



# Municipal wastewater reclamation by reverse osmosis—a 3-year case study

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Water is a scarce and valuable commodity in California because of the arid nature of the region and the uneven distribution of water with respect to both time and space. The northern third of the state receives the bulk of the rainfall. Moreover, the majority of the state's rainfall occurs during several winter months, often producing flood or near-flood conditions in winter. The late spring, summer, and fall are often devoid of any rainfall. The heterogeneous distribution of water resources has resulted in the development of the largest system of canals and reservoirs for water resources management in the world. In spite of this highly developed water resource management system, severe water shortages are predicted before the year 2000 unless changes are made in water-use policy or the state's agricultural economy.

To understand the potential need for wastewater reclamation in California, it is useful to review California water resources and demand. Asano *et al.*<sup>1</sup> have compiled some useful statistics. The total water demand in California in 1975 (an average rainfall year) was  $4.3 \times 10^{10} \text{ m}^3$  (35 million acre-ft), of which  $2.2 \times 10^9 \text{ m}^3$  (1.8 million acre-ft) was supplied by groundwater overdraft, which increased to  $7.6 \times 10^9 \text{ m}^3$  (6.2 million acre-ft) in the drought year of 1977, and is anticipated to reach this level during average rainfall years by the year 2000. A continued groundwater overdraft of this magnitude cannot be tolerated without drastic and unacceptable changes to the environment. Clearly, other sources of water and water conservation measures are needed.

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## Sufficient data were gathered to formulate a model that can determine an economically optimal design.

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One method of providing additional sources of water is through wastewater recycling. In 1975, approximately  $3.8 \times 10^9 \text{ m}^3$  (3.1 million acre-ft) of municipal wastewater was produced, of which 68% or  $2.6 \times 10^9 \text{ m}^3$  (2.1 million acre-ft) was discharged to saline

water.<sup>1,2</sup> Moreover, the quantity of wastewater discharged to saline water is anticipated to increase to  $4.6 \times 10^9 \text{ m}^3$  (3.7 million acre-ft) by the year 2000. It has been estimated that  $3.1 \times 10^9$  (2.5 million acre-ft) of this wastewater could be reclaimed and used for beneficial purposes.<sup>3,4</sup> Therefore, it is apparent that one excellent source of additional water during the predicted water shortage is wastewater recycling, which could reduce the projected year 2000 shortfall by approximately 40%.

Presently there are several full-scale wastewater recycling projects in California. Prominent among these are the surface spreading operations by the County Sanitation Districts of Los Angeles,<sup>1</sup> and the direct injection program at Water Factory 21 in Orange County.<sup>5</sup> During 1978, these projects and several smaller ones recycled a total of  $2.2 \times 10^8 \text{ m}^3$  (0.184 million acre-ft) of treated wastewater.

The majority of this reclaimed wastewater was produced by advanced, technologically sophisticated treatment plants, embodying the best available treatment technology. For example, Water Factory 21 uses high-lime coagulation, ammonia stripping, recarbonation, carbon adsorption, and reverse osmosis to treat activated sludge plant effluent prior to injection. The need for such advanced treatment arises from the potential public health hazards of partially treated wastewater.

There are other applications for recycled wastewaters where such advanced treatment may not be required. For example, industrial reuse for such applications as cooling tower make-up, irrigation of agricultural and non-agricultural lands (freeway medians and borders, parks, golf courses, etc.), and construction do not necessarily require such advanced treatment. In many cases the total dissolved solids (TDS) and bacterial quality (as measured by coliform counts) represent the only challenging treatment objectives. Therefore, some types of desalting and disinfection process may be the only requirement in recycling municipal secondary effluent. Moreover, simpler, lower-cost treatment plants will allow more widespread use of wastewater recycling, and will assist in meeting recycling goals.

This investigation is an example of the use of simplest technology to achieve recycling objectives. A 3-year case study of a small, tubular reverse osmosis (RO) pilot plant is presented, with associated plant experience and operating data. A systems analysis to determine the economically optimal water reclamation plant configuration is also provided.

## PILOT-PLANT DESCRIPTION

The pilot plant is located at the Las Gallinas Valley Sanitary District, north of San Rafael, in Marin County, Calif. The district operates a secondary treatment facility using 2-stage trickling filters and clarification. The total plant flow rate ranges from 0.065 m<sup>3</sup>/s to as high as 0.44 m<sup>3</sup>/s (1.5 to 10 mgd), depending on season and weather. The trickling filters are loaded at approximately  $1.17 \times 10^{-4}$  m<sup>3</sup>/m<sup>2</sup>·sec (11 mgd/acre) and 1.35 kg of 5-day biochemical oxygen demand (BOD<sub>5</sub> per cubic meter (84 lb BOD<sub>5</sub>/1 000 ft<sup>3</sup>) of media. The plant receives almost 100% municipal wastewaters.

