IMPROVEMENT OF REVERSE OSMOSIS THROUGH PRETREATMENT--PHASE II

by

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ABSTRACT

This report describes the final two years of a six year pilot scale investigation to desalt brackish wastewaters for reclamation and recycle using reverse osmosis. The work reported herein is only part of a large continuing development program sponsored by the California Department of Water Resources.

The work described herein was originally begun in April of 1976 at the Las Gallinas Valley Sanitary District, north of San Rafael, in Marin County, California. The site was originally selected because of the interest of two local agencies, the Marin Municipal Water District, and the Las Gallinas Valley Sanitary District. It was also selected in anticipation of the need for additional water supplies, which was latter demonstrated in the drought of 1976-77, when Marin County was one of the most severely affected areas.

This report describes the second phase of research, covering the period from January, 1980 to shut down in June, 1982. The prior report, <u>Improvement of Reverse Osmosis through Pretreatment</u>, (UCLA-Eng-8066) describes operation from the beginning in 1976 to 1980.

Originally a one inch tube-style reverse osmosis unit was assembled at Las Gallinas. The unit was originally operated using trickling filter effluent, and provided satisfactory effluent quality, but fouling rates were excessive, causing very poor recovery rates, and poor economics. To improve recovery rates, coagulation and filtration were added which increased average recovery from 25% with trickling filter effluent without additional pretreatment, to over 60% with ferric chloride coagulation and filtration. Operating costs declined from over $2.00 to $1.57 per 1000 gallons (in 1979 dollars).

At the conclusion of the first phase it was determined from analysis of the reverse osmosis fouling material that the major flux reducing substances were still organic in origin, and that further reduction of biological materials in the feed water was desirable. Additional biological materials in the feed water could only be reduced by removing soluble substances, since the total suspended solids in the feed water were less than 2 to 4 mg/l.

In order to remove additional soluble material, it was decided to use an activated sludge process in lieu of the trickling filter, and a search was made to find a suitable activated sludge plant. After some searching it was concluded that a suitable location with an activated sludge plant was not locally available, and that acquiring an activated sludge pilot plant at the Las Gallinas site would be more cost effective than moving the reverse osmosis facility.

A 15 GPM pilot activated sludge plant was designed from a commercially available package plant produced by the Clow corporation, and placed in service at the end of 1981. This unit was operated until shut down in June, 1982. The unit was installed in such a way that it could be operated in lieu of the trickling filter in order that all the previously installed pretreatment facilities could be reused.

The activated sludge plant provided additional organic material removal which reduced fouling and operating cost. The cost for the best activated sludge pretreatment system (activated sludge followed by filtration without chemical addition) was $1.11 per 1000 gallons which can be compared to $1.57 per 1000 gallons for lowest cost from the previous phase, using trickling filter effluent followed by ferric chloride coagulation, sedimentation, and filtration, in 1979 dollars. In first quarter 1983 dollars the cost comparison was $1.71 to $2.42 per 1000 gallons in favor of the activated sludge treatment system. In all cases the lowest cost operation was obtained with the highest level of pretreatment. Aluminum sulfate was always the poorest coagulant in reducing fouling properties of the feed water, but not always the poorest coagulant in reducing feed water turbidity.

1. INTRODUCTION

In an effort to develop future water resources for the state of California, the California Department of Water Resources (DWR) and others have funded a series of projects to develop technology to reclaim water from wastewater discharges. The development of additional water resources from wastewaters is one method of meeting the future water needs while reducing wastewater discharge problems. Previous projects have been described by Antoniuk and McCutchan (1973) and Speight and McCutchan (1979) for irrigation drainage wastewaters, by Argo and Moutes (1979), Wojcik, Lopez, and McCutchan (1980), and Stenstrom et. al. (1982a, 1982b) for domestic wastewaters, and by Johnson and Loeb (1969), Johnson, McCutchan, and Bennion (1969) for saline groundwaters. Other work has also been performed, and the review by Davis, et. al. (1980) should be consulted for additional information.

This report describes the Phase II results for the research facility located at the Las Gallinas Valley Sanitary District, north of San Rafael, in Marin County, California. The objective of the Phase II study was to further investigate pretreatment techniques and their effect on system preformance and cost, by adding an activated sludge plant. In the Phase I the economics and system design of a pilot scale tubular reverse osmosis plant treating coagulated and filtered trickling filter effluent were investigated. The Phase I work, through extensive investigation of coagulation/filtration techniques, including coagulation by organic polymers, ferric chloride, alum (aluminum sulfate), showed that pretreatment significantly reduced total costs. It was concluded from Phase I that total operating cost could be reduced from over \$2.00 to \$1.57 per 1000 gallons by employing optimum coagulation-filtration pretreatment, as compared to trickling filter effluent without additional treatment. Additionally the fouling materials removed from the RO membranes appeared to be organic materials, indicating that additional improvements in biological pretreatment, such as those provided by an activated sludge plant, would be beneficial.

This report describes the results of the second phase of research, using improved pretreatment provided by a pilot scale activated sludge plant, including revised system economics, followed by various filtration/coagulation alternatives. In writing this report no attempt was made to discuss results from the first phase, unless they were essential to interpret the results from the second phase.

2. EXPERIMENTAL APPARATUS

The reverse osmosis apparatus used in this study was the same as that used in the first phase of work at Las Gallinas and very similar to the units used in earlier investigations conducted by UCLA researchers (Johnson and Loeb, 1966; Johnson, et. al. 1969; Speight and McCutchan, 1979). The unit is very similar to the original design by Loeb and Sourirajan (1960, 1962). Table 2.1 lists the specifications for the unit.

MEMBRANE CONFIGURATIONS

There are three common membrane configurations used today. Each has advantages and disadvantages. The tubular membranes used in this study have the advantage of high flux rates, ease of cleaning, and simplicity. Unfortunately they have very low packing density. Spiral wound membranes have much higher packing density, but are more complicated to manufacture and often have lower fluxes. Hollow fine fiber membranes have the highest packing density, but are restricted in application to high quality feed water since they cannot be easily cleaned. Suspended solids are particularly bothersome and must be removed from feed water. Table 2.2 , taken from the the *Desalting Handbook for Planners* (Office of Saline Water, 1972), summarizes the advantages of each membrane type.

The tube style membranes have been used throughout the UCLA research projects in part because they allow for membrane development and testing without requiring extensive equipment and facilities. All the membranes used in both phases of this study were cast by DWR personnel at their Firebaugh, California facility. The tube style membranes are particularly useful in

Table 2.1: Reverse Osmosis Unit Specifications

Parameter (1)	Value (2)
Membrane Configuration	Tubular
Internal Diameter	0.88 inches
Membrane Material	Cellulose Acetate
Annealing Temperature	88 to 90 ^{o}C
Number of Membranes	160
Operating Pressure	600 PSIG
Feed Rate	6.4 GPM (Phase I)
Feed Rate	3.7 GPM (Phase II)

Table 2.2: Comparison of Reverses Osmosis Membrane Configurations*

Characteristic	Spiral Wound	Tubular	Hollow Fine Fiber
Membrane Surface Area per Volume (ft^2/ft^3)	100-300	40-100	5,000-10,000
Product Flux (gal/ft-day)	8-25	8-25	0.1-2
Typical Module Factors			
Brine Velocity (ft/sec)	0.7	1.5	0.04
Brine Channel Diameter (inch)	0.005	0.5	0.004
Method of Replacement	As a membrane module assembly-on site	As a tube, on site	As an entire module-on site, module returned to the factory
Membrane Replacement Labor	Medium	High	Medium, requires equipment
High Pressure Limits	Membrane Compaction	Membrane Compaction	Fiber Collapse
Pressure Drop, product water size	Medium	Low	High
Pressure Drop, feed to brine exit	Medium	High	Low
Concentration Polarization Problems	Medium	High	Low
Cleaning Methods			
Mechanical	No	Yes	No
Chemical	Yes-pH & solvent limited	Yes -pH & solvent limited	Yes-less restricted
Permissible Feed Water pH	5.5-7.5	5.5-7.5	2-10
Permissible Temperature (^{o}C)	<38	<38	<38

* After the Office of Saline Water and Bureau of Reclamation, 1972.

reclamation studies since they can be used with the largest range of feed water qualities. Figure 2.1 shows a cross section of the membrane configuration used at Las Gallinas. Each membrane is 0.88 inches in diameter and ten feet long, providing a total surface area of 2.24 ft^2. The entire RO unit contained 160 membranes, for a total area of 358 ft^2.

PILOT PLANT DESCRIPTION

The pilot plant was located at the Las Gallinas Valley Sanitary District north of San Rafael, California. The District operates a secondary treatment plant composed of primary clarification, two stage trickling filters, and secondary clarification. The flow rate to the plant ranges from the average value of 1.5 MGD (0.065 m^3/sec) to upwards of 8 MGD (0.35 m^3/sec) in wet weather. The trickling filters are loaded at a rate of 11 MGD/acre (1,17 x $10^{-4} m^3/m^2$-sec) and 84 pounds of five-day biochemical oxygen demand (BOD_5) per thousand cubic feet of filter media (1.35 kg BOD_5/m^3). This loading is considered to be a high loading rate according to current design standards, and at this loading rate the filters are expected to produce effluent BOD_5 ranging from 12 to 25 mg/l, and should not nitrify. (Reynolds, 1982, Metcalf and Eddy, 1979). This effluent BOD_5 concentration compares to 5 to 15 mg/l to be expected from a well designed and operated activated sludge plant(Metcalf and Eddy, 1979).

