Reverse Osmosis Treatment of Central Arizona Project Water—Brine Minimization Via Vibratory Shear-enhanced Processing

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Abstract

Central Arizona Project water was treated using slow sand filtration (SSF) and reverse osmosis (RO) in series. RO brine was recovered using vibratory shear-enhanced processing (VSEP®; New Logic, Inc.). SSF removed 90% of the turbidity in raw CAP water. Although SSF decreased total organic carbon by almost 20%, there was essentially no change in dissolved organic carbon concentration. After more than a year of continuous operation, performance of the RO system declined noticeably, as indicated by a rapid decrease in membrane permeation coefficient and an increase in salt flux. Foulant scrapings contained both clay material and large amounts of unidentified organics. Alternative hypothesis regarding major sources of membrane foulants are under investigation.

Water lost as brine was reduced to 2-4% via post-RO VSEP treatment. Estimated costs were compared to those of a no-VSEP option in which disposal of the entire RO brine flow was required. The total annualized cost of brine treatment was fairly insensitive to recovery during VSEP treatment in the range 80-90%, and period of VSEP operation between cleanings in the range 25-40 hrs. These values define a fairly broad window for near optimal VSEP operation under the conditions of the study. The cost of VSEP treatment to minimize brine loss was estimated at $1970 - 2150 per acre foot of CAP water treated. For a hypothetical 3 MGD RO brine flow, the use of VSEP to recover water and reduce the volume of brine for disposal results in a savings of more than $5M/year relative to the no-VSEP brine disposal alternative.

Introduction

Southwestern cities are among the fastest growing and driest in the United States. Satisfaction of projected water demand in Las Vegas and Phoenix, for example, will depend on conservation, use of water of impaired quality such as wastewater effluent or brackish ground water, importation of water and groundwater overdraft. Arizona law mandates achievement of a rough balance between groundwater withdrawal and replenishment rates by year 2025 in "active management areas" around major population centers (Pearson, R., et. at., 1999). In the Tucson Active Management Area (TAMA; Figure 1) projected compliance is based on full utilization of regional rights to Central Arizona Project (CAP) water and a degree of renovation/reuse of municipal wastewater effluent. Uncertainty regarding the long-term availability of Colorado River water to the CAP, however, is a major impediment to water resources planning (Barnett and Pierce, 2008).

For illustration, we focus momentarily on the TAMA. Native ground water in accessible aquifers of the TAMA contains 200-300 mg/L total dissolved solids (TDS) (Arnold, 2008). Traditionally, these waters have been directly served to the public following disinfection. When CAP water reaches Tucson, it contains ~650 mg/L TDS, a figure that will rise with water demand in the upper Colorado River Basin states. Full use of its CAP allocation in the TAMA will bring at least 200,000 metric tons of salt to the Tucson area annually (Arnold, 2008). Because Tucson is the southern terminus of the CAP canal, without salt management steps, essentially none of that salt will leave the region, and the average salinity of accessible TAMA ground waters will double over a 50-year period (Arnold, 2008).

The addition of CAP water to the TAMA water resource portfolio (Figure 1) has already increased TDS levels in delivered water, and membrane treatment necessary to maintain average TDS levels near 450 mg/L is under consideration. While essential to salt management, however, reverse osmosis (RO) will consume energy and produce brine. It has been estimated that recovery during RO treatment of CAP water will be limited to 75-80% to avoid membrane scaling (Malcom Pirnie and Separation Processes, 2008). If even a third of the regional CAP allotment is treated in this manner, the value of water lost as brine will be on the order of \$20 M·yr^{-1}, based on a unit water value of \$1000 per acre foot. Further, inland communities like Tucson have no ready sink for RO brines, so that disposal costs will add significantly to the cost of RO treatment.
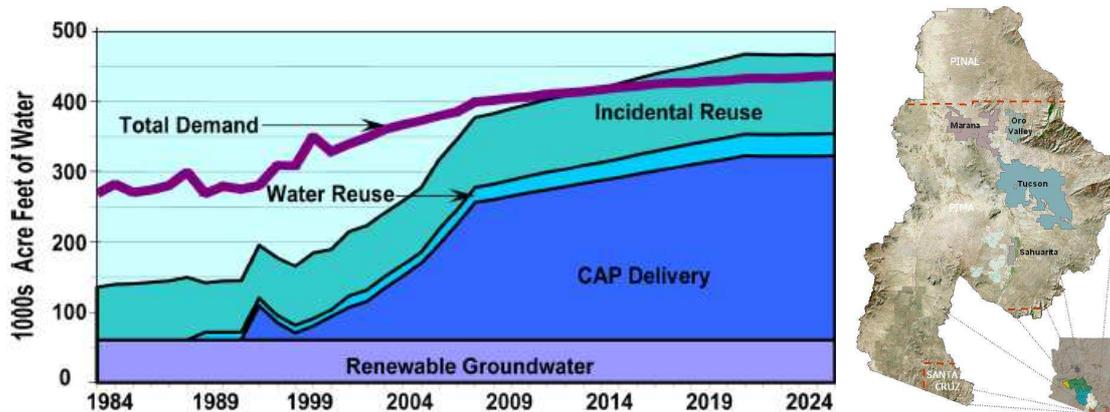
Figure 1. The water supply/demand projections for the TAMA. The geographic boundaries of the TAMA are shown in the map at right.

Strategies to minimize brine volume include (i) pre-RO softening to remove divalent cations that contribute to membrane scaling (so that greater recoveries can be achieved during RO treatment), (ii) two-stage RO interrupted by brine softening, and (iii) innovative technologies for water extraction from RO brines without softening steps. Softening processes are not considered here. We describe instead pilot-scale RO

desalination of CAP water followed by water extraction from brines using the vibratory shear-enhanced process (VSEP®; New Logic, Inc.). Pilot-scale data are developed and used to find economically optimal conditions for VSEP operation. The net benefits of VSEP treatment, relative to RO alone, are then calculated under optimal operating conditions. The sensitivity of results to assumed cost parameters is considered.

Materials and Methods

Pilot-scale facilities, located about 20 miles north of Tucson on I-10 (Figure 2), for desalination of CAP water consisted of slow sand filtration (SSF), chemical addition to prevent membrane scaling and fouling, reverse osmosis and VSEP treatment of RO brine. Design data for unit operations follow in abbreviated form (Table 1). A process schematic is provided (Figure 3).


Figure 2. The location of the field site

Table 1: Design Information for Pilot Plant

Reverse Osmosis	
Number of stages (-)	2
Number of vessels (3 elements per vessel)	6
Membrane area (sqft per element)	28
Water flux (gfd)	10.9
Flow rate (gpm)	4.74 (17.94 L/min)
Feed salinity (μS.cm^{-1})	1000-1100
Feed pressure (psi)	80
Recovery rate (%)	80.5
Slow Sand Filters	
Product flow per filter (gpm)	8.25
Filtration rate (gal/min.ft^2)	0.110 (0.27 m/h)
Filter area (sqft)	75
Number of filters (-)	2
Initial height of filter sand bed (ft)	3.0
Minimum height of filter sand bed (ft)	1.5
Sand uniformity coefficient, d60/d10	1.5
Height of underdrains, including gravel layers (ft)	3.0
Height of supernatant water (ft)	5.5
Free board (ft)	1.0
Total filter basin height (ft)	12.5
Vibratory Shear Enhanced Processing	
Conductivity of brine (μS.cm^{-1})	3000 – 5000
Operating pressure (psi)	500
Vibration (Hz)	52.0 – 52.5
Flow rate (gpm)	~1.0
Recovery rate (%)	75 – 90
Open valve period (min)	0.1
Closed valve period (min)	1.0 – 6.9

Two SSF units, each capable of providing filtrate at > 5 gallons per minute (gpm), were required for periodic filter cleaning without interrupting RO operation. Hydraulic control points were at filter outlet valves, which were adjusted to produce filtered water at a rate slightly greater than 5 gpm, the target influent flow rate for the downstream RO

unit. The operational strategy was designed to minimize excess SSF filtrate production and associated chemical demand.
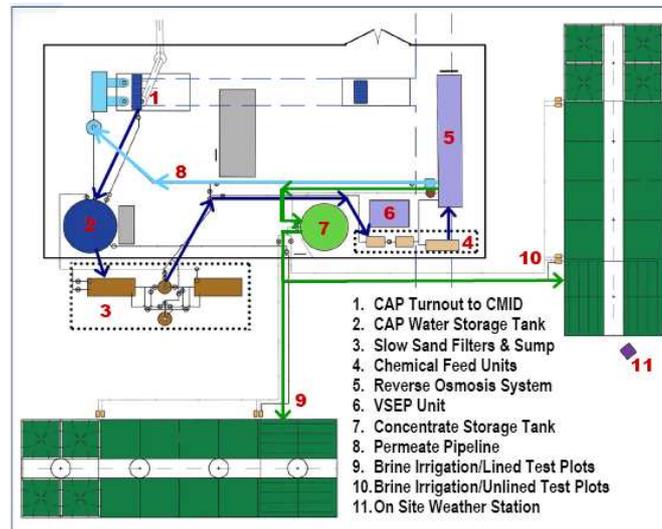
**Figure 3.** Site plan and water flow diagram for the Tangerine Road Field Site.

