.6
pH (valid from 2 to 12)		8.3
Ionic Strength		0.37478
Osmotic Pressure	psig	185.71
ION / Species	mg/L	mEq/l
Sodium - Na	4440.00	193.13
Potassium - K	114.00	2.92
Calcium - Ca	1220.00	60.88
Magnesium - Mg	260.00	21.39
Strontium - Sr	-	0.00
Barium - Ba	0.00	0.00
Ferric Iron - Fe+++ (as Fe)	0.57	0.03
Aluminum - Al	0.00	0.00
Ammonia - NH4 (as NH4)	0.00	0.00
Hydrogen Ion - H	5.01E-09	0.00E+00
<b>Sum all cations (incl metals):</b>	<b>6,034.6</b>	<b>278.3</b>
Bicarbonate - HCO3	498.00	8.16
Carbon Dioxide - CO2 (calc'd)	2.74	0.12
Carbonate - CO3 (calc'd)	20.36	-
Bromide - Br	-	0.00
Chloride - Cl	10100.0	284.88
Fluoride - F	-	0.00
Sulfate - SO4	2260.0	47.06
Nitrate - NO3	-	0.00
Phosphate - PO4 (as PO4)	1.26	0.04
Hydroxide - OH	3.39E-02	0.00E+00
<b>Sum Total Anions:</b>	<b>12,882.39</b>	<b>-340.1</b>
Silica (as SiO2)	228.00	7.59
Difference meq/l	-61.79	-9.99%
<b>TDS</b>	<b>mg/L</b>	<b>19145.0</b>

Osmotic Pressure calcs	
R_Gas	0.000082054
T Kelvin	299.75
RT	0.024595687
Epsilon	0.5
sqrt(IS_Treated)	0.612195476

	Valence	Molecular Weight	Ionic Strength	ActCoef	PI - atm Feed
Na_	1	22.990	193.129	0.94	4.47
K_	1	39.098	2.916	0.94	0.07
Ca_	2	40.080	121.756	0.78	0.59
Mg_	2	24.305	42.790	0.78	0.21
Sr_	2	87.620	0.000	0.78	-
Ba_	2	137.340	0.000	0.78	-
Fe3_	3	55.847	0.092	0.58	0.00
Al_	3	26.982	0.000	0.58	-
NH4_	1	18.038	0.000	0.94	-
H_	1	1.000	0.000	0.94	0.00
<b>Sum Cations</b>			<b>360.683</b>		<b>5.33</b>
HC03_	-1	61.017	8.162	0.94	0.19
CO2_	-2	44.010	0.249	0.78	0.00
CO3_	-2	60.009	1.357	0.78	0.01
Br_	-1	79.904	0.000	0.94	-
Cl_	-1	35.453	284.884	0.94	6.59
F_	-1	18.998	0.000	0.94	-
SO4_	-2	96.058	94.110	0.78	0.45
NO3_	-1	62.005	0.000	0.94	-
PO4_	-3	94.971	0.119	0.58	0.00
OH_	-1	17.000	0.002	0.94	0.00
<b>Sum Anions</b>			<b>388.884</b>		<b>7.24</b>
SiO2_	-2	60.085	15.179	0.78	0.07
<b>Ionic Strength</b>			<b>0.37478</b>		
<b>Osmotic Pressure</b>					<b>185.71</b>

Calculation By: A Kupp  
 Checked By: R Huehmer

Date: 11/21/11  
11/21/11



**Figure B.8 Osmotic Pressure Calculation for High-Salinity Brackish Concentrate**

Osmotic Pressure Calculation			
Project	Assessment of Osmotic Mechanisms		
Location	High Salinity Brackish Concentrate		
Temperature	°C	24.5	
pH (valid from 2 to 12)		8.1	
Ionic Strength		0.19366	
Osmotic Pressure	psig	100.82	


Osmotic Pressure calcs	
R_Gas	0.000082054
T Kelvin	297.65
RT	0.024423373
Epsilon	0.5
sqrt(IS_Treated)	0.44006873

ION / Species	mg/L	mEq/l		Valence	Molecular Weight	Ionic Strength	ActCoef	PI - atm	Feed
Sodium - Na	2701.19	117.50	0.27	Na_	22.990	117.495	0.95	2.73	
Potassium - K	79.13	2.02	0.01	K_	39.098	2.024	0.95	0.05	
Calcium - Ca	496.95	24.80	0.05	Ca_	40.080	49.596	0.82	0.25	
Magnesium - Mg	153.41	12.62	0.02	Mg_	24.305	25.248	0.82	0.13	
Strontium - Sr	0.00	0.00	0.00	Sr_	87.620	0.000	0.82	-	
Barium - Ba	0.48	0.01	0.00	Ba_	137.340	0.014	0.82	0.00	
Ferric Iron - Fe+++ (as Fe)	0.00	0.00	0.00	Fe3_	55.847	0.000	0.64	-	
Aluminum - Al	0.00	0.00	0.00	Al_	26.982	0.000	0.64	-	
Ammonia - NH4 (as NH4)	0.00	0.00	0.00	NH4_	18.038	0.000	0.95	-	
Hydrogen Ion - H	7.94E-09	0.00E+00	0.00	H_	1.000	0.000	0.95	0.00	
<b>Sum all cations (incl metals):</b>	<b>3,431.2</b>	<b>156.9</b>	<b>0.34</b>	<b>Sum Cations</b>		<b>194.377</b>		<b>3.15</b>	
Bicarbonate - HCO3	576.00	9.44	0.06	HCO3_	61.017	9.440	0.95	0.22	
Carbon Dioxide - CO2 (calc'd)	5.55	0.25	0.00	CO2_	44.010	0.504	0.82	0.00	
Carbonate - CO3 (calc'd)	10.87		0.00	CO3_	60.009	0.724	0.82	0.00	
Bromide - Br		0.00	0.00	Br_	79.904	0.000	0.95	-	
Chloride - Cl	4848.00	136.74	0.48	Cl_	35.453	136.744	0.95	3.18	
Fluoride - F	52.64	2.77	0.01	F_	18.998	2.771	0.95	0.06	
Sulfate - SO4	974.00	20.28	0.10	SO4_	96.058	40.559	0.82	0.20	
Nitrate - NO3	94.70	1.53	0.01	NO3_	62.005	1.527	0.95	0.04	
Phosphate - PO4 (as PO4)	7.10	0.22	0.00	PO4_	94.971	0.673	0.64	0.00	
Hydroxide - OH	2.14E-02	0.00E+00		OH_	17.000	0.001	0.95	0.00	
<b>Sum Total Anions:</b>	<b>6,568.87</b>	<b>-171.0</b>		<b>Sum Anions</b>		<b>192.944</b>		<b>3.71</b>	
Silica (as SiO2)	-	0.00		SiO2_	60.085	0.000	0.82	-	
Difference meq/l	-14.04	-4.28%							
<b>TDS</b>	<b>mg/L</b>	<b>10000.0</b>		<b>Ionic Strength</b>		<b>0.19366</b>			
				<b>Osmotic Pressure</b>				<b>100.82</b>	