The Las Gallinas plant showed fluctuations in treatment efficiency depending on season. In the winter the effluent was visibly poorer than summer effluent, with turbidities exceeding 20 NTU on occasions. The decrease in effluent quality in winter can be primarily attributed to the increase in

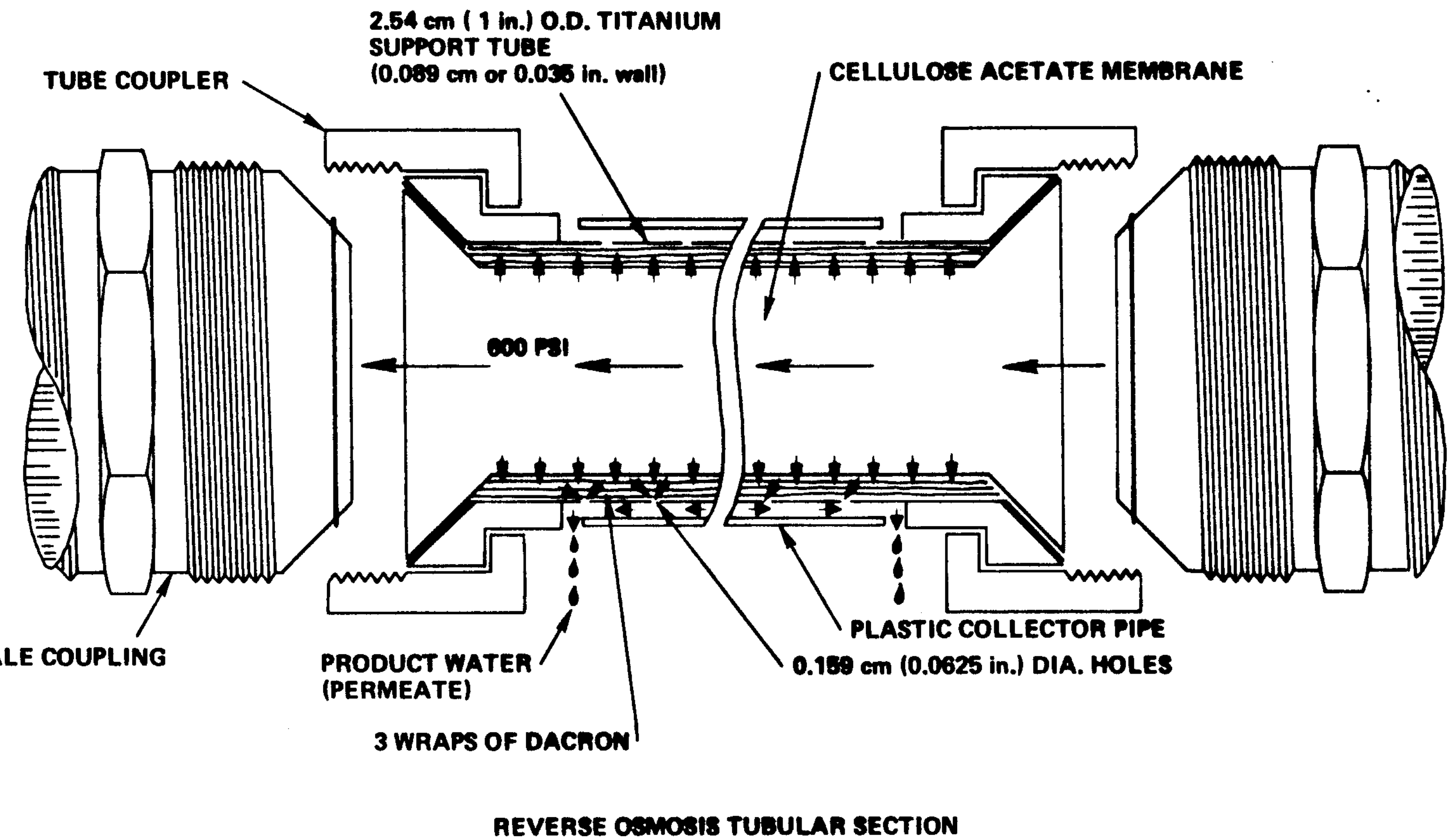

Figure 2.1: Membrane Cross Section

flow, although the filters would also perform more poorly at cooler wastewater temperatures.

The pilot plant included the RO plant, mixed-media filter, clarifier, pumps, pH control system, and chlorination facilities. The facility, excluding the activated sludge plant, is shown in Figure 2.2

The activated sludge plant was designed from a "package plant" available from the Clow corporation. The plant is designed to be a self contained unit which can be trucked to a site, unloaded, hooked up to utilities, and placed in service. Often package plants have reduced efficiency compared to full scale plants, and this results because of compromises in plant design to allow unattend, remote operation. For example the Clow plant, in the configuration used at Las Gallinas, did not have a mechanized skimmer, which occasionally allowed scum into the final effluent. Package plants often serve small subdivisions prior to the construction of sewers, or remote locations such as National Parks service facilities.

The plant used at Las Gallinas was constructed to provide dispersed flow operation ("plug flow" in the parlance of treatment plant operators). Provisions were made for tapering the aeration rate and for step feed operation (Torpey, 1952). The unit was operated as a conventional process through the entire operating period.

The secondary clarifier was constructed as a conical section welded to the rectangular aeration tank. Return sludge was pumped by an air lift pump. No rake was provided and skimming was provided by a open pipe skimmer located at the clarifier surface near the effluent weir. Figure 2.3 shows the

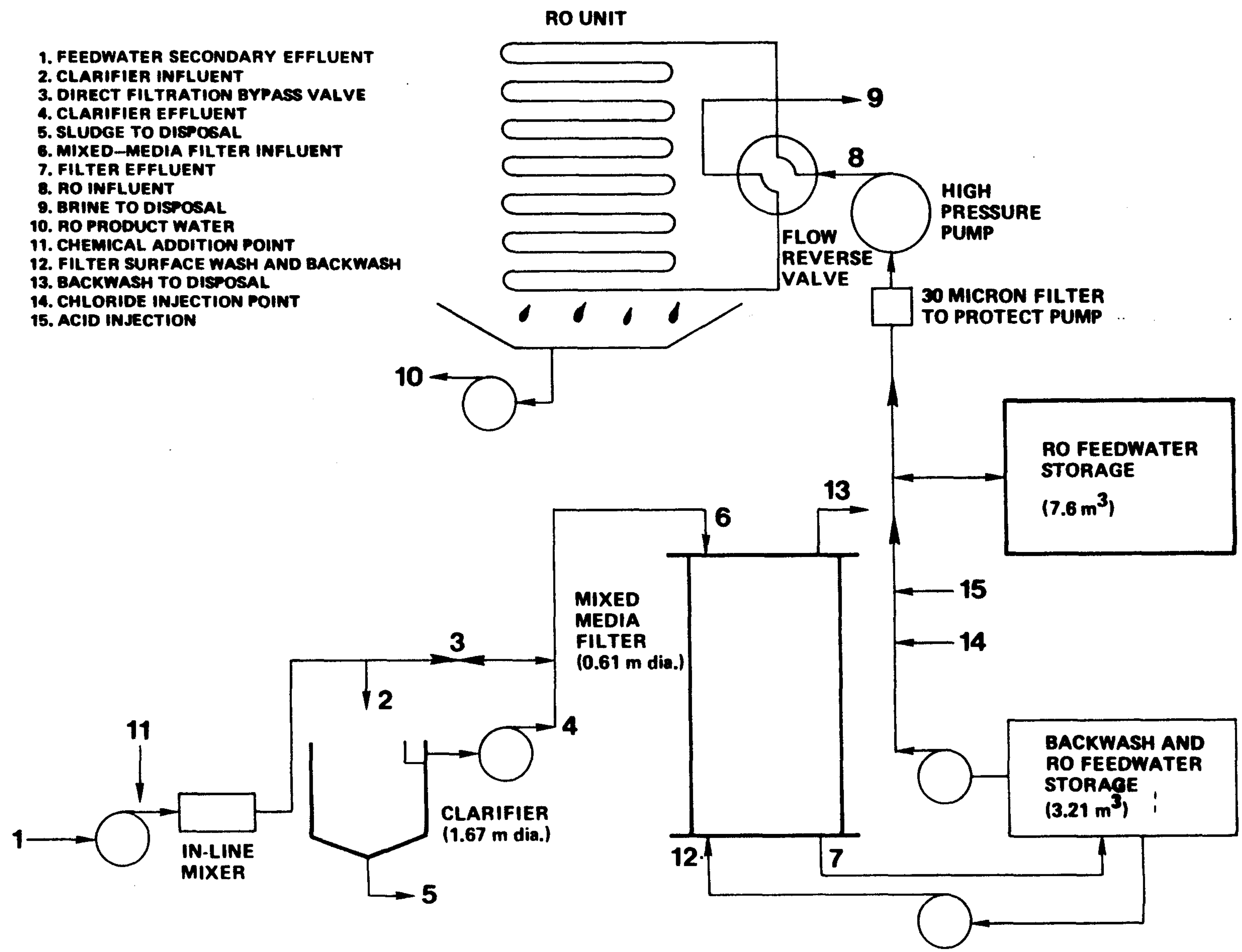

Figure 2.2: RO Pilot Plant and Pretreatment Facilities

activated sludge plant. The specifications and operating parameters for the plant are described in Table 2.3.

The filter was constructed from a 24 inch section of low carbon steel pipe, and was equipped with a rock underdrain structure and mixed media. The filter media used was a commercially available media (Neptune-Microfloc) consisting of a 1.0 to 1.2 mm size distribution of coal, a 0.42 to 0.55 mm size distribution of silica sand, and a 0.2-0.3 mm size distribution of garnet sand. The filter was backwashed using a hydraulic surface wash in addition to a normal backwash which fluidized the entire filter media.

CHRONOLOGY OF RO PLANT OPERATION

The RO pilot plant was originally placed in operation in April, 1976 treating trickling filter influent which was filtered through a 30 inch diameter multi-media (sand and coal) filter. This filter, besides providing feed for the RO unit, was operated by the Marin Municipal Water District (MMWD) to provide water for their reclamation activities. The RO unit operated on filtered trickling filter effluent from the Marin filter until May of 1979 when the 24 inch diameter mixed-media filter was installed and dedicated to pretreatment of RO feed water.

The initial period from April, 1976 to June, 1979 was dedicated to the development of membrane cleaning techniques and endurance testing of the RO membranes and equipment. The original cleaning technique was restricted to sponge ball cleaning without chemical cleaning agents (sponge ball cleaning was developed earlier by McCutchan and co-workers, and uses a sphere of flexible plastic or rubber which is forced along the tubular membranes by the brine