Table 2. Chemical feed operations.

Chemical	Source	Dosing Criteria (mg/L)	Dose (mg/L)
Antiscalant	Ecolab Flocon	3.6	0.4
Sulfuric Acid	J.T. Baker (Emos 9684-05)	pH = 7	15
Sodium Hypochlorite	Chlorex	1.0-1.5 (combine α chlorine)	1.4
Ammonia (ammonium sulfate)	Fisher A720 2-10 (technical grade)	Excess NH_3	2 (as NH_3)

Chemical addition points are shown in the facility schematic. Target chemical doses (Table 2) were selected to maintain a combined chlorine concentration of 1 - 2 mg/L and free chlorine level < 0.1.mg/L RO influent pH was maintained in the range 6.6 - 6.9, and antiscalant selection and dose were per manufacturer's recommendations. Bleach was added ahead of the 800-gallon filtrate storage tank to provide adequate

mixing and contact time prior to RO treatment. Ammonia, acid and antiscalant were added to the 300-gallon chemical mixing tank, from which water was recirculated to ensure adequate chemical dispersal ahead of the RO unit.

The RO treatment unit consisted of six 2.5-inch, 3-element pressure vessels (18 total membrane elements) in a 2:2:1:1 array (Figure 4). Feed pressure was maintained at 80 to 90 psi. The feed flow rate was ~5.0 gpm. Spiral-wound 2450, NF90 membrane elements (DOW, Film Tec) were selected based on the expected CAP water TDS concentration (650-700 mg/L). The unit was operated at 80 percent recovery, for an average permeate flux of 12.6 gfd (the assumed membrane area per element was 25.4 ft^2; DOW). The design water flux for this membrane is 12 gfd. Water fluxes in upstream and downstream elements were matched as nearly as possible via manipulation of permeate back pressures in respective pressure vessels. Refer to Appendix A for the full cleaning procedure of RO unit.
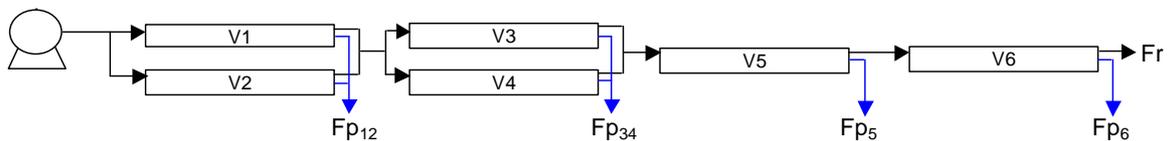
Figure 4. Schematic view of the vessel arrangement for RO

The VSEP apparatus was a commercially available pilot scale unit LP Series (Lab Pilot Scale; New Logic), containing 16.44 ft^2 of ESPA1. The feed flow (RO brine) was provided at 500 psi after preliminary testing to select an operating pressure. The unit was operated by automatically cycling the brine retention valve between its closed and open positions (Figure 5). With the valve closed, fluid left the reactor only as permeate. When open, brine was briefly flushed from the unit and replaced with reactor influent

(RO brine). During each cycle, the valve was opened (flush position) for six seconds.
The length of the closed valve period was calculated to yield target permeate recoveries
during VSEP operation. In general, permeate flow rates were inversely related to VSEP
recovery and elapsed time of operation following membrane cleaning.
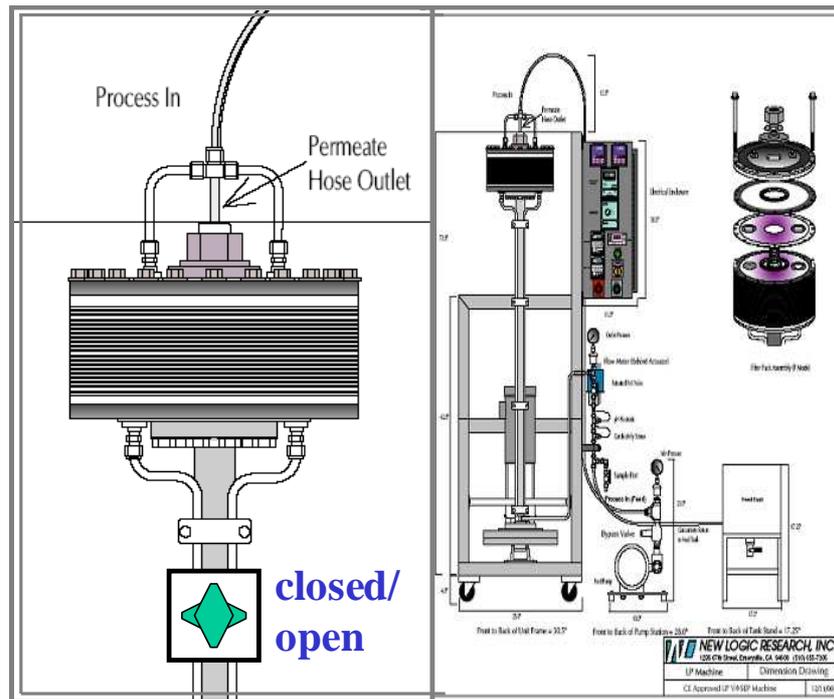
Figure 5. Schematic view of VSEP and the operation of brine retention valve
Refer to Appendix B for full cleaning procedure for VSEP unit.

Parameters measured included (among others) SSF influent turbidity and
conductivity, turbidity and silt density index (SDI) in SSF effluents, conductivity and
TDS concentration in the RO permeate and brine (VSEP influent), volume rate of flow
and conductivity in the VSEP permeate, and the volume and conductivity of the VSEP
reject water. Turbidity was measured daily with a Hach Turbidometer 2100P calibrated
between 20 NTU and 800 NTU. The instrument detection limit was estimated at 0.1
NTU, close to values that were generally obtained in SSF effluent.

The SDI provides a useful indicator of particle content in water that is prepared for RO treatment, overcoming the relative insensitivity of standard turbidity measurements. Samples for SDI measurements are passed at 30 psi through a 0.45-μm membrane filter. Particle accumulation reduces the filtration rate over the 15-minute standard test period. Flow data yield a plugging factor, defined as the percentage decrease in the flow rate that passes through the filter compare to the original

SDI values are calculated as follows:

$$\text{SDI} = \frac{PF}{t}$$

where,   SDI= silt density index (-)

PF= plugging factor (%)—the time-dependent percentage decrease in flow rate

t= time (minutes)

The SDI instrument was from Chemtec (Model FPA-3300). RO membrane warranty generally depends on maintenance of SDI < 3.0 in RO influent, although values below 5.0 are adequate to protect RO membranes from fouling due to influent particle accumulation (www.wateranywhere.com, 2009).

For waters of the lower Colorado River, TDS concentration is related to conductivity by:

$$\text{TDS} = 0.62 \, (1 + 0.00459 EC^{0.5}) \, EC \qquad \text{(Moody, 2002)}$$

where,   EC is conductivity ($mS \cdot cm^{-1}$)

TDS is total dissolved solid concentration (mg/L).

Reverse osmosis hydraulic performance is described in terms of a temperature-adjusted water transport coefficient (A) defined by:

$$A = q [0.5 (P_f + P_r) - P_P - \Pi_C + \Pi_p]$$

where, q is the permeate flux (Q_P/S)

Q_P is the volume rate of flow of permeate

S is the nominal membrane surface area

P_f, P_r and P_P are the pressures in the unit feed, reject and permeate streams, respectively, and Π_C and Π_p are the osmotic pressures in the unit concentrate and permeate.

Calculated values were adjusted to 20°C using:

$$A_{20} = A \times 1.026^{(20-T)}$$

where, T is temperature in °C

The salt transport coefficient (B) is defined as:

$$B = q_s C_P/(C_C - C_P)$$

where, q_s is the flux of salt through the membrane ($g/m^2.sec$)

C_P is the permeate salt concentration (mg/L)

C_C (mg/L) is the effective salt concentration on the feed (reject) side of the membrane. That is,

$$C_C = C_f \frac{\ln(C_r / C_f)}{[1 - (C_f / C_r)]}$$

C_f= salt concentration in the unit feed solution (mg/L)

C_r= salt concentration in the unit reject flow (mg/L)

Salt transport coefficients were adjusted to 20°C using:

$$B_{20}= B \times 1.026^{(20-T)}$$

Permeate flows in the VSEP reactor were similarly adjusted for temperature. That is, flows are reported as equivalent permeate flows adjusted to 20 or 25 °C, as indicated.

Results—Slow Sand Filtration.