Calculation By:	Name	Date
Checked By:	R. Huehmer	1/15/2010
	J Gomez	1/16/2010

**Figure B.9 Osmotic Pressure Calculation for High-Salinity Brackish Concentrate**

Osmotic Pressure Calculation		
Project	Assessment of Osmotic Mechanisms	
Location	El Paso, WWTP Effluent	
Temperature	°C	17.7
pH (valid from 2 to 12)		7.9
<b>Ionic Strength</b>		0.02398
<b>Osmotic Pressure</b>	psig	10.92


Osmotic Pressure calcs	
R_Gas	0.000082054
T_Kelvin	290.85
RT	0.023865406
Epsilon	0.5
sqrt(IS_Treated)	0.154862987

ION / Species	mg/L	mEq/l	Molecular		Ionic Strength	ActCoef	PI - atm Feed
			Valence	Weight			
Sodium - Na	330.00	14.35	1	22.990	14.354	0.98	0.34
Potassium - K	12.79	0.33	1	39.098	0.327	0.98	0.01
Calcium - Ca	65.38	3.26	2	40.080	6.525	0.92	0.04
Magnesium - Mg	13.02	1.07	2	24.305	2.143	0.92	0.01
Strontium - Sr	-	0.00	2	87.620	0.000	0.92	-
Barium - Ba	0.00	0.00	2	137.340	0.000	0.92	-
Ferric Iron - Fe+++ (as Fe)	-	0.00	3	55.847	0.000	0.82	-
Aluminum - Al	0.00	0.00	3	26.982	0.000	0.82	-
Ammonia - NH4 (as NH4)	0.47	0.03	1	18.038	0.026	0.98	0.00
Hydrogen Ion - H	1.26E-08	0.00E+00	1	1.000	0.000	0.98	0.00
<b>Sum all cations (incl metals):</b>	<b>421.7</b>	<b>19.0</b>	Sum Cations		23.375		<b>0.39</b>
Bicarbonate - HCO3	185.37	3.04	-1	61.017	3.038	0.98	0.07
Carbon Dioxide - CO2 (calc'd)	3.73	0.17	-2	44.010	0.339	0.92	0.00
Carbonate - CO3 (calc'd)	0.97		-2	60.009	0.064	0.92	0.00
Bromide - Br		0.00	-1	79.904	0.000	0.98	-
Chloride - Cl	278.0	7.84	-1	35.453	7.841	0.98	0.18
Fluoride - F	12.8	0.67	-1	18.998	0.673	0.98	0.02
Sulfate - SO4	294.0	6.12	-2	96.058	12.243	0.92	0.07
Nitrate - NO3	-	0.00	-1	62.005	0.000	0.98	-
Phosphate - PO4 (as PO4)	4.12	0.13	-3	94.971	0.390	0.82	0.00
Hydroxide - OH	1.35E-02	0.00E+00	-1	17.000	0.001	0.98	0.00
<b>Sum Total Anions:</b>	<b>778.99</b>	<b>-17.8</b>	Sum Anions		24.590		<b>0.34</b>
Silica (as SiO2)	35.00	1.17	-2	60.085	2.330	0.92	0.01
Difference meq/l	1.24	3.36%	Ionic Strength		<b>0.02398</b>		
<b>TDS</b>	mg/L	1235.6	Osmotic Pressure				<b>10.92</b>

Calculation By:	Name	Date
Checked By:	R. Huehmer	1/15/2010
	J Gomez	1/16/2010

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## **Appendix C    Review Comments**

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P.O. Box 13231, 1700 N. Congress Ave.  
Austin, TX 78711-3231, www.twdb.state.tx.us  
Phone (512) 463-7847, Fax (512) 475-2053

April 6, 2011

Nicole Sulzen  
Contracts Manager  
CH2M Hill, Inc.  
9191 South Jamaica Street  
Englewood, CO 80112-5946

Re: Contract between the Texas Water Development Board (TWDB) and CH2M Hill, Inc. (CH2M), TWDB Contract No. 0804830852, Draft Report Comments

Dear Ms. Sulzen:

Staff members of the TWDB have completed a review of the draft report prepared under the above-referenced contract. ATTACHMENT I provides the comments resulting from this review. As stated in the TWDB contract, the CH2M will consider incorporating draft report comments from the EXECUTIVE ADMINISTRATOR as well as other reviewers into the final report. In addition, the CH2M will include a copy of the EXECUTIVE ADMINISTRATOR'S draft report comments in the Final Report.

The TWDB looks forward to receiving one (1) electronic copy of the entire Final Report in Portable Document Format (PDF) and six (6) bound double-sided copies. The CH2M shall also submit one (1) electronic copy of any computer programs or models, and, if applicable, an operations manual developed under the terms of this Contract.

If you have any questions concerning the contract, please contact Dr. Saqib Shirazi, the TWDB's designated Contract Manager for this project at (512) 463-7932.

Sincerely,

Robert E. Mace, Ph.D., P.G.  
Deputy Executive Administrator  
Water Science and Conservation

Enclosures

c: Saqib Shirazi, Ph.D., TWDB

Our Mission	:	Board Members		
To provide leadership, planning, financial assistance, information, and education for the conservation and responsible development of water for Texas	:	Edward G. Vaughan, Chairman	James E. Herring, Member	Joe M. Crutcher, Member
	:	Jack Hunt, Vice Chairman	Thomas Weir Labatt III, Member	Lewis H. McMahan, Member
	:	Melanie Callahan, Interim Executive Administrator		



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*The CH2M HILL research team appreciates the detailed and thoughtful comments provided by the reviewers. Comments herein include those delivered as an attachment to the April 6, 2011 comment letter from the Texas Water Development Board on the draft report as well as subsequent comments on the pre-final draft. Significant drafting and reorganization of the report have resulted in changes to the pagination; therefore, page numbers in the comments may not match the final version. Responses to the comments are published in italic font.*

Attachment to April 6, 2011 TWDB comment letter

**Assessment of Osmotic Mechanisms Pairing Desalination Concentrate and Wastewater Treatment  
TWDB Contract No. 0804830852**

The study was performed to evaluate the feasibility of recovering fresh water from wastewater using a hybrid forward osmosis / reverse osmosis system.