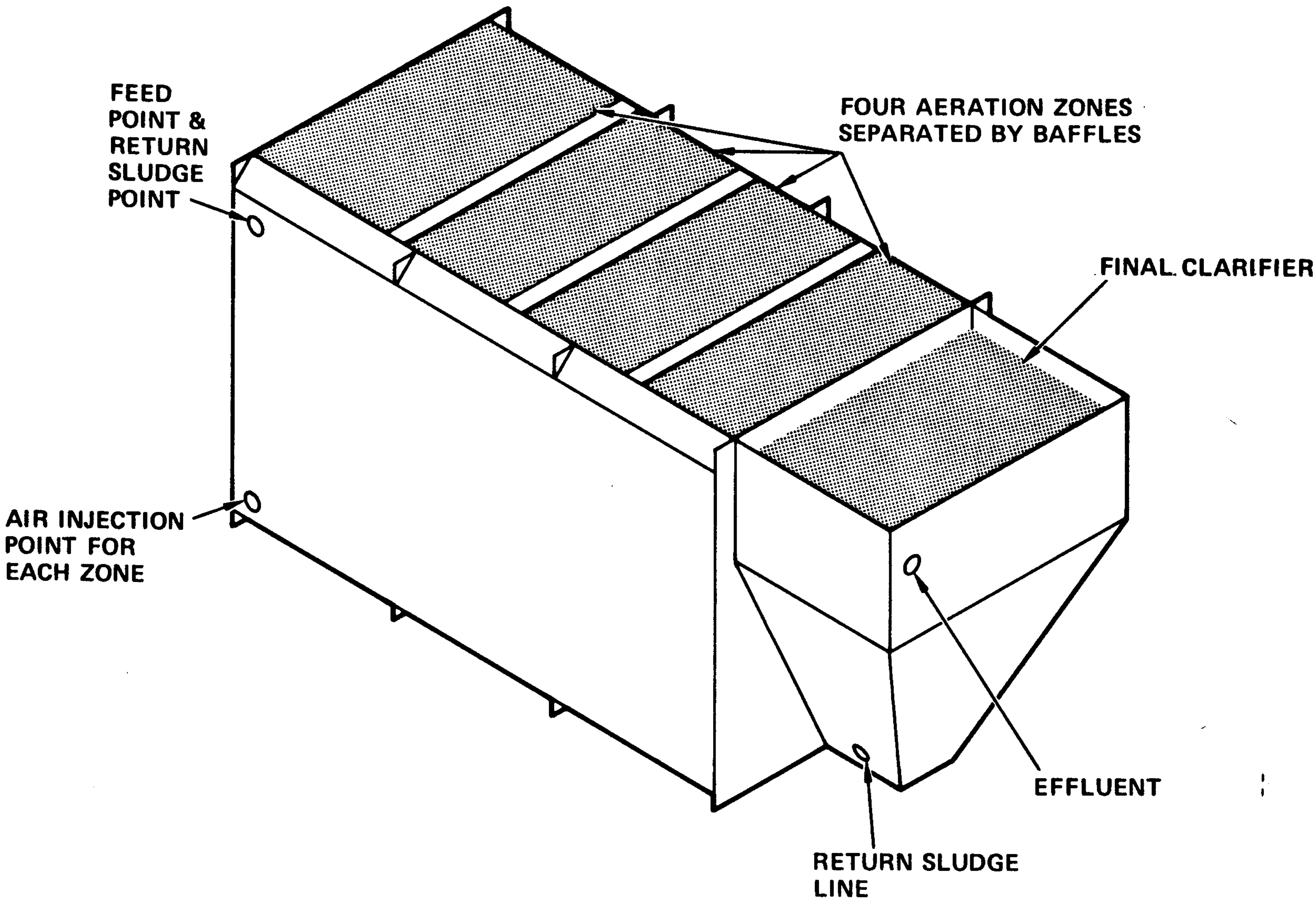

Figure 2.3: Activated Sludge Plant

Table 2.3: Specifications and Operating Parameters for the Activated Sludge Plant.

Parameter (1)	Value (2)
Aeration Section	
Depth	9 ft (2.75m)
Length	14 ft (4.25m)
Width	8 ft (2.43m)
Volume	7500 gal (28.4 m^3)
Clarifier Section	
Depth	9 ft (2.75 m)
Length	4.5 ft (1.37 m)
Width	8 ft (2.43 m)
Volume	1450 gal (5.5 m^3)
Surface Area	36 ft^2 (13.3 m^2)
Hydraulic Retention Time at 12 GPM*	10.4 hours
Overflow rate at 12 GPM	480 gal/ft^2day (4.9 m/day)
Mean Cell Retention Time	5 days

* 12 GPM = 0.045 m^3/min

or wash water pressure). During cleaning the unit was always depressurized and flushed with tap water or RO product water (later containing cleaning chemicals). Beginning in April of 1977 a two hour enzyme detergent flush was initiated. In June of 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 one hour flushes with citric acid and detergent followed by sponge ball cleaning, was developed. Table 2.4 summarizes the final cleaning procedure.

In March, 1978, chlorination of RO feed water was begun and in May of 1978 pH control of feed water was started. Operation continued in this fashion until June, 1979, when improved pretreatment facilities were placed in service.

From June until July, 1979 feed water was pretreated using direct filtration with a cationic organic polymer (Nalco 7134). In July, 1979 a 5.5 ft (1.7 m) diameter clarifier was installed and inorganic coagulants were used. At this time a protocol for operating the mixed-media filter was developed and continued throughout the remainder of the study. The filter was operated at 3.2 GPM/ft^2 (2.17 l/m^2 sec) filtration rate and backwashed at 18-20 GPM/ft^2 (10.2-13.6 l/m^2) for five minutes after a two minute surface wash. Backwashing was performed automatically on a timed cycle. Usually backwashes were performed every 12 hours. The filter was operated at the 3.2 GPM/ft^2 (2.17 l/m^2) rate independently of the RO feed water rate in order to provide uniform operation. Excess feed water was discharged with the Las Gallinas Valley Sanitary District effluent.

Table 2.4: Final Membrane Cleaning Procedure.

OPERATION (1)	PROCEDURE (2)
Citric Acid Flush	0.55 lbs (250 grams) of citric acid is added to 50 gallons (190 l) of tape water or RO product water at ambient temperature. This solution is circulated through the RO unit at approximately 5 GPM (0.315 l/sec) for one hour.
Enzyme Detergent Flush	1.10 lb (500 grams) of a commercially available detergent (Biz) is added to 50 gallons (190 l) of tap water at ambient temperature and circulated through the RO unit as before for one hour.
Sponge ball Cleaning	After completion of chemical cleaning, ten 1 1/2 inch (3.8 cm)sponge balls are introduced into the RO feed at approximately on minute intervals, and are allowed to pass through the unit at approximately 2.7 ft/sec (0.52 m/sec).
Sponge ball Cleaning	After approximately 70 hours of operation the unit is depressurized and the sponge ball cleaning is repeated.

The clarifier was operated at 10 GPM (0.63 l/sec) flow rate, giving an overflow rate of 610 gal/ft^{2}day (24.8 m^3/m^{2}day). Sludge was manually withdrawn on a regular basis.

The activated sludge plant was delivered to the Las Gallinas site in January, 1981, and placed in service in April, 1981. The plant was operated for two months to reach steady state. After reaching steady state in June, 1981, the entire pilot facility was placed in service, with the activated sludge plant providing feed water for the the RO unit. After a few hours of operation it was determined that the RO membranes were removing only 5 to 10 % of the influent total dissolved solids concentration (as measured by specific conductivity) and that the recovery rate was unusually high.

After examination of the membranes it was determined that they had deteriorated during storage. They were stored in the RO unit under approximately 50 PSIG of tap water pressure. The actual mode of deterioration remains unknown, but an analysis of a sample membrane indicated that the deterioration was consistent with hydrolysis and oxidation by chlorine. The destruction of the membranes was surprising since the replacement membranes were normally filled with tap water during storage. Also the membranes in normal operation were exposed to as much as 2.0 mg/l total chlorine residual. It was hypothesized that the destruction of the membranes occurred because of the chlorine contained in the Marin tap water. The Marin tap water usually contained a residual of less than 0.5 mg/l chlorine, but the residual was always in the form of free chlorine (HOCl or OCl^-). Since the Las Gallinas trickling filter never nitrified during the entire study, the residual chlorine in the RO feed water was always a combined residual (primarily monochloramine), which is know to affect cellulose acetate membranes less severely than free chlorine (Zachariah, 1982). An alternate hypothesis is failure due to bacterial degradation. In retrospect it is now known that the membranes should have been stored in the absence of free chlorine. One method of preserving the membranes would have been to inject ammonia into the feed water when the membranes were being preserved in tap water.

It was necessary to recast all the membranes at the Department of Water Resources' Firebaugh facility, were membrane casting equipment was located. This facility has recently been described by DWR (1983). Recasting and reinstallation was completed in January, 1982. During the period from June, 1981 to January, 1982 the activated sludge plant remained in operation, but data were not routinely collected.

The activated sludge plant was operated in the "conventional" mode (Metcalf & Eddy, 1979) during the entire study period. The feed water was pumped to the activated sludge plant from the launder of the Las Gallinas east primary clarifier through a 1 1/2 inch PVC pipe line using a submersible sump pump. Flow rate was monitored manually and ranged from 8 to 13 GPM. The large fluctuation in flow rate was caused by sliming of the PVC line. To reduce flow rate variation weekly cleaning was instituted by injecting several hundred milliliters of Chlorox bleach, followed by flushing with a 3 inch sponge ball. The flush was bypassed directly to a return sewer in order to prevent the chlorine from entering the activated sludge plant. The weekly cleaning helped control flow rate changes, but some variation still occurred.

The skimming device in the secondary clarifier was manually set to flow at approximately 0.25 GPM, but skimming was sporadic. Occasionally large quantities of scum would form and partially block the skimmer. Other times wind velocities or changes in flow rate would cause the skimmer to remove less than the desired rate. When the skimmer malfunctioned scum often was carried over to the downstream mixed-media filter, causing increased headloss and premature breakthrough of turbidity.

After repeated attempts to improve skimming it was decided to operate the 5.5 foot diameter clarifier which would act as a second skimmer. After placing this clarifier in service for skimming, no further scum problems occurred in the downstream operations. The clarifier removed very few suspended solids and rarely accumulated significant quantities of sludge, although it was periodically drained. The clarifier provided insurance against solids carry over. In an actual plant this problem would not occur, since mechanized skimmers would be provided, and operational intervention would be expected in the event of their failure. Therefore the clarifier was not included in later economic analysis where direct filtration following activated sludge treatment was used.

Sludge was wasted from the activated sludge plant directly from the mixed liquor using a Moyno pump with a variable speed DC motor. A sludge age of 4 to 6 days was maintained throughout the study. In this manner suspended solids determinations were not required to maintain sludge age control. Recycle sludge flow rate was maintained at approximately 8 GPM. Variations occurred due to occasional clogging of the return line (since gravity flow was used after an air lift pump).

It is useful to compare the activated sludge plant operation at the Las Gallinas test site to a typical full scale activated sludge plant. The Las Gallinas pilot plant did not receive as much operational attention as would be expected at a well operated full scale facility. The effluent turbidities for the pilot plant were more than the effluent turbidities routinely obtained at the activated sludge plants operated by the City of Los Angeles or Los Angeles County Sanitation Districts. This difference might be in part due to cooler operating temperatures at Las Gallinas. In contrast to most plants the Las Gallinas plant did not receive diurnal flow variation. The clarifier in the pilot plant did not operate as efficiently as a full scale clarifier, which was attributed to the lack of mechanical skimmer and rake. One would expect a full scale facility to provide equal or better quality effluent than the pilot plant.

Table 2.5 summarizes the period of operation and timing of significant events. Throughout the entire period of operation the units were maintained almost without day-to-day manual supervision. Perhaps 0.5 to 1.5 hours per day were spent on maintenance and operation (with the exception of membrane cleaning), and most of this time was spent data logging and mixing coagulants.