The SSF influent and effluent pH profiles (Figure 6) indicate that the pH of CAP water was stable at ~8.0 throughout the 3-year study period and was unaffected by the filtration process. As marked, although regular, seasonal temperature variation (7-35 °C) was apparent in the same record of measurements. As expected, TDS levels (460-750 mg/L) in delivered CAP water were not affected by the SSF process. Based on calculated mean values over the 3-year study (n ~ 10^3), SSF removed 90% of the SSF influent turbidity. SSF effluent turbidities all fell between 0.12-1.29 NTU. Influent turbidities, on the other hand were highly variable (Figure 7), with spikes in the range 0.4-21 NTU.
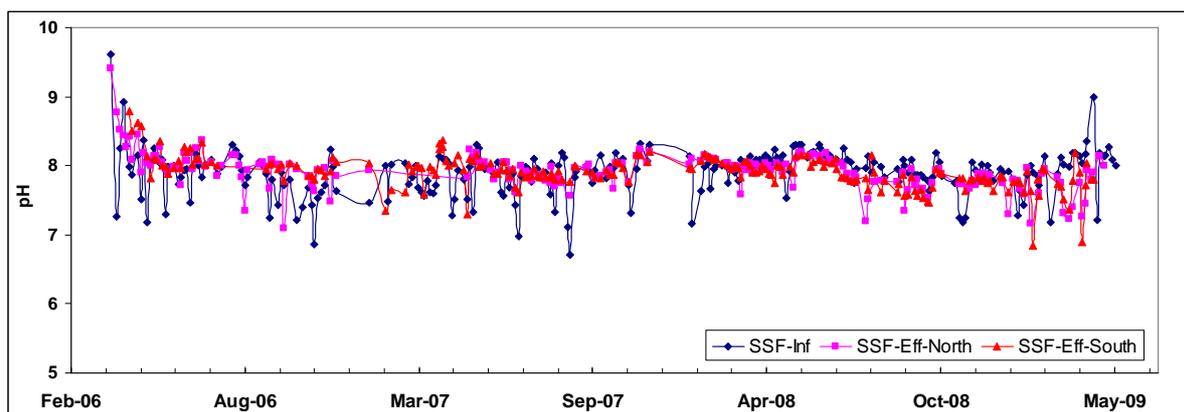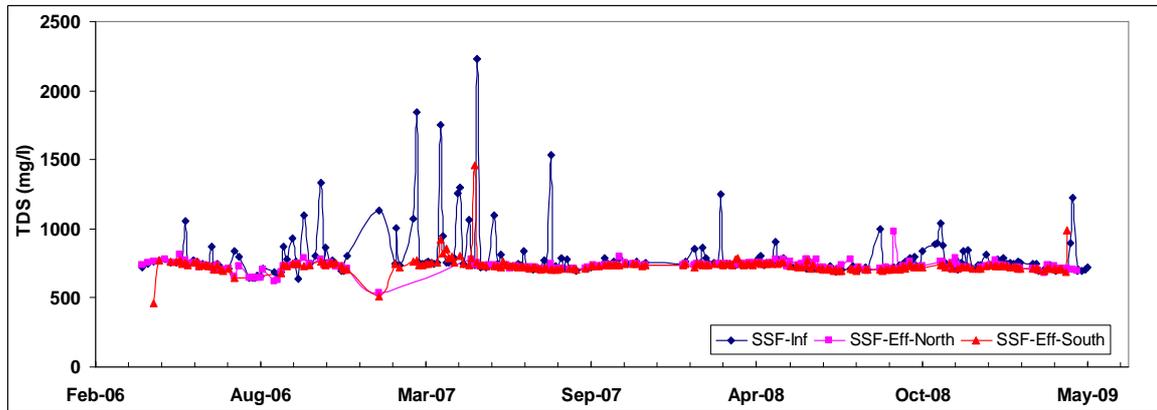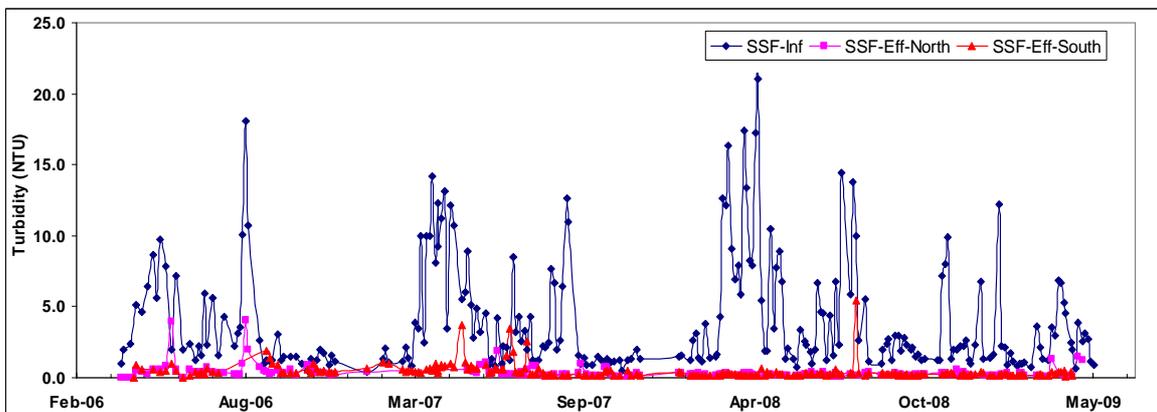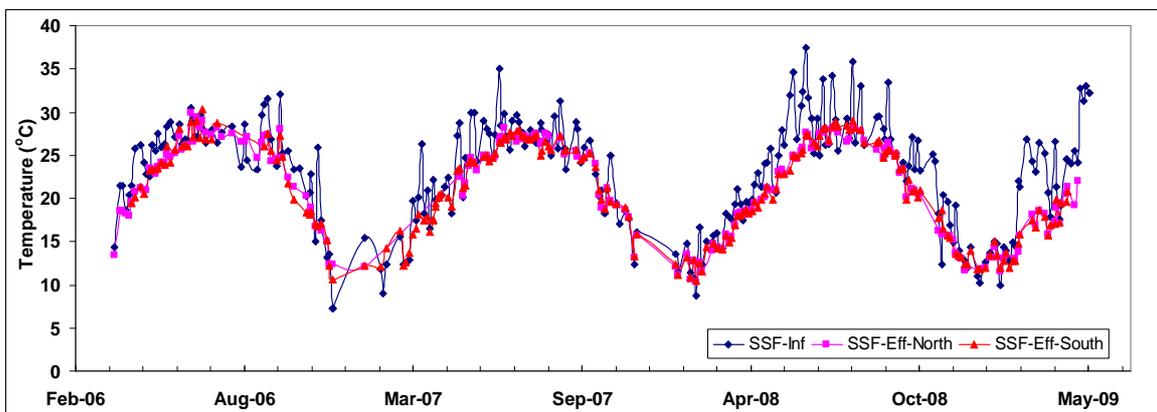

Figure 6a. SSF influent and effluent pH profiles

Figure 6b. SSF influent and effluent temperature profiles

Figure 6c. SSF influent and effluent total dissolved solids profiles
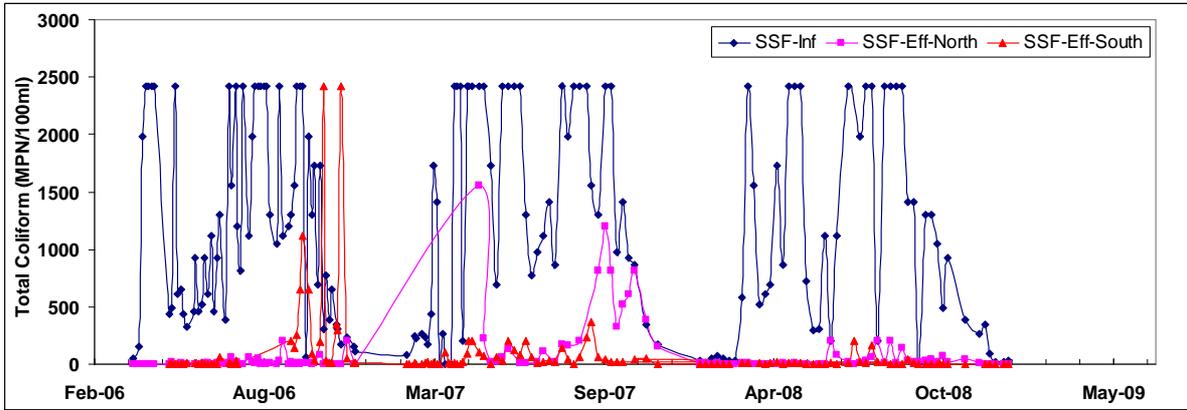

Figure 7. SSF influent and effluent turbidity profiles

Barium and calcium ion concentrations, which are important from the perspective of scale formation during RO treatment, were generally parallel to TDS concentrations in SSF influent and effluent (Table 3) (Moody, 2002). However, while calcium levels were independent of time, barium increased gradually from 2001 to 2007 and remained reasonable stabile thereafter. The source of the increase in barium concentration is not known with certainty but may be associated with barium leaching from agricultural soils in the upper Colorado River basin. SSF lowered the concentration of total organic carbon (TOC) but not dissolved organic carbon. Typical TOC removal was only 10-15% because dissolved organics were the primary component of TOC (Table 3). Total coliform counts, on the other hand, were much improved via SSF. Relatively high values that were occasionally observed in effluent from the southern SSF unit (Figure 8) were eliminated via filter redesign/construction at the beginning of 2007.