The 0.63-l/s (10-gal/min) RO pilot plant is composed of 160 3.05-m (10-ft)-long tubes, with an internal membrane diameter of 2.23 cm (0.88 in.). Each membrane has an approximate area of 0.208 m<sup>2</sup> (2.24 sq ft). Cellulose acetate membranes are used exclusively and are made by the pilot-plant operators using a cellulose acetate formulation, and membrane curing temperatures of 88 to 90°C. The total membrane area of the unit is 33.3 m<sup>2</sup>. Figure 1 is a schematic diagram of a typical membrane; the membrane casting procedure and construction techniques are no different than those used in earlier studies at the University of California, Los Angeles<sup>6-8</sup> and are similar to the original membranes reported by Loeb and Sourirajan<sup>9</sup> in 1960. The membranes are cast at the Department of Water Resources Firebaugh facility.

The membranes are all arranged in a series, single-stage configuration. A triplex positive displacement pump is used to pump feed water and can be operated

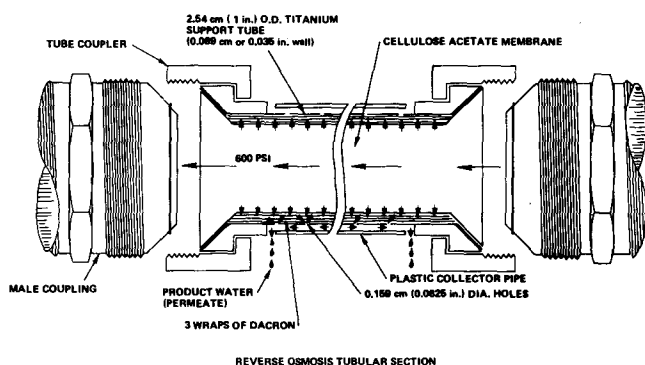


Figure 1—Schematic diagram of a typical membrane (after Goel and McCutchan<sup>8</sup>).

over a range of 14 to 55 atm with flow rates ranging from 0.13 to 0.75 l/s (2 to 12 gal/min). The work reported herein was all performed at 40.8 atm and the feedwater flow rate was normally 0.404 l/s (6.4 gal/min), providing an inlet Reynold's number of approximately 23 000 and an inlet velocity of approximately 1.05 m/s (3.4 ft/sec). The unit was also equipped with automatic flow-reversal valves and spongeball cleaning facilities.

The RO pilot plant was originally placed in operation in April 1976 treating trickling filter effluent, filtered by a 762-mm (30-in.) diameter, multimedia filter. This filter was installed and operated by the Marin Municipal Water District (MMWD) to provide a recycle water source. Subsequently, MMWD installed a second filter to increase their recycling effort. The RO unit operated in this manner until May 1979, when a smaller, 24-in. diameter, mixed-media filter was installed and dedicated to pretreatment of RO feedwater. No coagulating chemicals were used until after May 1979. The recycled water from MMWD filters was used for landscape irrigation and construction projects through the drought of 1977, and has continued until the present; MMWD should be consulted for further information regarding their work.

This initial period from April 1976 to June 1979 was devoted to the development of membrane cleaning techniques and endurance testing of the RO membranes and equipment. The original cleaning technique was restricted to spongeball cleaning without chemical cleaning agents. During cleaning, the unit was always depressurized and flushed with tap water or RO product water (later containing cleaning chemicals). Beginning in April 1977, a 2-hour enzyme detergent flush was initiated. In June 1977, the detergent flush was stopped and a citric acid flush was begun. Combinations of cleaning techniques were evaluated until March 1978, when a final cleaning procedure, consisting of 1-hour flushes with citric acid and detergent followed by spongeball cleaning, was developed. Table 1 is a summary of the final cleaning procedure.