ANALYTICAL MEASUREMENTS

Most of the analytical work was performed on site using the existing laboratory facilities. Turbidities were measured with a Turner Designs Model 40-005 turbidity meter. Flow rates were usually measured by clocking flows into vessels of known volume. Extensive analyses of the influent and effluent water quality parameters were performed periodically by the Department of

Table 2.5: Chronological Summary of Pilot Plant Operation.

Date (1)	Hour (2)	Event (3)	Comment (4)
4/27/76	0	Pilot plant started up on trickling filter effluent after multi-media filtration.	Weekly sponge ball cleaning without chemicals
4/18/77	8,500	Cleaning procedure changed by the addition of two hour enzyme detergent flush.	Various concentrations of detergent (up to 2.1 g/l) were used for flushing.
6/20/77	10,000	Citric Acid substituted for enzyme detergent	Concentrations between 0.04 and 0.53 g/l were used
9/26/77	12,400	Returned to enzyme detergent	Concentrations between 1.05 and .32 g/l were used.
1/1/78	14,700	Final cleaning procedure developed, using one hour citric acid flush, followed by one hour enzyme detergent flush, followed by sponge ball cleaning	0.66 g/l citric acid concentration and 1.32 g/l enzyme used for flush
3/23/78	16,700	Chlorination of multi-media filter effluent begun	Chlorine residual ranged from 0.5 to 6.0 mg/l, averaging 2.0 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 sponge ball cleaning started	Cleaning frequency set at six hours
6/1/79	27,100	Mixed-media filter cationic polymer coagulation initiated	Dosage set by Zeta potential measurements
7/6/79	28,000	Clarifier installed and operation with various coagulants and modes until shut down	Optimal concentrations of $FeCl_3$, $Al_2(SO_4)_3$ evaluated.
1/7/80	32,400	Unit shut down.	Membranes stored under pressurized tap water
4/1/81	43,200	Activated sludge plant started up	
6/1/81	44,600	RO unit started up and shut down	Membranes destroyed
1/22/82	49,400	RO restarted using activated sludge plant as feed water	Operated with various coagulants until shut down
6/23/82	53,000	Unit shut down and disassembled	

Water Resources Laboratories, using Standard Methods (1975) techniques. Total dissolved solids (TDS) were always measured at the Las Gallinas site using a specific conductivity meter, but were measured gravimetrically by the DWR laboratory.

FLUX DECLINE TESTS

In order to evaluate the effectiveness of various pretreatment techniques in preventing flux decline, twenty-four hour flux decline tests with close monitoring of flux and product TDS were performed. These tests were conducted during three periods of three to four tests each during the first phase of this project, and over the last two weeks of operation during the second phase. Tests were purposefully performed in clusters in order to prevent the effects of changing influent composition and temperature from obscuring the effects of pretreatment alternatives. The general procedure for performing the test is summarized as follows:

1. To prepare for the test, injection of the coagulant to be evaluated was begun at the clarifier influent pump discharge. In the event direct filtration was being evaluated, the clarifier was bypassed. The pretreatment system was allowed to operated for several hours in order to come to steady state before turning on the RO plant.

2. The RO unit was chemically cleaned with a one hour citric acid flush, followed by a one hour enzyme detergent flush, followed by cleaning with ten oversized sponge balls, introduced at one minute intervals.

3. During cleaning the 2,000 gallon feed tank was drained, flushed with water from the pretreatment system (now operating under test conditions), and allowed to fill.

4. The mixed-media filter was backwashed, and the pretreatment system was turned on.

5. The RO unit was started and adjusted to a feed rate of 3.7 GPM (6.4 GPM in the first phase) and a pressure of 600 PSIG. Data collection was initiated 30 minutes after start-up.

6. Brine and product flow rate were determined by timing 30 to 60 seconds of flow into 0.264 gallon (1000 ml) graduated cylinders and recording the results in milliliters per minute and gallons per minute. The feed flow was calculated by summing the brine and product flows. TDS was measured and recorded; also recorded were turbidities, power usage, operating pressure, and pH. A sample data collection sheet is enclosed in the appendix.

7. The measurements were repeated at hourly intervals for the first few hours of the test (usually seven hours) and then repeated again the next morning.

8. After the final morning measurements, the pretreatment system was shut down and preparations were begun for another 24 hour test.

In the first phase, using treated trickling filter effluent, twenty-four hours were sufficient to determine flux decline rates. In the second phase, using activated sludge plant effluent, the flux decline tests were conducted over 48 hours. This increase was necessary due to reduced fouling rate of the activated sludge plant effluent.

3. EXPERIMENTAL RESULTS

FLUX DECLINE AND THE EFFECTS OF CLEANING

The earliest results with the RO unit were disappointing in that very low recovery rates were obtained. The recovery averaged about 25% with fluxes in the range of 4.5 to 5.0 gal/ft^2 day (GSFD) or 7.6 to 8.5 l/m^2 hr. The earliest use of the sponge ball was effective in restoring the flux to 9 to 10 GSFD after cleaning. After about 8,000 hours operation the flux before cleaning decreased to approximately 3.5 GSFD while the flux after cleaning was restored to only 4.2 to 4.5 GSFD. The deterioration was due to the precipitation of insoluble salts on the membrane surface. These salts were not removed by the mechanical cleaning of the sponge balls.

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 sponge ball cleaning gradually increased from 4 to 4.5 GSFD to a maximum of 5 GSFD. In June, 1977 the first citric acid cleaning was performed, which restored membrane flux to 12.5 GSFD. This flux after cleaning was maintained until the end of September when flushing only with the detergent was begun again. The flux after cleaning gradually declined and by December, 1977 had declined to the previous levels of 4 to 4.5 GSFD. Beginning in March of 1978 the final cleaning procedure shown previously in Table 2.4 was consistently used, and flux after cleaning stabilized to 12.5 GSFD. The results of the improvements in cleaning techniques can be seen in Figure 3.1 which shows the entire period of investigation for Phase I. The increases in flux due to improved pretreatment are obvious. The increases in fluxes after

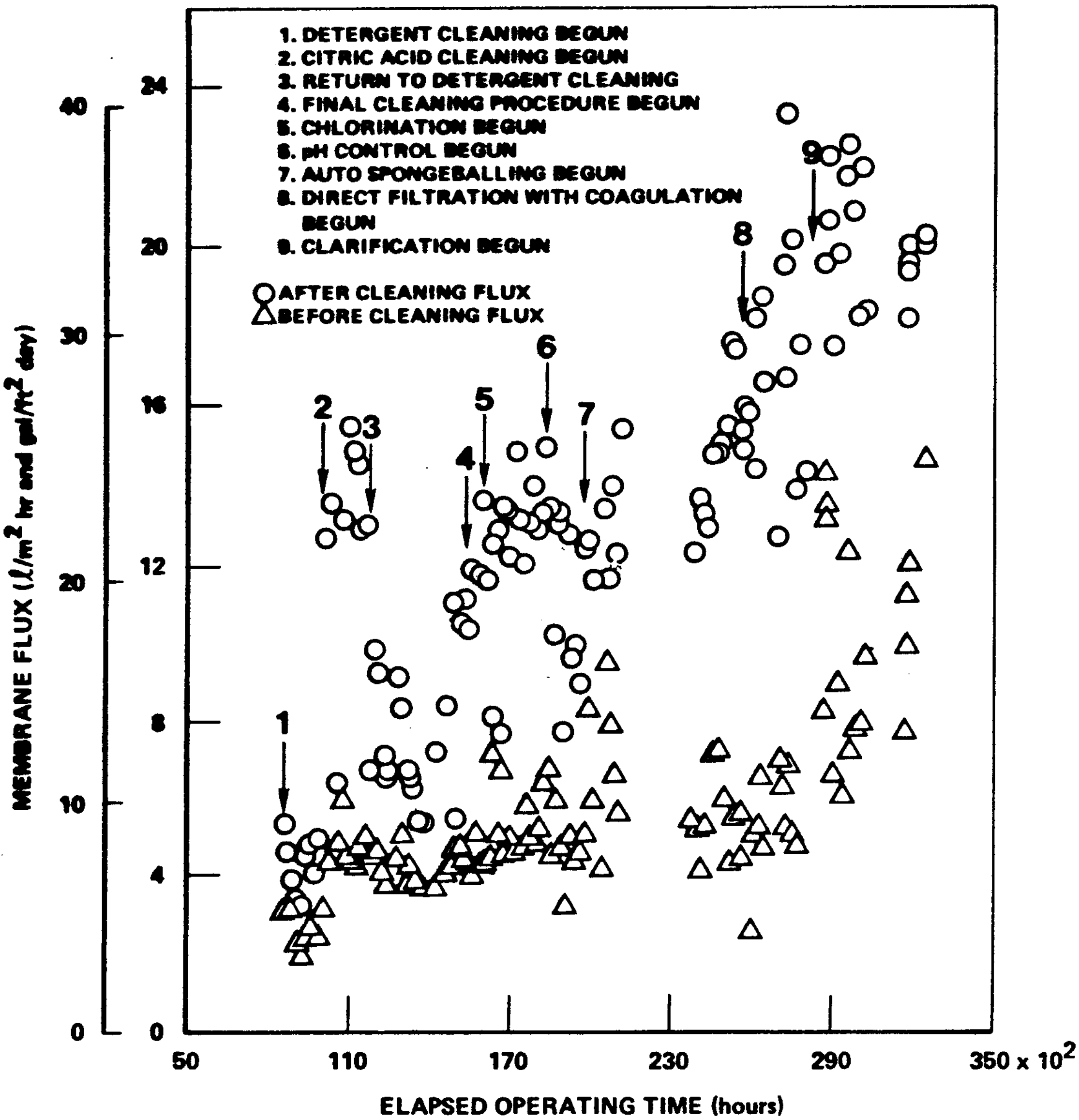

Figure 3.1: Membrane Fluxes Before and After Cleaning

January, 1978 were due to improved pretreatment techniques, rather than additional membrane cleaning techniques. This cleaning technique provides essentially complete membrane cleaning and was not changed for the remainder of the study.

FEED WATER QUALITY

Figure 3.2 shows the activated sludge plant effluent, clarifier effluent (second clarifier for scum control), and mixed-media filter effluent turbidities as a function of time. The activated sludge plant influent turbidity is not shown, but it averaged over 50 NTU and is typical of a primary effluent. The filter effluent averaged well below 2 NTU, which is significantly less than in Phase I when filter effluent turbidities ranged from 2 to 5, and were seldom less than 2.0.

It can be observed from Figure 3.2 than the clarifier effluent turbidity is little different than the activated sludge plant effluent. It was noted earlier that the clarifier was used primarily as a second scum control device, and that in the design analysis it was neglected. This figure supports the assumption that the clarifier would not be required in a full scale design, where mechanized skimming facilities would be available.

