

Pilot Investigation of Slowsand Filtration and Reverse Osmosis Treatment of Central Arizona Project Water

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13. ABSTRACT (Maximum 200 words) Because the August 2000 Bureau of Reclamation report, <i>Alternatives for Using Central Arizona Project Water in the Northwest Tucson Area</i> , identified slowsand filtration (SSF) as the lowest cost water treatment alternative for a proposed 40-MGD water treatment plant, this pilot investigation evaluated the technical effectiveness of SSF treatment of Central Arizona Project (CAP) water. Based on low turbidity levels in the SSF-treated water, subject to obtaining successful operation during the summer, the authors recommend SSF treatment. At a filtration rate of 0.11 gal/min/ft ² (0.27 m/h, 6.9 MGD/acre) the SSF filter run length measured 22 days. With 17 SSF cleanings per year, the revised SSF treatment cost estimate is \$0.16 per thousand gallons. For treatment with desalting, based on low fouling of low-pressure reverse osmosis (RO) polyamide membrane equipment operating on SSF-treated CAP water, the authors recommend either of two treatment combinations: (1) microfiltration or ultrafiltration - RO or (2) SSF - RO.				
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by
Charles Moody
Bruce Garrett
Eric Holler

Department of the Interior
Bureau of Reclamation



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Mission Statements

The mission of the Department of the Interior is to protect and provide access to our Nation's natural and cultural heritage and honor our trust responsibilities to tribes.

The mission of the Bureau of Reclamation is to manage, develop, and protect water and related resources in an environmentally and economically sound manner in the interest of the American Public.



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The Bureau of Reclamation's (Reclamation) Science and Technology Program, Reclamation's Phoenix Area Office, and the water provider organizations listed below contributed funds and services to conduct this study. Eric Holler (Reclamation) initiated and advocated the study and coordinated its implementation.

This report summarizes work undertaken by various agencies for the Southern Arizona Water Management Study to study the effectiveness of slowsand filtration (SSF) and reverse osmosis treatment of Central Arizona Project (CAP) water. The organizations and their participation follow:

Central Arizona Water Conservation District (CAWCD) – Hosted the pilot test at the Twin Peaks Pumping Plant Complex. The CAWCD provided water, installed wiring and provided power during operation, provided the earthen pad on which the slowsand tank was placed, and provided site security.

Town of Marana – Demobilized the SSF tank from Tucson Water's Hayden-Udall Treatment Plant and transported the tank to the Twin Peaks site. Marana representatives set the tank in place, provided equipment (including a crane) and labor to place the gravel underdrain and the specialty sand, provided a water meter to measure inlet flows, and demobilized and removed the tank upon completion of the testing.

Flowing Wells Irrigation District – Provided professional installation of the plumbing and piping between the SSF and the Mobile Treatment Plant, including overflows, drains, stands, supports, and SSF product water piping; and installed a float valve (at their suggestion) on the inlet to the SSF tank, which significantly improved operation.

Town of Oro Valley – Provided materials and installation of the inlet water piping, provided weekly sampling and water quality testing for the extensive monitoring that was required. Water quality testing included sampling at five locations for a large variety of constituents.

Metropolitan Domestic Water Improvement District (MDWID) – Provided comprehensive monthly water quality sampling and testing using the same methodology as Oro Valley. Acted as a clearinghouse for Oro Valley and MDWID to gather water quality information and input the data into the spreadsheet.

I thank the following individuals representing these water providers: Mark Stratton, Charlie Maish, Mike Block, and Stephanie Pranshke, MDWID; Bob Bradford, Dave Gunn, Mike Early, and Tom Plasay, CAWCD; Brad DeSpain, Town of Marana; Alan Forrest, Mary Kobida, and Charlie Soper, Town of Oro Valley; David Crockett and Jim Cavanaugh, Flowing Wells Irrigation District; Joe Crowson, Brad Prudhom, and Eric Holler, Reclamation. A special thank you is extended to Reclamation's consulting chemical engineer, Bruce Garrett, P.E., who made the pilot operational, operated and maintained the pilot, provided water quality testing, participated in presentations, and prepared the extensive data documentation and data reduction contained in appendices B and C. Without Mr. Garrett's expertise and dedication, the pilot would not have been possible.

It was a pleasure to work with this group of people. Their unflagging willingness to contribute to our common goals made this investigation possible. Together, they have continued to illustrate the positive impact a collaborative effort can make.

And, finally, I thank the others on the team from Reclamation who put together technical information and compiled and edited the report: Qian Zhang, water treatment; Sharon Leffel, editing and desktop publishing; Rodney Tang, technical support; Robert Michaels, management support; and Suzanne Sikora, financial management.

Charles "Chuck" Moody, Ph.D., P.E., Project Manager



Acronyms and Abbreviations

A

ADEQ	Arizona Department of Environmental Quality
AFY	acre-feet per year of water flow; multiply by 0.620 to convert to gallons per minute; divide by 724 to convert to cubic feet per second
ASR	aquifer storage and recovery
AWWA	American Water Works Association

C

°C	degrees Celsius
CA	cellulose acetate
CAP	Central Arizona Project
CASI	Central Arizona Salinity Interceptor
CAWCD	Central Arizona Water Conservation District
cfs	cubic feet per second
cfu	colony forming units
CO ₂	carbon dioxide
COU	coefficient of uniformity
CT	conventional treatment (coagulation, flocculation, and rapid sand filtration)
CT - RO	combination of conventional treatment and reverse osmosis

D

DBPs	disinfection byproducts
D/DBPR	Disinfectant/Disinfection Byproducts Rule
DO	dissolved oxygen
dP	difference in pressure between two locations

E

EPA	Environmental Protection Agency
ERT	energy recovery turbine

F

ft foot; feet
ft² square feet

G

gal/min gallons per minute
gal/min/ft² gallons per minute per square foot
gfd gallons of permeate per square foot of membrane per day
gal/f²/day gallons per square foot per day membrane water flux
gpd gallons per day
gpg grains per gallon
gpm gallons per minute
gpm/ft² gallons per minute per square foot

H

HAA haloacetic acid
HAA5 group of five haloacetic acids regulated by the Safe Drinking Water Act Stage 1 Disinfectants and Disinfection Products Rule with an MCL of 0.080 milligrams per liter
hp horsepower
HPC heterotrophic plate count

I

ICP inductively coupled plasma

K

kg kilogram(s)

L

LOI loss on ignition
L/m liters per minute

M

m meter(s)
m² square meters
MBD mass balance deviation

MCL	maximum contaminant level
MF	microfiltration
MF/UF	microfiltration or ultrafiltration
MF/UF - RO	combination of microfiltration/ultrafiltration and reverse osmosis
MGD	million gallons per day; multiply by 1,121 to convert to acre-feet per year
mg/L	milligrams per liter concentration
m/h	meters per hour
m ³ /h	cubic meters per hour
mm	millimeter(s)
MTP	Mobile Water Treatment Plant
μm	micrometer(s)
μg/L	micrograms per liter

N

NTU	nephelometric turbidity unit
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O

O&M	operation and maintenance
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P

PA	polyamide
pH	A measure of the relative acidity of water. pH depends on the composition of salts (electrolytes) dissolved in the water. Acid waters have pH values less than 7. Basic waters have pH values greater than 7.
ppm	parts per million
psi	pounds per square inch pressure; divide by 14.5 to convert to bars; multiply by 6.895 to convert to kilopascals
psid	psi differential pressure
psig	psi gage pressure (add atmospheric pressure to get absolute pressure)

R

Reclamation	Bureau of Reclamation
RO	reverse osmosis
ROSA	Reverse Osmosis System Analysis (DOW computer program)

S

SARWMS	Southern Arizona Regional Water Management Study
SDI	Silt Density Index

SDS	simulated distribution system
SDSDBP	simulated distribution system disinfection byproduct
SDSHAA	simulated distribution system haloacetic acid
SDSTHM	simulated distribution system trihalomethane
SDWA	Safe Drinking Water Act
SSF	slowsand filtration
SSF - RO	combination of slowsand filtration and reverse osmosis
SWTR	Surface Water Treatment Rule

T

TDS	total dissolved solids, milligrams per liter
THM	trihalomethane
THMFP	Trihalomethane Formation Potential
TOC	total organic carbon, milligrams per liter
TTHM	total trihalomethanes
t-value	statistical students' t-value

U

UF	ultrafiltration
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V

VFD	variable frequency drive
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Executive Summary

To address water supply and water quality issues in the northwest Tucson area, local, State, and Federal agencies prepared the appraisal study, *Alternatives for Using Central Arizona Project Water in the Northwest Tucson Area* (Bureau of Reclamation [Reclamation], 2000). The study estimated costs for a 40-million-gallon-per-day (MGD) municipal treatment plant to treat Central Arizona Project (CAP) water by three non-desalting processes: conventional treatment (coagulation and rapid sand filtration), slowsand filtration (SSF), and microfiltration (MF) or ultrafiltration (UF), as well as desalting by reverse osmosis (RO).

The 2000 study estimated that SSF had lower capital and unit costs than conventional treatment or MF/UF. The study estimated capital costs for the 40-MGD plant at \$12 million for SSF compared to \$47 million for conventional treatment and \$60 million for MF/UF. Comparing unit costs, SSF, at \$0.13 per thousand gallons, has less than one-fourth the cost of \$0.57 per thousand gallons for conventional treatment or MF/UF.

The 2000 study also estimated that the SSF - RO combination costs less than either conventional treatment with RO or MF/UF with RO by about \$0.30 per thousand gallons.

Slowsand filtration has previously been used extensively to treat Colorado River water, including plants at Yuma, Arizona; Calexico, California; and El Centro, California. With delivery of CAP water to central Arizona, recent pilot studies (Cluff et al., 1989; Cluff, 1993; Chowdhury et al., 2002) have also reported success with SSF. Because, to the authors' knowledge, no SSF water treatment plants presently treat Colorado River water in Arizona or southern California, the study recommended pilot tests to confirm the effectiveness of SSF to produce potable water and to serve as pretreatment to RO desalting.

To address the recommendation to conduct pilot tests, Reclamation, the Central Arizona Water Conservation District, Town of Marana, Flowing Wells Irrigation District, Town of Oro Valley, and Metropolitan Domestic Water Improvement District cooperated in funding, installing, and operating SSF and RO pilot equipment. In August 2001, a 200-square-foot slowsand filter began operation at the Twin Peaks Pumping Plant Complex near Marana, Arizona. After 2 months of slowsand filter conditioning, an RO pilot unit began operation in October 2001. Testing continued for 5 more months (until March 2002).

The pilot slowsand filter operated with minimal operation and maintenance requirements. At the design filtration rate of 0.11 gal/min/ft² (0.27 m/h, 6.9 MGD/acre), filter runs lasted 22 days, corresponding to 17 cleanings per year. Because the appraisal study assumed only six cleanings per year, the higher cleaning frequency increases the estimated cost of SSF to \$0.15 per thousand gallons.

The pilot slowsand filter meets the Surface Water Treatment Rule for turbidity by producing water with turbidity levels less than 1.0 nephelometric turbidity unit for over 95 percent of daily samples. Disinfection byproduct levels with chloramine disinfection appear to meet the Safe Drinking Water Act Stage 1 Disinfectants and Disinfection Byproducts Rule, but further tests are needed for confirmation.

SSF effectively removes particulates that foul RO equipment based on low-fouling operation of an RO pilot unit. In addition, cleaning the RO pilot unit after 3 months of low-fouling operation effectively restored RO production to new performance. This performance was maintained for the remaining 2 months of operation.

During the pilot tests, with an average feed salinity of 670 milligrams per liter (mg/L) total dissolved solids (TDS), low-pressure RO membranes (FilmTec NF90), and 80 percent water recovery, the RO product salinity measured 10 to 31 mg/L TDS. In the summer, higher water temperatures would result in an estimated RO product salinity of 50 to 60 mg/L TDS.

Low trihalomethane formation potentials of 0.001 to 0.003 mg/L in the RO product indicate that free chlorine disinfection can be used to meet the maximum contaminant level (MCL) for trihalomethanes (0.080 mg/L) and other disinfection byproducts. Chloramine disinfection can also be used, but is not required to meet the disinfection byproduct MCLs.

The pilot tests demonstrate that for the treatment of CAP water, SSF can be used to meet all primary drinking water standards, and the combination of SSF and low-pressure RO can be used to meet all primary and secondary drinking water standards. Because SSF effectively treats CAP water at one-fourth the cost of either conventional treatment or MF/UF, subject to future tests demonstrating successful year-round operation, the authors recommend SSF for CAP water treatment without desalting.

For CAP water treatment with desalting, the authors recommend two alternatives: one alternative is MF/UF in combination with low-pressure RO; the second alternative, recommended subject to the successful completion of additional pilot tests, is SSF in combination with low-pressure RO. Of the two alternatives, the SSF - RO combination costs less (by about \$0.30 per thousand gallons).



Pilot Investigation of Slowsand Filtration and Reverse Osmosis Treatment of Central Arizona Project Water

Introduction

This Slowsand Filtration and Reverse Osmosis Treatment pilot investigation was initiated as a result of recommendations in the August 2000 Southern Arizona Regional Water Management Study (SARWMS) report, *Alternatives for Using Central Arizona Project Water in the Northwest Tucson Area, Appraisal Study* (Bureau of Reclamation [Reclamation], 2000). To utilize Central Arizona Project (CAP) water as one of the water supply alternatives, SARWMS evaluated several CAP water treatment processes that meet primary and secondary drinking water standards. This pilot investigation exemplifies Reclamation's mission to evaluate, improve, and reduce the cost of desalting and other advanced water treatment technologies for the benefit of the water treatment industry, water utilities, and water users.

SARWMS explored the treatment of Colorado River water that is conveyed by Reclamation's CAP. For treatment without desalting, the study considered three alternatives: conventional treatment (CT), slowsand filtration (SSF), and microfiltration (MF) or ultrafiltration (UF).

Advanced treatment alternatives included the above treatments followed by low-pressure reverse osmosis (RO). Reclamation, 2000, included advanced treatment of 700 milligram per liter (mg/L) total dissolved solids (TDS) CAP water with RO desalting to produce low-salinity water because of its many health, aesthetic, and economic benefits. Health benefits include meeting all required primary drinking water standards plus all recommended secondary drinking water standards for inorganic contaminants. Health benefits include the physical disinfection afforded by the RO membrane serving as an additional barrier to disease micro-organisms (e.g., giardia, cryptosporidium, and viruses). Aesthetic benefits include improved water taste and the option of using free chlorine disinfection instead of chloramine disinfection in the water distribution system.

The economic benefits of using low-salinity waters are significant for the entire range of freshwater uses, including residential, commercial, industrial, utility, agricultural, and water reuse (Reclamation, 2000, appendix E, page E-70). In addition, for water reuse, an important economic benefit of using desalted CAP water is that low-salinity water produces low-salinity wastewater. Municipal wastewater is a valuable water supply in the southwest, but with CAP water, the high wastewater salinity of approximately 950 mg/L TDS reduces its value and limits its use. An RO-desalted water supply of 100 mg/L would result in a wastewater TDS of about 350 mg/L.

Costs for treatment without desalting were estimated for variable-production plants to supply water deliveries to meet maximum day deliveries of 40.14 million gallons per day (MGD) with an

average annual plant production of 26.76 MGD. Table 1 summarizes these estimated costs. The estimated SSF capital cost of \$12.4 million, corresponding to \$0.31-per-gallon-per-day capacity, is about one-fourth that of conventional treatment and one-fifth that of MF/UF. Comparing unit costs, SSF, at \$0.13 per thousand gallons, has less than one-fourth the cost of \$0.57 per thousand gallons for conventional treatment or MF/UF.

Costs for treatment with desalting were estimated for constant-production RO plants (with aquifer storage and recovery [ASR] of desalted water) producing 22.8 MGD throughout the year.

Table 1.—Cost comparison of three CAP water treatment plants¹ without desalting
(Reclamation, 2000; table E-41, p. E-96)

Treatment process	Cost category	Capital and operation and maintenance (O&M) costs	Annual costs	Unit costs
			million \$/yr	\$/1,000 gal
CT	Capital	\$47.0 million	3.93	0.57
	O&M	\$1.67 million/yr	1.67	
SSF	Capital	\$12.4 million	1.03	0.13
	O&M	\$0.24 million/yr	0.24	
MF/UF	Capital	\$59.8 million	5.01	0.57
	O&M	\$0.54 million/yr	0.54	

¹ Variable production plant capacity of 40.1 MGD (45,000 acre-feet per year [AFY]) meets peak-day demand. Average production is 26.8 MGD (30,000 AFY).

With a design RO water recovery of 85 percent, the pre-treatment process (SSF, CT, or MF/UF) operates at a constant of 26.8 MGD. The 15-percent concentrate that must be disposed of is 4.0 MGD.

Figure 1 summarizes the treatment costs in dollars per thousand gallons, including capital and operation and maintenance (O&M) costs, for three treatments without desalting (SSF, CT, and MF/UF) and three treatments with desalting (SSF-RO, CT-RO, and MF/UF-RO). Treatment using SSF has the lowest product water cost for treatment without desalting. Treatment of CAP water with the SSF-RO alternative has the lowest product water cost for producing desalted water in a constant-production plant with ASR by about \$0.30 per thousand gallons.