In March 1978, chlorination of RO feedwater was begun, and in May 1978, pH control of feed water was started. Operation continued in this fashion until June 1979, when improved pretreatment facilities were placed in operation. From June through July 1979, feed water was pretreated using direct filtration with cationic organic polymer. In July 1979, a 1.7-m (5.5-ft)-diameter clarifier was installed and inorganic coagulants were used. The mixed-media filter was operated at 2.17 l/m<sup>2</sup>·s (3.2 gal/min/sq ft) filtration rate and backwashed at 10.2 to 13.6 l/m<sup>2</sup>·s (15 to 20 gal/min/sq ft) after a 2-minute surface wash. Backwashing was performed automatically on a timed cycle. Usually backwashes were performed every 12 hours. The filter was operated at the 2.17 l/m<sup>2</sup>·s (3.2 gal/min/sq ft) rate indepen-

Table 1—Final weekly cleaning procedure.

Operation	Procedure
Citric acid flush	Add 250 g citric acid to 190 l tap water or RO product water at ambient temperature. Solution is circulated through RO unit at approximately 0.315 l/s for 1 hour.
Enzyme detergent flush	Add 500 g commercially available enzyme detergent to 190 gal tap water at ambient temperature and circulate through the RO unit as before for 1 hour.
Spongeball cleaning	After chemical cleaning, 10 spongeballs (3.8 = cm) are introduced into the RO feed at approximately 1-minute intervals and allowed to pass through the unit at approximately 0.52 m/s.
Spongeball cleaning	After approximately 70 hours of operation, the unit is depressurized and spongeball cleaning is repeated.

dently of the RO feed rate. Excess water was discharged with the Las Gallinas Valley Sanitary Districts effluent. The commercially available media consisted of 1.0- to 1.2-mm size distributions of anthracite coal, a 0.42- to 0.55-mm size distribution of silica sand, and a 0.2- to 0.3-mm size distribution of garnet sand.

The clarifier was operated at 0.63 l/s (10 gal/min) also, giving an overflow rate of 24.8 m<sup>3</sup>/m<sup>2</sup>·d (610 gal/sq ft/day). Sludge was manually withdrawn on a periodic basis.

The entire pilot plant, with the exception of cleaning, chemical makeup, and data collection, operated on an unattended, automatic basis. Operational attention was restricted to a daily check and daily data collection 5 days per week. Cleanings were normally performed on the remaining 2 days (usually Monday and Thursday).

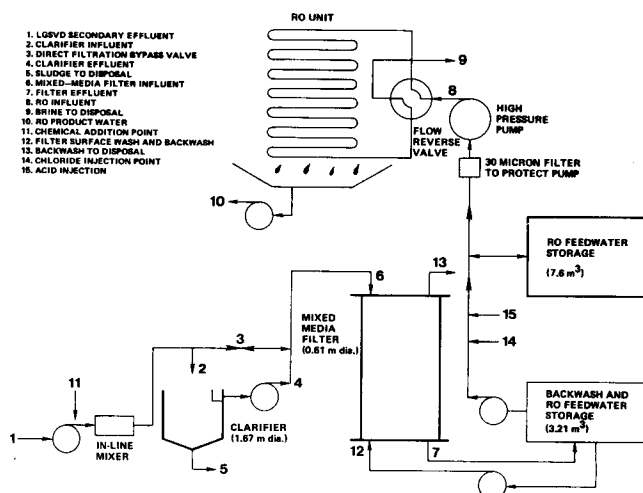


Figure 2—Schematic flow diagram of the reverse osmosis pilot plant with pretreatment facilities.

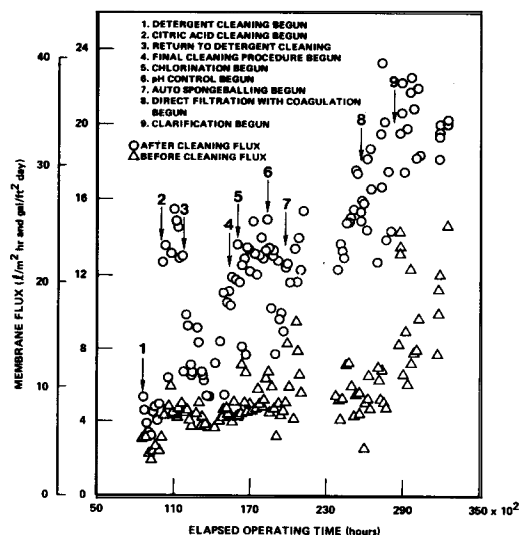


Figure 3—Membrane fluxes before and after cleaning versus operating time.

The total amount of operator time averaged about 2 hours per day.

Table 2 summarizes the various changes in operation of the unit.

Figure 2 is a schematic flow diagram for the entire RO plant with pretreatment facilities.