On April 5, 1982 samples of feed water and product water were collected and analyzed by the DWR Sacramento laboratory. Table 3.1 shows the results of these analysis, and two others performed in Phase I. The results shown in Table 3.1 for RO product water are very similar to the results obtained in Phase I. The feed water varies somewhat from Phase I as expected. The primary difference in feed water properties is the nitrogen compounds and its

Table 3.1 Chemical Analysis of RO Feed water and Product Water.

Parameter	Feed water			Product Water		
	3/19/79	9/17/79	4/5/82	3/19/79	9/17/79	4/5/82
(1)	(2)	(3)	(4)	(5)	(6)	(7)
Hardness+	216	241	163	13	19	6
Calcium	36	38	29	2	3	2
Magnesium	30	35	22	2	3	1
Sodium	136	218	70	28	57	7
Sulfate	77	251	75	1.5	8.5	1
Chloride	207	351	105	44	91	105
Boron	0.55	0.55	0.20	0.40	0.45	0.20
TDS	671	1090	434	98	203	41
TOC	27	23	36	1.5	1.2	1.9
Total N	26	39	19	4.1	4.7	2.6
Total P	9.3	12	7.0	0.21	0.83	0.05
Iron	0.09	0.36	0.51	0.0	0.08	0.04
Copper	0.01	0.15	0.13	0.0	0.01	0.02
Lead	<0.01	<0.01	0.015	<0.01	<0.01	<0.01

* All values reported in mg/l. Values represent averages of measured water quality before and after chemical cleaning.

+ As $CaCO^3$

In addition the following constituents were measured and less than 0.01 mg/l were found in both product and feed water: Arsenic, Cadmium, Chromium, Mercury, and Nickel.

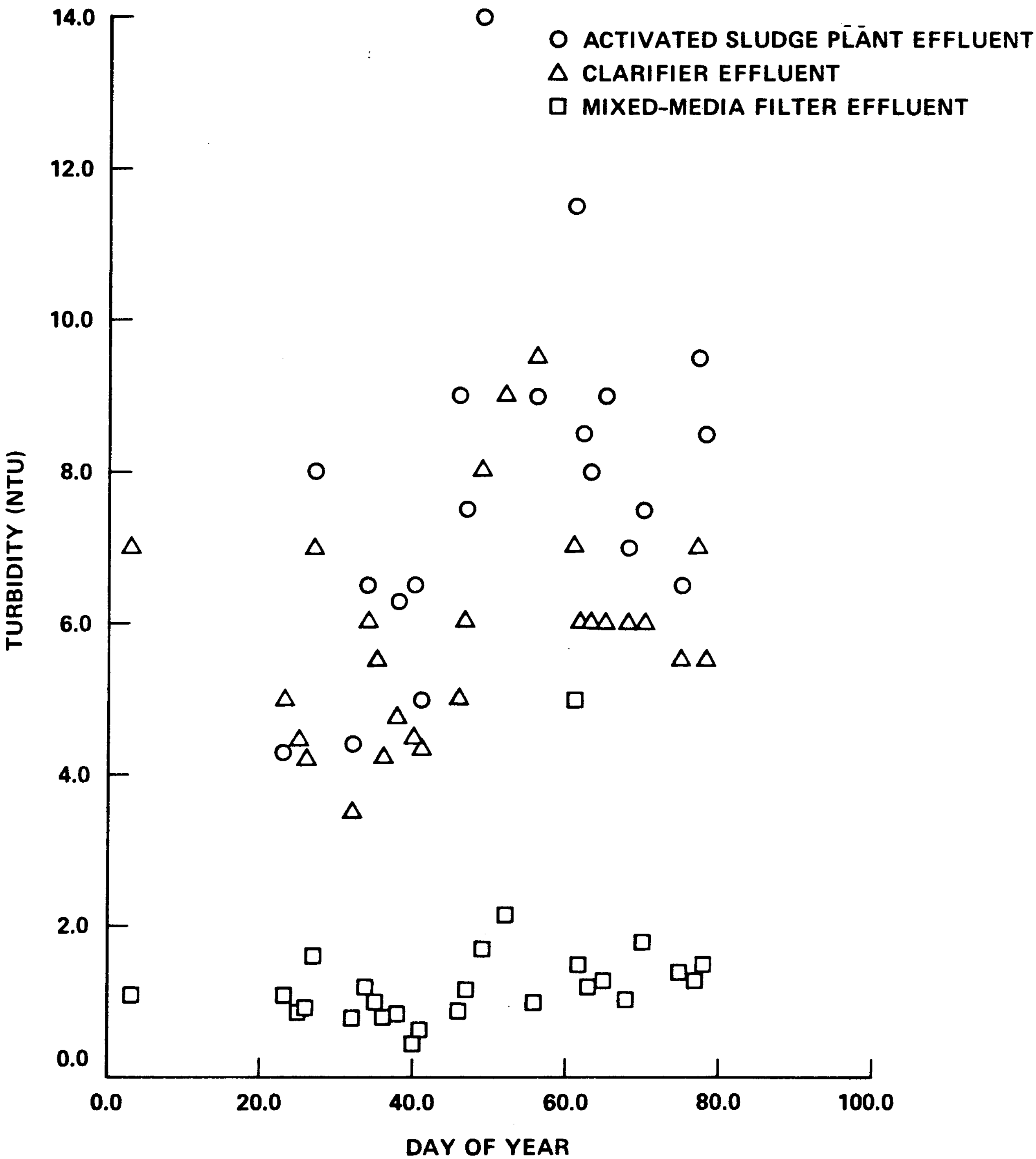

Figure 3.2: Turbidities as a Function of Time

forms. The nitrogen was almost 100% ammonia and organic nitrogen in Phase I, while in Phase II the activated sludge plant provided partial nitrification, reducing the ammonia concentration while increasing the nitrite and nitrate concentrations. This difference is also reflected in the product water, since with cellulose acetate membranes ammonia removal than nitrate and nitrite removal. In Phase I the total nitrogen of the product water was in the range of 4 to 7 mg-N/l, while in Phase II it was less than 2.0 mg-N/l.

FLUX DECLINE AND THE EFFECTS OF PRETREATMENT

Improvements made in recovery and flux maintenance after January, 1978 were largely due to improvements in RO feed water quality. Chlorination of RO feed water was begun in March, 1978, and feed water pH control (pH controlled to approximately 5.5) was begun in May, 1978. Both of these changes resulted in small increases in flux maintenance. The pH control improved flux maintenance due to the increased solubility of calcium sulfates and carbonates at the reduced pH, while chlorination prevented the growth of fouling slimes on the membranes. The actual purpose of pH control was to reduce membrane hydrolysis, but it also has this additional benefit.

The installation of the auto-sponge ball cleaning device in August, 1978, coincides with increases in before-cleaning fluxes to as high as 8 GSFD. Unfortunately the high before-cleaning fluxes declined to the level of 5 to 7 GSFD during the period of October, 1978 to May, 1979. No reason for this decline was determined.

The use of chemical coagulation and clarification had very large effects on both before and after-cleaning fluxes. Direct filtration with a cationic polymer which was begun on May 31, 1979 coincides with increasing trends in both before and after-cleaning fluxes. The before and after-cleaning fluxes increased to maximum values of about 14 and 25 GSFD, respectively, during the final periods of of Phase I, when the inorganic coagulants were used.

Flux Decline Tests

During various times in Phase I and at the end of Phase II a series of flux decline tests were made using various concentrations of ferric chloride, alum, and organic coagulants. During the first phase the tests were conducted over a twenty-four hour period, while in the second phase they were conducted over a 48 hour period. Flux decline curves for representative tests from Phase I are shown in Figure 3.3, along with a tap water flux decline test, which illustrates the flux decline caused by membrane compaction.

Unfortunately the flux decline tests performed in Phase II are not directly comparable to those shown in Figure 3.3, because of the differences in membrane characteristics. It was noted previously that the membranes were all replace beginning in July, 1981, due to deterioration during storage under tap water pressurization. The new membranes were cast using the same procedure as previously and cured at 88°C as previously; however the flux and sodium rejection properties of the new membranes were different than the old membranes. The new membranes were much "tighter" than the old membranes. The old membranes removed TDS to an average level of 200 mg/l, while the new membranes initially reduced the TDS to less than 50 mg/l and often less than 20