The reviewers of the Draft report found several deficiencies in the report, which are described below for your consideration and inclusion in the final report. Additionally, the reviewers provided general and specific comments to aid in revising the report.

**Deficiencies:**

1. Plotting erroneous data:

- a. Table 4-1 shows the composition of synthetic brackish water solution at 10 g/L. Based on the data, several key experiments were conducted, and the results of the experiments are plotted from Figures 4-5 to 4-12.

The reviewers performed an analysis (using the Reverse Osmosis System Analysis software) to verify the data plotted in Table 4-1. The analysis indicates that the actual concentration of the synthetic brackish water (plotted in Table 4-1) is 15 g/L, instead of 10 g/L as mentioned in the report. Since the concentration of the synthetic brackish water is shown incorrectly in the report, the results that are plotted from Figures 4-5 to 4-12 are questionable.

*Values are being confirmed and reported as used in the testing for data assurance; text revised as appropriate*

2. Lack of appropriate descriptions of experimental and model development procedures for several processes:

- a. The report does not clearly describe how the reverse solute flux was measured for the experiments in this study.
- b. The report does not mention clearly how ammonia rejection was measured in the study.
- c. The report does not describe clearly how the forward osmosis model was developed.
- d. The report provides a confusing description on the process of measuring water flux through the forward osmosis system.

*The equations used were added into an expanded experimental protocol section to address all of these comments; text revised as appropriate.*

3. Insufficient and/or inaccurate explanations for major findings:
  - a. The report does not properly describe why a non linear phenomenon was observed when high concentration of NaCl and synthetic brackish water were used as draw solutions in the bench scale system (Page 45).

*One can speculate that this is a result of internal concentration polarization at the membrane; however, no specific experiments were conducted to isolate other potential factors. The FO modeling work uses actual experimental data rather than equations. Text revised as appropriate.*

To explain the fact why water fluxes during the DI-SW experiment were higher than those observed in the DI-BW experiment, the report states that the behavior is attributed to the higher concentration of NaCl in seawater compared to brackish water (Page 46). The explanation conflicts with the information shown on Table 4-1 which lists a higher NaCl concentration for brackish water compared to seawater.

*The information in Table 4-1 was revised. At a fixed TDS, seawater typically has higher osmotic pressures due to the higher ratio of NaCl in seawater relative to brackish water. This is one reason why ASTM methods cannot be used to normalize RO performance in estuarial or littoral waters. So, for an equivalent TDS, water with an ionic balance consistent with seawater will have a higher osmotic pressure than most brackish waters. Text revised as appropriate.*

- b. The reason that the report provides for explaining higher specific solute flux for DI-SW experiment at 30 g/L DS in the bench scale system is not appropriate because the same reason could be true for 10 g/L and 20 g/L DS concentration (Page 46).

*The DI-SW flux was higher than DI-BW in all cases. We acknowledge this. Text revised as appropriate.*

The report does not explain why fresh water flux was much lower than expected at 60 g/L DS in the pilot experiments (page 71).

*The report, on page 71, Table 4-13 clearing indicates in the notes the reason why this occurred. Text revised as appropriate.*

- c. The report does not describe why ammonia was not rejected at 20 g/L seawater as DS when SPLT was used as feed (Page 52).

*This is not described, as we do not wish to speculate upon the causation.*

- d. To explain the results plotted in Figures 4-9 and 4-10, the report describes that the feed water quality has minimal effect on the FO water flux (page 48). The statement is inaccurate and needs to be revised.

*The statement refers to feed waters tested, which have an osmotic pressure of between ~8 and ~11 psi. The statement is appropriate for the work conducted; however, the text was revised to clarify.*

According to the Equation 2-1 [ $J_w = k_w(\sigma\Delta\pi - \Delta P)$ ]; the water flux is directly related to the osmotic pressure difference between the feed and the draw solution. The osmotic pressure of the feed, in turn, is dependent on the water quality.

- e. To explain the results plotted in Figures 4-11 and 4-12, the report describes that the reverse solute diffusion into the feed is almost independent of the draw solution used (page 51). The statement is inaccurate and needs to be revised.

According to the Equation 2-2 ( $J_s = k_s\Delta C$ ); the solute flux is directly related to the concentration gradient of solute between the feed and the draw solution.

*The point is taken by the authors. However, it is the intent of the authors to communicate, that without reasonable uncertainty (i.e. for engineering purposes) that the specific draw solution used in the experiments.*

- f. To explain the reason for adding approximately 20 liter of brine every day during the pilot experiments of FO, the report ignores the consequence of approximately 300 liter of fresh water that is added to the system through osmosis of impaired water.

*Draw solution is lost via reverse diffusion and was added during the experiment. Text revised as appropriate.*

#### 4. Confusing statements and description:

- a. The first sentence in page 38 states, 'Figure 3-5 indicates the equivalent flux...'. The next sentence states, 'The membrane flux shown in Figure 3-5 ...' and Figure 3-5 shown at the bottom of the page plotted draw solution concentration Vs. water flux. The description in the text did not clarify the relationship between the terms, 'equivalent flux', 'membrane flux', and 'water flux'.

Additionally, the second sentence in page 38 states that 'The membrane flux shown in 3-5 for lower osmotic potential is very low.' However, Figure 3-5 plots draw solution concentration vs. water flux. The text does not describe the relationship between the osmotic potential and draw solution concentration.

*Text revised as appropriate.*

- b. Section 4.4.2 plots the results of the effects of turbidity on FO flux. However, the description in the first paragraph in Section 4.4.2 is confusing because the paragraph states that the main objective of this experiment was to evaluate inorganic fouling during forward osmosis of impaired water. Additionally, instead of providing a description of inorganic fouling; the paragraph provides information on organic fouling.

Please address the objective of the experiment clearly in the first paragraph of Section 4.4.2.