## EXPERIMENTAL RESULTS

**Flux decline and effects of cleaning.** The early results with the unit were disappointing in that very low recovery rates were obtained. The recovery averaged about 25% with fluxes in the range of 7.6 to 8.5 l/m<sup>2</sup>·h (4.5 to 5.0 gal/sq ft/day). The spongeball cleaning was effective at first in that fluxes were increased from approximately 8.5 l/m<sup>2</sup>·h before cleaning to approximately 15 to 17 l/m<sup>2</sup>·h after cleaning. After approximately 8 000 hours operation, the flux before cleaning decreased to approximately 6 l/m<sup>2</sup>·h, whereas the flux after cleaning could only be restored to about 7.1 to 7.6 l/m<sup>2</sup>·h. This rapid deterioration resulted from the accumulation of insoluble salts on the membrane surface that could not be scrubbed with the spongeballs.

The use of the enzyme detergent partially restored the membrane fluxes, but results were still disappointing. Starting in April 1977, the fluxes after detergent and spongeball cleaning gradually increased from 6.8 to 7.6 l/m<sup>2</sup>·h to a maximum of about 8.5 l/m<sup>2</sup>·h. On June 20, 1977, the first citric acid cleaning was performed, which restored membrane flux to 21.2 l/m<sup>2</sup>·h. This flux after cleaning was maintained until the end of September when flushing with the enzyme detergent was resumed. The flux after cleaning gradually declined and by December 1977, had declined to the previous levels of 6.8 to 7.6 l/m<sup>2</sup>·h. Beginning in March 1978, the final cleaning procedures shown in

Table 2—Chronological summary of pilot-plant operation.

Date	Hour	Event	Comments
4/27/76	0	Pilot plant started on trickling filter effluent after multimedia filtration.	Weekly spongeball cleaning without cleaning chemicals.
4/18/77	8 500	Cleaning procedure changed by the addition of 2-hour detergent flush.	Various concentrations of detergent as high as 2.1 g/l were used in the flush water.
6/20/77	10 000	Citric acid substituted for detergent.	Various concentrations from 0.04 to 0.53 g/l were used.
9/26/77	12 400	Returned to detergent.	Concentration ranged from approximately 1.05 to 1.32 g/l.
3/1/78	16 150	Final cleaning procedure developed, using 1-hour citric acid flush, followed by 1-hour detergent flush, followed by spongeball cleaning.	Final concentrations for flush are: citric acid, 0.66 g/l; an detergent, 1.32 g/l.
3/23/78	16 700	Began chlorination of multimedia filter effluent.	Chlorine residual ranged from 0.5 to 6.0 mg/l, averaging approximately 2.5 mg/l.
5/15/78	18 000	Influent pH control initiated by addition of sulfuric acid.	Set point at pH = 5.5.
8/1/78	19 800	Automatic spongeball cleaning started using reverse flow to initiate spongeball cleaning.	Cleaning frequency set to 6 hours.
6/1/79	27 100	Mixed-media filter with cationic polymer operation begun.	Cationic polymer dosage optimized by zeta potential measurements.
7/6/79	28 000	Clarifier installed. Pilot plant operated with coagulation, clarification and filtration until shut down.	Optimal concentrations of alum, ferric chloride and cationic polymers tested during this period.
1/7/80	32 400	Unit shut down.	

Table 1 were consistently used and flux after cleaning again stabilized at about 21.2 l/m<sup>2</sup>·h. The results of the improvements in cleaning technique can be seen in Figure 3, which shows the before and after cleaning fluxes for the entire period of the investigation. The increases in flux from improvements in cleaning are obvious. The increase in fluxes after January 1978 are attributed to improved pretreatment, rather than improved membrane cleaning.

**Flux decline and effects of pretreatment.** Improvements made in recovery and flux maintenance after January 1978 are largely because of improvements in RO feedwater quality. Chlorination of RO feed water was begun in March 1978, and feedwater pH control (pH controlled to approximately 5.5) was begun in May 1978, resulting in small increases in the before-cleaning fluxes. Before-cleaning fluxes increased from approximately 6.8 to 10.2 l/m<sup>2</sup>·h. The installation of the automatic spongeball devices in August 1978 coincide with increases in before-cleaning fluxes to as high as

13.6 l/m<sup>2</sup>·h. Unfortunately, the high before-cleaning fluxes fell back to the 8.5 to 11.8 l/m<sup>2</sup>·h range during the period from October 1978, to May 1979. At present, there is no explanation of this decreasing trend.

The use of chemical coagulation and clarification had very large effects on both before- and after-cleaning fluxes. Direct filtration with a cationic polymer, begun on May 31, 1979, coincided with increasing trends in both before- and after-cleaning fluxes. Both fluxes increased to maximum values of about 23.7 to 42.4 l/m<sup>2</sup>·h, respectively, during the final phases of the study when the inorganic coagulants were used.