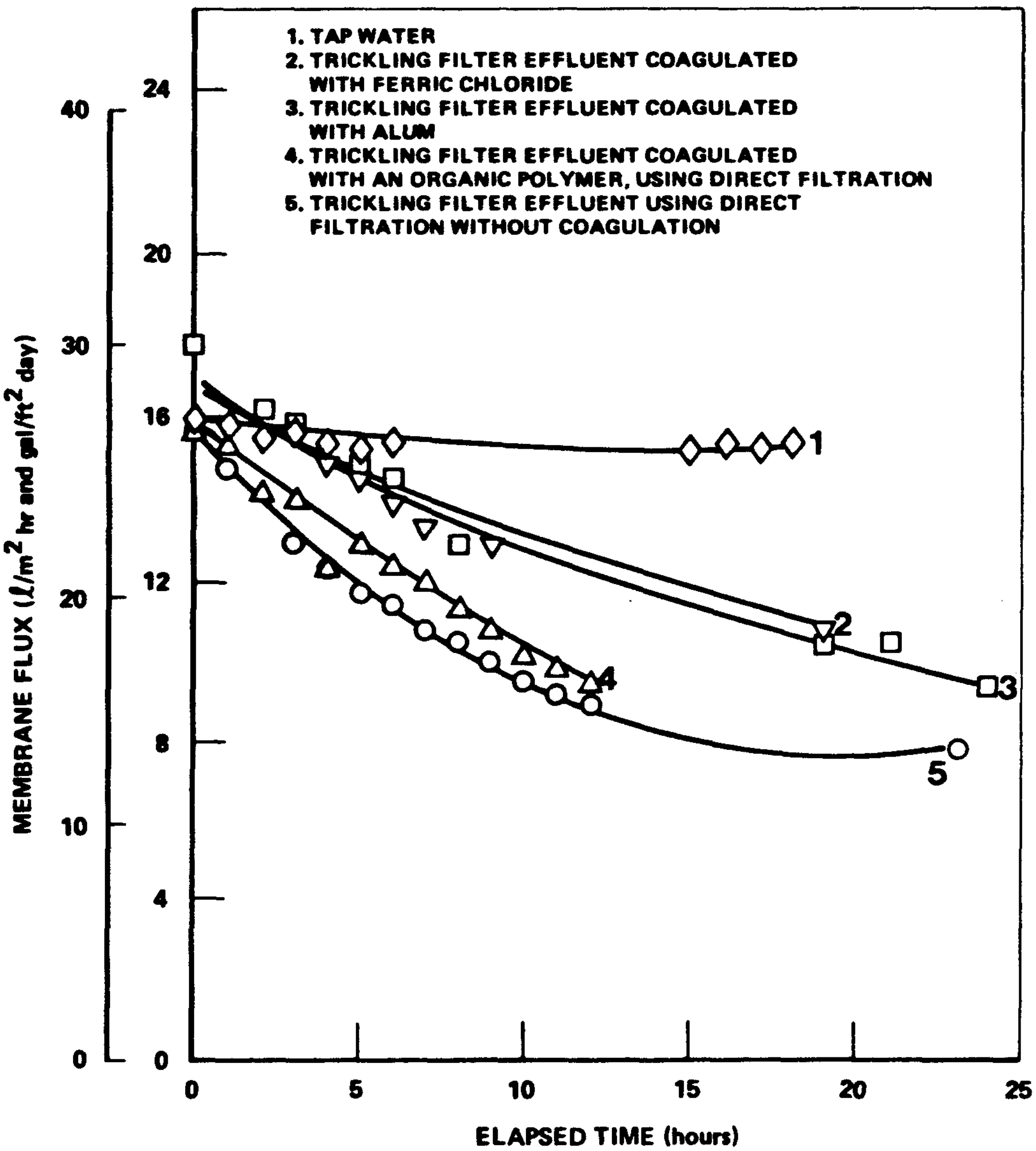

Figure 3.3: Twenty-Four Hour Flux Decline Tests: Phase I

mg/l. The after-cleaning flux of the old membranes averaged 15 to 20 GSFD, and was sometimes as high as 25 GSFD, while the after-cleaning flux of the new membranes was only 12 to 15 GSFD. This change in membrane properties is consistent with an increase in curing temperature, or may possibly be due to the newness of the membranes. At the conclusion of Phase I, when most of the twenty-four hour flux decline tests were performed, the membrane average age was 1.1 years, while the age of the membranes in the Phase II flux decline tests was six months or less. No explanation of the difference in membrane properties was determined.

In the design analysis performed later to determine the effects of pretreatment on design, a hypothetical condition was created which assumed the existence of a membrane which had the same flux decline properties as the new "tight" membranes, and the same salt rejection properties as the old membranes. This was a conservative assumption because a "looser" membrane should have higher fouling properties when biological materials are present (due to the high flux and resulting high throughput of fouling materials); therefore, flux decline should be higher with the old, "looser" membranes. The economic analysis described later shows that the new, "tight" membranes provide a more economical design; therefore the question of why the new membranes were different and what there flux decline properties were, does not effect the final conclusions of this study.

Figure 3.4 shows the three flux decline tests performed in June, 1982 using filtered activated sludge plant effluent, alum coagulated, filtered activated sludge plant effluent. and ferric chloride coagulated, filtered activated sludge plant effluent.

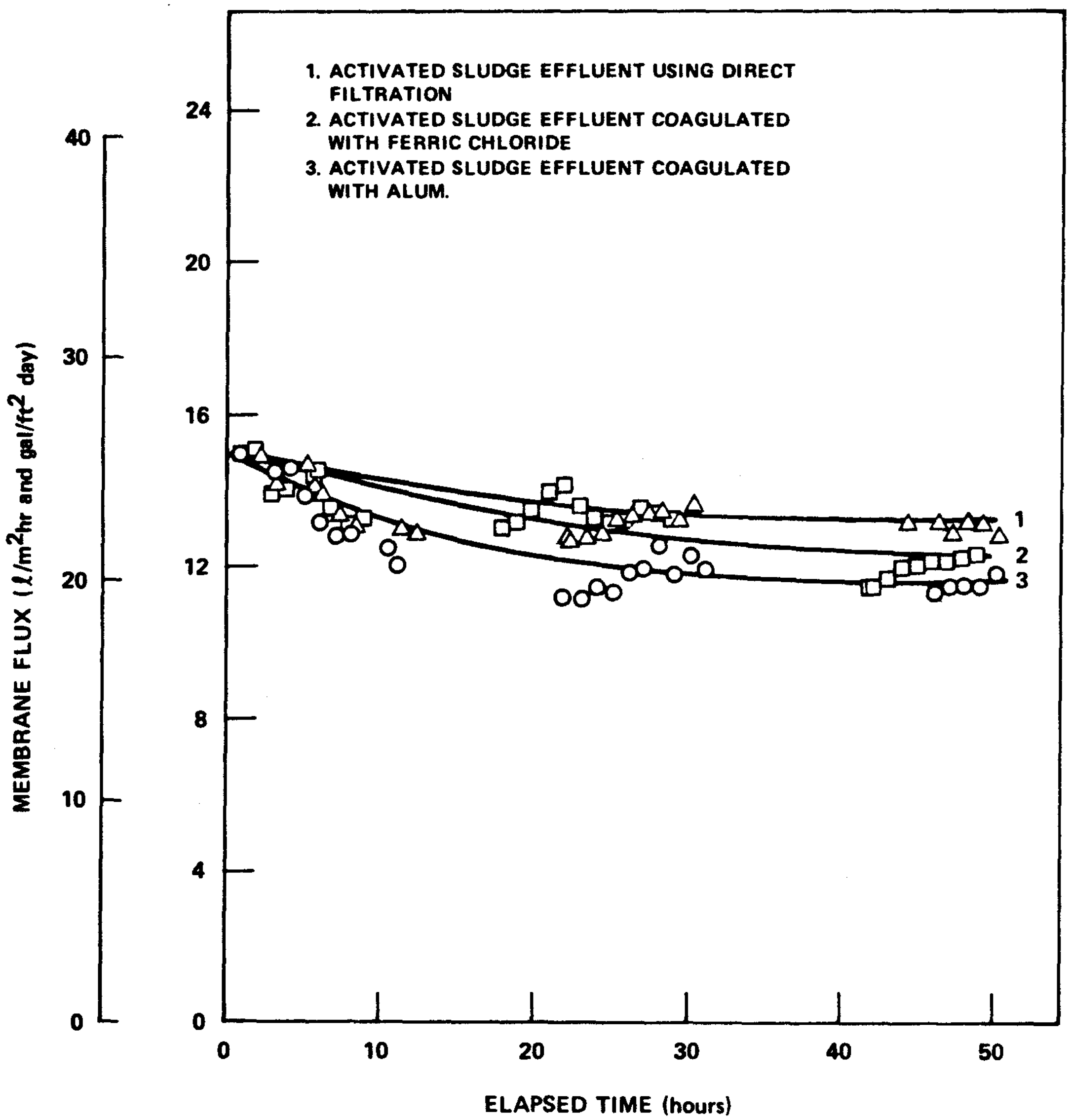

Figure 3.4: 48-Hour Flux Decline Tests-Phase II

In comparing the flux decline tests for Phase I and II, some observations can be made. The best case for Phase I, ferric chloride coagulation, showed a decline in flux from 16 GSFD to approximately 11.5 GSFD after 24 hours, or a decline to 72% of the after-cleaning flux. For the best case in Phase II, filtration without coagulation, the flux declined from 15 GSFD to 13.6 GSFD, or a decline to 91% of the after-cleaning flux. For alum coagulation, the worst case in Phase II, the decline was to 80% of the after-cleaning flux. The flux decline plots show dramatic evidence of the reduced fouling tendency of the activated sludge effluent.

The small fluctuations in product flow rate over the 48 hour period were due to changing feed water temperature. The actual Las Gallinas effluent temperature fluctuated very little during the day, but the activated sludge plant effluent, when stored in an above ground tank exposed to sunlight, varied in temperature by several degrees Celsius.

Flux Decline Parameter

In order to quantify the flux decline properties of a particular wastewater and at a particular condition, it was necessary to characterize the flux decline curves shown in Figures 3.3 and 3.4. The method of Thomas, et. al. (1973) was used. Thomas et. al. plotted the flux declines on log-log paper and found that the slopes were approximately linear, and called the slopes the Flux Decline Parameter. Flux decline parameters were calculated for the results from Phase I and II and are shown in Table 3.2 with several reported by Thomas et. al.(1973). The flux decline parameter (called the "B value" in the computer programs) is a useful method of comparing the fouling tendency of

Table 3.2: Flux Decline Parameters.

Flux Decline Parameter (1)	Feed Water Type (2)	Reference (3)
0.243	Trickling Filter Effluent with Dual Media Filtration	This Study, Phase I
0.202	Trickling Filter Effluent with Alum coagulation, clarification, and mixed-media filtration	This Study, Phase I
0.204	Trickling Filter Effluent with organic polymer coagulation, clarification, and mixed-media filtration	This Study, Phase I
0.146	Trickling Filter Effluent with ferric chloride coagulation, clarification, and mixed-media filtration	This Study, Phase I
0.0136	Tap Water (TDS=100 mg/l)	This Study, Phase I
0.059	Activated Sludge Effluent with ferric chloride coagulation, clarification, and mixed-media filtration	This Study, Phase II
0.037	Activated Sludge Effluent with direct mixed-media filtration	This Study, Phase II
0.075	Activated Sludge Effluent with alum coagulation, clarification, and mixed-media filtration	This Study, Phase II
0.9	Raw Wastewater	Calculated by Thomas, et. al. (1973) from Feuerstein and Bursztynsky (1970).
0.56	Primary Effluent	Calculated by Thomas, et. al. (1973) from Feuerstein and Bursztynsky (1970).
0.35	Secondary Effluent	Calculated by Thomas, et. al. (1973) from Feuerstein and Bursztynsky (1970).
0.14	Carbon-treated Secondary Effluent	Calculated by Thomas, et. al. (1973) from Feuerstein and Bursztynsky (1970).

a wastewater. The lower the flux decline parameter, the lower the fouling tendency of the feed water. The small positive value of the flux decline parameter for Marin tap water is probably due to membrane compaction. It is obvious from Table 3.2 than pretreatment significantly reduced the fouling tendency of the feed water.

4. DESIGN ANALYSIS

The design approach used in this report draws heavily upon the technique developed in Phase I. The Phase I technique was developed in part from previous work by Hatfield (1967), Hatfield and Graves (1970), Fan, et. al. (1970), and McCutchan and Goel (1974). The Phase I report (Davis, et. al. 1979)) should be consulted for a more complete discussion of the design and optimization techniques.

FACILITY SIZE

Using the data collected in this study and the previous phase, the cost data compiled by the Oak Ridge National Laboratory (1980), and the EPA (1979), an economic model was developed. When the model was developed in Phase I, these two sources of cost data were current. Unfortunately these two sources are still the most current cost data available. They have been updated through the use of the Engineering News-record (ENR) cost updates to the present in order to keep the financial calculations current. Table 4.1 shows the ENR index, along with other indices frequently used.