**Table 3.** Statistical summary of the major parameters measured for SSF evaluation

Parameter	Influent		North SSF Effluent			South SSF Effluent		
	Mean	*# of Samples*	*Mean*	*# of Samples*	*P-value*	*Mean*	*# of Samples*	*P-value*
pH (-)	**7.88**	256	**7.93**	203	2.6E-1	**7.92**	220	3.4E-1
Temperature (^{o}C)	**22.85**	255	**21.36**	200	7.0E-3	**20.83**	220	1.9E-4
Conductivity (μS/cm)	**1126.1**	255	**1046.8**	201	2.5E-6	**1049.9**	219	4.1E-6
TDS (mg/L)	**786.8**	247	**730.8**	195	1.3E-5	**736.0**	216	6.0E-6
Turbidity (NTU)	**3.97**	246	**0.36**	191	4.7E-32	**0.44**	211	1.7E-33
Mg^{+2} (mg/L)	**30.43**	168	**29.68**	133	4.6E-4	**29.98**	136	3.2E-2
Ca^{+2} (mg/L)	**74.99**	168	**70.76**	133	5.5E-9	**71.90**	136	2.4E-5
Ba^{+2} (mg/L)	**0.152**	140	**0.138**	37	1.4E-7	**0.150**	45	2.8E-1
Na^{+} (mg/L)	**101.77**	168	**99.64**	61	5.9E-2	**102.53**	64	4.5E-1
SO_4^{-2} (mg/L)	**255.7**	168	**252.3**	64	2.9E-1	**267.4**	64	1.2E-4
Cl^{-} (mg/L)	**93.71**	168	**92.17**	64	9.5E-2	**96.91**	64	7.8E-5
TOC (mg/L)	**2.84**	79	**2.58**	64	2.9E-5	**2.45**	47	7.1E-9
Total Coliform (MPN/100ml)	**1279.2**	162	**81.83**	131	4.0E-37	**94.17**	135	3.9E-36

Figure 8. SSF influent and effluent total coliform profiles

Results—Reverse Osmosis.

The time dependent water and salt transport coefficients for pressure vessels 1/2,

3/4, 5 and 6 show deterioration of performance over the period of study (Figure 9).
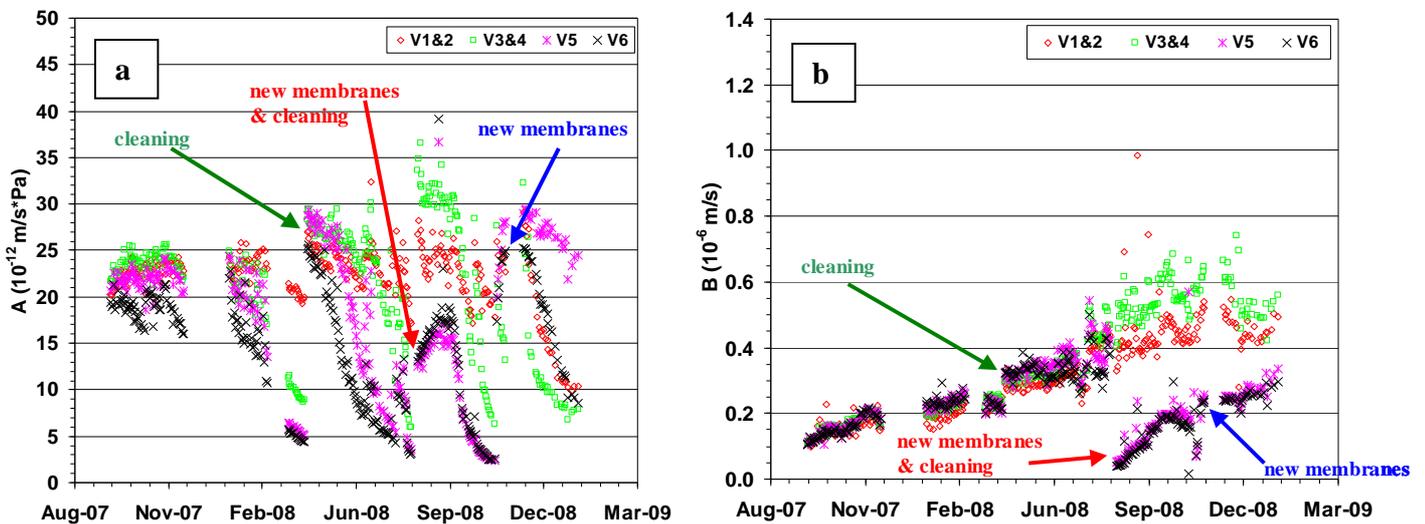
Figure 9: Water (a) and salt (b) transport coefficients of RO

Overall recovery was maintained as nearly as practicable to 80% by gradually

relaxing the backpressures and downstream brine discharge points ($F_{p1/2}$, $F_{p3/4}$, F_{p5}, F_{p6}

and F_r in Figure 4). A reasonable salt balance ($\pm$ ~10% and normally much less) was maintained across the RO unit throughout the period of study.
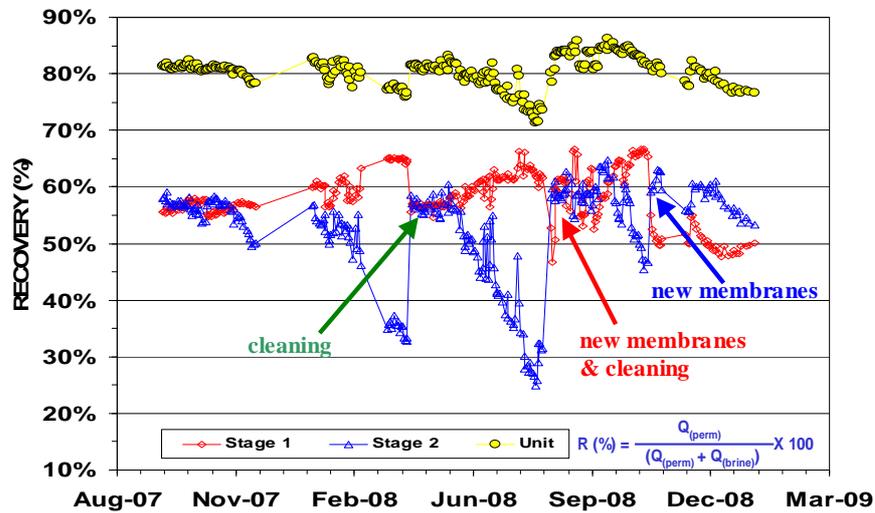
**Figure 9c.** Recovery rate for RO operation

Rapidly, declining water transport coefficients (A) were observed in all membrane stages after about 1 February 2008. As expected, effects were most severe in downstream pressure vessels V5 and V6, where ion concentrations and scaling potential were the greatest. There was a coincident increase in the salt transport coefficients (B) in all pressure vessels. When the mass flux of total dissolved solids in permeate from the unit as a whole reached 4% of the influent mass flux (mid March 2008), RO treatment was interrupted for membrane cleaning, leading to temporary restoration of membrane permeability to water and a modest discontinuity (increase) in the salt transport coefficient. Unfortunately, restoration of RO performance was followed by a second rapid loss of membrane permeability (to water) and a steady increase in salt transport over the four months following the cleaning procedure. Again, the deterioration of performance was most rapid in the downstream pressure vessels, suggesting that scaling

was the primary causal agent. In August 2008, the membranes in pressure vessels 5 and 6 were replaced with ESPA3 RO membranes (Hydranautics) and the entire RO reactor was again cleaned before service was restored. The new membranes proved to be much more resistant to water and salt transport. Membrane permeability improved during the first month of operations. Thereafter, however, the permeability of all membranes in all pressure vessels again decreased rapidly. Results suggest that the cause of declining membrane performance was reversible, through membrane cleaning or replacement, but that the source of difficulties was not eliminated, and the problem was quickly able to reassert itself. During the final four months of the record of performance, the reverse osmosis system was operated to produce 85 percent recovery, an operational condition that may have accelerated membrane scaling.

Results RO Membrane Autopsies

In July 2008, Membranes from the stage 2 RO units were removed for analysis of foulant or scalant and replaced with new membranes. In November, the stage 2 membranes were again replaced and suspect via direct visual observation, scanning electron microscopy (SEM) and energy dispersive X-ray (EDX) analysis. Material scraped from retired pilot-scale membrane surfaces was analyzed for culturable heterotrophic bacteria (heterotrophic plate count, HPC) and mass fractions of organic and inorganic carbon as indicators of a bacterial role in membrane fouling.