Because SSF has a significantly lower estimated cost than conventional treatment or MF/UF, and because recent operating experience with SSF treatment of Colorado River water is limited,

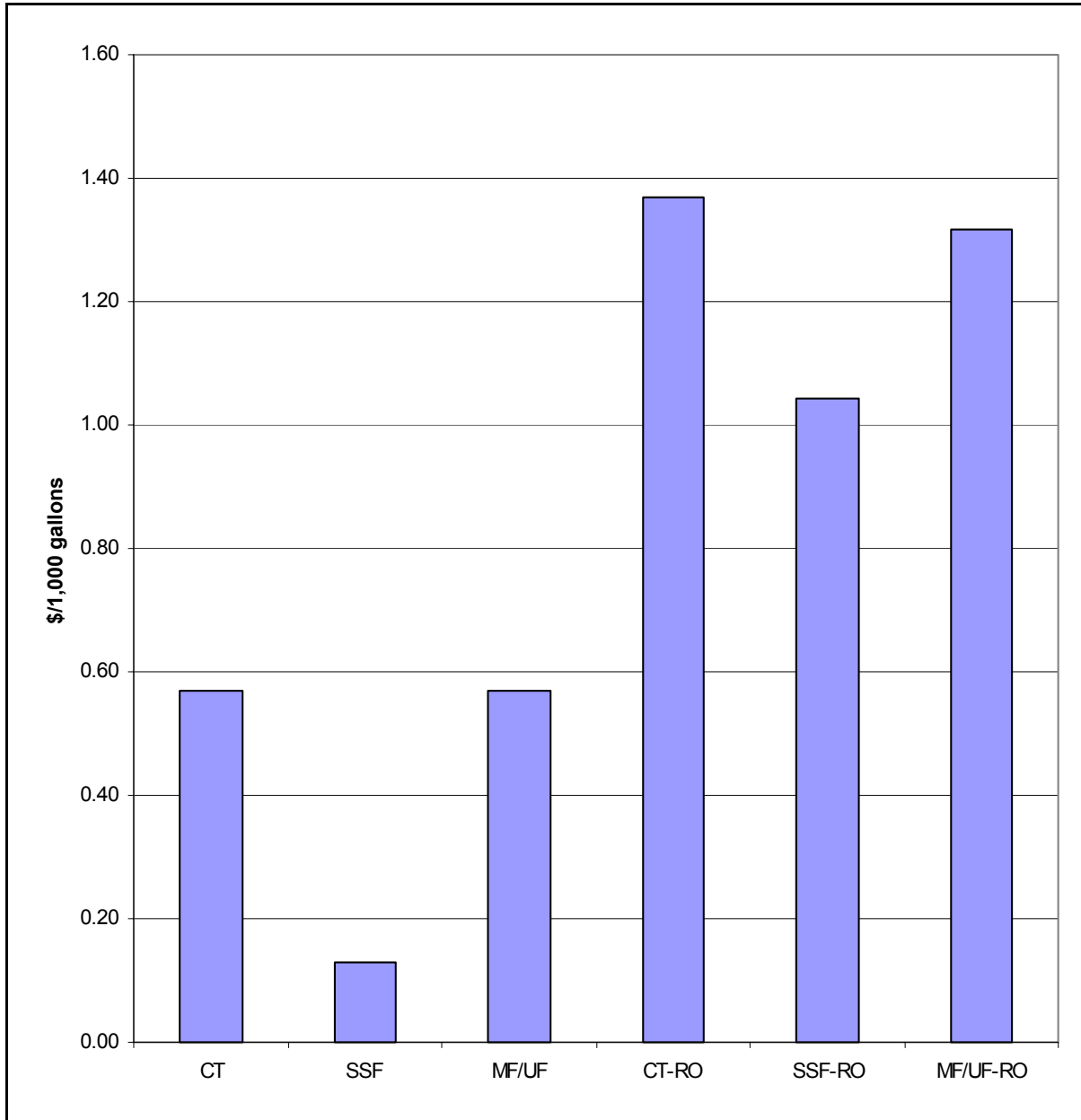


Figure 1.—Estimated CAP water treatment costs. For treatments with RO desalting, concentrate disposal costs are based on a Central Arizona Salinity Interceptor (CASI) pipeline to Yuma conveying 37.6 MGD from Tucson and the Arizona Municipal Water Users Association Sub-Regional Operating Group (SROG) (Reclamation, 2000, p. II-14.)

the August 2000 report recommended pilot testing of SSF and SSF - RO. The pilot tests address the effectiveness of SSF with respect to the following questions:

1. Can SSF be used to treat CAP water to primary drinking water standards? Specifically:
 - A. What turbidity levels does SSF produce? To meet the Surface Water Treatment Rule (SWTR) for turbidity, does SSF reduce turbidity to less than 1 nephelometric turbidity unit (NTU) for 95 percent of daily samples in a month?
 - B. What are the disinfection byproduct (DBP) levels in SSF product water that have been disinfected by chlorination - chloramination? Expected in a water treatment plant is post-SSF chlorination disinfection to meet the SWTR for giardia and virus removal and chloramination disinfection for the water distribution system. With this disinfection, are the 7-day simulated distribution system (SDS) disinfection byproduct levels less than the maximum contaminant levels (MCLs) listed by the Safe Drinking Water Act (SDWA) Stage 1 Disinfectants and Disinfection Byproducts Rule (D/DBPR) of 0.080 mg/L total trihalomethanes (TTHM) and 0.060 mg/L of five haloacetic acids (HAA5) ?

In addition, because total organic carbon (TOC) levels affect levels of DBPs produced during chlorination or chloramination, and because TOC levels may affect biological fouling of RO membranes, what is the TOC removal of SSF?

2. How effective is SSF as a pretreatment to RO? Specifically, does a slowsand pilot system provide adequate removal of particulates that foul RO membranes and reduce RO productivity?
3. What is the salinity and composition of CAP water treated by SSF and RO? Does the RO product water meet all primary and secondary drinking water standards for inorganic contaminants?
4. For disinfection of RO product, can free chlorine without ammonia be used to meet the SDWA Stage 1 D/DBPR of 0.080 mg/L TTHM and 0.060 mg/L of HAA5?
5. What changes to SSF costs estimated in the August 2000 SARWMS are indicated by the pilot tests?

Education and Public Information

An additional goal of performing the pilot tests was to familiarize Reclamation and local water provider staff, managers, and policy makers with the objectives, procedures, and technologies

used. To achieve this goal, tours were performed while the pilot was operating. One activity during the tour was a PowerPoint presentation that outlined the reasons for doing the study, and it included a description of the pilot setup and preliminary results. Following the presentation, the tour included a step-by-step visit to each part of the operation. The tours ended with a taste test of the finished RO product.

The groups were limited in size to 10 people. Participants included staff, senior management from the utilities and towns, as well as elected policy makers and representatives from other interested agencies. Over 12 tours were conducted for representatives from partners in the study and other entities interested in water resources, including the Arizona Department of Water Resources, Pima County Wastewater Management, University of Arizona Water Resources Research Center, Community Water Company of Green Valley, Arizona Wetlands Research, City of Goodyear, West Maricopa Combine, Inc., the Tohono O’odham Nation, San Xavier District, and the Bureau of Indian Affairs. The feedback on the tours was very positive, indicating a high interest level from the participants.

The pilot also provided an opportunity for technology transfer. Partners in the study participated in assembling the pilot components and provided valuable expertise and suggestions. In addition, partners' staff worked on the pilot—operating it, providing water quality testing, and even assisting with maintenance. The partners were kept up-to-date on the progress of the pilot by “Progress Reports” e-mailed out on a biweekly basis.

Slowsand Filtration

Slowsand filtration water treatment is considered applicable primarily for relatively high-quality water supplies with turbidities less than 10 NTU (American Water Works Association [AWWA], 1991). Because of its low capital and operating costs and low requirement for operator attention, SSF is particularly attractive to small communities. SSF does not require chemical coagulation or backwashing. Operation requires only the adjustment of water flow, the monitoring of headloss and turbidity, and the periodic (ca. monthly or longer) removing of the “schmutzdecke,” a thin layer of particulates deposited on top of the filter. Slowsand filters remove turbidity and biological particles such as Giardia cysts, Cryptosporidium oocysts, algae, bacteria, and viruses (AWWA, 1991).

The slowsand filter system can be constructed of reinforced concrete, ferro-cement, stone/brick masonry, or earthen berms lined with high-density polyethylene geomembrane liner. A slowsand filter system consists of the following:

- A supernatant layer of raw water
 - A bed of fine sand, with a depth of 1.6 to 3.3 feet (0.5 to 1.0 meter)
-

- Gravel layers and perforated piping (filter underdrain) to collect the filtered water
- Inlet and outlet structures
- Flow control instrumentation and valves
- Drain and overflow components for controlling the supernatant water level during operation and filter cleaning

The water flow into a slowsand filter can be controlled at either the inlet or outlet of the filter. Under “inlet control,” an inlet valve controls the inlet flow, and the height of the water above the filter increases during the filter run. Under “outlet control,” the supernatant water is fixed near its maximum level, and the filter flow is controlled by gradually opening an outlet valve during a filter run. Figure 2 shows the basic components of an outlet-controlled slowsand filter.

The water in the filter slowly passes through the porous sand bed. The Arizona Department of Environmental Quality (Arizona Department of Environmental Quality, 1978) specifies a range of 0.032 to 0.16 gal/min/ft² [0.08 to 0.4 m/h, 2.0 to 10.0 MGD/acre]. This is in contrast to conventional treatment with rapid sand filters that operate at filtration rates of 2 to 5 gal/min/ft² (5 to 12 m/h, 130 to 310 MD/acre). At a filtration rate of 6.9 MGD/acre, 6 acres of slowsand filter area (not including access roads, piping galleries, and other components of a water treatment plant) produce 40 MGD of treated water.

During this passage, the physical and biological quality of raw water improves through a combination of biological assimilation and physical filtration. A thin layer forms on the surface of the sand bed as the bed matures. This thin layer (schmutzdecke) consists of retained organic and inorganic materials and micro-organisms that may consume some of the natural organic matter (NOM) in the raw water. As these materials are collected, the resistance to filter flow increases. The filtration capacity can be restored by cleaning the filter, which involves scraping or washing off the top 1/4- to 1-inch depth of the sand filter bed, including the retained organic and inorganic material filter skin. In contrast with CT's rapid sand filters, the SSF is never backwashed.

Slowsand filters remove particles and micro-organisms, but do not reduce hardness or salinity (TDS) levels in the water.

Slowsand filters have been extensively used to treat Colorado River water in the past—including plants at Yuma, Arizona; Calexico, California; and El Centro, California. As reported in 1918, the Calexico SSF operated at 0.22 gal/min/ft² (0.55 m/h, 14 MGD/acre), the El Centro SSF operated at 0.13 gal/min/ft² (0.32 m/h, 8.3 MGD/acre), and the Yuma SSF operated at as high as 0.46 gal/min/ft² (1.1 m/h, 29 MGD/acre) (*Engineering New-Record*, 1918).

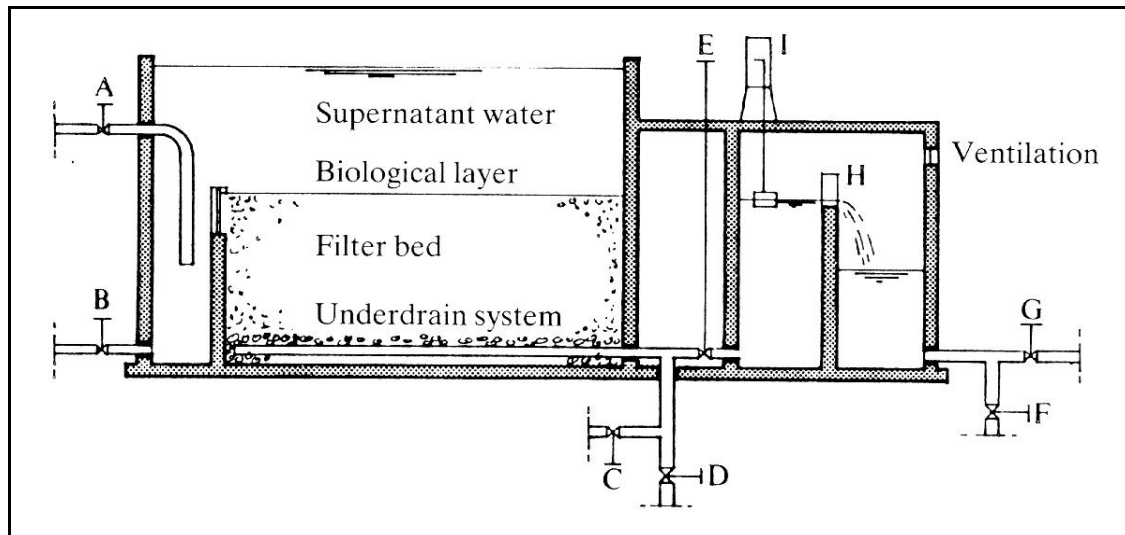


Figure 2.—Basic components of an outlet-controlled slowsand filter (Visscher et al., 1987, p. 21).

- A — Raw-water inlet valve
- B — Valve for drainage of supernatant water layer
- C — Valve for back-filling the filter bed with clean water
- D — Valve for drainage of filter bed and outlet chamber
- E — Valve for regulation of the filtration rate
- F — Valve for delivery of treated water to waste
- G — Valve for delivery of treated water to the clear-water reservoir
- H — Outlet weir
- I — Calibrated flow indicator

Hiriam W. Blaisdell developed, patented¹, and built several slowsand filter washing machines (see figure 3), the first of which was installed in Yuma and operated until 1954 when SSF was replaced with conventional treatment with rapid sand filtration (Doyle & Associates and Carollo Engineering, ca. 1995).

In 1908, William F. Fuller described use of the Blaisdell machine (see figure 3) and slowsand filter operation at Yuma:

Such a machine has already been in operation at Yuma, Arizona. For 4 or more years, on a slowsand filter, clarifying Colorado River water, having an average turbidity of over 2,000 at a rate of 3,000,000 gallons per acre per day, through sand having an effective size of 0.13 millimeter, without the use of any coagulants, at a very small maintenance cost and with satisfactory results (Fuller, 1908).

¹ Slowsand filter cleaning patents by Hiriam W. Blaisdell from 1903 to 1909 include patent numbers: 729,718; 729,719; 729,720; 729,721; 729,722; 752,196; 763,354; 12,488 (reissue); 840,104; 842,850; 845,744; 845,746; 864,151; 867,003; 873,010; 882,738; 894,873; and 12,932 (reissue).



Figure 3.—Blaisdell slowsand filter washing machine in Yuma, Arizona (photo by Paul McAleese, June 2002).

In 1946, C.G. Ekstrom described the Yuma slowsand filters: “The filters of the Yuma water plant are of the slowsand filter design with provisions for continuous mechanical washing of the sand. The four filters, which are 25 feet wide, are arranged end-to-end and total 411 feet in length, for a combined filtering area of 10,275 square feet. Each filter consists of 1 foot of coarse gravel, 6 inches of pea gravel, and 18 inches of sand. The filters are kept in continuous operation except during the washing process, when the particular filter being washed is out of service for approximately 1-1/2 hours. Washing is accomplished by the use of a Blaisdell machine (Ekstrom, 1947).”

For a survey of slowsand filter cleaning technologies, see appendix H and Huisman and Wood, 1974, p. 78-95.

Following CAP delivery of Colorado River water to Central Arizona, C. Brent Cluff tested pilot SSF equipment on Colorado River water (Cluff et. al, 1989; Cluff, 1993). In 1999 and 2000, the American Water Works Research Foundation and the City of Tucson sponsored a pilot-scale evaluation of treatment alternatives for CAP water. The study included an evaluation and comparison of SSF, bio-optimized rapid sand filtration, and MF as pretreatment processes for nanofiltration and low-pressure RO (Chowdhury et al., 2002).

SSF Pilot Tests

SSF pilot tests were conducted to determine the effectiveness of SSF in treating CAP water and to get design estimates for SSF filter run lengths and associated filter cleaning frequencies. The SSF - RO pilot combination provides information on operation of the integrated system with specific focus on determining if SSF effectively removes particulates that foul RO membranes.

Test Site

The Mobile Water Treatment Plant (MTP) semi-tractor trailer was located on the east bank of the Central Arizona Canal at the Twin Peaks Pumping Plant near Marana, Arizona. Figure 4 shows the general MTP site layout. Raw water was delivered by gravity flow from the upper Twin Peaks pool to the pilot SSF tank.

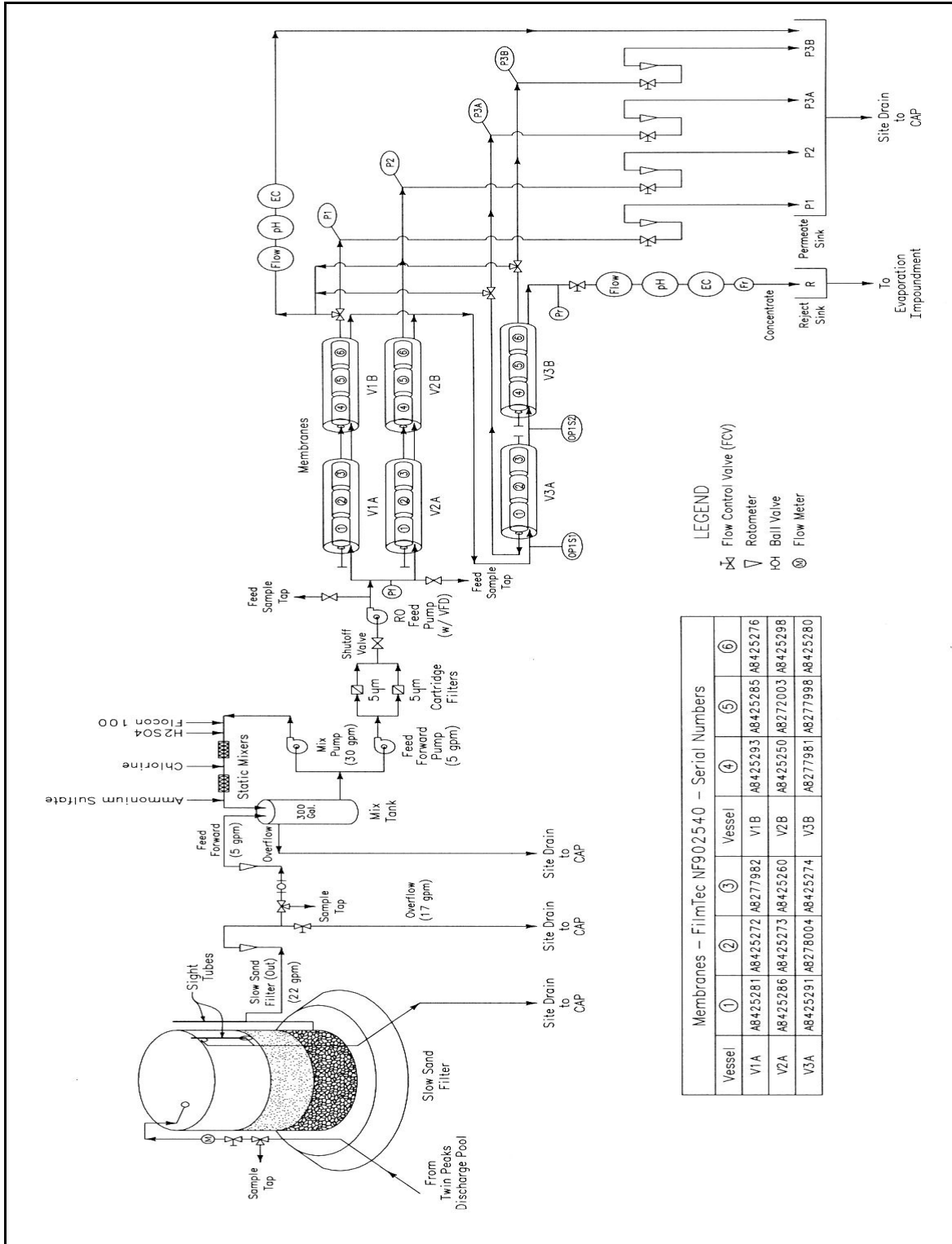


Figure 4.—SARWMS Pilot Plant, process and instrumentation drawing (Mobile Water Treatment Plant).