The model is based upon using the RO product water in conjunction with a specific quantity of RO feed water, to provide water for recycle with a specified water quality. The calculation procedure is to determine the minimum quantity of RO product water for blending with feed water to meet the specified water quality standards such as TDS, TOC, turbidity, and biochemical oxygen demand. Pretreatment level and cleaning frequency are considered as variables, while RO operating pressure, membrane characteristics, and velocity are considered constant.

Table 4.1: Cost Indices from Various Sources.

Year (1)	ENR[1] Construction (2)	ENR[1] Building (3)	EPA[2] (4)	CE Plant[3] (5)	M & S[4] (6)
1978	2869	1734	145	218	545
1979	3140	1909	158	239	599
1980	3376	2017	169	261	660
1981	3705	2184	180	297	721
1982	3931	2294			746

[1] Engineering News-record index for the fourth quarter of each year.
[2] EPA national average index for 5 MGD plants for the fourth quarter of each year.
[3] CE plant cost index, published in Chemical Engineering
[4] M & S equipment cost index published in Chemical Engineering

BASIS FOR COST ESTIMATES

The least cost pretreatment system was found by simulating the RO unit with various pretreatment alternatives. This basis of comparison for the simulation was the flux decline parameter (B), for each pretreatment alternative, and the associated processes which provided the pretreatment. The flux decline parameters shown previously Table 3.2 were used in the simulation program. The cleaning frequency was evaluated for each pretreatment alternative and the value producing the least cost was found. The cost using this cleaning frequency for each alternative was then compared, and an optimal alternative was selected.

The costs for the pretreatment system processes were considered to be log-linear functions using parameters calculated from the appropriate cost reference. This is the technique of the EPA (1979) method. The assumption of log-linearity corresponds well with the data for plants in the range of one to ten MGD. The cost equations take the following form:

$$\log_{10} (\text{ Cost Variable }) = a * \log_{10} (\text{ Size Variable }) + b. \qquad (4.1)$$

and

$$\text{Cost} = \text{Index} * \text{Cost Variable} \qquad (4.2)$$

where

a and b are parameters from the EPA (1979) document,

and Index is the appropriate cost update index shown in Table 4.1.

The functions were calculated for each of the five categories specified in the EPA (1979) report which allows for variation of the costs for labor and energy. Labor and energy costs are assumed independently of the EPA cost figures. For the analysis presented in the Phase I report, the cost of labor was assumed to be $12.00/hr, while energy costs were assumed to be $0.05/kWhr. It was assumed for this report that these costs gradually increase to $16.84/hr and $0.07/kWhr. The interest rate was assumed to be 8% in the Phase I report and has been increased to 12% for this analysis.

Implicit in this analysis are assumptions about scale up in technical parameters such as fouling rate, and that costs per unit area for tubular and spiral wound membrane modules will be similar. Brine disposal costs have not been included, nor have any costs been assumed for the secondary treatment system. Table 4.2 summarizes the design and size variables, while Table 4.3 shows the cost coefficients based upon the 1979 data. Sample calculations using the technique describe herein were present in the Phase I report.

Table 4.4 shows the optimal design for the 1 MGD hypothetical reclamation plant using the new information generated in Phase II. Several differences should be noted. No coagulants are required, and in fact coagulants not only add to the operating cost but reduce recoveries as well. This was not true of the results of the Phase I. Energy consumption is less due to the increased recovery rate and reduced number building and processes. Operational requirements are less since no coagulating chemicals are required. The optimal results for Phase I are shown in the appendix. The results of all alternatives, including the four pretreatment alternatives evaluated in Phase I, in terms of cost per thousand gallons, are presented in Table 4.5. Also

Table 4.2: Size Variables and Design Basis.

Process (1)	Size Variable (2)	Design Basis (3)	Rate (4)
Filter			
Vessels, Tanks etc	Filter Area	Loading Rate	5 GPM/ft^2
Surface Wash	Filter Area	Loading Rate	5 GPM/ft^2
Filter Media	Filter Area	Loading Rate	5 GPM/ft^2
Backwash	GPM	Upflow Velocity	2 ft/sec
Clarifier			
Vessels	Clarifier Area	Loading Rate	1000 GPD/ft^2
Polymer	lbs/day	Dosage	5 mg/l
$FeCl_3$	lbs/day	Dosage	50 mg/l
Alum	lbs/day	Dosage	60 mg/l
Reveres Osmosis			
Vessels	Feed Flow	Flux Decline Index (B)	0.05-0.23
Chlorine	lbs/day	Dosage	2-9 mg/l
Acid	lbs/day	Dosage	15 mg/l

Table 4.3: Cost Coefficients for the Log-Linear Functions.

Cost Section	Total Capital ($)		Energy(kWhr/yr)				Operation and Maintenance			
			Lights, Heating & Cooling		Process		Materials ($/yr)		Labor (hr/yr)	
	a	b	a	b	a	b	a	b	a	b
(1)	(2)	(3)	(4)	(5)	(6)	(7)	(8)	(9)	(10)	(11)
Filter										
Tanks & Vessels	0.32	4.72	21,000		0.97	2.47	0.79	2.47	0.30	2.51
Surface Wash	0.24	3.96	NA	NA	0.89	1.45	0.14	2.00	0.49	0.69
Filter Media	0.65	2.55	NA	NA	NA	NA	NA	NA	NA	NA
Backwash	0.37	3.49	NA	NA	1.00	1.38	0.28	2.24	0.062	2.15
Clarifier										
Vessels	0.32	4.01	NA	NA	0.17	3.02	0.64	1.57	0.15	1.74
Polymer	20,200[1]		8,210[1]		17,300[1]		270[1]		198[1]	
$FeCl^3$	0.28	4.00	0.57	3.20	4,900[1]		0.067	2.19	0.062	0.067
Alum	0.23	4.08	0.57	3.22	4,900[1]		200[1]		0.62	3.97
Reverse Osmosis										
System	0.85	5.89	0.90	5.02	0.96	6.38	0.19	3.27	0.89	4.99
Acid	0.12	3.82	3,680[1]		1,630[1]		0.33	1.53	0.22	1.56
Chlorine	0.36	3.75	0.52	3.45	0.17	2.58	0.11	2.53	0.18	2.57
Cleaning	0.28	3.31	NA	NA	1.0	1.08	0.62	2.13	0.28	2.16

[1] Value is approximately constant for the range of considered values.

Table 4.4: Optimal Design for a 1 MGD Facility. (3 pages)

THE LIMITING PARAMETER IS TDS FOR B(8)= 0.037

THE FOLLOWING WATER QUALITY RESULTS

	FILTER	RO	REQUIRED	BLENDED
TDS	1200.	250.0	500.0	500.0
TOC	18.	1.0	15.0	5.5
NTU	3.	0.5	2.0	1.2
TSS	4.	0.0	5.0	1.1

THE RATIO OF BLENDED RO PRODUCT WATER TO TOTAL PRODUCT FLOW = 0.737

ASSUMPTIONS

LABOR RATE = $ 16.84 PER HOUR

ELECTRICAL RATE = $ 0.07 PER KWH

INTEREST RATE = 12.00%

LIFE OF PROJECT = 20 YEARS

INFLATION RATIO = 1.370

PROJECT YEAR = 1983

COSTS PER KGALS FOR VARYING VALUES OF B AND THE CLEANING INTERVAL

CLEANING INTERVAL (HOURS)

B	8	16	24	32	40	48	56	64	72	80
0.04	2.16	2.03	2.00	1.98	1.98	1.98	1.98	1.98	1.99	1.99

Table 4.4 Continued

ENERGY AND LABOR ANALYSIS

	BUILDING ENERGY KWH/YR	BUILDING ENERGY COST($/YR)	PROCESS ENERGY KWH/YR	PROCESS ENERGY COST($/YR)	LABOR HRS/YR	LABOR $/YR
FILTER UNIT	41176.	3088.	46072.	3455.	1556.	26205.
SURFACE WASH	0.	0.	2951.	221.	62.	1041.
MEDIA	0.	0.	0.	0.	0.	0.
BACKWASH	0.	0.	4425.	332.	193.	3257.
CLARIFIER UNIT	0.	0.	3641.	273.	165.	2776.
COAGULANT	0.	0.	0.	0.	0.	0.
REVERSE OSMOSIS	21883.	1641.	2560645.	192048.	1863.	31368.
SULFURIC ACID	722.	54.	1630.	122.	70.	1182.
CHLORINE	5867.	440.	487.	36.	395.	6644.
CLEANING	0.	0.	308.	23.	164.	2759.

UNIT COSTS

	TOTAL CAPITAL($/YR)	ANNUAL CAPITAL	O & M ($/YR)	WATER COSTS($/KGAL)
VESSELS	381790.	51114.	34933.	0.25
SURFACE WASH	43206.	5784.	1545.	0.02
MEDIA	14727.	1972.	0.	0.01
BACKWASH	79405.	10631.	4619.	0.04
CLARIFIER	142067.	19020.	3560.	0.07
COAGULANTS	0.	0.	0.	0.0
RO UNIT	1121885.	150197.	365472.	1.52
H2SO4	11759.	1574.	1531.	0.01
CHLORINATION	12983.	1738.	7777.	0.03
CLEANING	18733.	2508.	8972.	0.03

COAGULATION/CLARIFICATION/FILTRATION $/KGAL= 0.39
REVERSE OSMOSIS SYSTEM $/KGAL= 1.59
TOTAL SYSTEM $/KGAL= 1.98

THE PLANT PRODUCES 1 MGD PER DAY OF THE SPECIFIED QUALITY WATER
THE TOTAL ANNUAL COSTS = $ 673065. COST/KGAL = $1.98

Table 4.4 Continued

THE OPTIMAL DESIGN REQUIRES FILTRATION ONLY

OPTIMAL SYSTEM SPECIFICATIONS

FILTER

INFLUENT FLOW:	1.33	MGD
LOADING RATE:	5.00	MGD
FILTER AREA:	184.	SQUARE FEET
DIAMETER:	15.	FEET
BACKWASH VELOCITY:	2.	FEET/MIN

REVERSE OSMOSIS

INFLUENT FLOW:	1.07	MGD
PRODUCT FLOW:	0.74	MGD
PERCENT RECOVERY:	69.18	%
FLUX DECLINE INDEX (B):	0.04	
AVERAGE FLUX:	17.12	GPD/FT2
NUMBER OF MEMBRANES:	18709	
TOTAL AREA:	43031.37	SQUARE FEET
TIME REQUIRED FOR CLEANING:	2.0	HOURS
CLEANING INTERVAL:	40.0	HOURS
SULFURIC ACID INJECTED:	8.19	PPM
CHLORINE INJECTED:	9.00	PPM

Table 4.5: Treatment Cost in Dollars per Thousand Gallons for a 1 MGD Reverse Osmosis Wastewater Reclamation Facility.