*Text revised as appropriate.*

- c. The terms 'impaired water', 'feed water', 'brine' and 'reject water' are used interchangeably in the description of the pilot system operation in Section 4.5.4. It is difficult to identify each term separately from the description.

*Text revised as requested. The literature, depending on the specific application, routinely uses these terms as synonyms where appropriate.*

- d. Line 6 of the last paragraph in page 58 states that 'stage 1 permeate flow rate had to be greater than the flow rate of water diffusing through the FO membrane...'. Line 10 in the same paragraph states 'water flux through the FO membrane was calculated from the accumulation of RO permeate in the permeate tank'. If RO permeate flow rate was greater than the FO flow rate, how the FO flux was then calculated using RO permeate. The process for measuring FO flux needs to be described clearly in the report.

*Text revised to explicitly indicate how values were calculated.*

- e. The first sentence of the second paragraph in Page 59 states that pilot-scale experiments were conducted with either tertiary treated feed water or the same stream dosed with bentonite or activated sludge. The statement is confusing because the report does not contain any experimental data for the tertiary treated feed water with bentonite or activated sludge.

*Text revised as appropriate.*

- f. Second sentence of the second paragraph in Page 59 states that all experiments with MBR effluent were conducted with a draw solution concentration of 35 g/L or 60 g/L synthetic sea salt. The statement is inaccurate because Figures 4-20 through 4-24 plot the results for the experiments that were conducted with seawater draw solution having concentration of 25 g/L TDS, and Figures 4-25 through 4-29 plot the results for the experiments that were conducted with seawater draw solution concentration of 30 g/L TDS.

*Additional data sets were developed. Text revised as appropriate.*

## 5. Major Editorial Problems:

- a. ES.2 Guide to This Document (Page 2): This is a fairly complex report. There are a lot of paths that stop abruptly at some point and resume a number of pages later. Consequently the idea of a guide to the document is a very good one. However, this guide falls short on execution. This is not much more than a list of the chapters, which already exists in the Table of Contents.

*Text revised as appropriate.*

- b. Section 2 in the report is not properly organized. The paragraphs are not linked to each other, discussion within a paragraph jumps from one topic to another frequently. The proper organization of the section will help the readers understand the topics clearly.

*Text reorganized and/or revised as appropriate.*

- c. A significant portion of discussion in Section 2.2 is not related to the rest of the report.

*Text reorganized and/or revised as appropriate.*

- d. The pilot scale forward osmosis / reverse osmosis hybrid system is described in two different sections in the report; in Section 4.1.2, and in Section 4.5.1. Please consider combining the descriptions of these two sections into one.

*Text reorganized and/or revised as appropriate.*

- e. Section 4.3.2 describes the pilot scale experimental procedure for K<sub>2</sub>SO<sub>4</sub>-NaCl experiments. However, the report does not provide any experimental results for the processes described in these two sections.

*Text reorganized and/or revised as appropriate.*

- f. Several incomplete sentences are present in the report;
  - i. Page 59, 2<sup>nd</sup> paragraph, 4<sup>th</sup> sentence ('The FO membrane was replaced twice during the.')
  - ii. The last sentence of the foot note 'b' of Table 4-13 (Page 71).
  - iii. Page 72, 2<sup>nd</sup> paragraph (Section 5.1 Forward Osmosis Model Development), second sentence ('The model assumes counter-current flow into ----- upon the initial').

*Text reorganized and/or revised as appropriate.*

The second paragraph in Section 7.1 is redundant to the last two sentences of the first paragraph.

*Text reorganized and/or revised as appropriate. Second paragraph eliminated.*

## 6. Scope of Services:

The following task of the contracted "Scope of Services" has received minimal attention in the report:

- Compare cost models with existing reclamation costs obtained from literature

*At the end of the cost modeling section, we have added an additional section bench-marking the data.*

*Text revised as appropriate.*

**General Comments:**

- Please update the report's front matter regarding TWDB board members and executive administrator. Please refer to
- <http://iweb/Agency/Executive/Admin/boardmembers.asp>
- Please use a standard font size and format for all the equations used in this report.
- Fonts in equations should match the text of the report.
- Ionic constituents of several Tables in Sections 3-2, 3-3, and 4-12 lack proper valence. Please make the correction.

*Text revised as appropriate.*

**Specific Comments:**

*Section 2.1, Fundamentals of Forward Osmosis (Page 6 -12)*

- a. Figure 2-1: Please re-label different solutions shown in the Figure. For the feed to RO to be labeled "brine" is unusual and for the product from RO to be labeled "feed" is atypical. This difficulty could be avoided by designating the left cylinder as "low osmotic pressure solution" and the right cylinder "high osmotic pressure solution" or just "low concentration" and "high concentration." This nomenclature should be correct across all processes.

*Figure revised.*

- b. Page 7, 1<sup>st</sup> paragraph, last line: Please correct the spelling of the word 'centrate'.

*No typographical error exists. The cited work refers to "Forward osmosis for concentration of anaerobic digester centrate".*

- c. Last sentence in Page 7: Please add the word, 'hydraulic' between the words 'not' and 'pressure'.

*Text revised as appropriate.*

- d. Second line in Page 8: Please explain the concept of hydrophilic membrane.

*Text revised as appropriate.*

- e. Last paragraph in Page 8: Please elucidate the validity of the finding of the other researchers where they identified that the theoretical minimum energy at 50% recovery is  $1\text{kWh/m}^3$ .

*The claims regarding a minimum theoretical energy have been properly referenced from multiple sources. It is not in our scope of supply to specifically review and validate the work conducted by other researchers. No change made.*

Section 2.1.1 General Theory of Forward Osmosis (Page 9 -10)

- a. Equation 2-1: Please define the term 'reflection coefficient'.

*Text revised as appropriate.*

- b. Last paragraph of Section 2.1.1 (Page 10): Please add in the discussion that the principal reason why water flux is lower than anticipated is that the values typically used for  $\Delta\pi$  are those in the bulk solution rather than those at the surfaces of the membranes for which the equation was derived. The explanation that follows will make more sense to readers unfamiliar with this particular technology if "concentration polarization" is defined when it is first introduced. The first paragraph in 2.1.2 is important information, but it is out of place, diverting attention away from concentration polarization.

*Text revised and reorganized as appropriate.*

Section 2.1.2, Impacts of Concentration Polarization (Page 10 - 12)

- a. Third paragraph (Page 10): Please define the term 'concentrative ICP' before discussing the impact of concentrative ICP in the FO process.