Pilot Equipment

The pilot SSF consists of a 16-foot-diameter by 10-foot-high galvanized tank (previously used in the study by Chowdhury et al., 2002). The SSF tank was placed on a 4-foot high soil pad to elevate the discharge and provide sufficient head for gravity flow to downstream equipment (refer to figure 4). The tank contains perforated polyvinyl chloride (PVC) collection piping at the bottom of 1.5 feet of three (1-inch, 3/4-inch, and 3/8-inch) gravel layers, 3.0 feet of 0.3-millimeter sand, 4.5 feet of water being filtered, and 1.0 foot of freeboard (see table 2, figures 5(a) and 5(b), and appendix A for a detailed description of the slowsand filter assembly). A geotext fabric layer was installed as a precaution to prevent sand from penetrating into the gravel.

Table 2.—Slowsand filter design¹ and pilot operating conditions

Design criteria	Full-scale design	August 7, 2001 - January 2, 2002	January 2, 2002 - March 18, 2002
Period of operation	24 hours/day	24 hours/day	24 hours/day
Product flow	40.14 MGD	22 gal/min (31,680 gal/day)	16 gal/min (23,040 gal/day)
Filtration rate	0.112 gal/min/ft ² 0.274 m/h 7.0 MGD/acre	0.110 gal/min/ft ² 0.269 m/hr 6.9 MGD/acre	0.080 gal/min/ft ² 0.196 m/hr 5.0 MGD/acre
Total filter area	330,450 ft ² or 7.6 acres	200 ft ² (16-foot diameter)	200 ft ² (16-foot diameter)
Number of filters	1 with 4 cells at 13.38 MGD each	One	One
Initial height of filter sand bed	3.0 ft	3.0 ft	3.0 ft
Minimum height of filter sand bed	1.5 ft	1.5 ft	1.5 ft
Sand effective size, d ₁₀	0.27 to 0.33 mm	0.27 to 0.31 mm	0.27 to 0.31 mm
Sand uniformity coefficient ² , d ₆₀ /d ₁₀	Less than 2.5	1.5	1.5
Height of underdrains, including gravel layers	1.5 ft	1.5 ft (0.5 ft of 1-in, 0.5 ft of 3/4-in, and 0.5 ft of 3/8-in gravel)	1.5 ft (0.5 ft of 1-in, 0.5 ft of 3/4-in, and 0.5 ft of 3/8-in gravel)
Height of supernatant water	5 ft	4.5 ft	4.5 ft
Free board	2 ft	1.0 ft	1.0 ft
Total filter basin depth	11.5 ft	10 ft	10 ft

¹ Reclamation, 2000, table E-2, p. E-7.

² The ratio of the sieve size through which 60 percent of the sand will pass to the size through which 10 percent will pass.

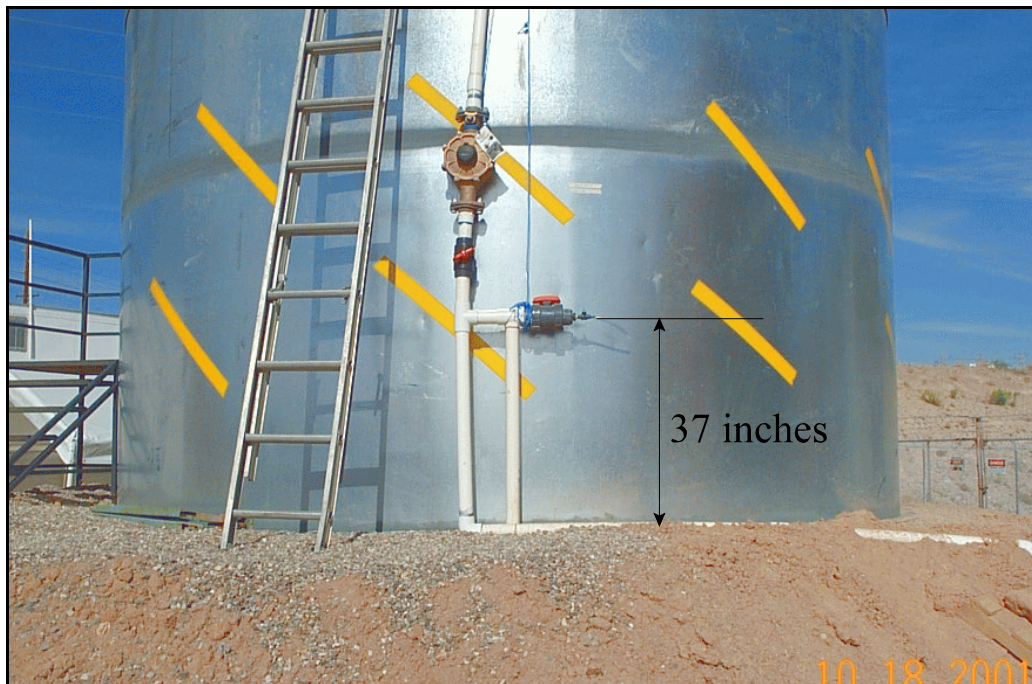
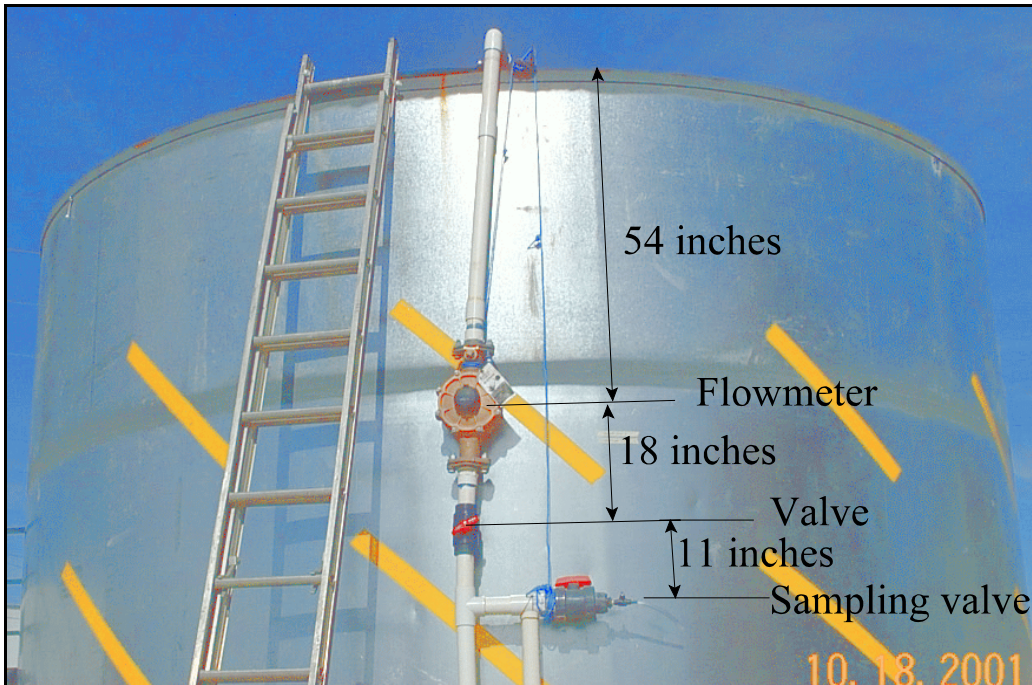


Figure 5(a).—Inflow piping of slowsand tank. The tank is 10 feet tall and has a diameter of 16 feet.

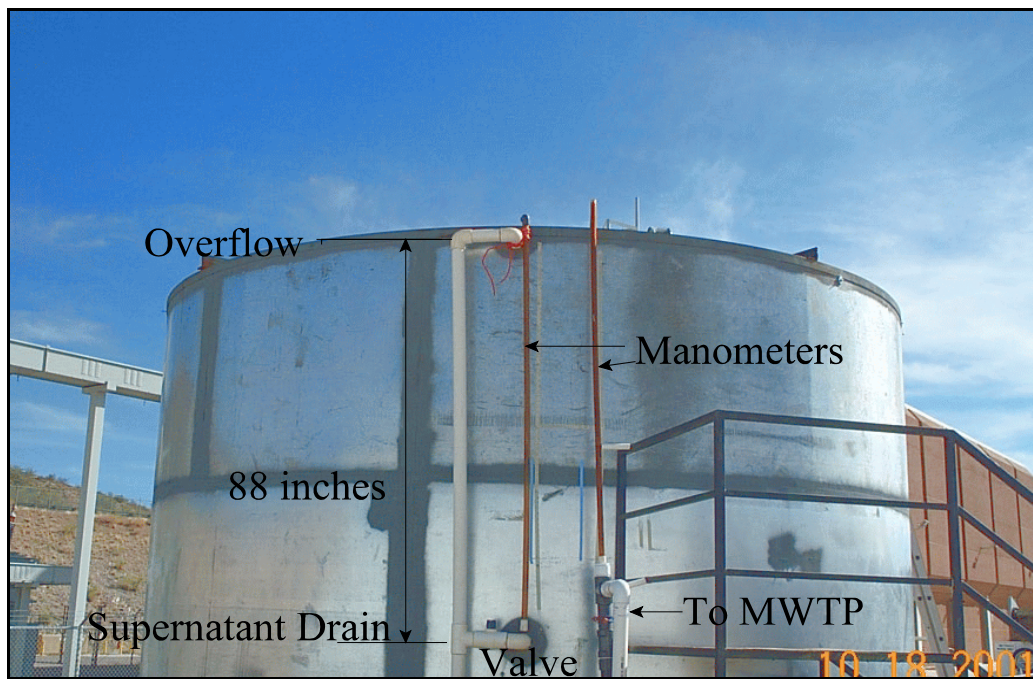
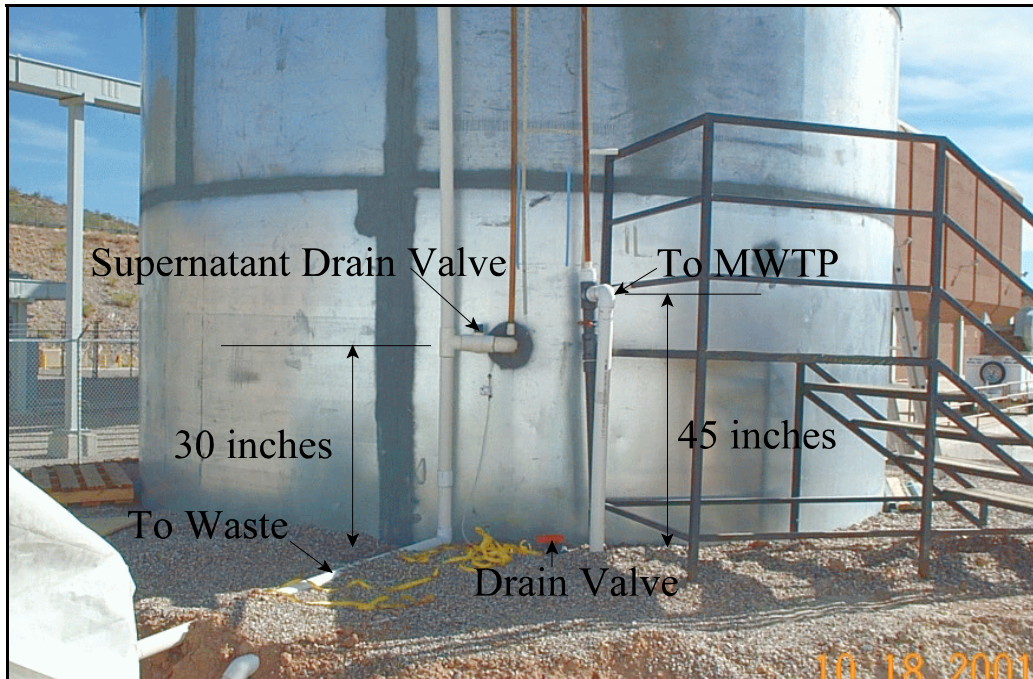


Figure 5(b).—Outflow piping of slowsand tank.

The SSF pilot was operated with “outlet control.” The tank water level was held constant at maximum height with an inlet float valve. The water height above the initial sand level was monitored with a sight tube manometer and is shown as the “SSF inlet level” in appendix B, figure B-8. With an outlet valve, the filter flow was controlled, generally to either 22 or 16 gal/min. A second sight tube on the filter outlet (upstream of the control valve) served as a manometer to measure the outlet pressure. This pressure is shown as the “SSF exit level” in appendix B, figure B-8. The difference in water heights of the two manometers is the filter pressure drop and is shown as “SSF dP” in appendix B, figure B-7.

Figure 6 is a photo of the SSF outlet flow control structure. The total flow of 16 or 22 gal/min of SSF-filtered water enters from the 2-inch hose in the center of the photo between the two pallets. The total flow is

measured by the first rotameter flowmeter. The flow then splits, with flow to the right controlled at 6 gal/min by a ball valve at the base of the second rotameter.

The rotameters have opaque green sleeves to minimize the growth of algae in the clear tubes. To the left of the 6-gal/min rotameter flowmeter is the sample tap (location SSF-Out) for water quality analyses of the SSF effluent. The 6-gal/min flow from the rotameter outlet flows into the chemical mixing tank and becomes the “RO feed.”



Figure 6.—SSF outlet flow control structure.

SSF production in excess of that needed for RO operation flows to the left of the tee. This excess flow (10 or 16 gal/min of filtered water) was returned by natural drainage to the CAP Canal. As the SSF outlet head decreased during a filter run, the gate valve located to the left of the tee was gradually opened to maintain SSF flow at 16 or 22 gal/min.

Test Period

The SSF pilot equipment operated 5,000 hours, from August 2001 through March 2002. The SSF began operation August 7, 2001. After four filter runs for SSF “conditioning,” RO equipment was loaded with membranes, and RO operations began on October 11, 2001. Both SSF and RO equipment operated until March 19, 2002. The SSF equipment operated continuously with planned outages of approximately 8 hours every 3 or 4 weeks at the end of a

filter run when the SSF surface was scraped. The RO equipment was offline during SSF cleanings, during a few short power outages in December 2001, and during two RO cleanings in January and February 2002.

Water Supply

The water supply to the SSF came from the Central Arizona Project Canal, Twin Peaks Pumping Plant discharge pool, gravity siphoned back to the inlet of the SSF tank. CAP water is dominated by water from the Colorado River, which is pumped from Lake Havasu to Phoenix and then to Tucson.

During the pilot study, the SSF inlet water temperature range was 9 to 24 degrees Celsius (°C) (48 to 75 degrees Fahrenheit) (see appendix B, figure B-4). The TDS of the canal is expected to be about 700 mg/L (SARWMS, August 2000, table III-7, p. III-10). During the pilot study, the average TDS was 670 mg/L (see appendix B, figure B-5). The TOC levels, expected to be about 3.5 mg/L (Reclamation, 2000, table III-7, p. III-10) during the pilot study, measured from 2 to 7 mg/L, with an average of 3.3 mg/L (see appendix B, figure B-11).

In November 2001, the canal was taken out of service for 3 weeks of maintenance. The pilot equipment was not without water, however, as the upper pool was filled before the shutdown. The quality changed (as indicated by color changing from blue toward green and increased odor) as the canal operations contributed to stagnant water. On December 12, 2001, the turbidity increased markedly when Twin Peak pumps began to deliver fine black mud in suspension to the upper pool and to the SSF. Because of the proximity of the SSF supply intake, the discharge from the Twin Peaks Pumping Plant probably caused localized mixing and increased turbidity levels (nevertheless, SSF outlet turbidity rarely changed, remaining near 0.2 NTU throughout the study). The SSF inlet turbidity levels during the pilot study are shown in appendix B, figure B-9.

Operating Conditions

The pilot SSF was operated at two filtration rates. From August 7, 2001, to August 2002, the 200-ft² pilot SSF operated at 22 gal/min, corresponding to the “high-level” filtration rate of 0.110 gal/min/ft² (0.27 m/h, 6.9 MGD/acre). From January 2, 2002, to March 18, 2002, the pilot SSF operated at 16 gal/min at the mid-level filtration rate of 0.080 gal/min/ft² (0.20 m/h, 5.0 MGD/acre). Table 2 describes the pilot SSF construction and operating conditions and how these compare with the proposed full-scale design (SARWMS, 2000).

Filtration Rates

During the conditioning phase (August 6, 2001 through September 11, 2001), the filter was operated with manual controls by adjusting inlet and outlet valves to achieve constant water level and to maintain 22 gal/min (0.11 gal/min/ft² through the 200-ft² filter). For more consistent level control, a simple inlet float was installed that maintained the inlet level at 55 inches above the initial sand level. On September 11, the outlet rate was set at 22 gal/min, and it was checked frequently. Few (approximately weekly) adjustments to the outlet control valve were made during the first 2 weeks of the filter run.

On January 3, 2002, the rate was adjusted to the mid-level filtration rate of 16 gal/min (0.080 gal/min/ft²) to determine if the filtration rate significantly affected the filter run period and water production.

The SSF pilot was not operated at a planned optional low-level filtration rate of 10 gal/min (0.050 gal/min/ft², 0.122 m/h, 3.1 MGD/acre) because filtrate quality and filter run lengths were found acceptable at the mid-level and high-level filtration rates.

Filter Cleaning

The progress of a filter run was tracked by monitoring the filter pressure drop. The filter pressure drop in inches of water head is calculated as the inlet water level (55 inches above sand as controlled by a float valve) minus the outlet sight tube manometer level.

The SSF component that primarily contributed to the filter pressure drop was the growth and eventual plugging of the schmutzdecke. After each filter scraping removal of the schmutzdecke, the filter pressure drop returned to 5 to 7 inches of water (see appendix B, figure B-8).