During the last months of the study, a series of 24-hour flux-decline tests were made using various concentrations of ferric chloride, alum, and organic coagulants. Flux-decline curves for representative 24-hour tests for each coagulant, and uncoagulated, filtered trickling filter effluent are shown in Figure 4. The flux-decline coefficients (slope of a log-log plot of flux and time) are shown in Table 3 and are compared with

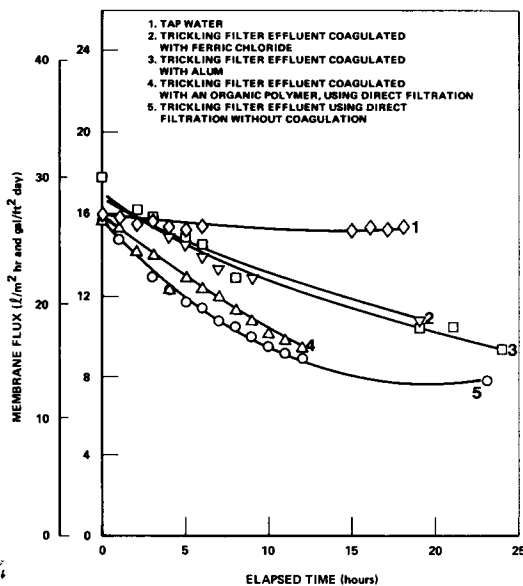


Figure 4—Flux decline versus time for various levels of pre-treatment and for tap water.

decline coefficients calculated by Thomas *et al.*<sup>10</sup> The flux decline coefficients show that the ferric chloride coagulant produced feed water with the least tendency to foul the membranes. The organic coagulant using direct filtration and alum with clarification were about equal. The uncoagulated water, as expected, produced the highest fouling rate.

**RO product water quality.** RO product water quality was routinely measured for TDS and turbidity, and measured quarterly for a broad spectrum of contaminants, including trace metals, boron, total organic carbon (TOC) and nutrients. Table 4 shows two such analyses of RO feed water and product-water quality. The levels of contaminants shown in the feed water on September 17 are typical of the feed water produced by direct filtration with an organic coagulant. The best-quality feed water was not analyzed for all the contaminants shown in Table 4, only for a small subset of contaminants, including TOC, total suspended solids, and turbidity, which averaged approximately 15 mg/l, 4 mg/l, and 3.0 NTU, respectively.

The product-water quality shown in Table 4 is excellent for most recycle water applications, and in fact is much higher than needed for many applications. The only contaminants which might be of concern for non-potable recycle uses are boron and sodium. The boron might be of concern when irrigating boron-sensitive plants. The sodium concentration is not prohibitively high, except that the balance between it and the divalent cations, such as calcium or magnesium, is poor. If the RO product-water were used as the only source of irrigation water for sensitive crops, it might be advantageous to adjust the sodium/calcium ratio by adding calcium to the product water.

Table 3—Flux-decline coefficients for various types of feed waters.

Flux-decline coefficient	Feedwater type	Reference
0.243	Trickling effluent with dual-media filtration.	This study
0.202	Trickling filter effluent with alum coagulation, clarification, and mixed-media filtration.	This study
0.204	Trickling filter effluent with organic polymer coagulation and direct mixed-media filtration.	This study
0.146	Trickling filter effluent with ferric chloride coagulation, clarification, and mixed media filtration	This study
0.013 6	Tap water (TDS = 100)	This study
0.9	Raw wastewater	Calculated by Thomas <i>et al.</i> <sup>10</sup> from the data of Feverstein and Bureztynsky. <sup>11</sup>
0.56	Primary effluent	Calculated by Thomas <i>et al.</i> <sup>10</sup> as previously.
0.35	Secondary effluent	Calculated by Thomas <i>et al.</i> <sup>10</sup> as previously.
0.14	Carbon-treated secondary effluent	Calculated by Thomas <i>et al.</i> <sup>10</sup> as previously.

**Membrane life.** It has been reported by Kimura and Nakao<sup>12</sup> that frequent spongeball cleaning of membranes may tend to reduce membrane life. In the present study, there is no way to determine the decrease in life expectancy from cleaning; however, it is interesting to note membrane-replacement statistics. From the pilot-plant start-up until November 1979, 363 membranes were replaced, not including membranes that failed within 30 days of insertion. Membranes that failed in less than 30 days were judged to be defective because of the casting or insertion technique. The average membrane life was 420 days, with a standard deviation of 262 days. Membranes were replaced when product TDS was above 500 mg/l or when after-cleaning flux declined to less than 8.5 l/m<sup>2</sup>·h-d. The majority of membranes were replaced because of high product-TDS concentration.