Micrographs corresponding to a clean membrane surface and a representative post mortem membrane fragment (pressure vessel 6, membrane element 3) show that the spent membrane was heavily fouled. The general background (gray) material was fletched with small whitish grains over much of the surface (Figure 10 a & b).
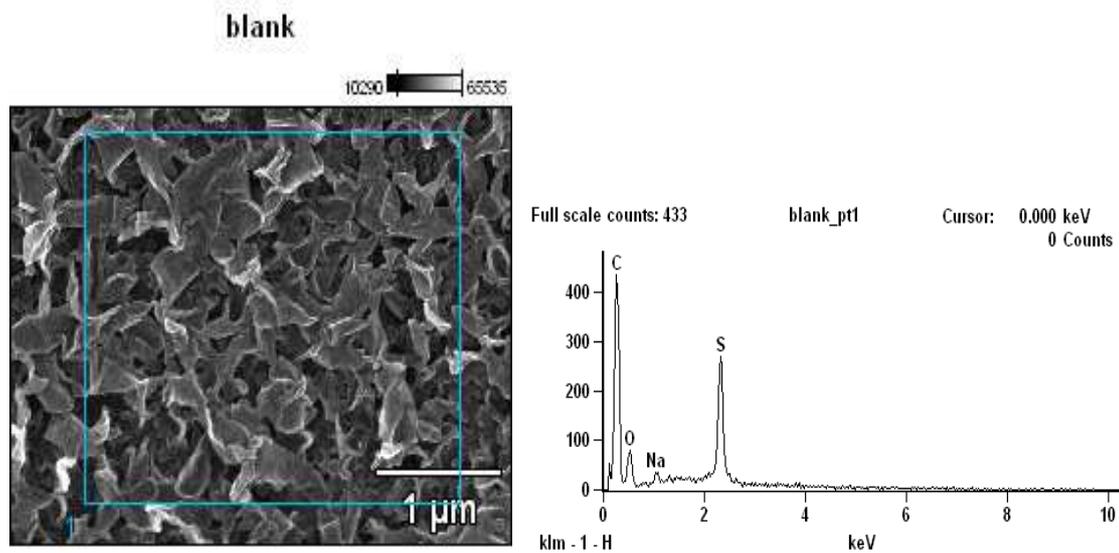
Figure 10a. SEM images and EDX analysis of control membrane
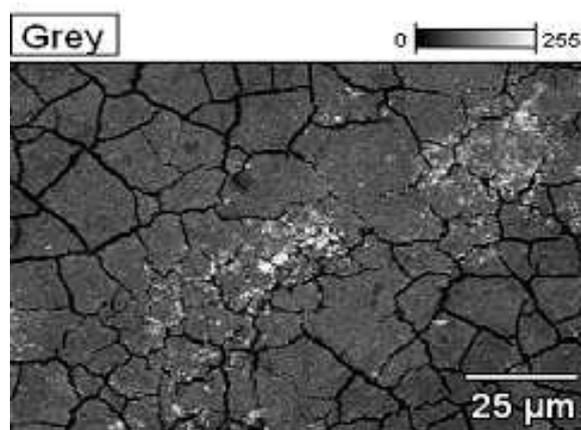
Figure 10b. SEM image of CAP water fouled RO membrane.

The elemental composition of the clean surface was demonstrated by carbon, sulfur and oxygen, conforming to expectations based on polymer structure. SEM and EDX analyses conducted on membrane fragments from vessel 5 and 6 were similar. EDX results suggest that the whitish spots consist of $BaSO_4(s)$. Major EDX peaks in the gray zones were dominated by silicon and aluminum, with noticeable quantities of carbon, oxygen and calcium (Figure 11).
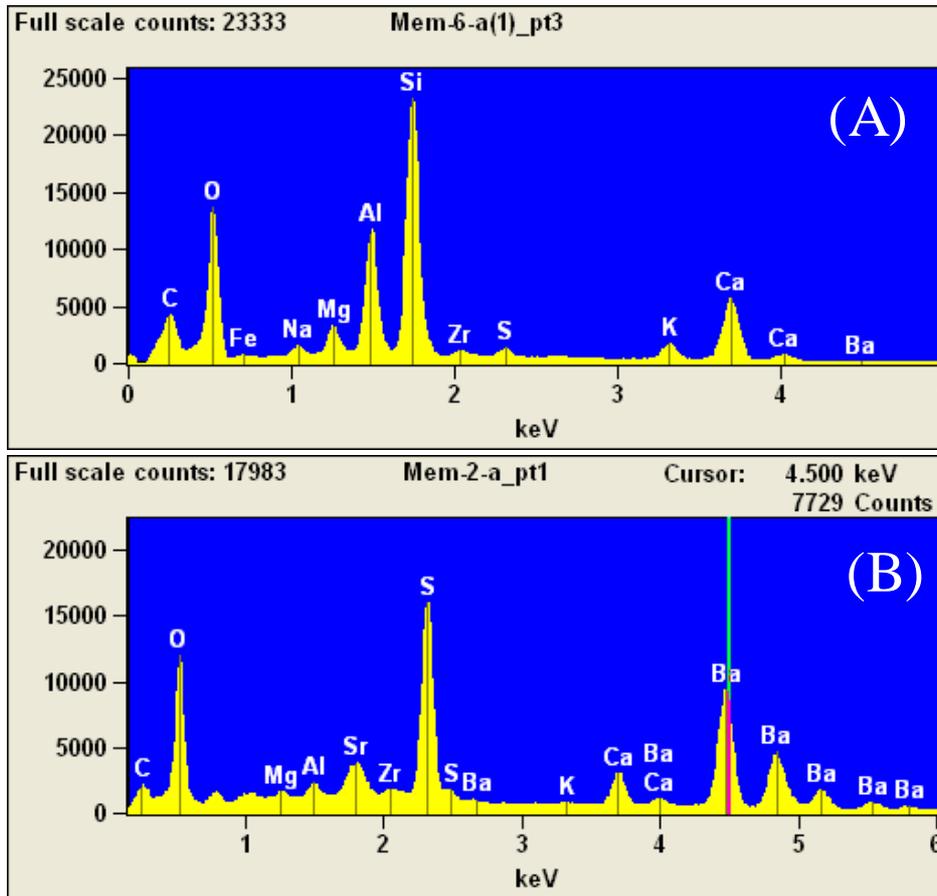
Figure 11. EDX peaks produced by the foulant in grey area (A) and the small white area (B) shown in the SEM image (Figure 10b).

Summary EDX reports containing relative amounts of the materials present (atom fraction as percentages) indicated that silicon-to-aluminum ratios were relatively constant from 2.0 to 2.4 (Table 4). The near – constant elemental ratio suggests that deposited clay particles were at least partially responsible for membrane fouling. However, Si/Al ratios in common Si-Al clay materials do no t fall immediately in this range (Table 5), so that a mixture of aluminum-silicate clays was probably present.

Table 4. Atomic percentage of elements in the foulant at surface of spent membrane

	C-K	O-K	Na-K	Mg-K	Al-K	Si-K	S-K	Cl-K	K-K	Ca-K	Fe-K	Zr-L	Ba-L
Mem4(1)pt1	37.09	43.50	0.77	1.23	2.70	4.63	2.61	0.29	0.37	5.64	0.90	0.20	0.07
Mem4(1)pt2	33.36	34.20	1.37	1.01	2.30	5.05	9.47		0.29	3.76	0.64	0.08	8.48

The broad-scale distribution of specific elements in micrographs (Figure 12) also provides diagnostic information. That is, Si, AL and K appear to be co-distributed and indicative of clay deposition. Finally, Ba an S are co-distributed and probably represent precipitated barium sulfate.
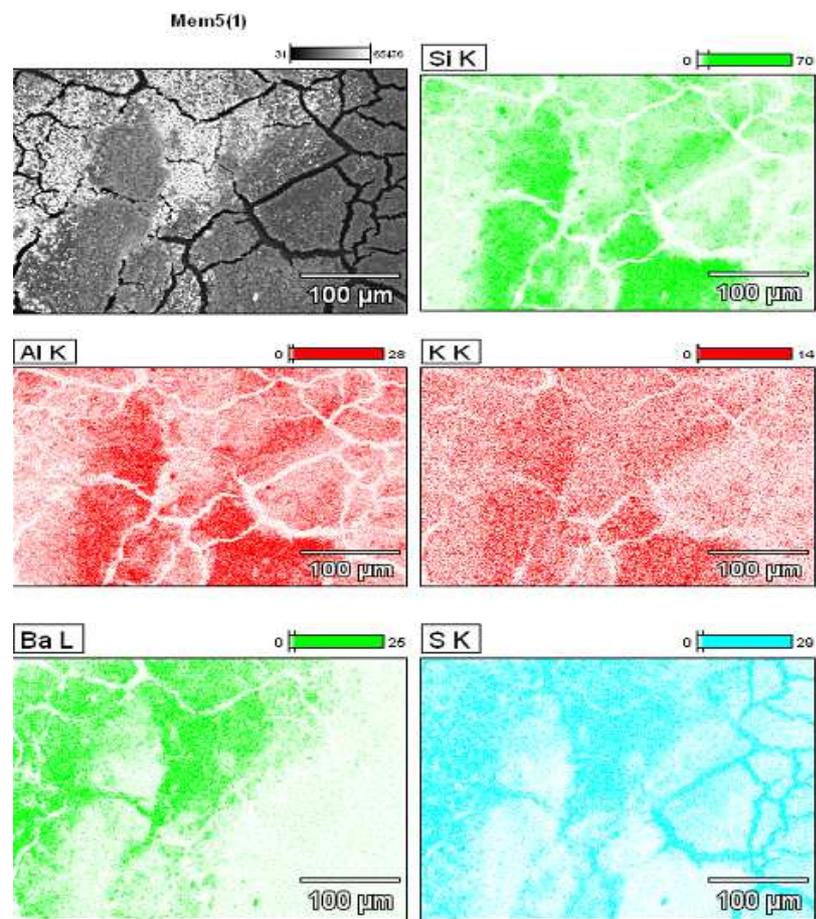
Figure 12. Distribution of critical elements on the membrane surface. Data from EDX scans.