When the outlet sight tube manometer level, which was 48 to 51 inches above the sand at the start of a filter run, approached the level of the sand surface, the SSF was drained to a water level 5 to 20 inches below the sand surface. The supernatant water drained via a drain port to a level of 2 inches above the sand. Draining the last 2 inches through the bed required several (4 or more) hours. Sometimes areas of water remained perched on top of the schmutzdecke although the sand bed below had drained.

Filter restoration was done by manually scraping the top 3/8 to 1 inch of the filter surface with a hand shovel. The required scraping depth was made obvious by the color contrast of the greenish-black schmutzdecke layer versus the white sand beneath.

Scraping was performed by entering the tank via ladders and shoveling the schmutzdecke into buckets. Typically, it required 15 buckets and 1 hour of scraping to complete the job. A few

variations in the scraping methods were tested. Generally, the schmutzdecke was raked into windrows and then picked up with a shovel. On other occasions, it was shoveled directly into the buckets. In both cases, the scraped surface was leveled with a shovel and left scored with a rake. Raking tended to leave remnants of the schmutzdecke. Shoveling removed more of the schmutzdecke material, but possibly removed excess sand. No effect of the above scraping variations was observed on the time required for filter-to-drain following each scraping, water quality, or filter run length. Except at the end of run 7, permeability upon return to service was restored to that the original levels (see the “Filter Run Lengths” section below), indicating that the performance was restored to “new” and was independent of method of scraping or rate of filtration. Appendix H describes several mechanized techniques used to clean slowsand filters.

SSF Results

Filter Run Lengths

Filter run lengths and filter run water productions for filter runs 4 - 10 are shown in appendix B, figures B-1 and B-2. Runs prior to run 7 were terminated before the complete head of water was utilized (see appendix B, figure B-8) to avoid filter cleaning on weekends and holidays.

Table 3 documents the SSF filter run periods, filter run water production, and filter permeabilities for runs 7 through 10, when the SSF pilot filter was operated to utilize the complete head of water. For comparison with values reported for other plants, the SSF parameters are listed in two systems of units.

At the high filtration rate of 0.11 gal/min/ft² (0.27 m/h, 6.9 MGD/acre), the average filter run period for filter runs 7 and 8 was 22 days. At the mid-level filtration rate of 0.08 gal/min/ft² (0.20 m/h, 5.0 MGD/acre), the average filter run period for filter runs 9 and 10 was 30 days.

Letterman (Logsdon, 1991, p. 151) reports that filter run periods are site specific and have a range of 1 week to 1 year, with an average of about 1.5 months. The August 2000 SARWMS report (Reclamation, 2000; table E-3 on p. E-10) anticipated a filter run period of 2 months and six filter cleanings per year at a filtration rate of 0.112 gal/min/ft² (0.28 m/hr, 7.0 MGD/ acre). At the high filtration rate, the filter run period of 22 days observed in runs 7 and 8 is half the 1.5-month average reported by Letterman and one-third the 2-month period anticipated in the August 2000 SARWMS report. The costs associated with the shorter filter run periods and more frequent cleanings are discussed in the “Design and Cost Revisions” section.

Although the run period was shorter with the high-level filtration rate compared to the mid-level filtration rate, filter run water production was the same or higher. At the high-level filtration

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Table 3.—SSF filter run periods, filter run water production, and filter permeabilities with 200-ft² filter

	High filtration rate			Mid-level filtration rate		
Filter run number	7	8	7 and 8 average	9	10	9 and 10 average
Operating dates	Nov. 20, 2001- Dec. 14, 2001	Dec. 14, 2001- Jan. 3, 2002	—	Jan. 3, 2002- Feb. 4, 2002	Feb. 4, 2002- March 5, 2002	—
Filtration rate						
gal/min	22	22	22	16	16	16
gal/min/ft ²	0.110	0.110	0.110	0.080	0.080	0.080
m/h	0.27	0.27	0.27	0.20	0.20	0.20
MGD/acre	6.9	6.9	6.9	5.0	5.0	5.0
Filter run period, days	24	21	22.5	32	29	30.5
Initial filter dP						
in H ₂ O	6.5	9	—	5	4	—
m H ₂ O	0.16	0.23	—	0.13	0.10	—
Final filter dP						
in H ₂ O	45.5	45	45.2	55.5	54.5	55
m H ₂ O	1.15	1.14	1.15	1.41	1.38	1.40
Water production						
gallons	750,600	624,200	687,400	690,900	648,200	669,600
million gallons/acre	164	136	150	151	141	146
m = m ³ /m ²	153.1	127.3	140.2	140.9	132.2	136.6
ft = ft ³ /ft ²	502.3	417.6	460.0	462.3	433.7	448.0
Initial permeability						
gal/min/ft ² / inch H ₂ O	0.020	0.012	—	0.020	0.020	—
h ⁻¹	1.6	1.2	—	1.5	1.9	—
Final permeability						
gal/min/ft ² / inch H ₂ O	0.0024	0.0024	—	0.0015	0.0016	—
h ⁻¹	0.23	0.24	—	0.14	0.14	—

rate, the average filter run water production was 140 meters versus 137 meters at the mid-level filtration rate (in addition, the 140-meter production occurred with 10 inches less final head (dP) of water across the filter). Both values are at the lower end of the filter run production range of 112 to 650 meters reported by Letterman (Logsdon, 1991, p. 151).

Filter Permeability

After each cleaning, the filter permeability was consistently restored to 0.016 to 0.018 gal/min/ft² per inch of water head (see appendix B, figure B-7). The exception was the December 14, 2001, cleaning after filter run 7, indicating incomplete cleaning. Subsequent cleanings after filter runs 8, 9, and 10, however, restored the initial filter permeability to its previous initial levels.

At the high filtration rate of 0.11 gal/min/ft² (0.27 m/h, 6.9 MGD/acre), the initial SSF pressure drop at the start of each filter run measured 6 to 7 inches of water (the exception was the start filter run 8, when on December 15, the pressure drop measured 10 inches of water). At the mid-level filtration rate of 0.08 gal/min/ft² (0.20 m/h, 5.0 MGD/acre), the initial SSF pressure drop measured 4 to 5 inches of water. With a maximum of 4 feet of water above the filter, these pressure drops for the SSF without schmutzdecke represent 8 to 15 percent of available head.

SSF Filtrate Quality

Twenty six sets of water samples of the SSF inlet and SSF outlet were collected for water quality analyses by off-site laboratories (see appendix C, tables C-2 to C-27).

To evaluate the particulate-removal effectiveness of SSF, daily measurements of turbidity and silt density index (SDI) were recorded. In addition, the rate of fouling of a downstream pilot RO unit was observed over a 6-month period.

The SSF treatment process does not remove inorganic solutes and TDS. For documentation of water compositions, appendix C (tables C-2 to C-7) lists the inorganic water compositions in the SSF feed and product water measured during the study. Levels of TDS, hardness, and pH are discussed below.

Because of potential oxygen depletion problems at low SSF filtration rates, dissolved oxygen concentrations were measured daily.

To evaluate the physical disinfection afforded by SSF, heterotrophic plate counts were measured weekly.

To evaluate the TOC removal and disinfection byproduct levels that are produced in the SSF product, weekly TOC, monthly Trihalomethane Formation Potential (THMFP), and one simulated distribution system Disinfection Byproduct (SDSDBP) chemical analyses were conducted.

Turbidity

Turbidity, based on light-scattering, was measured daily with a Hach Turbidimeter 2100P calibrated between 20 NTU {clear} and 800 NTU {translucent}. For SSF outlet water, the instrument operated near its generally considered detection limit of 0.1 NTU. Measurement errors included use of a scratched sample cell before October 30, 2001, fingerprints on the sample cell, a cold sample (relative to the air temperature), causing condensation on the vial, and the refractive index silicon oil film on the cell being too thick or too thin. These measurement errors may have contributed to more of the data scatter than any real change in SSF product water quality.

Appendix B, figure B-9 shows the turbidity levels observed during the study. SSF inlet values ranged from 0.3 to 24 NTU. The maximum values of 12 and 24 NTU measured at the end of the test are higher than the maximum value of 10.2 NTU from 4 years of monthly turbidity readings of CAP water near the Twin Peaks site.

At Brady (42 miles upstream) and San Xavier (21 miles downstream), from 1998 - 2001, the average monthly turbidity was 1.7 NTU and the maximum was 10.2 NTU (Central Arizona Water Conservation District, 1999-2002). Because these levels are generally less than the 10-NTU maximum turbidity level recommended for SSF (AWWA, 1991), CAP water appears to meet the recommended turbidity criterion for treatment with SSF.

SSF outlet turbidity levels were consistently below 1.0 NTU. Values above 1.0 NTU were measured on eight days: five days in October, two days in December, and one day in January (a scratchy sample cell sometimes used before October 30, 2001, is the likely cause of the four high October readings). For the period of October 1, 2001, through March 19, 2002, 170 daily turbidity measurements were recorded (of which seven were greater than or equal to 1 NTU, and 162, or 95.2 percent of the daily measurements, were less than 1.0 NTU). Omitting October data due to probable measurement errors for the period of November 1, 2001, through March 19, 2002, of 140 measurements, three greater than 1 NTU, and 137, or 97.9 percent, were less than 1 NTU. Therefore, during this study, SSF treatment of CAP water met the SWTR for turbidity requiring less than 1 NTU for 95 percent of daily samples in a month.

Silt Density Index

The Silt Density Index (SDI) measurement provides data on concentration and characteristics of particulate materials in water. The measurement involves passing water maintained at a constant pressure of 30 pounds per square inch (psi) through a membrane filter with 0.45-micrometer (μm) pore size. As the measurement proceeds, the progressive blockage of the filter pores by particulates causes the filtration flow rate to decrease. The relative flow decrease yields a percent plugging factor value after a given test duration. SDI is obtained by dividing plugging factor by the filtration duration. A duration of 15 minutes is standard. If the 15-minute SDI value is greater than 5, then shorter times of 5 and 10 minutes are used. In this study, SDI values were recorded with a Chemtec Filter Plugging Analyzer model FPA-3300.

Compared with turbidity, SDI measures lower concentrations of particulates and is generally used only on filtered water. Therefore, during this study, no SDI measurements of SSF feedwater were conducted.

Appendix B, figure B-14 shows the SDI values for SSF filtrate and RO feed, where the RO feed location is downstream of chemical additions (for a description of chemicals additions, see the “Chemical Feed Systems and Operating Conditions” section) and a 5- μm cartridge filter (Osmonics II GX-05-20). In appendix B, figure B-14 is a horizontal line at SDI of 5. To avoid fouling RO membranes with a coating of particulates, a 15-minute SDI of 5 is the maximum value recommended for RO feedwater.

The SDI values were highest immediately after cleaning the SSF filter, sometimes being near 5.0. These anticipated high values generally required a filter-to-drain period of 2 hours (two sand pore volumes) before returning the SSF to RO service. In later cycles, for convenience in starting up the RO equipment during the day, the filter-to-drain period lasted overnight. During individual filter runs, the SDI values improved toward approximately 2.5 at the end of a filter run. On December 9, 2001, a second SDI unit began measuring SSF outlet upstream of chemical addition.

Table 4 lists summary statistics for the SDI measurements. Because all statistics presented assume no dependence between daily samples, and because some dependence is expected, the results may indicate tighter 95-percent intervals and higher significance than warranted.

For 96 SDI measurements of SSF outlet water from December 8, 2001, through March 19, 2002, the average is 3.59. For 162 SDI measurements of RO feed water from October 9, 2001, through March 19, 2002, the average is 3.90. The average SDI of 3.6 of SSF outlet water is within acceptable limits for RO operation.

The chemical additions caused an increase in SDI levels (no biases appeared to be introduced by the different instruments and two sample lines when checked by switching flows to the

Table 4.—SDI

Dates	SSF outlet	RO feed (SSF outlet + chemicals)
	December 8, 2001 - March 19, 2002	October 5, 2001 - March 19, 2002
Number of measurements	96	162
Mean	3.59	3.90
Standard deviation	1.60	1.06
Standard deviation of the mean	0.126	0.84
95 percent upper bound	3.84	4.06
95 percent lower bound	3.33	3.73

instruments and also by operating with both flows to the same instrument). For 15-minute SDI levels of 5 and less, the average SDI increase, the SSF outlet, and RO feed for 96 measurements was 0.45 and was significantly different from zero at the 95-percent confidence level (table 5).

Table 5.—Comparison of 15-minute SDI measurements of less than 5 at two locations

Dates	SSF outlet	RO feed (SSF outlet + chemicals)	Difference from chemical addition
	October 5, 2001 - March 19, 2002	October 5, 2001 - March 19, 2002	October 5, 2001 - March 19, 2002
Number of measurements	96	96	96
Mean	3.29	3.74	0.45
Standard deviation	0.84	0.67	0.78
Standard deviation of the mean	0.086	0.068	0.080
95 percent upper bound	3.46	3.88	0.61
95 percent lower bound	3.12	3.60	0.29

RO Fouling Rate

Operation of the pilot RO unit provided information on the effectiveness of SSF in removing particulates that foul RO membranes. The most sensitive measure of RO fouling is the RO water transport coefficient, “A.” “A” is the proportionality coefficient between the membrane water flux and the driving pressures (hydrostatic and osmotic; see appendix D).

Because “A” varies with temperature (at approximately 3 percent per °C), “A” is reported at a reference temperature of 25 °C using the following temperature correction equation (Beardsley, 2001):

$$t_{\text{corr}} = \exp (-3020 \times (1 / 298.15) - (1 / (273.15 + T))) \tag{1}$$

where T is the temperature in °C.

Temperature compensation is critical to evaluating the change in “A” because during the first 3 months of RO operation, the temperature decreased 11.3 °C, from 22.4 °C on October 18, 2001, to 11.1 °C on January 21, 2002 (see appendix B, figure B-4). Even in the absence of fouling, by equation 1, the temperature decrease alone would cause the nontemperature-compensated “A” values to decrease by 33 percent.

Appendix B, figure B-33 shows the “A” values during pilot tests. Table 6 summarizes the values of “A” for the RO pilot unit on October 18, 2001 (1 week after startup), before cleaning on January 21, 2002, and at the end of the test on March 19, 2001. During the first 3 months, the average of the “A” values for the pilot unit decreased 17 percent, or 8 percent per 1,000 hours of operation.

Table 6.—RO water transport coefficient “A” and change in “A” during the pilot test

Description	Date	Operating hours (approx.)	Parameter	RO vessel					Average
				1		2		3	
				Stage 1	Stage 2	3a	3b	(avg.)	
1 week after RO startup	10/18/01	168	A, 10 ⁻¹² m/s/Pa	23.3	23.4	23.6	22.1	22.8	23.2
Before first cleaning	1/21/02	2,328	A, 10 ⁻¹² m/s/Pa	20.7	19.1	18.5	17.1	17.8	19.2
			Change in “A” after 3 mos, % / 3 mos	-11.0%	-18%	-22%	-23%	-22%	-17%
			Change in “A” after 3 mos, %/1,000 hrs	-5.2%	-8.5%	-10%	-10.5%	-10.2%	-8%
At end of test	3/19/02	3,684	A, 10 ⁻¹² m/s/Pa	24.5	21.7	21.7	19.9	20.8	22.3
			Change in “A” after 5 mos, % / 5 mos	5%	-7%	-8%	-10%	-9%	-4%
			Change in “A” after 5 mos, %/1,000 hrs	1.4%	-2.1%	-2.3%	-2.9%	-2.6%	-1.1%

Because of the decline in “A” values, on January 21, 2002, and February 7, 2002, the RO equipment was cleaned at pH 12. The cleaning restored the “A” values to approximately their original values. The RO pilot test ended after 5 months on March 19, 2002 (with a water temperature of 13.1 °C). During the 5-month test period, the average of the “A” values for the pilot unit decreased 4 percent, corresponding to a decrease of 1.1 percent per 1,000 hours of operation. The 4-percent change between the 1-week and 5-month performances is within the expected accuracy of flow instrumentation and temperature compensation (see equation 1) and is not considered significantly different from zero.

A second measure of RO fouling is the pressure drop across each RO stage. Because pressure drop varies with flow rate and inversely with temperature, an element flow coefficient (C_e ; analogous to the valve coefficient C_v) was calculated (see appendix B, figure B-37). After October 19, 2001, (when interstage pressure lines were connected), C_e for stage 1 remained steady at 1.5 gal/min/psid^{0.6} per element, and C_e for the first three elements in stage 2 (vessel 3a) remained at 1.4 gal/min/psid^{0.6} per element. C_e for the tail three elements of stage 2 (vessel 3b) decreased approximately 20 percent, from 1.5 to 1.2 gal/min/psid^{0.6} per element, over the 5-month period. Thus, based on C_e values, the RO equipment appeared to experience fouling (and/or scaling) in the tail three elements of stage 2.

In summary, although the initial decline of “A” at the rate 8 percent per 1,000 hours indicates particulate fouling, the fouling was apparently effectively removed with two simple high-pH cleanings.

Five weeks after cleaning, and 5 months after startup, the average of the “A” values was only 4 percent less than observed at 1 week after startup. Because the initial decline in “A” was removed by cleaning, and because after 5 months the values of “A” remained at or near their 1-week level, the authors conclude that the pilot SSF served as an effective pretreatment of Colorado River water for removing particulates that foul RO membranes.

Total Dissolved Solids

The TDS (estimated daily from conductivity) of the SSF inlet and outlet waters is shown in appendix B, figure B-5. No difference in TDS was observed between the SSF inlet and outlet. From October 19, 2001 (when the conductivity instrument was calibrated) to March 19, 2002, the average TDS was 673 mg/L. TDS values based on monthly laboratory analyses are listed in appendix C, tables C-2 through C-7.

Hardness

Total hardness is the sum of the calcium and magnesium ion concentrations. For hardness expressed as mg/L CaCO₃, the relationship is:

$$\text{Total hardness [mg/L as CaCO}_3\text{]} = 2.5 \times \text{Ca}^{2+} \text{ [mg/L]} + 4.1 \times \text{Mg}^{2+} \text{ [mg/L]}$$

Total hardness can also be expressed as grains per gallon (gpg), where 17.12 mg/L as CaCO₃ equals 1 gpg.

Appendix C, tables C-2 through C-7 list monthly values of hardness, calcium, and magnesium. These are summarized in table 7.