*Text revised and reorganized as appropriate.*

- b. Third paragraph in Page 11: This paragraph explains Figure 2-5. Please mention clearly that the figure is a schematic representation of the FO process. Additionally, please mention that the primary reason for running wastewater next to the active "dense" layer is to prevent contamination of the membranes and to ease in cleaning out the elements.

*Text revised as appropriate.*

- c. Second sentence, last paragraph, Page 11: The statement "salt diffusion . . . is driven by concentration gradient . . . not by pressure gradient" is questionable. It is certainly driven by concentration gradient, but unless a very good reference can be cited that indicates ion transport is unaffected by pressure gradient or by interaction with water flow, please consider leaving off the part about the pressure gradient.

*Technically, most authors are defining FO as conceptually occurring at atmospheric pressure. In practical engineering terms, we know that some pressure is required to drive the hydraulics. We concur that there is a small pressure impact – however the pressures involved are significantly lower order of magnitude than the osmotic pressure, and for most practical applications are insignificant. A footnote was added for clarification.*



- d. Last paragraph of Section 2.1.2 (Page 12): Please clarify the statement, “By examination ----- FO membrane”.

*Text revised as appropriate. The sentence has been edited to better reflect the concept of concentration polarization of the non-linearity of  $dPI$  vs  $dflux$ .*

Section 2.2, Fundamentals of Reverse Osmosis (Page 12-16)

- a. Figure 2-6 (Page 13): Please provide a courtesy credit for the photo.

*Courtesy credit for the photo has been added.*

- b. Line 7, Page 15: Please replace the sentence, “ $K_w$  = water mass transfer coefficient (see Eq 2-6)” with the sentence, “ $K_w$  = water mass transfer coefficient (see Eq 2-9)”

*Text revised as appropriate.*

- c. Equation 2-8 (Page 15): Please maintain consistency in identifying different parameters through the symbols. The previous equations identified the solute flux as  $J_s$ , instead of  $F_s$ .

*Text revised as appropriate.*

Section 3.1, Desalination in Texas (Page 19-23)

Please provide updated information on the desalination facilities in Texas.

*Text revised as appropriate using the updated data as provided by the TWDB.*

Section 3.2.1, Kay Bailey Hutchison Desalination Plant, El Paso, TX (Page 24-25)

Line 2, 2<sup>nd</sup> paragraph: TDS concentration of concentrate stream (3,600 mg/L) seems too low. Please check the source data and provide a correct value for the concentrate stream.

*The concentrate stream concentration shown is consistent with data provided to us by EPWU.*

Section 3.2.3, Regional Desalination Plant, Brownsville, TX (Page 26-27)

Table 3-4: Please add a blank line between min of one set and avg of the next to separate the parameters more clearly. Also, please line up the avg, max and min with the first three rows of numbers.

*Text revised as appropriate.*

Section 3.2.5 (Page 30 – 32)

Please correct the spelling of Corpus Christi in the Title of Section 3.2.5.

*Text revised as appropriate.*

Section 3.4, Characterization of Osmotic Potential of Various Waters in Texas (Page 36 – 38)

a. Table 3-14:

- Please change the term 'primary wastewater' to 'primary effluent of wastewater'; and 'secondary wastewater' to 'secondary effluent of wastewater'.
- Please provide the source (the name of the wastewater treatment plant) of the raw wastewater, primary effluent of wastewater and secondary effluent of wastewater.
- Please provide the source of the high salinity brackish water.
- Appendix B (Page 107) shows that the TDS concentration of primary wastewater is 1079 mg/L. Please correct the value on the Table.

*Text revised as appropriate.*

- Table 3-15: Please consider changing the Title of the second column from 'waste solution' to 'wastewater'.

*For consistency, Impaired Water has been adopted.*

- The values of the osmotic potential show that the El Paso WWTP Effluent (avg) (not the secondary wastewater) was used as 'waste solution' for estimating the osmotic potential. In the second column of the Table, please replace the term 'secondary wastewater' with the term 'El Paso WWTP Effluent (avg)'.

*Text revised as appropriate.*

- b. Figure 3-4: The plot provides an equation for the straight line. At 0 Instant Ocean Concentration, the osmotic pressure should be 0; therefore, the straight line should pass through the point of origin (0, 0).

*The model calculates an approximation of PI. We have forced the line through the origin and eliminated the equation on the figure.*

- c. First sentence, first paragraph, Page 38: The phrase "resulting in very high capital equipment costs" sounds like it was drawn from experience. It would be more accurately phrased as "which would result in very high capital equipment costs."

*Text revised as appropriate.*

Section 4 Performance of Spiral Wound Forward Osmosis Membranes (Page 39)

Since this section covers experiments both with plate-and-frame membranes and spiral wound elements, the title seems either inaccurate or misleading.

*The section title revised to read "Performance of Forward Osmosis Membranes"*

Section 4.1.1 Bench-scale FO System (Page 39)

- a. Because this section describes an experimental procedure, please provide a little more description on the membrane and how it was installed. The description of the membrane may include information on the active surface area of the membrane, membrane thickness, symmetry (homogeneous or asymmetrical), and information on the spacer and the support layer used in the membrane. Please mention if the membrane is commercially available.

*Text revised as appropriate.*

- b. Please replace the word '3-L' with the word, '3 liter'.

*Text revised as appropriate.*

Section 4.1.2 Pilot-scale forward osmosis/reverse osmosis hybrid system (Page 40 – 41)

- a. In this section, please provide detailed description about the spiral wound membrane. This description would include the diameter and length of the element, how many leaves it contains, and the active area of membrane per leaf and total. The description of the internals of the element should be understandable to someone familiar with the construction of a spiral wound RO element. In the spiral wound configuration, chambers alternate between draw solution chambers (shown in Figure 4-3) and donating, which are called "feed" chambers (very similar to the feed/reject chambers in an RO element). Presumably, there is a spacer of some sort in both types of chambers. In assembling the element, are the boundaries of the draw solution chambers glued on assembly as the boundaries of a product chamber in an RO element would be and are the ends of the donating chambers open? Is the center barrier in the draw solution chambers formed in the same way as the boundaries?

*Much of the detailed membrane information is considered proprietary by the manufacturer. The nominal information for the membrane has been provided. Published manufacturer data for RO membranes, in comparison, do not include number of leaves, active area per leaf, not the internal construction of the element. The remaining information requested is not required to achieve the scope of report. No change made.*

- b. Figure 4-3: The text says in several places that flows are counter-current but the drawing shows them as co-current.