Table 4—RO feed and product water quality on March 19, and September 17, 1979.<sup>a</sup>

Contaminant	Feed water		Product water	
	3/19/79	9/17/79	3/19/79	9/17/79
Hardness <sup>b</sup>	216	241	13	19
Calcium	36	38	2	3
Magnesium	30	35	2	3
Sodium	136	218	28	57
Sulfate	77	251	1.5	8.5
Chlorides	207	351	44	99
Boron	0.55	0.55	0.4	0.45
TDS	671	1090	98	203
TOC	26.5	22.5	1.5	1.2
Total nitrogen	26	39	4.1	7.6
Total phosphorous	9.3	12	0.21	0.83
Iron	0.085	0.36	0.0	0.08
Copper	0.01	0.15	0.0	0.01

<sup>a</sup> All units are in mg/l. Values represent averages of measured water quality before and after chemical cleaning.

<sup>b</sup> As CaCO<sub>3</sub>.

Note—In addition, the following constituents were measured and less than 0.01 mg/l were found in both product and feedwater: arsenic, cadmium, chromium, lead, mercury, and nickel.

It seems that the major source of membrane failure is from the corrosion of end fittings and not from hydrolysis or other types of membrane failure. The end fittings of the membranes were brass, and apparently, an electrochemical corrosion reaction gradually reduced the bevel at the coupling end that eventually allowed the high-pressure flow to reach the dacron backing material, producing very rapid failure. An alternate type of PVC-lined coupling manufactured to alleviate the corrosion problems, and after installation, there was essentially no corrosion and reduced membrane failure.

## DISCUSSION AND SYSTEMS ANALYSIS

The cleaning technique obviously is very important to maintaining membrane flux. Spongeball cleaning alone seems effective only for a limited period of time, after which the build-up of insoluble material on the membrane surfaces prohibitively reduces membrane flux. Enzyme-detergent cleanings in conjunction with spongeball cleaning are only slightly more effective than spongeball cleaning alone. The most significant cleaning agent is citric acid, which appears to be several times more effective than either spongeball cleaning or enzyme-detergent cleaning. Membrane fluxes using the final cleaning procedures have been restored to approximately the original flux levels. It appears that the cleaning procedure does not improve the flux decline characteristics of the membranes. The membrane flux observed in this study declines to approximately the same value independent of the cleaning technique.

The value of the spongeball cleaning, especially au-

tomatic spongeball cleaning, seems to be in retarding the initial rapid rate of flux decline. The value of spongeball cleaning in reducing gradual flux decline appears to be very limited.

The use of chlorination and pH control appears to have a modest beneficial effect on membrane-flux decline. This result is similar to the results of Winfield<sup>13,14</sup> with hollow-filter membranes, Hanagi and Mori<sup>15</sup> with tube membranes, and Cruver and Nusbaum<sup>16</sup> and Nusbaum, Cruver, and Sleigh<sup>17</sup> with spiral-wound membranes. The improved flux appears to be related to inhibition of bacterial growth on the membranes and reduced pH, where many of the scaling salts are more soluble. The fate of the chlorination by-products in the pilot plant and the effect of chlorine on the membranes have been investigated by Cooper *et al.*<sup>18,19</sup>

Pretreatment with coagulation, clarification, and filtration made the most dramatic effect on flux maintenance. Before cleaning, fluxes increased from 13.6 to as high as 20 to 24 l/m<sup>2</sup>·h. This flux increase is attributed to removal of the submicron colloidal particles, as evidenced by relatively ineffective results of filtration without coagulation. The amount of potentially fouling colloidal and suspended material carried to the RO unit, as measured by recovering the material during cleaning, declined dramatically with enhanced pretreatment.

The ferric chloride coagulant provided the highest quality effluent, as indicated by flux-decline rates. Alum and organic polymers were less effective. Both alum and ferric chloride efficiently removed suspended and colloidal material, as indicated by decreases in the total suspended solids (TSS) and turbidity of untreated and treated water from 20 to 30 mg/l and 10 to 20 NTU, to 3 to 5 mg/l and 1 to 4 NTU, respectively.