Table 7.—SSF hardness

Date	Location	Ca ²⁺ , mg/L	Mg ²⁺ , mg/L	Total hardness, mg/L as CaCO ₃
August 28, 2001	SSF inlet	68	27	280
	SSF outlet	69	27	280
September 25, 2001	SSF inlet	74	28	300
	SSF outlet	74	27	300
October 25, 2001	SSF inlet	66	28	280
	SSF outlet	65	28	280
November 19, 2001	SSF inlet	66	29	280
	SSF outlet	66	29	280
December 17, 2001	SSF inlet	55	26	240
	SSF outlet	55	26	240
January 15, 2002	SSF inlet	65	27	270
	SSF outlet	64	27	270

The hardness levels of 240 to 300 mg/L as CaCO₃ (14 to 17.5 gpg) correspond to a qualitative classification of “hard” water where the four classifications of soft, moderately hard, hard, and very hard, are listed in table 8. No change in hardness levels was observed between SSF inlet and outlet.

SSF pH

In CAP water with calcium and bicarbonate solutes, the natural high pH of about 8.5 reduces corrosion in water pipes. For RO treatment with polyamide (PA) membranes, acid is added to RO feedwater to lower the pH to about 7 to prevent calcium carbonate scaling in the tail RO elements. Because of its low TDS, the pH of the RO product is readily increased to about 8.3 by adding a small amount of base (e.g., lime).

Table 8.—Qualitative classification of waters according to level of hardness
(adapted from Sawyer, 1994)

Description	Total hardness	
	mg/L as CaCO ₃	Grains per gallon
Soft	< 50	< 3.0
Moderately hard	50 - 150	3.0 - 9.0
Hard	150 - 300	9.0 - 17.5
Very hard	> 300	> 17.5

Appendix B, figure B-12 shows the pH levels of the SSF inlet and outlet. Table 9 summarizes the pH statistics from November 5, 2001, through March 18, 2002.

Table 9.—SSF pH

Dates	SSF inlet	SSF outlet	SSF reduction
	November 5, 2001 - March 19, 2002	November 5, 2001 - March 19, 2002	November 5, 2001 - March 19, 2002
Number of measurements	126	126	126
Mean	8.45	8.20	0.25
Standard deviation	0.09	0.25	0.25
Standard deviation of the mean	0.008	0.022	0.023
95 percent upper bound	8.47	8.24	0.30
95 percent lower bound	8.44	8.15	0.21

The average SSF inlet pH was 8.45. The average SSF outlet pH was 8.20. The SSF reduced the measured pH values by an average of 0.25. This reduction is significantly different from zero at the 95-percent confidence level.

Similar pH reductions of 0.13 and 0.18 were observed during 1987 pilot SSF tests in New Hampshire. Collins suggested the respiratory production of carbon dioxide or organic acid intermediates from NOM in the water as a possible cause for the pH reduction (Collins and Graham, 1994).

The pH reduction could also be caused by the deposition of calcium carbonate solids in the filter bed. This possibility can be explored in future tests with pre- and post-test sand analyses.

Dissolved Oxygen

Dissolved oxygen (DO) levels of the SSF inlet and outlet streams were measured daily beginning October 1, 2001, with a YSI model Y55012 dissolved oxygen meter to monitor oxygen depletion in the SSF outlet. Depletion of the DO within the SSF may create a variable and unstable operating environment.

The oxygen content is important; if it falls to zero during filtration, anaerobic decomposition occurs, with consequent production of hydrogen sulfide, ammonia, and other taste- and odor-producing substances together with dissolved iron and manganese. . . . Thus, the average oxygen content of the filtered water should not be allowed to fall below 3 mg/L if anaerobic conditions are to be avoided throughout the whole area of the filter-bed (Huisman and Wood, 1974).

Appendix B, figure B-15 shows the daily measurements of DO in the SSF inlet and outlet. Table 10 summarizes DO statistics.

At the high-level filtration rate from October 1, 2001, to January 2, 2002, DO was measured on 85 days. The DO concentration in the SSF inlet averaged 7.78. The DO concentration in the SSF outlet averaged 7.34. The water experienced an average reduction of 0.43 mg/L.

At the mid-level filtration rate from January 3, 2002, to March 19, 2002, the DO concentration in the SSF inlet averaged 7.78 mg/L. The DO concentration in the SSF outlet averaged 7.34. The water experienced an average reduction of 0.78 mg/L. At the 95-percent confidence level, this is significantly greater than the 0.43 mg/L reduction observed at the high-level filtration rate.

The DO depletions of 0.4 mg/L observed at the high-level filtration rate and 0.8 mg/L at the low-level filtration rate indicate that DO depletion was minor and presented no problem during the pilot study.

Operation with warmer summer water at low filtration rates may result in lower DO concentrations in the SSF outlet because the solubility of water is less in warm water, and the longer residence time permits more time for biological activity to deplete the oxygen. Because the study operated at mid- and high-level filtration rates, the SSF did not operate under conditions where significant oxygen depletion is likely to occur.

Heterotrophic Plate Count

Weekly water samples were collected for heterotrophic plate count (HPC) analysis of the SSF inlet, SSF outlet, RO feed, RO product, and RO reject streams. The results are shown in appendix B, figure B-10. The primary drinking water standard for HPC is a MCL of 500 bacteria colony forming units (cfu) per mL. Out of 26 sample collection times, this MCL was exceeded

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Table 10.—SSF dissolved oxygen
(mg/L)

	SSF inlet	SSF outlet	SSF reduction
High-level filtration rate (0.11 gal/min/ft ² , 0.27 m/h, 6.9 MGD/acre)			
Dates	October 1, 2001 - 2002	October 1, 2001 - 2002	October 1, 2001 - 2002
Number of measurements	85	85	85
Mean	7.78	7.34	0.43
Standard deviation	0.96	0.78	0.95
Standard deviation of the mean	0.104	0.084	0.103
95 percent interval upper	7.98	7.51	0.64
95 percent interval lower	7.57	7.18	0.23
Mid-level filtration rate (0.08 gal/min/ft ² , 0.20 m/h, 5.0 MGD/acre)			
Dates	January 3, 2002 - March 19, 2002	January 3, 2002 - March 19, 2002	January 3, 2002 - March 19, 2002
Number of measurements	77	77	77
Mean	8.45	7.68	0.78
Standard deviation	0.58	0.42	0.61
Standard deviation of the mean	0.066	0.047	0.069
95 percent upper bound	8.6	7.8	0.91
95 percent lower bound	8.3	7.6	0.64

only once (in September 2001) during the pilot study, and that was in the untreated CAP feedwater (SSF inlet), not the SSF product. Table 11 summarizes the geometric means and 95-percent confidence intervals for the SSF inlet and outlet. A geometric mean and log transformations of the data were used to help create a normal distribution of the data.

The geometric mean HPC values of the SSF inlet and outlet are 48 cfu/mL and 27 cfu/mL for a 44-percent reduction. The 95-percent confidence intervals and the t-value of 2.1 indicate that the HPC reduction by the SSF is significantly different from zero at the 95-percent confidence level.

Table 11.—SSF heterotrophic plate counts (cfu/mL)

	SSF inlet	SSF outlet	SSF reduction	t-value of reduction	Is reduction different from zero at the 95-percent confidence level?
Dates	August 28, 2001 - February 19, 2002	August 28, 2001 - February 19, 2002	August 28, 2001 - February 19, 2002		
Number of measurements	26	26	26		
Geometric mean	48	27	21	2.1	Yes
95 percent upper bound	76	44	37		
95 percent lower bound	31	17	12		

SSF Removal of Total Organic Carbon

Although an on-site TOC instrument (Anatel A-2000) was used initially, its use was abandoned on October 9, 2001. Because replicate analyses from the same sample vials produced increasingly higher TOC values—increasing from 0.5 parts per million (ppm) to 7.0 ppm.

Differences between SSF inlet and outlet TOC values were dwarfed by this analytical variation. Additionally, the instrument sample lines were constructed of very fine capillary tubes—so small as to be easily blinded by the algae and canal debris in the inlet water samples. After using the instrument for over 1 month and obtaining inconsistent values, the instrument was taken off line, and samples were sent to an off-site lab for analysis. The off-site lab seemed to have similar troubles, including intermittent measurements of high TOC values in the RO product, where very low TOC levels are expected. In summary, TOC measurements did not achieve the expected levels of accuracy during the study.

Table 12 summarizes the statistics of the weekly SSF TOC measurements from October 9, 2001, to February 19, 2002. The average SSF inlet TOC concentration was 3.3 mg/L. The average SSF outlet TOC concentration was 2.9 mg/L. The SSF reduced the measured TOC values by an average of 0.4 mg/L, or 12 percent. This reduction is significantly different from zero at the 95-percent confidence level.

SSF Disinfection Byproduct Levels

Trihalomethane Formation Potential

The THMFP performed per Standard Methods 5710B measures the trihalomethane (THM) concentrations formed during 7 days in the presence of chlorinated water. In the procedure, the

Table 12.—SSF TOC

	SSF inlet	SSF outlet	SSF reduction
Dates	October 9, 2001 - March 19, 2002	October 9, 2001 - March 19, 2002	October 9, 2001 - March 19, 2002
No. of measurements	20	20	20
Mean	3.32	2.91	0.41
Standard deviation	1.05	1.08	0.70
Standard deviation of the mean	0.23	0.24	0.16
95 percent upper bound	3.80	3.42	0.73
95 percent lower bound	2.83	2.40	0.08

free chlorine level at the end of the 7-day period is 3 to 5 mg/L. This study uses the THMFP measurement to evaluate SSF and RO removals of DBP (specifically THM) precursors. THM and THMFP measurements were performed monthly.

The THM levels of the SSF inlet and outlet (operated without chlorination) were measured as 0.0007 mg/L on August 28, 2001, and less than the detection limit of 0.005 mg/L in the other five monthly analyses.

Table 13 summarizes the SSF inlet and outlet THMFP levels and the THMFP reduction afforded by SSF. The average SSF inlet THMFP concentration was 0.133 mg/L. The average SSF outlet THMFP concentration was 0.121 mg/L. The SSF reduced the measured THMFP values by an average of 0.013 mg/L, or 10 percent. This reduction is *not* significantly different from zero at the 95-percent confidence level.

SSF Simulated Distribution System Disinfection Byproducts

Expected in a water treatment plant is post-SSF chlorination for the contact disinfection needed to meet the SWTR for giardia and virus removal and chloramination disinfection for the water distribution system. The SWTR requires a combination of filtration and chemical disinfection to provide 3-log (99.9 percent) reduction of Giardia and 4-log (99.99 percent) reduction of viruses. One way to meet these criteria is SSF and disinfection of the SSF product for 35 minutes with a free chlorine residual of 1.0 mg/L before adding ammonia to form chloramines. The expected inactivation of Giardia and viruses is shown in table 14.

To estimate the concentrations of DBPs that would be delivered to the consumer with this treatment, a SDSDBP analysis (Standard Methods 5710C and D) was conducted. For the

Table 13.—SSF THMFP levels and removals
(mg/L)

Date	SSF inlet	SSF outlet	SSF reduction
August 28, 2001	0.236	0.159	0.077
September 25, 2001	0.120	0.114	0.006
October 25, 2001	0.110	0.144	-0.034 (increase)
November 19, 2001	0.133	0.125	0.008
December 17, 2001	0.115	0.103	0.012
January 15, 2002	0.090	0.081	0.009
Number of measurements	6	6	6
Mean	0.133	0.121	0.013
Standard deviation	0.052	0.028	0.036
Standard deviation of the mean	0.021	0.011	0.015
95 percent upper bound	0.188	0.151	0.051
95 percent lower bound	0.079	0.091	-0.025 (increase)

Table 14.—Expected log-removals of Giardia and viruses
(AWWA, 1991)

Treatment	Organism	
	Giardia	Viruses
SSF	2.0	2.0
Free chlorine (35 minutes at 1.0 mg/L)	1.0	> 4.0
Total	3.0	> 6.0
SWTR requirement	3.0	4.0
Compliance expected	Yes	Yes

SDSDBP analysis, water samples were collected February 25, 2002, and delivered to the analytical laboratory (ATEL, Melmore OH, lab. no. MEL02-02127). The laboratory added 1.0 mg/L of free chlorine, added 0.2 mg/L ammonia after 35 minutes to form chloramine disinfection, and then incubated the sample for 7 days. At the end of the 7-day period, the residual total chlorine was 0.43 mg/L.

Tables 15 and 16 list the SDSDBP results. At the end of the 7-day period, the TTHM concentration (per Environmental Protection Agency [EPA] 524.2) was 0.0147 mg/L (14.7 micrograms per liter [$\mu\text{g/L}$]), and the total HAA concentration (per EPA 552.2) was 0.0074 mg/L (7.4 $\mu\text{g/L}$).

Table 15.—SDSTHM
($\mu\text{g/L}$)

Parameter	Concentration	Typical report limit
Chloroform	3.2	0.5
Bromoform	1.7	0.5
Bromodichloromethane	4.4	0.5
Dibromochloromethane	5.4	0.5
Total THM	14.7	2

Table 16.—SDSHAA
($\mu\text{g/L}$)

Parameter	Result ($\mu\text{g/L}$)	Typical report limit
Bromochloroacetic acid	3.1	1
Dibromoacetic acid	2.6	1
Dichloroacetic acid	1.7	1
Monobromoacetic acid	<1.0	1
Monochloroacetic acid	<1.0	1
Trichloroacetic acid	<1.0	1
Total HAA	7.4	6

These concentrations are less than the MCLs listed by the SDWA Stage 1 D/DBPR. The measured SDSTHM of 0.0147 mg/L is 18 percent of the 0.080 mg/L MCL for THM. The measured SDSHAA of 0.0074 mg/L is 12 percent of the 0.060 mg/L MCL for HAA5.

Because the SDSDBP procedure had less than the intended 1.0-mg/L free chlorine concentration at the end of the initial 35 minutes, the measured SDSDBP concentrations reported above underestimate the DBP levels associated with the intended disinfection. With higher chlorine dosages of 2.0 to 2.5 mg/L (instead of 1.0 mg/L) and contact times longer than 35 minutes, the SDSDBP concentrations are expected to be higher but probably still less than the DBP MCLs. To confirm or correct the SDSDBP concentrations, SDSDBP analyses should be repeated in future tests.

SSF Post-Test Sand Inspection

Upon completion of the pilot test, the sand was inspected. Several core samples were obtained near the center of the tank. Test pits were dug at the center and near the outside circumference. The test pits and core samples were visually examined, and gradations were run on several of the samples. See appendix A for pictures taken during the examination.

The visual analysis of both core samples and test pits indicated that the sand cross-section below the schmutzdecke appeared unaltered. The schmutzdecke layer, about one-half inch thick, was clearly visible due to the slightly darkened appearance. No signs of piping or channeling were seen in any of the test pits.

Gradations were performed at the near-surface (schmutzdecke layer), at a depth of 2 feet, and at a depth of 3 feet. The surface layer gradation contained a slightly higher fines content of 1.3 percent compared to a pretest gradation reading of 0.5 percent fines. The 2- and 3-foot depth gradations indicated a fines content of 0.3 percent, compared with a pretest average reading of 0.5 percent.

In summary, there appeared to be essentially no change in the fines content of the sand based on pre- and post-pilot gradation analyses. The fines content in the top schmutzdecke layer increased slightly as would be expected. The gradations at depth indicated fewer fines than at pretest, possibly due to flushing action during the test. No piping or channeling of the flow through the filter cross-section was observed. The flow through the filter matrix was evidently very uniform.

The sand used in the pilot test provided a very effective filter. Very little (if any) migration or penetration of fines was observed.

Reverse Osmosis Description

RO pilot equipment was operated to evaluate the effectiveness of SSF for removing particulates that foul RO membranes. The RO pilot equipment was also operated to demonstrate operation of the complete SSF - RO process, document RO solute removals and product water compositions, and determine if disinfection of the RO product can use free chlorine without ammonia and still meet the SDWA Stage 1 D/DBPR of 0.080 mg/L TTHM and 0.060 mg/L of HAA5.

RO is a barrier process that uses hydraulic pressure to overcome the osmotic pressure of the feed flow. Osmosis is a natural process whereby water is transported through a semipermeable membrane from a solution of low concentration to one of high concentration. For example, if fresh water and salty water are separated by a semipermeable membrane, the fresh water will tend to move through the semipermeable membrane in an attempt to equalize the salt concentrations of the waters on both sides of the membrane. This tendency produces a driving force that operates in a manner similar to pressure. The term “osmotic pressure” is used to describe it.

RO operates by applying sufficient pressure on the feedwater (salty) side of the membrane to force water through the membrane to the fresh water side, thus reversing the osmotic process. The pressure required depends on several factors, including the TDS salinity of the feedwater, temperature of the feedwater, the membrane water transport coefficient, the design membrane flux (gallons of permeate per square foot of membrane per day [gfd]), and the water recovery. Recovery is the percentage of feedwater recovered as RO product. RO membranes permit very little passage of dissolved salts so that the RO product TDS is much lower than the feedwater TDS.

Another aspect of RO as a barrier process is it rejects other contaminants in the feedwater in addition to dissolved salts. RO rejects not only *Giardia* and *Cryptosporidium*, but also viruses, dissolved salts, and many organic solutes (often measured as TOC). Because many organic solutes form DBPs when free chlorine is used for disinfection, RO removal of these DBP precursors greatly reduces the levels of DBPs in the finished water. Present and proposed regulations for THM, HAA, and DBPs are readily met with RO treatment. Because of the very low DBP levels, use of free chlorine (instead of chloramine) becomes an option for disinfection of the finished water.

There are several types of RO membranes that can be used. The selection of the appropriate membrane for a particular case depends on the TDS level of the feedwater and the permeate quality goals. “Nanofiltration” is a form of RO that is useful in softening hard water. Nanofiltration membranes are most effective in rejecting divalent ions (ions with a positive or negative charge of two) such as calcium, magnesium, and sulfate. Typically, rejection rates for divalent ions are in excess of 95 percent, whereas the rejection rates for monovalent ions (ions with a

single charge) are much lower (approximately 50 to 70 percent). Nanofiltration typically operates at feedwater pressures of about 80 to 100 psi. Depending on the feedwater quality, permeate recoveries of 80 percent or more are common.