*The figure clearly indicates that it is not representative of the actual membrane elements used. No change made.*

Section 4.2 Solution Chemistry (Page 42)

Table 4-1: There's a nomenclature problem here. The idea is acceptable that mixtures of Instant Ocean with various amounts of water, above 35 g/L can represent seawater and its concentrates. But to call a 10 g/L solution "seawater" just gives the wrong picture. One needs an abbreviated way of saying it, and SW is a good abbreviation for water with an ion balance characteristic of that in seawater. So maybe 10

g/L SW works, or maybe 10 g/L Instant Ocean or even 10 g/L IO. Given the alternatives, please avoid the use of “seawater” when referring to such a solution.

Section 4.3.1, Bench-scale Experimental Procedure (Page 43 – 44)

DI water feed – NaCl draw solution experiments: second line, add the word ‘40’ in between ‘30’ and ‘50’.

*No experiments were conducted at a concentration of 40 g/L. No change made.*

Section 4.4 (Page 45 – 54):

- a. To help the readers in understanding the results of the experiments accurately, please consider making the following changes:
  - Please consider changing the Titles of the sub-sections:
    - Please change the Title of the Section 4.4. from ‘Results’ to the ‘Results of Bench-scale FO System’
    - From ‘Performance testing of RO Osmosis elements’ to ‘Performance testing of FO elements using BW and SW as draw solutions’
    - From ‘Ion Analysis for the DI-SW experiments’ to the ‘Rejection criteria of FO elements’
    - From ‘4.4.2 Bench-scale tests’ to the ‘Effect of turbidity on FO performance’
    - From ‘Rejection of ammonia’ to ‘Rejection of ammonia from secondary effluent’
    - From ‘Rejection of other constituents’ to ‘Rejection of ammonia from SPLT’
  - Please consider adding the following sub-section numbers in front of the revised Titles;
    - 4.4.1 Performance tests with DI water feed and NaCl draw solution
    - 4.4.2 Performance testing of FO elements using BW and SW as draw solutions
    - 4.4.3 Rejection criteria of FO elements
    - 4.4.4 Effect of turbidity on FO performance
  - Please consider dividing the performance testing of RO Osmosis elements (which will be re-named as ‘Performance testing of FO elements using BW and SW as draw solutions’) into three sub-sections:
    - Performance of FO using DI as feed
    - Performance of FO using secondary effluent as feed
    - Performance of FO using SPLT as feed
  - Please consider dividing the ‘Ion Analysis for the DI-SW experiments’ (which will be re-named as ‘Rejection criteria of FO elements’) into two sub-categories
    - Rejection of dissolved organic matter
    - Rejection of ammonia

*Text revised as appropriate to address all these comments.*

- b. Figures 4-4 to 4-12: FO process is driven by osmotic pressure difference between the feed and the DS. Plotting the data as the DS concentration vs. flux barely provides any information on determining the appropriate cause for variation in the flux. To identify the effect of feed and

draw solution on the flux in FO process, it is highly recommended that the data are plotted as osmotic potential (i.e., the osmotic pressure difference between the feed and the DS) vs. flux.

*Text and figures revised as appropriate.*

- c. Figure 4-4: Please explain the reason in the Text, why  $J_s$  does not increase in proportion to the draw solution concentration.

*Text revised as appropriate.*

- d. First paragraph, Page 46: To aid the readers in interpreting the results completely, please provide a Table showing specific solute fluxes for DI-BW and DI-SW experiments.

*Figures revised as appropriate.*

Figure 4-7: In the Title of the figure, please mention the name of the facility from where the secondary effluent was collected.

*Text revised as appropriate.*

- e. First paragraph, Page 48: Please provide an explanation why the calculated specific solute flux values were higher for the DI-BW experiments at all draw solution concentrations.

*There was a typographical error in this sentence. On data review, the DI-SW flux was greater, due to higher dPI. Text revised as appropriate.*

- f. Second paragraph, Page 48: Please provide a Table showing the constituents of SPLT.

*The SOW did not include use of river water as a feed and all data and reference to SPLT have been eliminated from the draft report. Text revised as appropriate.*

- g. Last paragraph, first line, Page 48: The words, 'Figure 4-13' should be replaced with the words 'Figure 4-10'.

*This section of the report was eliminated in its entirety at CSM's request as it represents out of scope work.*

- h. Second paragraph, Page 51 (Rejection of dissolved organic matter by FO): Please discuss why the rejection was not measured using SPLT as feed.

*Work on SPLT water is deemed by CSM as out of scope and reference is being eliminated.*

- i. Tables 4-2 and 4-3 (Page 51): These tables indicate that under the conditions studied most organic carbon, 97% or more, is removed. The implication is that the water extracted by FO is "pure" to use the description that appears several times in this report. The problem is that this may not be correct. A study funded by the WateReuse Foundation currently being concluded

has shown that while most dissolved organic compounds are highly rejected in RO, some specific compounds are not. One can expect the same conclusion for FO. Please address the issue in the report.

*Text revised to clarify, as appropriate. As this study did not address fractionation and examination of the organics, and the report referenced to by the commenter does not address FO, we cannot report this claim. We have no data in the study to confirm or deny the statement.*

Section 4.5.1 Test System (Page 55-56)

- a. Third sentence, last paragraph, Page 55: The sentence states, “The three membranes ..... in the second pass.” Should the words, ‘the second pass’ be deleted?
- b. Figure 4-17: Please show the direction of solute flux (through a broken arrow) in the FO system.

*Text revised as appropriate.*

Section 4.5.4 Pilot system Operation (Page 58 – 59)

- a. Page 58, second paragraph (4.5.4 Pilot system Operation), last sentence: Please revise the sentence to read ‘Average properties of impaired water streams used as feed to the forward osmosis sub-system is summarized in Table 4-9’.

*Text revised as appropriate.*

- b. In the third paragraph of Page 58, please mention that the process is schematically described in Figure 4-17 and the operating parameters of RO sub-system is plotted in Table 4-8.

*Text revised as appropriate.*

Section 4.5.5 Pilot Experiments (Page 59)

- a. Table 4-10: Please replace the parameter ‘NO3’ with  $\text{NO}_3^-$ .