Foulant that was scraped from the membrane surface was analyzed for mineral content via X-ray diffraction (XRD; Figure 14) The major parts in the figure was derived from silicon dioxide. Clays identified by SRD included potassium aluminum silicate and sodium calcium aluminum silicate. Neither calcite ($CaCO_3$) nor barite ($BaSO_4$) was detected. The sensitivity of the technique was estimated at 5% mass of analyte phase / total mass.
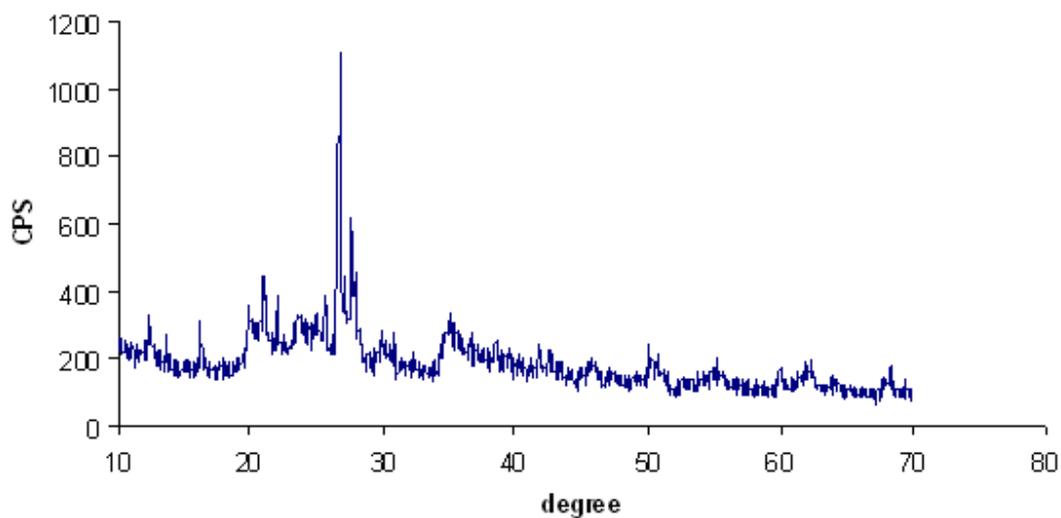
Figure 14. XRD peaks of the foulant scraped off the spent membrane.

Foulant scrapings contained, however, relatively high levels of organics and inorganics carbon (Table 5), suggesting that biomass, natural organic matter or even the anti-scalant chemical added was a primary contributor to membrane fouling. HPC measurements, on the other hand , were low. To illustrate, if the footprint of an attached bacterium is taken as 2 x 10^{-8} cm^2 (1μm x 2 μm), even the largest HPC result (5.1 x 10^{-4} cm^{-2}) would account for coverage of only 0.1% of the membrane surface. This is supported by virtual evidence (SEMs) since no recognizable bacterial morphologies were

immediately apparent and suggests that the source of organic carbon in scrapings was not culturable bacteria. Alternative hypothesis are under investigation.

Table 5. Heterotrophic plate counts and TOC/TIC content in the foulant samples scraped off the spent membrane.

Sample ID	TOC (%)	Total C (%)	TIC (%)	HPC (cells/cm^2)
V5-1	11.00	13.78	2.78	4.40×10^4
V5-2	ND	ND	ND	1.73×10^3
V5-3	11.30	14.29	3.00	ND
V6-3	ND	ND	ND	5.10×10^4

Results—VSEP Operation

The operating pressure for the VSEP reactor was selected based on a preliminary experiment in which the RO brine was treated at pressures from 200 to 500 psi. The highest stabile recovery pressure (500 psi) was selected for subsequent experiments. Since none of the pressures selected produced a declining permeate flow over a standard 100-minute observation period, the highest pressure tested was selected for reactor operation. The open-valve period selected (6 seconds) was sufficient to purge the feed (brine) chamber of the reactor.

To explore the operational capabilities of the VSEP reactor, the relationship between reactor performance and (i) water recovery (controlled by manipulating the length of the closed-valve period of operation cycles—see above) and (ii) operating time between cleaning steps on reactor performance was explored. Target VSEP recoveries ranged from 75 to 90%. At each recovery, reactor performance was assessed by measuring the time – dependent permeate flux (Figure 11) and the permeate and brine salinities (Figure 15) over 20 hours of continuous operation. Permeate fluxes gradually declined at each experimental recovery. The rate of decline was directly related to

recovery, so that there may exist an optimal recovery—dependent on the value of water saved, VSEP capital and operating costs, and the cost of residual brine disposal.
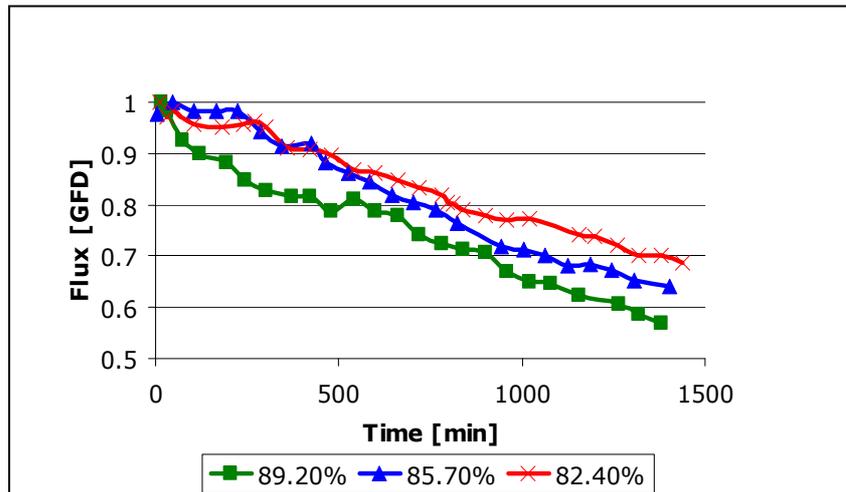

Figure 15: Permeate flux vs. time for different recovery rates

Analysis of VSEP Results.

An economically optimal VSEP operational period (period between successive interruptions of service for filter cleaning) was determined for each target water recovery as follows: The Figure 15 data were used to develop functional relationships between permeate flow rate and time of operation at each recovery curve. The total volume of permeate produced, divided by the operational period plus cleaning time (Table 6), provides the average permeate production rate for single VSEP device (a function of time of operation). Not surprisingly, both the instantaneous and average permeate fluxes decreased over the length of the operational period. The average rate at which water (RO brine) is treated by a single VSEP machine is then the average rate of permeate

production divided by the fractional water recovery, and the number of machines required is total rate of RO brine production divided by the average rate of water treatment in a single VSEP machine. Manufacturer's data related to the cost of full-scale VSEP reactors and VSEP operational costs are as shown (Table 6). Operational costs that were considered included power costs for flow through the pressure vessel at 500 psi and generation of torsional vibration (manufacturer's data), membrane cleaning and replacement costs and personnel costs. The value of water lost as VSEP brine was also taken as a system cost, estimated as $1000 per acre foot of brine produced. Other costs considered in the analysis were the capital and operational costs of brine disposal via enhanced evaporation. All unit costs, methods for cost annualization, etc., are as summarized (Table 6). The service life for VSEP reactors was assumed to be 10 years. That of evaporation ponds and related equipment was assumed to be 30 years. System costs were estimated using a discount operator (r) equal to 0.06 per year.

At each recovery rate for which there was a sufficient set of pilot operating data, the annualized cost for treating 3 MGD of RO brine was then calculated as a function of the period of VSEP operation (time after membrane cleaning) to establish the economically optimal period between cleanings. The procedure was repeated at each recovery level, to find a near optimal recovery during VSEP operation. Results were compared to the economics of a no-VSEP option in which disposal of the entire RO brine flow was required.

The annualized costs of construction and operation for a VSEP treatment system are summarized as a function of recovery and time of operation (Figure 16). There are fairly broad regions of near optimal operation. That is, total annualized cost is fairly

insensitive to recovery in the range 80-90 percent and period of operation in the range of
25-40 hrs.
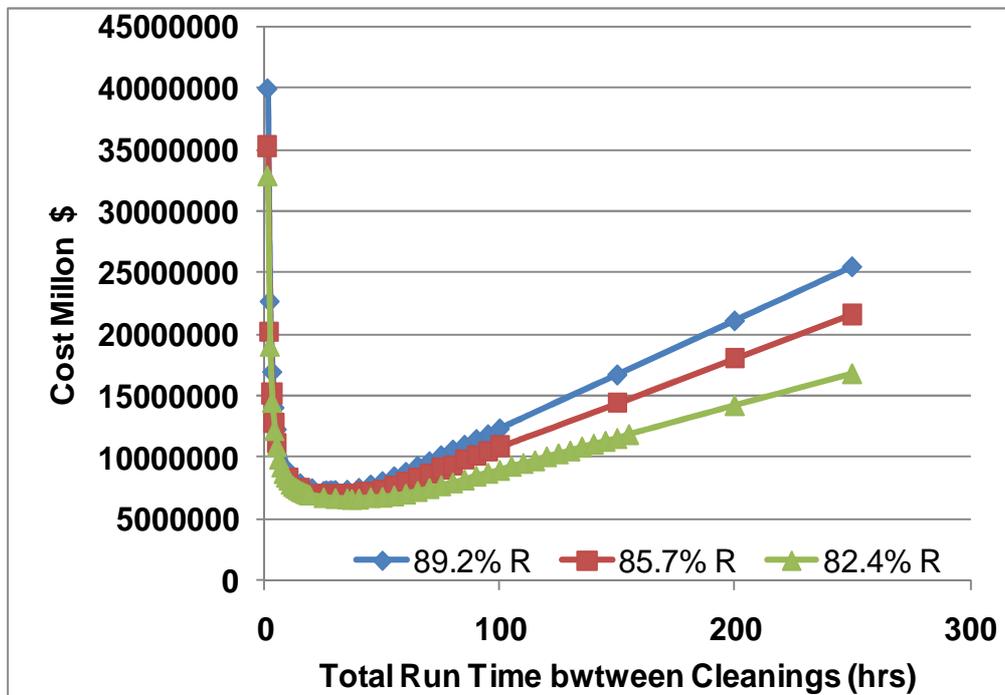
Figure 16. Annualized costs of construction and operation for a VSEP treatment system
as a function of recovery and time of operation