For treatment of Colorado River water with a TDS of about 700 mg/L, low-pressure RO membranes appear to have the appropriate combination of low to moderate operating pressures and moderate salt rejection. Applicable commercial RO membranes include DOW FilmTec model NF90 (where NF indicates nanofiltration RO), Fluid Systems model TFC ULP (where ULP indicates ultra-low pressure RO), and Hydranautics model ESPA 3 (a.k.a. ESNA). The recommended RO design in the SARWMS report was based on the FilmTec NF90 membrane projections with fouling allowances estimated from pilot tests conducted by Reclamation with the City of Tucson (Reclamation, 2002).

Particulates in feedwater, as well as microbiological growth in the RO equipment, can coat or “foul” RO membranes. Fouling causes loss of flux and/or an increase in operating pressure. Fouling can sometimes be removed by cleaning, but at additional costs. If the fouling coating severely restricts the passage of water and cannot be removed, the resulting low-water flux and/or high operating pressures require premature replacement of the RO membranes.

The RO membrane chemistry type selected in the August 2000 SARWMS design and operated in the pilot test is PA. PA membranes operate at lower pressures and have correspondingly lower energy costs than cellulose acetate (CA) membranes.

PA membranes are much more sensitive than CA membranes to fouling by particulates in the feedwater. Because membrane fouling results in higher operating pressures, it can partially offset some of the energy savings associated with PA membranes and may require frequent cleaning.

For pilot tests evaluating the effectiveness of SSF in removing particulates that foul RO membranes, the fouling sensitivity of PA membranes provides a stricter standard than would CA membranes. If PA membranes can operate with low fouling with SSF pretreatment, then one would expect that CA would operate with much lower fouling. Membrane manufacturers continue to develop improved PA membranes with increased resistance to chlorine degradation and fouling.

For treating Colorado River water, less acid is need with PA membranes than with CA membranes, which require water acidification to about pH 5.5. PA membranes, however, degrade in the presence of free chlorine.

A critical design parameter for inland desalting plants is water recovery, the percentage of product water that is recovered from the feedwater. Achieving the highest water recovery possible without precipitation (scaling) of sparingly soluble solutes in the concentrate is generally the goal for inland sites and for brackish (waters with TDS of 3,000 to 10,000 mg/L) and lower

salinity waters. Water recovery is a critical design parameter because high recovery corresponds to high utilization of the feedwater supply and to low volumes of waste concentrate to be discharged.

The design water recovery is 85 percent. Recovery appears to be limited by barium sulfate solubility (barium and sulfate concentrations in the design CAP Canal water are 10 times barium sulfate solubility). A conservative estimate of maximum barium sulfate concentration in the reject stream is 60 times solubility. When a mix of Colorado River water and Aqua Fria River water (from Lake Pleasant) is delivered with a TDS of 700 mg/L, 60 times solubility corresponds to a water recovery of 72 to 79 percent. At the design 85 percent recovery and feed pH adjusted to 7.2 with sulfuric acid, the barium sulfate concentration is about 120 times solubility.

The CAP Canal operation includes periods when 100 percent of Colorado River water is delivered (Central Arizona Water Conservation District, 1997, 1998). With the Colorado River water composition, barium sulfate remains the principal scaling compound of concern. At 85-percent recovery and feed pH of 7.2, the barium sulfate concentration is about 130 times solubility.

Based on pilot test results at the Hayden-Udall Water Treatment Facility and the Water Quality Improvement Center, operation at 85-percent water recovery (at 120 to 130 times the solubility of barium sulfate) appears feasible with the use of antiscalants. Higher water recoveries may be achievable but require confirmation with additional high-recovery pilot tests on Colorado River water.

The design water flux is 12 gfd, and the intent was to operate the pilot RO equipment at 12 gfd based on FilmTec Product Information listing a membrane area of 23 square feet for 2.5-inch-diameter by 40-inch-long RO elements. The membrane area, however, has apparently been increased because the DOW FilmTec website (<http://www.dow.com/liquidseps/pc/nfe.htm>) as of June 6, 2002, lists an active element area of 28 square feet for NF90-2540 elements. Post-test dissections of two NF90-2540 test elements measured an active membrane area of 25.4 square feet. With 25.4 square feet of membrane area per element, the RO equipment operated at 10.9 gfd.

SSF - RO Pilot Process Description

A flow schematic of the pilot SSF - RO process is shown in figure 4. The pilot SSF is contained in a 16-foot-diameter tank and has an area of 200 square feet. SSF feedwater was drawn from the CAP Canal at the discharge of the Twin Peaks pumping station in Marana, Arizona. No chemical treatments (e.g., disinfection or coagulation) were used on the SSF feed.

Between the SSF outlet and feed to the RO processes, four chemicals were added. Antiscalant (Ecolab Flocon 100) was added to prevent scaling by barium sulfate. Sulfuric acid (J.T.Baker CMOS 9684-05; 95 - 97 percent H₂SO₄) was used to adjust the RO feed pH from about 8.3 to 7.0 to avoid scaling by calcium carbonate. Bleach (Clorox, 6 percent) was used for chlorination disinfection. Ammonium sulfate (Fisher A702-10 technical grade) was added to combine with the free chlorine to produce chloramine disinfection.

The August 2000 SARWMS report (Reclamation, 2000) describes a 24-MGD (27,000 acre-feet per year) RO design with a water recovery of 85 percent. Because three-stage RO equipment is recommended for achieving greater than 80-percent recovery, and because the MTP pilot system has only two stages, a water recovery of 80 percent was used during the pilot tests.

The pilot test was not designed to address biological fouling of commercial PA membranes operated with no disinfectant in the RO feedwater. Because commercial PA membranes are considered susceptible to damage and loss of salt-rejection with chlorinated water, they are not presently recommended for operation on chlorinated or chloraminated water. For this short-term study, however, to obtain an accurate picture of particulate fouling in the absence of biological fouling, chloramine disinfection was used in the RO feedwater.

Chemical Feed Systems and Operating Conditions

The chemical feed systems consist of four ProMInent diaphragm metering pumps (ProMInent Gala and g/4b), two static mixers (Chemineer Kenics HEV-6) in a 30-gal/min mixing loop (refer to figure 4). On the suction side of the RO feed pump, four chemicals are added in a specific order: antiscalant, acid, bleach, ammonia (see figures 7 - 10). These were added in a rapidly flowing mixing loop. The order of addition was considered important. Because adding sulfuric acid adds sulfates, which could begin forming barium sulfate scale, the antiscalant was added upstream of the acid injection point. This mixture was then passed through a static mixer. Sulfuric acid was added to lower the pH from 8.3 to 7.0 to prevent calcium carbonate scaling. Next, Clorox brand bleach was injected. For operation with polyamide membranes, ammonium sulfate is added to convert free chlorine to chloramine disinfection. The pumped mixing loop was used to provide adequate flow through the static mixers and served to externally mix the 150-gallon RO feed tank (table 17).

The process flow is 6.0 gal/min. Dilution water is RO product with less than 25-mg/L TDS and with chloramine residual.

Post-test review of the doses listed in table 17 revealed that 1/10th the intended dose was used for the antiscalant because of a dilution error. The dilution error resulted in an antiscalant dose of 0.4 mg/L instead of the intended 3.6 mg/L and may have contributed to scaling indicated by the analyses of solids scraped from the tail element (see the “Scaling” section below).

Table 17.—Chemical feed operating criteria and doses

Chemical	Source	Concentration (percent)		Criteria and dose of 100 percent chemical	Dose	
		Neat	Diluted		Criteria	mL/min
Antiscalant	Ecolab Flocon 100	100	0.5	3.6 mg/L	1.8	0.4
Sulfuric acid	J.T.Baker CMOS 9684-05	95 - 97	5.2	pH = 7.0	5.0	15
Sodium hypochlorite	Clorox	6	1.0	1.0 to 1.5 mg/L residual chlorine	3.15	1.4
Ammonia (ammonium sulfate)	Fisher A702-10 technical grade	26 as NH ₃	1.2	Excess NH ₃	4.0	2.2

RO Pilot Plant Operation

Reverse Osmosis Equipment and Operating Conditions

The MTP RO equipment consists of a series of six 2.5-inch, 3-element pressure vessels that contain a total of 18 membrane elements in a 12:6 element (and 2:1 6-element vessel) array (refer to figure 4). A VFD pump boosts the feed pressure to about 90 psi, where the feedwater splits to feed vessel 1a & 1b (V1) and vessel 2a and 2b (V2). The pressurized concentrate from V1 and V2 feeds V3. V1 and V2 each contain six 2.5-inch-diameter by 40-inch-long “2540” spiral-wound NF90 elements by Dow FilmTec. Water which passes through the RO membrane travels through a spiral path and collects into a center product water tube. The feedwater which does not permeate through the first element leaves the annular space of the first element and enters the annular passages of the second element. This inter-element water is at slightly lower pressure than seen by the first element and slightly more concentrated—being deprived of the

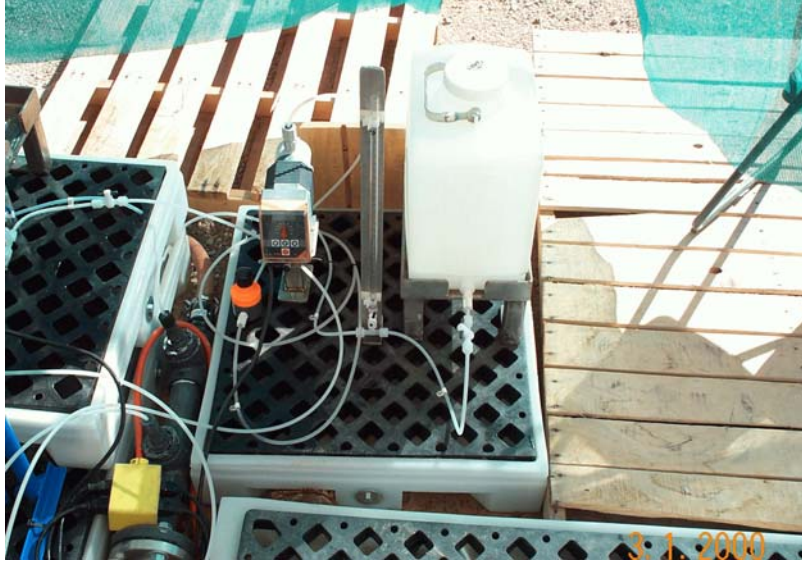


Figure 7.—Antiscalant feed system (Flocon 100) with spill pallet, 5-gallon supply tank, calibration cylinder, and back pressure valve.



Figure 8.—Acid feed system (sulfuric acid), with spill pallet, 17-gallon supply tank, calibration cylinder, and back pressure valve.

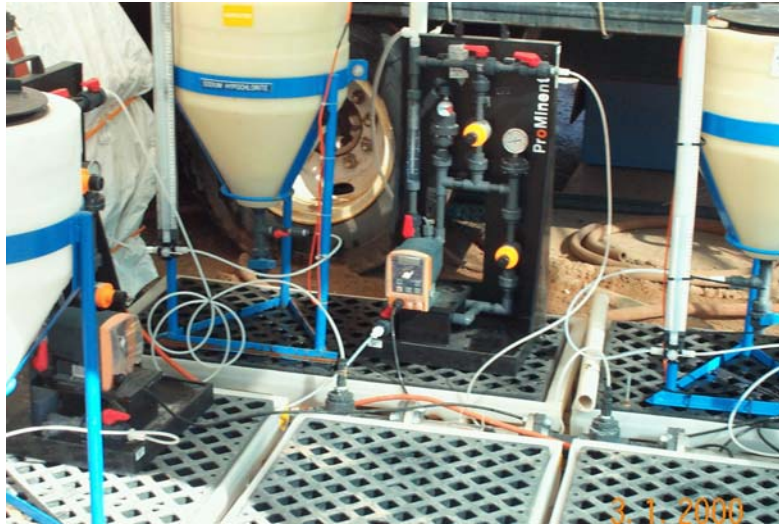


Figure 9.—Chlorine feed system (bleach), with spill pallet, 17-gallon supply tank, calibration cylinder, bypass relief line, pulsation dampener, and back pressure valve.



Figure 10.—Ammonia feed system (ammonia sulfate), with spill pallet, 17-gallon supply tank, calibration cylinder, bypass relief line, pulsation dampener, and back pressure valve.

permeate water produced by the first element. At slightly lower hydrostatic pressure, ca. 11 percent higher osmotic pressure, and ca. 11-percent higher feed salinity, each succeeding identical element produces slightly less permeate water and with higher salinity than the permeate water of the preceding element.

As RO product water is produced from the six elements of V1, it is collected into a common product tube and directed to a sample sink. Just prior to entering the sink, it encounters a needle valve which controls the RO product back pressure and hence the volume of RO product. The discharge from the needle valve is measured by a rotameter flowmeter.

The rotameter calibration was checked by graduated cylinder, stopwatch, and mass scale. From this mass measurement, a rotameter float set point was determined, and frequent RO product flow rate checks could be made quickly and accurately. In the pilot RO unit, the product valves were used to approximately equalize element product flows for both stages. In a full-scale RO plant, interstage booster pumps may be used to increase pressures by about 30 psid per stage to compensate for the hydrostatic pressure drop across each stage and for the higher osmotic pressure of the feedwater in each succeeding stage.

The flow through V2 also encountered six elements and produced P2 RO product. The interstage concentrate from V1 and V2 enters V3a, which has three elements and produces RO product as P3a. The second stage concentrate from V3a enters V3b and produces P3b RO product and the final rejected concentrate as R.

The RO reject and four product flows were controlled by manually adjusting the reject valve (a flow control valve on V3b discharge), the four RO product valves, and the VFD speed. When the system production was affected by water temperature, fouling, or cleaning, flow rates were maintained approximately constant by daily adjustments to the RO reject and four RO product valves and infrequent (three times during the 5-month test) adjustments to the RO feed pump VFD.

A portion of the RO product was diverted and collected in a RO product tank. During normal operations, RO product was measured and then discharged back to the CAP Canal.

RO Startup, Cleaning, and Shutdown Procedures

When starting the RO equipment after installing new elements, or restarting after being down for more than 4 hours, the process started at low pressure and flow to purge the RO equipment of air bubbles. Initially, the feed-forward pump operated at about 30 psi with the reject valve fully open and the RO product valves closed. After about 5 minutes, the reject and RO product valves were brought to about 80 percent of their normal operating positions. After rechecking the chemical metering systems, the VFD pump was ramped up (taking about 5 minutes to reach

full speed). Flow control valves and pump speed were adjusted in rapid rotation to achieve the specified product and reject flows (table 18). As flows and pressures increased, the sound of remaining air in the equipment was noticeable. After this 15-minute startup sequence, the system would be nearly stable. Minor changes continued to occur during the next 4 hours. After this startup, the system usually remained stable for weeks before the next adjustment was required.

Table 18.—Reverse osmosis pilot equipment flow rate setpoints for 80-percent water recovery

Stream location	Flow rate setpoint (L/min)
Feed	16.32
Reject	3.26
Total product	13.06
Product, stage 1	
Vessel 1 (six elements)	4.35
Vessel 2 (six elements)	4.35
Product, stage 2	
Vessel 3a (three elements)	2.39
Vessel 3b (three elements)	1.96

Because of the observed decline in water transport coefficient “A” during the first 3 months of the test, the RO equipment was cleaned in January and February. With SSF pretreatment and the absence of iron or aluminum coagulants, high-pH cleaning solutions were used to address fouling by organic matter.

On January 21, 2002, an 18-gallon batch of chlorine-free RO product was heated to 20 °C and adjusted to pH 12 with sodium hydroxide (Red Devil lye). This solution was circulated through the RO system at low pressure and high velocity.

Because the target water temperature of 30 °C (see appendix E) was not achieved, on February 7, 2002, a second cleaning was conducted with RO product at pH 12 and 30 °C.

Prior to both cleaning events, the RO equipment operated without chloramine disinfection to produce RO product water free of chloramines. The cleaning solutions were circulated for several hours, while maintaining the original temperature with stainless steel immersion heaters (Cole Parmer U-03046-50). During the January 21 cleaning, foaming and minor amounts of particulates (pepper flake in appearance) appeared in the return line (these particulates could have come from the action of the cleaner on dirty rubber circulation hoses). During the February 7 cleaning, the water was slightly foamy but remained clear with a slight tea color.

When the RO equipment was shut down for SSF cleaning, it was flushed with collected RO product until the reject concentration was essentially equal to inventoried RO product concentration. This flush took approximately 15 minutes and three system volumes at 5 gal/min. (one system volume is approximately 15 gallons). The RO equipment operated at lower pressure after each RO flush, as described in the “Reverse Osmosis Results” section.

The RO reject (concentrate) water was discharged into an evaporation bed. Figure 4 shows the SARWMS pilot plant process and instrumentation drawing for the MTP.

Reverse Osmosis Results

Operating conditions and performance are shown in figures in appendix B. Water compositions, mass balance checks, and solute removals are listed in tables in appendix C.

Membrane Equations

To monitor and evaluate RO equipment performances under varying feedwater and operating conditions, mathematical descriptions of the RO process can be used to try to eliminate operational variations to get a clearer description of the performance of the RO equipment at constant operating conditions. Sometimes called “normalization” equations, the equations used in this report, as well as a symbol list and sample calculations, are described in appendix D WQIC Reverse Osmosis Equations.

Appendix D also describes mass balance equations used to check RO flows and water composition listed in appendix C.

Water Flows, Compositions, and Operating Pressures

To maintain RO product membrane fluxes of 10.9 gfd and 80-percent water recovery, the RO feed and RO product pressures were adjusted. The RO feed flow was maintained at 16.3 L/min

(4.31 gal/min). The total RO product was maintained at 13.06 L/min (3.45 gal/min). The RO reject (concentrate) flow was maintained at 3.26 L/min (0.86 gal/min), which is slightly less than the minimum 3.8 L/min (1.0 gal/min) recommended by the manufacturer.

The RO pilot equipment operated with a maximum hydrostatic pressure difference of 644 kilopascals (90 psid) across the RO membranes (see appendix B, figure B-31). This maximum pressure occurred December 14, 2001, the coldest day of the test, with a feedwater temperature of 8.9 °C.

Tables 21, 22, and 23 summarize the RO performance at the start of the test, after 3 months, and at the end of the test. For comparison, the tables also show the DOW FilmTec Reverse Osmosis System Analysis (ROSA) version 5.0 prediction for the RO pilot unit. Although the ROSA program lists an active membrane area of 28 square feet per element, post-test dissection of two test elements measured an active membrane area of 25.4 square feet per element.