*Text revised as appropriate.*

- b. Table 4-11: Please clearly mention in this section of the report how the osmotic pressures of different DS solutions were measured.

*Text revised to indicate that the osmotic pressure was calculated – not measured.*

Section 4.5.7 Long term FO performance test with MBR effluent FO feed (Page 60)

- a. First sentence of the last paragraph in Page 60: Usually, the flux that is declined due to fouling does not increase automatically without cleaning the membrane. Therefore, the only reason for the initial flux decline could be due to the adjustment of operating conditions.

*The figure has been replaced.*

- b. In Figures 4-21, 4-26, and 4-31, please replace the terms 'back pressure' with the term 'feed pressure'.

*Text revised as appropriate.*

- c. Figure 4-22: In the Title, please spell out the term, 'DS' and add the term in the abbreviation list.

*Text revised as appropriate.*

- c. Figure 4-28: In the legend, please replace the term 'Feed' with the term 'Impaired Water'.

*Text revised as appropriate.*

- d. Figure 4-32: In the legend, please replace the term 'Reject Flow' with the term 'DS'.

*Text revised as appropriate.*

- e. Figure 4-33: Please add a legend for 'Feed'.

*Text revised as appropriate.*

- f. Figure 4-34: Please delete the words, 'A1' from the Title of the y-axis.

*Text revised as appropriate.*

- g. In the Text, please explain the reason for wide variation of draw solution flow for the 60-hours pilot testing with MBR effluent as FO feed and 60 g/L sea salt as DS.

*Text revised as appropriate.*

Section 5.1 Forward Osmosis Model Development (Page 72 – 75)

- a. Figure 5-1 (Page 72): Please provide a narrative description of various steps and logics used in this flow chart.

*Text and figure revised as appropriate.*

- b. Figure 5-2 (Page 74): Please explain the information provided in Figure 5-2.

*Text and figure revised as appropriate.*

- c. Figure 5-4 (Page 77): Please explain why the colors of the leading elements are different than the rest of the elements.

*A footnote has been added to clarify that the colors represent the placement of high rejection and lower rejection elements in a single pressure element.*

- d. Figure 5-7 (Page 80): Please add notation to the various components of this Figure.

*Text revised as appropriate.*

- e. Table 5-1: Please clarify the term 'Recovery/Rejection' that is present in the last row of the Table because recovery and rejection are completely two different terms.

*Text revised as appropriate. This was a typographical error that has been corrected to define both recovery and rejection appropriately.*

Section 5.4 Model Output (Page 80)

- a. The CH2M Hill model may be appropriate for RO; it is clearly no worse than the Toray model. However, it is not shown to be appropriate for FO.

*Thank you for the comment*

- b. Figure 5-8 (Page 81):
  - Please define various terms (NDP, Feed Q, Perm Q, Bypass Q, Conc Q) used in this Figure.
  - Please explain why the feed TDS is such an odd but precise value. Additionally, please describe why the dP of element 1 is so large compared to all the others. Please identify if recovery of element  $n$  is a percentage of feed to element  $n$  or to element 1.

*The original figure was generic – which has since been replaced. The specific issues have been addressed. Text revised as appropriate.*

Section 6 Development of a Forward Osmosis /Reverse Osmosis System Cost Model (Page 82 – 95)

Please consider deleting the words 'for a given application' from the last sentence of the first paragraph in Page 82.

*Text revised as appropriate.*

Section 6.1.1 Capital Cost Development (Page 82 – 86)

- a. Figure 6-2: Please define the term CD.

*Text revised as appropriate.*

- b. Figure 6-3: Please provide the units of Building Footprint.

*Text revised as appropriate.*

- c. Figure 6-4: In the Title, please include a statement that says that the cost curve was prepared excluding the equipment cost.

*Text revised as appropriate.*

- d. Figure 6-6:



- Please spell out 'I&C' at the first instance.
- Before showing the graph of capital cost curves for other engineering disciplines, please explain various impacts of other engineering disciplines on the cost of the project.

*Text revised as appropriate to include expanded discussion on costs.*

Section 6.1.2 Operations and Maintenance Costs (Page 86 – 87)

Please mention if the membrane and cartridge filter replacement cost for FO was same as RO.

*Text revised as appropriate.*

Section 6.2 Reverse Osmosis Cost Model Development (Page 87 – 90)

Please explain the cost implication of the assumption that the RO system will be based upon a seawater RO system.

*Given the osmotic pressure of the diluted draw solution, a brackish water system is not appropriate in terms of pump sizing or materials of construction. The assumption is valid; estimating cost implications associated with the assumption is beyond the scope of this research.*

Section 6.2.2 Operation & Maintenance Cost (Page 90)

Figure 6-10: x-axis is missing from the figure.

*The figure has been revised.*

Section 6.3 Comparison of FO-RO to Advanced Wastewater Treatment (Page 90 - 95)

- a. Given the O&M cost for a given capacity and the amortized capital cost (using a reasonable interest rate) for the same capacity, it is possible to calculate the life cycle cost per thousand gallons of product water. These would be very useful numbers, but they do not appear in the report.

*Text revised as appropriate to discuss estimated life-cycle costs.*

- b. What is missing in this analysis is optimization of the draw solution concentration. Apparently in order to make the FO process look cheap, a high draw solution concentration is used. The effect of this is to make the RO system very expensive. As the draw solution concentration goes down, the cost of FO increases and the cost of RO decreases. What the potential user is interested in is the total cost. This is the stuff that optimizations are made of. It should be possible with the analytical tools that were used on this project to calculate the cost of water for various draw solution concentrations at some fixed capacity. If lower draw solution concentrations work out, then the concentrate from brackish plants like those that a lot of space in Section 3 was devoted to, may be useful.

*Optimization of the draw solution concentration is beyond the scope of this research.*

Section 6.3.3 Comparison of FO/RO versus AWWTP (Page 93 – 95)

Figures 6-13 and 6-14: Please explain the reason for showing the cost estimate of standalone FO process in the Figure because a standalone FO does not produce any good water.

*Standalone FO processes may well not provide “good” water, but can be used to minimize waste volumes of other waters. We believe that this is very relevant.*

Section 7 Conclusion (Page 96 – 97)

Somewhere in the conclusions, the problem brought up in Figure 2-5 and the surrounding text needs to be brought up. It seems like economic suicide to lose more than half of the available driving force to concentration polarization.