Furthermore, the total annualized cost for the VSEP system is significantly lower
($6.6 M · yr^{-1} versus $11.6 M · yr^{-1}) than the cost of the no-VSEP option. Finally, major

cost contributions arise from the VSEP capital cost (~20%), equipment for augmented,

post-VSEP brine disposal (~10%), energy for VSEP operation (~15%), chemicals for

cleaning the VSEP membranes (~16%), VSEP membrane replacement (~20%), operation

of augmented evaporation ponds (~8%), operations personnel (~5%) and the value of

water lost as the residual brine (~7%). Clearly, the total annualized cost of VSEP system

construction and operation is broadly distributed among capital, energy and so forth.

Cost savings in any single area, or even a minor computational error, is unlikely to affect

the general conclusion of the exercise—that VSEP treatment to minimize brine disposal in inland areas is cost effective relative to the no-VSEP option. In the RO/VSEP combined system that was proposed and evaluated here, only 2-4% of the CAP water treated would be lost as brine. The incremental cost of VSEP treatment to minimize brine loss is estimated at $390 - $430 per acre foot of CAP water treated. The cost per unit of water recovered from the RO brine is $2379 - $2412 per acre foot. Comparison to the costs of softening alternatives for minimizing water losses during desalination is important but lies beyond the present scope.

Table 6. Cost Summary Table

Item	Basis of Calculation	Assumptions	Contribution to Annualized Cost
Capital costs:			
VSEP units	$(d \times n) + \dfrac{d \times n}{(1+r)^{10}} + \dfrac{d \times n}{(1+r)^{20}}$	All the VSEP units are going to be renewed at the end of each 10 years.	18%-21%
Disposal of brine cost (augmented evaporation ponds)	$59\text{M\$} \times (Q_{VSEP,total}/3\text{MGD}) \times F_{30}$	There is a linear relation between the capital cost of evaporation ponds and the flow rate of the brine to be disposed.	6%-12%
Energy costs:			
Pumping cost	$\left[\left(\dfrac{Q_{VSEP,total} \times \rho \times g \times h}{3.6 \times 10^6} \right) \Big/ \varepsilon \right] \times 0.12\text{\$}/kWh \times 24hrs/d \times 365days/yr$	The shaft efficiency of the pump is 80%.	8%-9%
Vibration cost	$P_{vibration} \times n \times 0.12\text{\$}/kWh \times 24hrs/d \times 365 \ days/yr$	Power requirement for vibrating each VSEP unit is 12 hp	5%-6%
Cleaning costs:			
Chemical cost	$C_{chem} \times [24hrs/d \times 365days / (\check{T}+T_C)] \times n$		12%-21%
Water cost	$C_{w,cleaning} \times [24hrs/d \times 365days / (\check{T}+T_C)] \times n \times (C_{w,unit} / 3.3 \times 10^5 \ gal/AF)$	Unit value of water ($C_{w,unit}$) is 1000\$/AF.	0.4%-0.8%
O&M Costs:			
Membrane replacement cost	$(C_{memb.set} / 2 \ yrs) \times n$	A single set of membranes lasts for 2 years.	18%-22%
O&M cost for evaporation ponds	$3.9\text{M\$} \times (Q_{VSEP,total}/3\text{MGD})$	There is a linear relation between the O&M cost and the flow rate of the brine to be disposed.	5%-11%
Personnel cost	$400\text{K\$}/yr$	There are 4 personnel with 8 hours shifts around the clock at the facility.	5%-6%
Miscellaneous Costs:			
Brine cost (due to loss of water)	$(C_{w,unit} / 0.33 \ Mgal/AF) \times [Q_{VSEP,total} \times (1 - R_{VSEP})] \times 365days/yr$	Unit value of water ($C_{w,unit}$) is 1000\$/AF.	5%-9%

Summary:

The need for desalination of potable water in the southwestern United States arises form projected reliance on brackish and reclaimed water resources to make a water balance. Pilot studies were conducted to establish the long-term feasibility of RO treatment of CAP water in Tucson. Deterioration of membrane performance after more than a year of continuous field operation was indicated by a decrease in water permeability and an increase in salt flux. Membrane autopsies suggested that poor performance was caused primarily by membrane fouling with clays and unidentified organic material. Numbers of culturable bacteria in membrane foulants were low.

Water loss during desalination was minimized via post-RO VSEP treatment. Under optimal conditions, VSEP treatment reduced the volume of water lost as brine by >80%. Projected annual savings from VSEP treatment of a hypothetical 3 MGD brine stream was $6.6 million dollars per year due primarily to reduced brine disposal costs.

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**APPENDIX A:** Membrane Cleaning for MTP at the Tangerine Road Site Project

The cleaning procedure was carried out following an increase in the product TDS concentrations for all RO pressure vessels (Figure 1). The water transport coefficient (A) also declined, suggesting that membrane fouling was occurring (Figure 2). An increase in the differential pressures (dP) for both of the stages of the RO unit was also observed, as well as an increase in salt transport coefficient (B) (Figure 3).
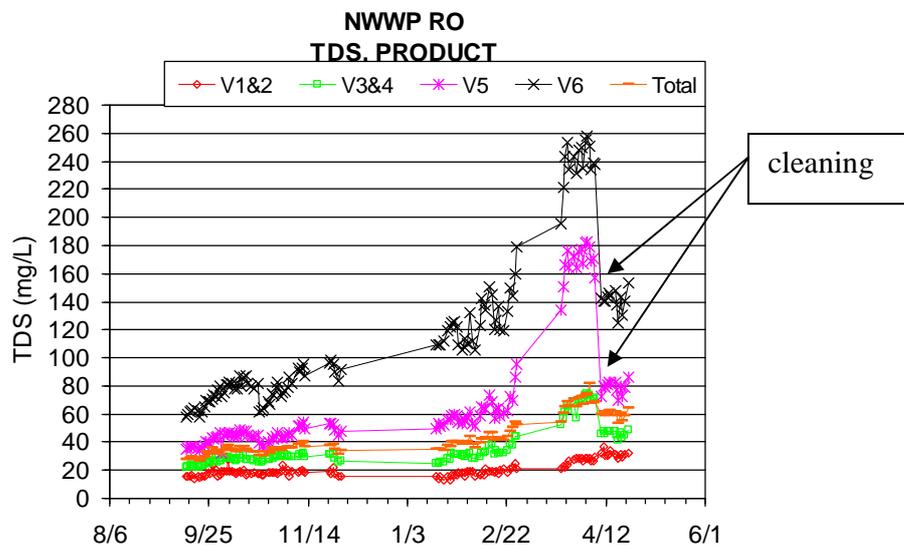
Figure 1: TDS concentrations for the product water
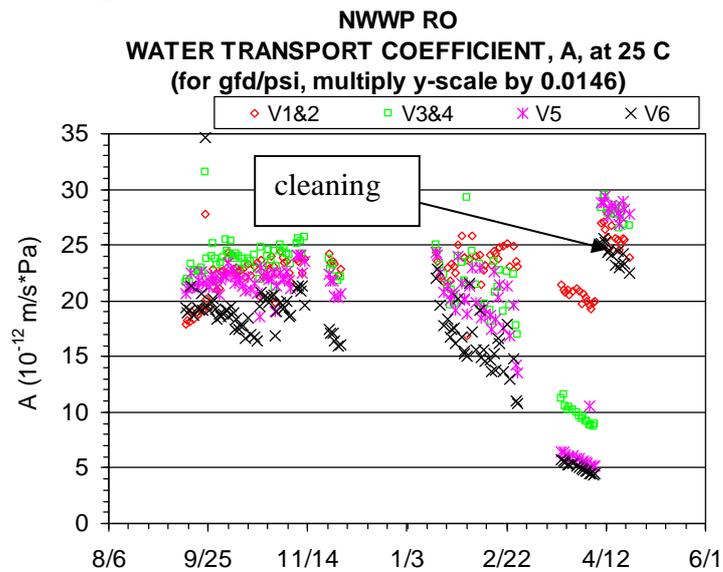
Figure 2: Water transport coefficient (A) for the RO operation
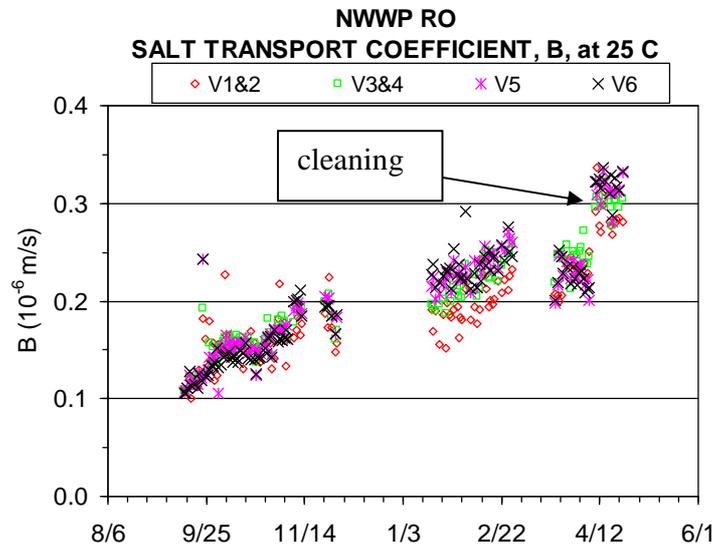
Figure 3: Salt transport coefficient (B) for the RO operation