The RO product TDS estimated from conductivity on the three dates ranged from 15 to 23 mg/L, approximately 30 percent of the 55- to 79-mg/L levels predicted by ROSA. The individual solute concentrations in the RO product were frequently less than the laboratory detection limit, and none exceeded the ROSA predictions.

RO pilot equipment pressure measurements were used to estimate the ROSA fouling factor where a fouling factor of 1.0 corresponds to “new” membrane elements and a positive value less than 1.0 corresponds to fouled elements. At 2 weeks after startup, the estimated fouling factor is 0.81 (see table 19). After 3 months and observed fouling, the estimated fouling factor is 0.70 (see table 20). At the end of the test after 5 months of operation and 5 weeks after a pair of cleanings, the estimated fouling factor is 0.86 (see table 21). The final fouling factor of 0.86 indicates that the high-pH cleanings conducted in the 4th month (January and February) were effective in removing deposited foulants and restoring RO operation to “near new” performance.

Fouling – RO Fouling Rate

The ROSA fouling factors estimated above (see tables 21, 22, and 23) of 0.87 at 2 weeks after start and 0.75 at 3 months after start indicate that the RO equipment required higher operating pressures due to fouling over the first 3 months. The estimated fouling factor of 0.93 at 5 months after start (and 5 weeks after cleaning) indicates that the RO performance was restored to better than at 2 weeks after start.

As described above under the SSF “RO Fouling Rate,” section, the most sensitive measure of RO fouling is the RO water transport coefficient “A” at a reference temperature of 25 °C. “A” is the proportionality coefficient between the membrane water flux and the driving pressures,

Table 19.—Comparison of observed and projected RO performance at beginning of test

Date	October 25, 2001					
Test period	2 weeks after start					
Configuration	Eighteen NF90-2540 elements in a 2:1 array of 6-element vessels					
Feed temperature, °C	20.9					
Product flow, gal/min	3.46					
Feed flow, gal/min	4.31					
Water recovery	80.4%					
	Measured			ROSA prediction		
Water composition	RO feed	RO product	RO reject	RO feed	RO product	RO reject
Bicarbonate - HCO ₃ (mg/L)	61	8.54	281	61	16.8	240
Calcium - Ca (mg/L)	65	<4	290	65	1.1	324
Chloride - Cl (mg/L)	65	7.8	310	147	26.1	457
Magnesium - Mg (mg/L)	28	<3	130	28	0.3	140
Potassium - K (mg/L)	6	<5	30	6	1.5	26
Silica - SiO ₂ (mg/L)	8	0.4	39	8	1.2	35
Sodium - Na (mg/L)	88	7.3	390	88	22.0	356
Strontium - Sr (mg/L)	2	<0.2	9	2	0.02	8
Sulfate - SO ₄ (mg/L)	260	<5	1,200	260	3.0	1,302
Total dissolved solids - TDS (mg/L)	581	<41.9	2,666	685	72.0	2,891
TDS (residue at 180 °C) (mg/L)	690	130	3,000	685	72.0	2,891
TDS estimated from conductivity (mg/L)	694	22.8	2,943	685	72.0	2,891
Lab pH (S.U.)	6.8	5.2	7.3	6.8	4.6	7.5
	Measured			Adjusted to match measured pressures		
Pressures, psig	82	32.5 (stg. 1)	69	85	32.5 (stg. 1)	61
		21.5 (stg. 2)			21.5 (stg. 2)	
ROSA fouling factor estimated from measured pressures	0.81					

Note: To achieve ion balance in the RO feedwater, the ROSA program added 46 mg/L of chloride.

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Table 20.—Comparison of observed and projected RO performance midway through test

Date	January 15, 2002					
Test period	3 months after start					
Configuration	Eighteen NF90-2540 elements in a 2:1 array of 6-element vessels					
Feed temperature, °C	13.3					
Product flow, gal/min	3.45					
Feed flow, gal/min	4.35					
Water recovery	79.5%					
	Measured			ROSA prediction		
Water composition	RO feed	RO product	RO reject	RO feed	RO product	RO reject
Bicarbonate - HCO ₃ (mg/L)	61	1.22	262	61	13.1	247
Calcium - Ca (mg/L)	64	4	300	64	0.9	309
Chloride - Cl (mg/L)	59	3.8	270	110.8	19.7	464
Magnesium - Mg (mg/L)	27	3	120	27	0.2	131
Potassium - K (mg/L)	6.1	5	27	6.1	1.1	25
Silica - SiO ₂ (mg/L)	6.9	0.6	30	6.9	0.8	31
Sodium - Na (mg/L)	86	5	380	86	16.6	355
Strontium - Sr (mg/L)	1.7	0.2	9	1.7	0.01	8
Sulfate - SO ₄ (mg/L)	250	5	1,100	250	2.1	1,211
Total dissolved solids - TDS (mg/L)	560	28.6	2,486	614	55.0	2,784
TDS (residue at 180 °C) (mg/L)	520	20	2,600	614	55.0	2,784
TDS estimated from conductivity (mg/L)	685	14.8	2,905	614	55.0	2,784
Lab pH (S.U.)	6.8	4.8	7.3	6.8	6.2	7.5
	Measured			Adjusted to match measured pressures		
Pressures, psig	94	26 (sig 1) 6 (sig. 2)	78	98	26 (sig. 1) 6 (sig. 2)	70
ROSA fouling factor estimated from measured pressures				0.70		

Note: To achieve ion balance in the RO feedwater, the ROSA program added 52 mg/L of chloride.

Table 21.—Comparison of observed and projected RO performance at end of test

Date	March 19, 2002					
Test period	5 months after start					
Configuration	Eighteen NF90-2540 elements in a 2:1 array of 6-element vessels					
Feed temperature, °C	13.1					
Product flow, gal/min	3.44					
Feed flow, gal/min	4.34					
Water recovery	79.3%					
	Measured			ROSA prediction		
Water composition	RO feed	RO product	RO reject	RO feed	RO product	RO reject
TDS est. from conductivity (mg/L)	694	18.8	2,499	685	60.0	3,055
Field pH (S.U.)	7.0	5.9	7.3	7.0	6.3	7.7
	Measured			Adjusted to match measured pressures		
Pressures, psig	93	33 (stg. 1)	76	97	33 (stg. 1)	69
		15 (stg. 2)			15 (stg. 2)	
ROSA fouling factor estimated from measured pressures	0.86					

Note: Because the RO feed conductivity values were the same on October 25, 2001, and March 19, 2002, the October 25, 2002, RO feed water composition was used in the ROSA prediction for March 19, 2002.

both hydrostatic and osmotic, and its calculation is described in appendix D. High values of “A” correspond to high membrane water flux and/or low feed pressure. For the pilot test operated at constant RO product flow, high values of “A” correspond to low feed pressure. Appendix B, figure B-33 shows the values of “A” during the pilot test.

During the first 3 months (see table 6), the average of the “A” values for the pilot unit decreased 17 percent, or 8 percent per 1,000 hours of operation.

Because of the decline in “A” values, on January 21, 2002, and February 7, 2002, the RO equipment was cleaned at pH 12. The cleaning restored the “A” values to approximately their original values. The RO pilot test ended after 5 months on March 19, 2002. During the 5-month test period, the average of the “A” values for the pilot unit decreased 4 percent, corresponding to a decrease of 1.1 percent per 1,000 hours of operation (see table 6). The 4-percent decrease in performance between the beginning of the test (1-week after start) and the end of the test 5 months later is within the expected accuracy of flow instrumentation and temperature compensation (see equation 1) and is not considered significantly different from zero.

Another measure of RO fouling is the pressure drop across each RO stage. Because pressure drop varies with flow rate and inversely with temperature, an element flow coefficient (C_e ; analogous to the valve coefficient C_v) was calculated (see appendix B, figure B-37). After October 19, 2001, (when interstage pressure lines were connected), C_e for stage 1 remained steady at 1.5 gal/min/psid^{0.6} per element, and C_e for the first three elements in stage 2 (vessel 3a) remained at 1.4 gal/min/psid^{0.6} per element. C_e for the tail three elements of stage 2 (vessel 3b) decreased approximately 20 percent, from 1.5 to 1.2 gal/min/psid^{0.6} per element, over the 5-month period. Thus, based on C_e values, fouling was not observed in stage 1 or the lead three elements of stage 2. Based on a 20-percent decrease in C_e values for vessel 3b, the RO equipment appeared to experience fouling in the tail three elements of stage 2.

In summary, fouling was evident during the first 3 months of operation. The estimated ROSA fouling factor decreased from 0.81 to 0.70, for a relative decrease of 14 percent. This decrease is consistent with the change in calculated “A” values, which decreased by 17 percent.

Although the initial declines of fouling factor and “A” at the rate of 7 percent and 8 percent per 1,000 hours indicates particulate fouling, the fouling was apparently effectively removed with two simple high-pH cleanings. Five weeks after cleaning and 5 months after startup, the estimated fouling factor was 0.93 (higher than the 0.87 value at 2 weeks after start), and the average of the “A” values was only 4 percent less than observed at 1 week after start. Because the initial decline in fouling factor and “A” was removed by cleaning, and because after 5 months the values of fouling factor and “A” remained at or near their initial 1- and 2-week levels, we conclude that the pilot SSF served as an effective pretreatment of Colorado River water for removing particulates that foul RO membranes.

Scaling

In the RO process, the feedwater becomes more concentrated between the lead and tail membrane elements. The high concentrations in the tail elements may cause precipitation out of solution of sparingly soluble solutes. This precipitation is called “scaling.” Whereas fouling coats the membrane with particulates already in the feedwater, scaling coats the membrane with particulates formed within the RO equipment. Scaling and fouling have similar operational effects, but initially, scaling generally occurs only in the tail elements. For CAP water, scale-forming compounds of concern are barium sulfate, calcium sulfate, strontium sulfate, and calcium carbonate. Organic solutes from dissolved natural organic matter and organic antiscalant may also possibly precipitate out of solution and coat the RO membrane.

The values of “A” declined during the first 3 months, and the rate of decline was somewhat higher in stage 2 than in stage 1 (see table 6 and appendix B, figure B-33). The stage 2 decrease in “A” could be from either fouling or scaling.

The value of C_c for stage 2 decreased 20 percent during the test while the value of C_c for stage 1 did not decrease (see appendix B, figure B-37). The stage 2 decrease in C_c could be from either fouling or scaling.

Mass balance deviations (MBDs) with high positive values may indicate deposition of solutes within the RO equipment and are another indication of scaling. The MBDs for mineral and organic solutes in four monthly chemical analyses are listed in appendix C, tables C-4, C-5, C-6, and C-7.

Barium levels are generally below detection in the RO feed, so the MBD for barium cannot be calculated.

The highest MBD for calcium was calculated for October 25, 2001, to be between 7.5 and 12.4 percent. This is within the accuracy of the chemical analysis and the flow measurements and is not considered significant.

All MBDs for sulfate are less than 10 percent.

None of the MBDs for strontium are positive.

Organic solutes from dissolved natural organic matter and organic antiscalant may possibly precipitate out of solution and coat the RO membrane. The MBDs for TOC are listed in appendix C, figures C-2 through C-25 and C-27. On six dates (October 25, 2001; November 19, 2001; December 4, 2001; December 17, 2001; January 15, 2002; and February 19, 2002), the MBDs for TOC are positive (with an average bracket of 13 to 29 percent). On the other dates,

the TOC is zero (within the accuracy of the measurements) or negative. With both positive and negative MBDs for TOC, the MBD does not indicate a deposition of TOC in the RO equipment.

Appendix F describes post-test dissections of one lead element (from vessel 1) and one tail element (from vessel 3b). The tail element operated at approximately 50 percent lower water flux and processed approximately 50 percent less water during the pilot test. Scrapings of one leaf (with an area of $12.7 \text{ ft}^2 = 1.18 \text{ m}^2$) measured 5.6 times the mass of solids on the tail element than on the lead element. After drying at $105 \text{ }^\circ\text{C}$, the density of solids calculated was $70.1 \text{ mg/m}^2 (= 82.7 \text{ mg}/1.18 \text{ m}^2)$ for the lead element and $393 \text{ mg/m}^2 (= 464 \text{ mg}/1.18 \text{ m}^2)$ for the tail element.

Eleven times the mass of carbon compounds was found on the tail element compared to the lead element. The difference in mass between drying at $105 \text{ }^\circ\text{C}$ and $600 \text{ }^\circ\text{C}$ is called “loss on ignition (LOI)” and provides an estimate of the carbon compound content of the solids. The LOI measured 31 percent for the lead element and 62 percent for the tail element (see appendix F, “Membrane Foulant Analysis”). Assuming that all of the LOI consists of carbon compounds, then the density of carbon compounds scraped from the membranes can be calculated as $22 \text{ mg/m}^2 (= 0.31 \times 70.1 \text{ mg/m}^2)$ for the lead element and $244 \text{ mg/m}^2 (= 0.62 \times 393 \text{ mg/m}^2)$ for the tail element.

Possible sources of carbon compounds include NOM in the water and antiscalant. Calcium carbonate may also be present, but a high concentration of calcium carbonate is considered unlikely because of the success in cleaning the RO membranes at pH 12.

The inorganic analyses indicate possible scaling with barium sulfate, calcium sulfate, and strontium sulfate. Appendix F, figures F-1 and F-2 summarize the results of inductively coupled plasma (ICP) analyses of the solids after drying at $600 \text{ }^\circ\text{C}$. Scaling of the tail element is indicated because the barium, calcium, and strontium mass concentrations in appendix F, figure F-1 are greater than ten times higher for the tail element than for the lead element. The sulfur concentration is four times higher for the tail element than for the lead element.

In summary, ICP analyses of solids scraped from the lead and tail elements indicate possible scaling of the tail element by barium sulfate, calcium sulfate, and strontium sulfate. This is consistent with the observations of somewhat higher rates of decline for “A” and C_c in stage 2 than in stage 1. In contrast, the MBDs do not indicate scaling.

Because ICP analysis of solids scraped from the tail element indicates that scaling may have occurred at a water recovery of 80 percent and because incorrectly low levels of antiscalant were used, further tests are needed to evaluate RO water recovery levels that are achievable without scaling. Monitoring of the performance of the individual tail element in future tests can provide more sensitive and rapid detection of changes in “A” due to scaling.

Salt Passage

Appendix B, figure B-29 shows the RO salt passage calculated from TDS values estimated from daily conductivity measurements. Because the RO equipment operated at constant water flow, salt passage varies with water temperature. The salt passage of the FilmTec NF90 elements measured 2.4 to 3.0 percent with a water temperature of 21.8 °C at the start of the test and 1.3 to 1.9 percent with a water temperature of 13.1 °C at the end of the 5-month test. With the coldest water temperature of 8.6 °C on December 14, 2001, the salt passages measured 1.0 to 1.5 percent.

Salt passage measures membrane desalting performance, and significant increases in salt passage would indicate degradation, for example from chloramine disinfection or cleaning. The stability of the salt passage values do not indicate degradation.

A similar RO performance parameter that does not vary with temperature is the ratio of the salt and water transport coefficients. The parameter “B/A” (see appendix B, figure B-35) exhibited a temporary increase following cleaning. This increase could indicate membrane degradation. Nevertheless, the final values of B/A approximately equal the initial values and do not indicate degradation. Further tests with repeated cleanings are recommended to determine if cleaning adversely increases salt passage and B/A.

RO Solute Removal

Appendix C, tables C-4 through C-7 list the concentrations of solutes in the RO feed, product, and reject waters and the calculated RO removals of solutes for four dates: October 25, 2001, November 19, 2001, December 17, 2001, and January 15, 2002. Appendix B, figure B-26 also shows TDS levels estimated from daily conductivity measurements. Discussed below are TDS, hardness, TOC, and THMFP. See appendix C, tables C-4 through C-7 for other composition measurements.

Based on measured concentrations listed in appendix C, tables C-4 through C-7 and on expected RO removal of contaminants not analyzed, RO product concentrations meet all primary and secondary drinking water standards for inorganic contaminants.

TDS Removal

The RO product TDS measured 31 mg/L at the start of the test, 10 mg/L on December 14, 2001, and 19 mg/L at the end of the 5-month test (see appendix B, figure B-26). At constant

production, higher TDS levels of 50 to 60 mg/L would occur in the summer. These TDS levels readily meet the secondary drinking water standard of 500 mg/L for TDS and are less than the approximately 200-mg/L TDS levels in area groundwater.

With a feed TDS of 670 mg/L and a reject TDS of approximately 3,000 mg/L (see appendix B, figure B-27), the RO removal of TDS during the test ranged from 97 to 99 percent (note that RO removal of TDS equals 100 percent minus salt passage discussed above).

Appendix C, tables C-4 through C-7 list TDS levels measured by evaporation at 180 °C. Because of the low TDS levels in the RO product, however, the authors consider TDS estimated from conductivity to provide more accurate values for the RO product.

Hardness

With 250 to 280 mg/L hardness in the RO feed and 1,200 to 1,400 mg/L hardness in the RO reject, the RO product hardness measured less than the 22-mg/L laboratory detection limit on all four dates (see appendix C, tables C-4 through C-7). These levels correspond to greater than 95.4 to 96.0 percent removal of hardness (based on average feed and reject concentrations). The RO product hardness level places the RO product in the “soft” water classification (see table 8).

RO Removal of Total Organic Carbon

With 2.8 to 3.9 mg/L TOC in the RO feed and 9.3 to 13 mg/L in the RO reject, the RO product TOC measured less than the 0.5-mg/L laboratory detection limit on all four dates (see appendix C, tables C-4 through C-7). These levels correspond to greater than 90.3 to 93.2 percent removal of hardness (based on average feed and reject concentrations).

Disinfection Byproduct Measurements

With chloramine disinfection, the TTHM concentrations in the RO feed, RO product, and RO reject measured less than the 1.0- μ g/L laboratory detection limit on all four dates (see appendix C, tables C-4 through C-7).

To obtain estimates of RO removal of THM precursors and to determine if disinfection with free chlorine alone (without ammonia addition to form chloramines), THM formation potential (THMFP) measurements were conducted. THMFP measures the THM concentrations after 7 days of storage with a final free chlorine concentration of 3 to 5 mg/L.

With 85 to 138 $\mu\text{g}/\text{L}$ total THMFP in the RO feed and 148 to 208 $\mu\text{g}/\text{L}$ in the RO reject, the RO product total THMFP measured 0.9 to 3 $\mu\text{g}/\text{L}$ (see appendix C, tables C-4 through C-7). These levels correspond to 98.9 to 99.6 percent removal of hardness (based on average feed and reject concentrations).