The superiority of ferric chloride over alum can, in part, be attributed to the pH used in clarification/filtration and in the RO unit. The pH during clarification/filtration is approximately 7.2, which is reduced after filtration to approximately 5.5 with the addition of sulfuric acid. The pH for minimum solubility of the ferric hydroxide is approximately 8.3, and solubility increases with decreasing pH. Therefore, one would expect any ferrous ions remaining in the filter effluent to be more soluble, tending to precipitate less on the RO membranes. Aluminium hydroxide has just the opposite solubility chemistry. Coagulation occurs at a pH where alum is more soluble, and ironically, the RO unit is operated near the pH for minimum aluminum hydroxide solubility. Undoubtedly, the chemistry is more complex than the overview presented here, especially since the ionic strength and coagulant concentration are increased in the RO unit. However, the large accumulation of aluminum hydroxide flocs found in the cleaning flush when using alum coagulation supports this theory.

Using the data collected in this study and current

cost data compiled by the EPA,<sup>20</sup> an economic simulation model has been developed. The model is based on using the RO product water in conjunction with a specific quantity of RO feed water to provide water for recycle with specific water quality. The calculation procedure is to determine the minimum quantity of RO product water for blending with feed water and meet specified water quality standards, such as TDS, total organic carbon (TOC), and turbidity. Pretreatment level and cleaning frequency are considered variables, whereas RO operating pressure, membrane characteristics, and velocity are considered constant.

Table 5 shows a hypothetical analysis for a plant producing 0.044 m<sup>3</sup>/s (1 mgd) of reclaimed water. The water quality required for recycle has a TDS of less than 750 mg/l, turbidity of less than 1 NTU, TSS less than 5 mg/l, and TOC less than 5 mg/l. Using the values obtained in the experimental study (RO product water: TDS—250 mg/l, Turbidity—0.1 NTU, TOC—1.0 mg/l, and TSS—0 mg/l, with feedwater quality; TDS—1 200 mg/l, Turbidity—3.0 NTU, TOC—18.0 mg/l, and TSS—4.0 mg/l), a blending ratio of 0.033 m<sup>3</sup>/s (0.76 mgd) RO product flow to 0.01 m<sup>3</sup>/s (0.24 mgd) of feed-blending flow is required. The size and cost of each pretreatment system, with associated flux-decline coefficients, are calculated along with the optimal cleaning frequency. The optimal cleaning frequency affects RO plant size because cleaning requires the RO unit to be down for 2 hours.

Table 5—Optimal 1 mgd reclamation plant design.

<b>Filter</b>		
Influent flow	0.071 m <sup>3</sup> /s	(1.63 mgd)
Loading rate	3.4 l/m <sup>2</sup> ·s	(5.00 gpm/sq ft)
Filter area	20.6 m <sup>2</sup>	(222.00 sq ft)
Diameter	5.2 m	(17.00 ft)
Backwash velocity	0.010 m/s	(2.00 ft/min)
<b>Clarifier</b>		
Influent flow	0.071 m <sup>3</sup> /s	(1.63 mgd)
Loading rate	40.7 m <sup>3</sup> /m <sup>2</sup> ·d	(1 000 gal/sq ft/day)
Clarifier area	148. m <sup>2</sup>	(1 602 sq ft)
Diameter	13.8 m	(45 ft)
Coagulant	Ferric chloride	
<b>Reverse osmosis</b>		
Influent flow	0.059 m <sup>3</sup> /s	(1.37 mgd)
Product flow	0.033 m <sup>3</sup> /s	(0.76 mgd)
Percent recovery	56%	
Flux decline index (B)	0.15	
Average flux	23 l/m <sup>2</sup> ·s	(13.58 gal/day/sq ft)
Number of membranes	24 484	
Total area	5 231 m <sup>2</sup>	(56 314 sq ft)
Cleaning time	2.0 hr	
Optimal cleaning interval	16.0 hr	
Sulfuric acid injected	8.19 ppm	
Chlorine injected	2.00 ppm	

The optimal reclamation plant size and cost breakdown are also shown in Table 5. In determining the economics in the table, the capital cost data from the EPA survey<sup>20</sup> was used and operating costs for labor and electricity were estimated to be \$12/h and \$0.05/KWh. The interest rate was assumed to be 8% and project life was assumed to be 20 years.

The influent to the RO unit contains 1 200 mg/l TDS, 18 mg/l TOC, and 4 mg/l TSS. The RO effluent contains 250 mg/l TDS, 1 mg/l TOC, and 0 mg/l TSS. The blended water quality requirements are 750 mg/l TDS, 5 mg/l TOC and TSS. The blended water also has a turbidity requirement, but it is not active for this case.