*We have added additional narrative on this topic. In discussions with manufacturers, control of ICP is a complicated issue that may not be adequately resolved.*

Section 7.1 Mechanics of forward osmosis-reverse osmosis treatment (Page 96)

- a. The last sentence of the first paragraph of Section 7.1 states that simulated seawater or seawater concentrate was used as the draw solution. The statement is not accurate because the results in Section 4 show that seawater and brackish water were used as draw solutions.

*Based upon streams available in Texas, no municipal brackish water was identified that has sufficient osmotic pressure to serve as a reasonable draw solution. As a result, the conclusion that seawater or seawater concentrate be used is a reasonable conclusion, based upon the scope of work.*

- b. Third paragraph, Section 7.1, Page 96: The statement provided in this paragraph is not sufficient to explain the process. Please clearly mention that three different feeds (DI, secondary effluent and SPLT) were used to determine the performance of FO elements.

*Reference to SPLT has been removed. The conclusions have been elaborated and refined.*

- c. First paragraph (Page 97): Before including a discussion in the conclusion on the use of ammonia as a draw solution, please discuss the option in the Text of the report.

*Text revised as appropriate. Additional discussion of various draw solutions has been added into Section 2. Only work with saline streams has been conducted as part of this study.*

Section 7.3 Cost-effective application of forward osmosis to wastewater treatment with desalination concentrate as a draw solution (Page 97)

Page 97, last paragraph, second line: Please replace the word ‘and’ with the word ‘an’.

*Text revised as appropriate.*

Section 8 Recommendations (Page 98)

Last line in third bullet: Please add the word 'forward osmosis' between the words hollow-fiber and membrane.

*Text revised as appropriate.*

References (page 100 – 101)

- a. Cath et al. (2008) is not present in the reference section.
- b. Please provide the URL address for Henthorne L. (2007) reference. The URL is:  
<http://www.usbr.gov/pmts/water/publications/reports.html>
- c. Reference is made to the Water Desalination Report, 2008. There were 45 issues of WDR issued in 2008. It would be a courtesy to the reader to indicate the number or date of the one of these that contains the information about the Gibraltar Desalination plant.

*Text revised as appropriate.*

Appendix B (Page 102)

Second paragraph, last sentence: Please replace the words 'Figure A1' with the words 'Figure B1'.

*Text revised as appropriate.*

Appendix C (Page 113)

There is no practical way to review the Table that is provided in Appendix C.

*No change made.*

**Subsequent Comments on the Pre-final draft**

1. Various editorial comments (e.g., fonts, spacing, figure numbering) throughout the document.

*Text revised as appropriate.*

2. If possible, add back Appendix B.

*Appendix B added back to the final report.*

3. Add an appendix with reviewer comments and responses to the comments.

*Appendix C added to the final report.*

4. Please cross-check all references and correct any inconsistencies.

*Text/citations revised as appropriate.*

5. Section 2.1: Define osmotic pressure.

*Text revised as appropriate.*

6. Figure 2.2: Label should read ammonia/ carbon dioxide rather than ammonium bicarbonate as in reference cited for the figure.

*This refers to ammonium bicarbonate draw solution which is made by dissolving ammonium bicarbonate salt ( $\text{NH}_4\text{HCO}_3$ ) in water. Ammonia and carbon dioxide are the decomposition gases of this ammonium bicarbonate solution. The process could be referred to as an "ammonia-carbon dioxide" FO process, but the solution is not. No change to the figure or text made.*

7. Figure 2.4: Add a reference.

*Reference added.*

8. Table 3.2: Replace data with updated values for the concentrate at KBH plant.

*Table 3.2 and Appendix B updated with new data provided by TWDB.*

9. Section 3.4: To obtain the reason for a non-linear flux, please provide a statement similar to the last paragraph of Section 2.1.2.

*Text revised as appropriate.*

10. Table 4.2: Column labeled BW Concentration at -15 g/L TDS" should be 10 g/L.

*Table revised.*

11. Figure 4.6: Should “Osmotic pressure, bar” be “osmotic pressure differential”?

*Figure revised.*

12. Page 45: The statement regarding “performance testing of FO elements” is confusing. Please consider deleting the statement.

*Text revised as appropriate.*

13. Page 47: Since the results of specific solute flux for DI-BW and DI-SW are confusing; please consider not discussing specific solute flux for the experiments.

*Text revised as appropriate.*

14. Page 47: Last sentence in the paragraph discussing Figures 4-11 through 4-13. The statement contradicts a basic equation:  $J_s = K_s DC$ . Please consider revising the statement.

*Text revised as appropriate.*

15. Page 50: Rejection of Ammonium, Add “Using Secondary Effluent as Feed” to title of section.

*Text revised as appropriate.*

16. Page 51: Rejection of Other Constituents, Delete “of other constituents” and replace with “Using South Platte River Water as Feed”

*Text revised as appropriate.*

17. Page 52: First paragraph of FO rejection of turbidity and flux effects, Should “deionized water feed” be “deionized water” or “deionized feed water”?

*Text revised as appropriate.*

18. Page 52: Both “colloidal fouling” and “inorganic fouling” are used in the discussion. Which is true?

*Text revised for consistency.*

19. Page 52: Please consider deleting final statement before Figure 4.15.

*Text revised as appropriate.*

20. Page 54 and Figure 4.18: Please explain why permeate is returned to the brine tank.

*Permeate from stages 2 and 3 was returned to the feed tank with the diluted FO permeate stream to facilitate the measurement and calculation of water flux and solute rejection in the FO process. For example, the final permeate from Stage 1 was equal to water flux in the FO subsystem. The purpose of the final RO stages was to supply and reconcentrate draw solution for the FO system. The text was revised to provide additional explanation.*

21. Section 4.5.4: The reviewers raised questions about the accuracy of 20 L brine for the process.

*Text revised as appropriate.*

22. Figure 5.3: Why are these two elements colored as light green?

*Footnote 5 added: Increasingly, the use of different models of membrane in a single pressure vessel is being adopted. By installing high rejection membranes in the lead elements, and lower rejection, higher productivity elements in the later positions, it is possible to produce better flux balancing within the vessel, with a slight reduction in energy consumption. The green and blue elements of the figure (if viewed in color) represent the higher and lower rejection elements, respectively.*

23. Figure 6.1: Where is the permeate stream?

*There are two streams coming from the "Draw solution re-concentration system", one is the recovered draw solution (RED), and the other is the permeate (BLACK).*