The total THMFP levels of 0.9 to 3 $\mu\text{g}/\text{L}$ in the RO product are much less than the MCLs listed by the SDWA Stage 1 D/DBPR of 80 $\mu\text{g}/\text{L}$ TTHM. In addition, the THMFP 7-day free chlorine levels of 3 to 5 mg/L are much higher than the 0.5 to 1.0 mg/L required for disinfection of water distribution systems. For these two reasons, with RO product water-free chlorine alone (without ammonia addition to form chloramines) can be used as the disinfectant and meet the SDWA D/DPR MCL for TTHM.

Design and Cost Revisions

Based on these pilot tests, no major changes are recommended to the SSF design described in appendix E of the August 2000 report (Reclamation, 2000).

During pilot tests, the SSF was run to exhaustion and cleaned when the schmutzdecke became plugged. Draining the bulk of supernatant was accomplished quickly through a drain valve, but the final dewatering of the top few inches of sand media required several hours. To avoid this delay, the supernatant filter drain should be designed to drain all of the supernatant water.

The appraisal-level estimates of O&M labor costs (Reclamation, 2000, tables E-5 and E-6) are based on using a single operator for a 42-MGD facility. Many activities at a water treatment facility, however, require at least two persons for safe and efficient operations. For this reason, and to accommodate the increased SSF cleaning frequency from 6 to 17 times per year (see below), the revised estimate includes increased labor costs for a staff of three operators. This change increases the labor cost from \$41,600 per year to \$124,800 per year.

The August 2000 SARWMS report (Reclamation, 2000, table E-3, p. E-10) estimates a sand filter cleaning frequency of six times a year. Based on pilot results with 22-day filter runs (refer to appendix B, figure B-1, runs 7 and 8), the estimated sand filter cleaning frequency has been increased to 17 times per year. The higher SSF cleaning frequency is estimated to increase costs of sand replacement by a factor of 17/6 per year.

Appendix G incorporates the increased costs of labor and sand replacement which, in turn, increase the total O&M costs. For SSF treatment without desalting, the estimated total of the O&M costs for administration and chemistry laboratory, SSF, and finished water reservoir increases from \$ 0.24 million per year to \$ 0.41 million per year (see table 22 and appendix G, figure G-3). The increased O&M cost changes the estimated unit cost of SSF-treated water from \$0.13 per thousand gallons to a revised estimate of \$0.15 per thousand gallons.

Pilot Investigation of Slowsand and Reverse Osmosis Treatment

Table 22.—Treatment alternatives summary

		Treatment		Annual and unit costs	
		Configuration	Costs	\$/million/year	\$/1,000 gallons
Capital costs (\$ million)		CAP-CT	46.97	3.93	0.57
O&M costs (\$million/year)			1.67	1.67	
Capital costs (\$ million)		CAP-SSF	12.35	1.03	0.15
O&M costs (\$million/year)			0.41	0.41	
Capital costs (\$ million)		CAP-MF/UF	59.80	5.01	0.57
O&M costs (\$million/year)			0.54	0.54	
Constant-production plant with ASR for 23.95-MGD average, 26.76-MGD peak capacity, and 95-percent plant factor (30,000 AFY)					
		Treatment	Concentrate disposal	Annual and unit costs	
		Configuration	Costs	Total costs	\$/1,000 gal
Capital costs (\$ million)		a. Pipeline to Puerto Penasco (with no partners)	75.02	132.35	10.51
O&M costs (\$million/year)			0.20	3.53	3.53
Capital costs (\$ million)		b. Evaporation ponds	62.96	120.29	9.59
O&M costs (\$million/year)			0.07	3.40	3.40
Capital costs (\$ million)		c. Pipeline to Puerto Penasco with 17.6 MGD from Tucson	15.50	72.82	5.98
O&M costs (\$million/year)			0.04	3.36	3.36
Capital costs (\$ million)		d. CASI pipeline to Yuma with 37.6 MGD from Tucson and SROG	48.64	105.96	8.50
O&M costs (\$million/year)			0.13	3.45	3.45
Capital costs (\$ million)		e. CASI canal to Yuma with 37.6 MGD from Tucson and SROG	28.47	85.79	6.96
O&M costs (\$million/year)			0.23	3.56	3.56
Capital costs (\$ million)		f. CASI canal to Yuma with 272-MGD total flow	13.72	71.04	5.84
O&M costs (\$million/year)			0.01	3.33	3.33
Capital costs (\$ million)		a. Pipeline to Puerto Penasco (with no partners)	75.02	109.95	8.63
O&M costs (\$million/year)			0.20	2.72	2.72
Capital costs (\$ million)		b. Evaporation ponds	62.96	97.89	7.71
O&M costs (\$million/year)			0.07	2.59	2.59
Capital costs (\$ million)		c. Pipeline to Puerto Penasco with 17.6 MGD from Tucson	15.50	50.42	4.10
O&M costs (\$million/year)			0.04	2.55	2.55
Capital costs (\$ million)		d. CASI pipeline to Yuma with 37.6 MGD from Tucson and SROG	48.64	83.56	6.62
O&M costs (\$million/year)			0.13	2.64	2.64
Capital costs (\$ million)		e. CASI canal to Yuma with 37.6 MGD from Tucson and SROG	28.47	63.40	5.09
O&M costs (\$million/year)			0.23	2.75	2.75
Capital costs (\$ million)		f. CASI canal to Yuma with 272-MGD total flow	13.72	48.65	3.97
O&M costs (\$million/year)			0.01	2.52	2.52

Table 22.—Treatment alternatives summary (continued)
 Constant-production plant with ASR for 23.95 MGD average, 26.76 MGD peak capacity, and 95-percent plant factor (30,000 AFY) (continued)

	Treatment		Concentrate disposal		Annual and unit costs		
	Configuration	Costs	Alternative	Costs	Total costs	\$/million/year	\$/1,000 gal
Capital costs (\$ million)				75.02	134.76	10.71	1.55
O&M costs (\$million/year)			a. Pipeline to Puerto Penasco (with no partners)	0.20	2.88	2.88	
Capital costs (\$ million)			b. Evaporation ponds	62.96	122.70	9.79	1.43
O&M costs (\$million/year)				0.07	2.75	2.75	
Capital costs (\$ million)			c. Pipeline to Puerto Penasco with 17.6 MGD from Tucson	15.50	75.24	6.18	1.02
O&M costs (\$million/year)				0.04	2.72	2.72	
Capital costs (\$ million)		59.74	d. CASI pipeline to Yuma with 37.6 MGD from Tucson and SROG	48.64	108.38	8.70	1.32
O&M costs (\$million/year)		2.68		0.13	2.81	2.81	
Capital costs (\$ million)			e. CASI canal to Yuma	28.47	88.21	7.17	1.15
O&M costs (\$million/year)			with 37.6 MGD from Tucson and SROG	0.23	2.91	2.91	
Capital costs (\$ million)			f. CASI canal to Yuma with 272-MGD total flow	13.72	73.46	6.04	1.00
O&M costs (\$million/year)				0.01	2.69	2.69	

For treatment with desalting (SSF-RO), the estimated total of the O&M costs for administration and chemistry laboratory, SSF, RO, post-treatment, ASR, and finished water reservoir increases from \$2.37 million per year to \$2.51 million per year (refer to table 22 and appendix G, figure G-4). The increased O&M cost changes the estimated unit cost of [SSF-RO]-treated water, from a range of \$0.73 to \$1.28 per thousand gallons to a revised range of \$0.74 to \$1.30 per thousand gallons (refer to table 22).

Figure 11 summarizes the treatment costs in dollars per thousand gallons, including capital and O&M costs, for three treatments without desalting (SSF, CT, and MF/UF) and three treatments with desalting (SSF-RO, CT-RO, and MF/UF-RO). Treatment using SSF has the lowest estimated product water cost for treatment without desalting. Treatment by SSF, at \$0.15 per thousand gallons, remains much more economical than treatment by CT or MF/UF, at \$0.57 per thousand gallons.

The pilot study indicated no changes to the RO design or cost estimates. The only changes in the costs of treatment with desalting by SSF - RO are the increased SSF O&M costs for labor and sand replacement.

For treatment with desalting by SSF and RO, the SSF O&M cost estimate increases from \$131,000 to \$275,000 per year (see appendix G, table G-2). Appendix G, table G-4 includes this revised SSF O&M cost in the revised total treatment O&M cost estimate of \$2.51 million per year.

The revised cost estimates for SSF-RO are shown in table 22 and figure 11. With concentrate alternative d (CASI pipeline to Yuma conveying 37.6 MGD from Tucson and SROG), the revised estimate for SSF - RO is \$1.06 per thousand gallons, about \$0.30 less than for CT - RO or MF/UF - RO.

Conclusions

Based on 2001 - 02 test results from 7 months of operation with a 200-square foot slowsand filter and 5 months of operation with a RO pilot unit equipped with eighteen 2.5-inch-diameter by 40-inch-long spiral-wound polyamide low-pressure RO membranes with CAP water in Marana, Arizona, the study offers the following conclusions:

1. Can SSF be used to treat CAP water to primary drinking water standards?

The study pilot SSF operated with low operational and maintenance requirements and effectively treated CAP water.

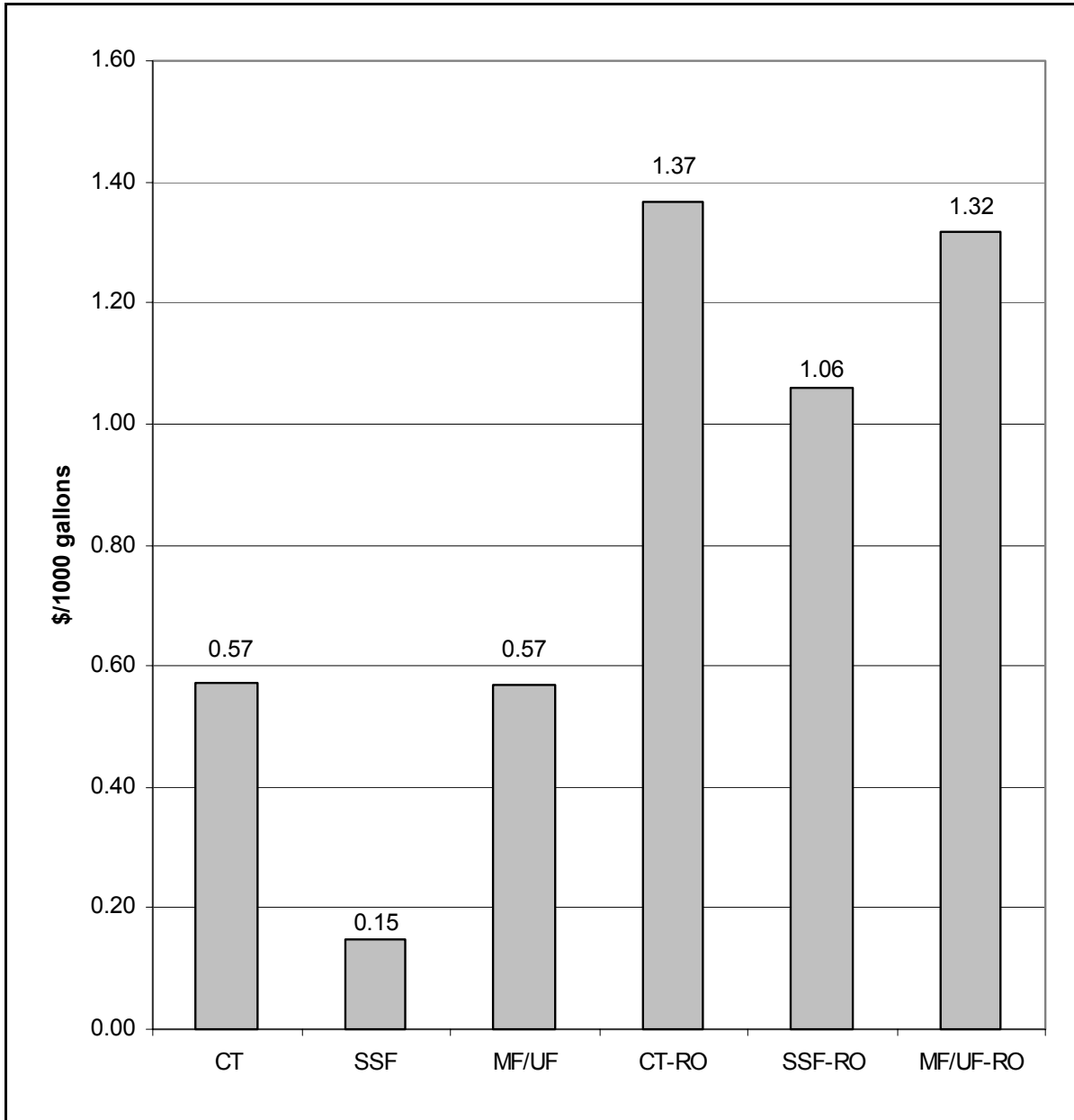


Figure 11.—Estimated CAP water treatment costs revised to include increased SSF labor and sand replacement cost estimates. For treatments with desalting, concentrate disposal costs are based on a Central Arizona Salinity Interceptor (CASI) pipeline to Yuma conveying 37.6 MGD from Tucson and the Arizona Municipal Water Users Association Sub-Regional Operating Group (SROG).

During this study, SSF treatment of CAP water met the SWTR for turbidity requiring less than 1 NTU for 95 percent of daily samples in a month.

Expected disinfection in a water treatment plant is post-SSF chlorination disinfection to meet the SWTR for giardia and virus removal and chloramination disinfection for the water distribution system. With lower free chlorine concentration than specified to achieve 4-log giardia removal, the 7-day SDSDBP levels from one analysis measured 0.015 mg/L for TTHM and 0.007 mg/L for HAA5. These levels are much less than the SDWA Stage 1 D/DBPR MCLs of 0.080 mg/L TTHM and 0.060 mg/L HAA5. With higher free chlorine concentrations, we expect the SDSDBP levels to be higher but still meet the TTHM and HAA5 MCLs.

Based on TOC and THMFP measurements, SSF does not greatly reduce DBP precursor concentrations. SSF removed 0.4 mg/L, or 12 percent of the TOC. SSF reduction of TTHMFP was not statistically significant at the 95-percent confidence level.

2. How effective is SSF as a pretreatment to RO?

SSF treatment of CAP water appears adequate for removing particulates that foul reverse osmosis polyamide membranes. SSF should also provide adequate pretreatment for cellulose acetate membranes, which were not tested but generally foul much less than polyamide membranes. Although the initial decline of membrane water transport coefficient “A” at the rate of 8 percent per 1,000 hours indicates particulate fouling, two simple high-pH cleanings effectively restored the “A” values. Five weeks after cleaning, and 5 months after startup, the average of the “A” values was only 4 percent less than observed at 1 week after startup.

3. What is the salinity and composition of CAP water treated by SSF and RO?

During the pilot test, the RO product TDS measured 10 to 31 mg/L, hardness measured less than 22 mg/L, and TOC measured less than 0.5 mg/L. Based on results listed in appendix C, water treated by SSF - RO meets all primary and secondary drinking water standards for inorganic contaminants.

4. For disinfection of RO product, can free chlorine without ammonia be used to meet the SDWA Stage 1 D/DBPR MCLs of 0.080 mg/L TTHM and 0.060 mg/L of HAA5?

Based on very low measured total THMFP levels of 0.001 to 0.003 mg/L in the RO product, free chlorine without ammonia be used to meet the MCL for TTHM and most probably also meet the MCL for HAA5.

5. What changes to SSF costs estimated in the August 2000 SARWMS are indicated by the pilot tests?

The estimated cost of SSF treatment is revised to \$0.15 per 1,000 gallons, increased from the \$0.13 per 1,000 gallons estimated in the August 2000 SARWMS report (Reclamation, 2000). The increase is based on cleaning the SSF 17 times per year (every 21 days) as required with the pilot SSF at the high filtration rate of 0.11 gal/min/ft² (0.27 m/h, 6.9 MGD/acre). The August 2000 SARWMS report assumed six cleanings per year.

Recommendations

Because SSF effectively treats CAP water at one-fourth the cost of either CT or MF/UF, subject to future tests demonstrating successful year-round operation, the authors recommend SSF for CAP water treatment without desalting.

For CAP water treatment with desalting, the authors recommend two alternatives. One alternative is MF/UF in combination with low-pressure RO. The second alternative, recommended subject to the successful completion of additional pilot tests, is SSF in combination with low-pressure RO.

Because SSF has significantly lower costs than CT, MF treatment, or UF treatment, and because the present pilot test has demonstrated the effectiveness of SSF filtration during 7 months of operation and SSF has little operational flexibility once constructed, the authors recommend following the advice of Leland and Logsdon to conduct pilot tests for at least 1 year (Logsdon, 1991, p. 194). Because SSF cleaning costs represent 60 percent of the estimated O&M costs, the authors also recommend evaluating SSF cleaning technologies. The recommended pilot tests and evaluation of cleaning technologies should address the following questions:

- (1) What are the effects of a range of seasonal (i.e., spring and summer) CAP water conditions on SSF filtrate quality, filter run lengths, and filter run production volumes?
 - (2) What are the effects of using locally available sand with a different size and uniformity coefficient than the sand used in the present pilot study (shipped from California) on SSF filtrate quality, filter run lengths, and filter run production volumes?
 - (3) What SSF filtration rates provide filter runs lasting at least 1 month over a range of CAP water conditions, and what are the associated filter run production volumes?
 - (4) How effective are mechanized SSF cleaning technologies, and which technologies appear most cost-effective?
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- (5) What are the effects of changes in water conditions and SSF loading rates on SSF effluent DO, HPC, pH, TOC, and SDSDBP levels and on effluent particulate levels as indicated by turbidity, SDI, and fouling rates of RO pilot equipment?
- (6) What range of combinations of filtration rates and cleaning frequency appear most practical when used with mechanized SSF cleaning technologies?

For RO treatment of CAP water, the authors recommend pilot tests to address the following questions:

- (7) To ensure that the apparently successful RO cleanings conducted during the pilot tests do not damage the RO membranes, what are the effects of multiple high-pH, high-temperature RO cleanings on RO salt passage?
 - (8) Can RO pilot equipment be operated at the plant design recovery of 85 percent without scaling? To provide more sensitive and rapid detection of changes in “A” due to scaling, include instrumentation to monitor the performance of the tail RO element.
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