Sodium hydroxide (NaOH) was used to increase the pH to the cleaning solution (pH ~ 12.0), and hydrochloric acid (HCl) was use to readjust the pH to the neutral level (pH ~ 7.0). The chemicals were ordered through the stores at the University of Arizona. Base/acid was added to chlorine free water that was produced by running the RO temporarily without chlorine addition. Consequently, sodium sulfite (Na_2SO_3) was omitted from the cleaning solution recipe.

As the first step in the production of the chlorine free water, the bleach and ammonium sulfate chemical feed pumps were shut down for about 18 hours of operation immediately preceding the cleaning steps. Both the free chlorine and the total chlorine levels in the RO influent were 0.01 mg/l before the chlorine-free permeate was accepted. The permeate tank was emptied before collecting the chlorine-free permeate. Then, 250 gallons of chlorine-free permeate was collected in that tank. Another 250 gallons of permeate was stored in an additional tank. This would provide the flushing solution at

the end of the cleaning process. The pH of the cleaning solution was adjusted to 12 by adding 260 g of NaOH to the cleaning solution. Unfortunately, the heater stopped working during the operation, so that the cleaning solution could not be adjusted to 30 $^{\circ}$C, as specified for membrane cleaning. The highest temperature recorded was 26 $^{\circ}$C. Initial (500-mL) samples of the basic cleaning solution were taken for analysis of specific ions and TOC levels.

The cleaning process consisted of four steps: (i) a low-flow flush with the basic solution, in recycle mode (ii) soak (iii) a high-flow flush with the basic solution, again in recycle mode (iv) neutral pH flush, without recycle. The low-flow flush and the high-flow flush were supposed to be at 16 L/min and 16-40 L/min (reject flows), respectively. However, the low-flow flush was performed at a flow rate of 8.09 L/min at 25.9 $^{\circ}$C and the high-flow flush was performed at a flow rate of 11.4 L/min at 24.7 $^{\circ}$C. The high-pressure pump shut off when the high-pH solution was provided, because the pH was outside the acceptable range for pump operation. Shut down at high pH is a safety factor in the control program. The inaccessibility to the control program did not allow us to bypass this control feature. Thus the only source of water input was the RO forwarding pump, which was unable to run higher than 8.09 L/min.

A new 1-μm cartridge filter was installed on the RO forwarding pump before the cleaning process was initiated. The cleaning solution had a conductivity of 1568 μS/cm and a pH of 12.12. It should be noted that the pH meter could not be calibrated properly, however, because the slope of the voltage vs solution pH for the test solution was not within the specified limits for acceptance.

The low-flow flush was performed for 1 hour. For the first 2 minutes, water was flushed to the drain. The reject and permeate flows were 7.23 L/min and 0.86 L/min, respectively, for a total flow of 8.09 L/min. The initial turbidity of the cleaning solution (before the run) in the permeate storage tank was 1.26 NTU. During the low-flow flush, a cloudy effluent was observed, especially between the 20th and the 45th minutes of the run. The turbidity of the reject water was 62.2 NTU at the 30th minute. The operating pressures were as follows:

Table 1: Operating pressures during low-flow flush step of the cleaning process

OPERATING PRESSURES DURING LOW-FLOW FLUSH		
Vessels	Influent (psi)	Product (psi)
1 & 2	43.3	42.6
3 & 4	37.6	37.7
5	32.5	27.5
6	17.8	0.4
Reject	6.1	-

The soaking period started right after the low-flow flush run. The membranes were soaked in the cleaning solution for 2 hours. The high-flow flush followed (conductivity 1446 μS/cm and pH 11.7). The temperature 24.7 ^{o}C. The flow rates of the reject and product (vessel 6) waters were 9.84 L/min and 1.56 L/min, respectively, during the run, for a total flow rate of 11.4 L/min. Reject water turbidity was 21.8 NTU during the run. Operating pressures follow:

Table 2: Operating pressures during high-flow flush step of the cleaning process

OPERATING PRESSURES DURING HIGH-FLOW FLUSH		
Vessels	Influent (psi)	Product (psi)
1 & 2	68.1	66.3
3 & 4	59.9	59.1
5	52.5	43.3
6	29.1	-0.2
Reject	10.8	-

At the end of the high-flow flush period, the cleaning solution had a pH of 12.02 and a conductivity of 1483 µS/cm. The temperature was 23.6 ^{o}C and the turbidity was 19.9 NTU. The spent cleaning solution was temporarily stored in the permeate storage tank. After adjusting the pH to 7.0 by addition of HCl, it was discharged to the CMID irrigation canal. The conductivity was 1000 µS/cm and temperature was 24.2 ^{o}C.

The high-flow permeate flush was performed for half an hour as the last step of the cleaning process. During the run, the flow rates of the reject and product water were 9.80 L/min and 1.51 L/min, respectively, for a total flow of 11.3 L/min. The turbidity of the permeate water increased to 1.23 NTU in the reject line while flushing.

The RO was reinitiated (pH = 6.72; conductivity = 1103 µS/cm; temperature = 20.9C; free chlorine = 0.17 mg/L; total chlorine = 0.52 mg/L, eventually increasing to 1.95 mg/L) at 11:30 PM on 04.07.2008. (3189.9 hours was the reading on the hour meter.) Recovery was 81.44 % at a total permeate flow rate of 15.27 L/min. The reject flow rate was 3.48 L/min. Additional details are available on the data sheet.

The cleaning process can be summarized as:

Table 3: Summary of the cleaning process

STEPS OF THE CLEANING PROCESS				
	Flowrate (L/min)	Temperature (^{o}C)	pH (-)	Time (hours)
Low-flow flush with recycle	8.09	25.9	12.12	1
Soaking	-	25.9	11.9	2
High-flow flush with recycle	11.4	24.7	11.7	2
High-flow flush with permeate water	11.3	24.2	~7.0	0.5

APPENDIX B: VSEP Cleaning Procedure

Flush the filter pack with cold permeate for 15 minutes. Use 50 psi and ¾" amplitude.

1. Increase the pressure to 300 psi and keep the amplitude ¾". Record the permeate flow rate in order to measure the cleaning success.

2. Prepare the feed tank with 3% solution of NLR 404, use 1.14 liters of NLR 404 in 10 gallons of warm permeate (50 °C). Run in recycle mode for 45 minutes. It is important to maintain the temperature constant during the cleaning.

3. Flush the filter pack with cold permeate for 15 minutes.

4. Repeat the same procedure as before but this time prepare the feed tank with 3% NLR 505 solution use 1.14 liters of NLR 505 in 10 gallons of warm permeate (50 °C).

5. Flush the filter pack with cold permeate for 15 minutes.

6. Increase the pressure to 300 psi and keep the amplitude ¾". Measure the flow rate at the end of the cleaning to determining the cleaning success.

7. Finally, neutralize the cleaning solutions for appropriate disposal of the residual.

CHEMICAL INFORMATION

NLR 404

NLR 404 is revolutionary acidic, liquid membrane cleaner formulated to effectively remove metallic-based foulants and scaling components. It is proven to target metallic salts such as iron sulfate, aluminum sulfate, barium sulfate, calcium sulfate and calcium carbonate. In addition, it can also remove dyes and inks. This formula uses a non-foaming formulation that reduces cleaning time. NLR 404 provides the cleaning performance you desire at a fraction of the time. At the recommended cleaning concentration of 3% the solution has a pH of 3.5 making it

compatible with a wide range of membranes from microfilters to reverse osmosis. This cleaner is often paired with a more alkaline cleaner in a two-stage process to successfully remove a wider range of foulants (www.membranecleaner.com)

NLR 505

NLR 505 is a unique blend of surfactants and chelating agents in a caustic liquid membrane cleaner. This unique cleaner targets organics, biological components, lignins, dyes and oil & grease.

The NLR 505 cleaner is usually used in a 3% concentration resulting in a pH of 11.5. It is compatible with most microfiltration, ultrafiltration, nanofiltration and reverse osmosis membranes. The formula can be customized to control the pH to within the limits of your specific membrane.

This cleaner is often paired with NLE 404 when a more comprehensive two-stage cleaning is needed to remove a wider range of foulants. This is strategic for waste streams as they tend to contain a variety of fouling bodies.

The liquid formula is ideal for automated CIP systems with chemical dosing and allows for ease of handling (www.membranecleaner.com).