For this hypothetical case, ferric chloride with clarification/filtration was found to be optimal and the optimal cleaning frequency was found to be every 16 hours. The total system cost was found to be \$0.41/m<sup>3</sup> (\$1.55/1 000 gal) of reclaimed water produced. This cost is relatively insensitive to cleaning frequency over the range of once per 12 hours to once per 26 hours.

In all hypothetical cases considered thus far, the most advanced pretreatment system of those evaluated experimentally is always optimal. In this case, the cost varies from a low of \$0.41/m<sup>3</sup> (\$1.55/1 000 gal) to a high of \$0.59/m<sup>3</sup> (\$2.23/1 000 gal). The results of this program are considered to be tentative.

## CONCLUSIONS

An experimental and economic analysis of a (10-gal/min) RO pilot-plant has been presented. From the experimental study, the following conclusions are made:

- Flushes of citric acid, followed by enzyme detergent and spongeball cleaning are effective at maintaining essentially initial membrane-flux levels. The citric acid is the major cleaning agent. Enzyme detergent and spongeball cleaning without citric acid are relatively ineffective.

- The automatic spongeball cleaning technique appears to have promise for maintaining membrane flux between chemical cleanings.

- The major factor in determining membrane life for the type of membrane construction shown in Figure 1 is corrosion, resulting in average membrane life of 420 days (10 000 hours).

- For the feed water investigated in this study, ferric chloride coagulation with clarification and filtration provided the highest-quality feed water with the least fouling tendency.

- For the optimal conditions investigated, reclaimed water containing approximately 500 mg/l for TDS can be produced for approximately \$0.42/m<sup>3</sup> (\$1.60/1 000 gal). The costs estimated are considered tentative.

- For all the conditions tested and analyzed, the

Table 6—Energy and labor analysis.

Treatment steps	Energy					
	Building		Process		Labor	
	(kWh/yr)	cost (\$/yr)	(kWh/yr)	cost (\$/yr)	(hr/yr)	(\$/yr)
Filter unit	41 176	2 059	55 220	2 761	1 646	19 756
Surface wash	0	0	3 484	174	68	813
Media	0	0	0	0	0	0
Backwash	0	0	5 336	267	196	2 384
Clarifier unit	0	0	3 760	188	170	2 036
Coagulant	2 114	106	4 900	245	306	3 674
Reverse osmosis	27 387	1 369	3 253 701	162 685	1 952	23 424
Sulfuric acid	722	36	1 630	82	76	914
Chlorine	2 988	149	388	19	343	4 119
Cleaning	0	0	2 267	113	185	2 226

greatest level of pretreatment provided the most economical operation.

• The optimal energy requirements are approximately 2.6 KWh/m<sup>3</sup> (10 KWh/1 000 gal).

An important aspect of this investigation is the relatively uncomplicated levels of pretreatment that were used. High lime coagulation and carbon adsorption were not investigated because of the complexity of operating a lime recovery/disposal system and carbon regeneration system. It is hoped that continued development will permit less-expensive operation without using the higher-technology pretreatment processes.

It is anticipated that the existing pilot-plant will be operated for at least 2 more years. During this time, several major modifications are planned. Installation of an activated sludge pilot-plant is planned for the fall of 1980, and investigations of velocity effects (which have been largely ignored in this project) are planned. Additionally, several novel flux-control techniques, such as those proposed by Belfort and Marx,<sup>21</sup> are being considered.

Table 7—Unit costs.

Treatment units/steps	Total capital (\$/yr)	Total capital	O&M (\$/yr)	Water costs (\$/m <sup>3</sup> )
Vessels	295 873	30 135	26 436	0.045
Surface wash	32 986	3 360	1 199	0.003
Media	12 144	1 237	0	0.003
Backwash	62 114	6 326	3 408	0.008
Clarifier	110 138	11 218	2 645	0.018
Coagulants	25 312	2 578	4 219	0.005
RO Unit	1 011 413	103 015	3 15 734	0.325
H <sub>2</sub> SO <sub>4</sub>	8 841	900	1 181	0.003
Chlorination	5 904	601	4 668	0.005
Cleaning	32 496	3 310	6 038	0.008
Coagulation/clarification/filtration		\$0.07/m <sup>3</sup>	(\$0.27/1 000 gal)	
Reverse osmosis system		\$0.34/m <sup>3</sup>	(\$1.28/1 000 gal)	
Total system		\$0.41/m <sup>3</sup>	(\$1.55/1 000 gal)	

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