	TYPE OF PROCESS AND PRETREATMENT*							
Year	TF $FeCl_3$	TF Nalco	TF Alum	TF None	ASP $FeCl_3$	ASP None	ASP Alum	ASP+ None
1979	1.57	1.81	2.09	1.70	1.22	1.11	1.65	1.29
1980	1.84	2.11	2.41	1.99	1.42	1.30	1.89	1.50
1981	2.04	2.33	2.67	2.21	1.58	1.45	2.09	1.67
1982	2.22	2.53	2.91	2.40	1.72	1.57	2.29	1.82
1983	2.42	2.74	3.17	2.61	1.87	1.71	2.49	1.98

*CODES FOR PROCESS AND PRETREATMENT

TF = Trickling Filter Secondary Treatment

ASP = Activated Sludge Process Secondary Treatment

$FeCl_3$: Coagulation, Sedimentation, and Filtration using ferric chloride as a coagulant.

Alum: Coagulation, Sedimentation, and Filtration using alum as a coagulant.

Nalco: Coagulation, Sedimentation, and Filtration using Nalco cationic polymer as a coagulant.

None: Filtration with no coagulation or sedimentation.

+ Hypothetical membrane having the same fouling properties as the tight membranes used in the second phase, but having flux and TDS removal properties similar to the membranes used in the first phase.

Included in Table 4.5 is the hypothetical membrane, described earlier, which was used to evaluate the potential effects of the "tight" membranes used in the Phase II. It is observed that this membrane did not provide the least cost alternative. The cost values are considered tentative due to the ambiguities of scale-up.

5. CONCLUSIONS

The results of an experimental and theoretical analysis of a 10 GPM pilot plant for producing reclaimed water has been presented. The results were applied to the design of a full scale 1 MGD facility. In addition to the conclusions presented in the Phase I report (none were contradicted in this phase of research), the following additional conclusions are made:

1. The activated sludge plant effluent had significantly less tendency to foul the membranes than trickling filter effluent, indicating that a major source of fouling material in Phase I was organic material. This corroborates the predictions from the fouling material analyses performed in Phase I. The activated sludge plant followed by direct filtration produced feed water which had only one-third the fouling tendencies of the best trickling filter effluent which could be obtained from the Las Gallinas facilities.

2. Cleaning using flushes of citric acid, followed by enzyme detergent and sponge ball cleaning were effective at maintaining membrane flux to essentially the initial flux levels. The citric acid was the major cleaning agent. Enzyme detergent and/or sponge ball cleaning without citric acid were relatively ineffective.

3. The automatic sponge ball cleaning technique appeared to have promise for maintaining membrane flux between chemical cleanings. Further testing is desirable.

4. The major factor contributing to membrane degradation for the type of membranes used in this study was corrosion of the end couplings.

The average membrane life during Phase I was 10,000 hours.

5. The chance production of tighter membranes reduced the total operating cost by \$0.27 /1000 gallons.

6. The greatest level of pretreatment again produced the least cost alternative.

7. Alum was always the poorest coagulant, which was probably due to carry over into the RO unit of aluminum hydroxide which has minimum solubility at pH=5.5.

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APPENDICES

Appendix 1. Data Summary from 1/1/82 to 6/30/82

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1981-82)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT -- FLUX (GSF)	RECOVERY
04/27/82	650.	1400.	238.	3.67	10.53	0.734
04/28/82	675.	950.	160.	3.67	9.08	0.633
04/29/82	700.	2000.	50.	3.59	9.35	0.666
+04/29/82	675.	2200.	44.	3.52	12.44	0.904
04/29/82	700.	2900.	66.	3.57	12.60	0.903
05/03/82	650.	1600.	36.	3.53	8.37	0.607
*05/03/82	650.	1600.	105.	3.51	13.07	0.952
05/04/82	625.	650.	202.	4.64	13.70	0.755
05/05/82	650.	4100.	82.	3.47	11.39	0.839
05/06/82	650.	4000.	85.	3.40	10.17	0.765
+05/06/82	650.	2500.	70.	3.72	13.97	0.960
05/10/82	625.	1425.	35.	3.54	7.67	0.554
*05/10/82	650.	2000.	92.	3.54	12.99	0.939
05/11/82	650.	950.	215.	3.70	10.64	0.736
05/12/82	650.	2300.	44.	3.57	9.94	0.712
+05/13/82	625.	2700.	66.	3.58	11.23	0.802
05/14/82	600.	900.	35.	3.50	8.53	0.623
05/17/82	575.	1475.	52.	3.51	6.85	0.499
*05/17/82	550.	3000.	80.	3.58	13.81	0.987
05/18/82	600.	2000.	167.	3.54	10.76	0.777
05/19/82	600.	1300.	88.	3.61	10.41	0.737
05/20/82	575.	1900.	35.	3.54	9.47	0.684
+05/20/82	575.		78.	3.59	12.80	0.912
05/21/82	650.	2700.	40.	3.66	10.49	0.733
05/24/82	725.	2000.	42.	3.57	8.69	0.622
*05/24/82	775.	2800.	155.	3.59	13.85	0.987
05/25/82	825.	2200.	88.	3.61	9.39	0.665
05/25/82	800.	2250.	128.	3.56	12.68	0.911
05/27/82	1100.	7600.	72.	3.54	8.57	0.619
+05/27/82	1150.	8400.	190.	3.37	11.90	0.903
05/28/82	900.	5000.	88.	3.47	11.27	0.831
06/01/82	875.	1950.	55.	3.45	7.79	0.577
*06/01/82	850.	0.	180.	3.53	13.81	1.001
06/03/83	750.	5000.	102.	3.57	11.58	0.830
06/03/82	800.	3200.	69.	3.49	10.33	0.757
+06/03/82	1700.	0.	112.	3.52	13.77	1.001

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

Appendix 1. Continued

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1981-82)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT-- FLUX(GSF)	RECOVERY
03/19/82	2000.	5000.	288.	4.92	9.86	0.513
03/22/82	1550.	3550.	225.	4.85	9.08	0.479
*03/22/82	700.	1900.	100.	4.83	12.29	0.651
03/23/82	625.	1150.	46.	4.80	11.74	0.625
03/24/82	650.	1800.	78.	4.81	10.02	0.533
03/25/82	650.	1300.	24.	4.81	9.74	0.518
+03/25/82	650.	1550.	42.	5.10	12.01	0.602
03/26/82	700.	1200.	120.	5.05	10.84	0.549
03/29/82	700.	1225.	28.	5.05	8.65	0.438
*03/29/82	700.	1100.	102.	3.14	11.11	0.905
03/30/82	600.	800.	215.	3.31	8.33	0.644
03/31/82	550.	750.	112.	3.22	7.71	0.612
04/01/82	450.	1100.	45.	3.20	7.55	0.604
+04/01/82	420.	800.	38.	3.59	9.70	0.691
04/02/82	360.	435.	90.	3.65	9.12	C.639
04/02/82	420.	1200.	24.	3.58	9.20	0.657
04/05/82	440.	875.	20.	3.69	6.65	0.461
*04/05/82	490.	2600.	60.	3.67	11.74	0.818
04/07/82	600.	1100.	41.	3.10	10.84	0.894
04/08/82	600.	1250.	75.	3.66	11.35	C.793
04/09/82	600.	1300.	44.	3.67	8.53	0.594
+04/09/82	600.	1300.	68.	3.75	13.34	0.910
04/10/82	600.	1750.	39.	3.54	9.08	0.656
*04/12/82	400.	1700.	50.	3.59	12.80	0.912
04/13/82	385.	1300.	205.	3.53	9.94	0.720
04/14/82	450.	1050.	42.	3.57	9.43	0.676
04/14/82	850.	2250.	70.	3.53	11.15	0.808
04/15/82	1100.	4700.	85.	3.39	7.83	0.590
+04/15/82	600.	1150.	67.	3.64	11.90	0.836
04/16/82	525.	600.	62.	3.59	11.27	0.803
04/19/82	500.	1200.	70.	3.62	8.49	0.600
04/19/82	550.	1350.	55.	3.58	8.45	0.604
*04/19/82	550.	2450.	42.	3.57	13.34	0.956
04/20/82	550.	6700.	78.	3.48	12.37	0.909
04/21/82	590.	950.	248.	3.62	10.33	0.730
04/22/82	550.	4850.	62.	3.53	11.11	0.805
+04/22/82	575.	10250.	102.	3.79	13.62	0.919
04/23/82	600.	4500.	72.	3.70	12.40	0.857
04/26/82	800.	1950.	38.	3.53	8.61	0.624
*04/26/82	650.	12500.	90.	3.57	13.07	0.936

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

Appendix 1. Continued

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1981-82)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT-- FLUX(GSF)	RECOVERY
01/22/82	510.	950.	40.	5.25	10.10	0.492
01/23/82	600.	1150.	26.	5.25	9.82	0.478
01/25/82	600.	775.	20.	5.73	9.00	0.402
01/25/82	550.	1050.	26.	6.25	9.27	0.379
01/28/82	700.	1450.	29.	6.13	9.16	0.382
01/27/82	650.	1000.	16.	6.14	8.73	0.363
02/02/82	575.	1025.	15.	6.06	8.69	0.367
+02/02/82	700.	1450.	49.	6.31	11.58	0.469
02/04/82	750.	1250.	19.	6.26	10.02	0.409
02/05/82	775.	2350.	62.	3.72	10.02	0.689
02/06/82	825.	2550.	33.	3.70	9.67	0.668
02/08/82	900.	2150.	45.	3.72	8.18	0.562
02/08/82	900.	1900.	34.	4.75	8.61	0.464
02/10/82	850.	1600.	24.	4.68	7.79	0.426
02/11/82	900.	1500.	23.	4.72	7.12	0.386
+02/11/82	850.	1800.	34.	4.78	8.49	0.454
02/16/82	500.	950.	15.	4.83	8.77	0.464
+02/16/82	575.	1075.	30.	4.85	10.33	0.545
02/17/82	725.	1450.	29.	4.98	9.43	0.484
*02/17/82	550.	1100.	47.	5.25	13.97	0.681
02/19/82	650.	1675.	27.	4.40	9.67	0.562
02/22/82	700.	1325.	23.	4.97	8.92	0.459
*02/22/82	700.	1850.	41.	5.50	13.97	0.650
02/26/82	725.	1500.	28.	5.04	9.82	0.498
+02/26/82	725.	1825.	50.	5.06	11.86	0.599
03/01/82	725.	1350.	21.	5.04	8.88	0.451
*03/01/82	725.	1350.	34.	4.94	11.93	0.618
03/02/82	675.	850.	29.	4.85	9.90	0.522
03/03/82	625.	1600.	25.	4.98	8.53	0.438
03/05/82	675.	1650.	24.	5.04	9.00	0.457
+03/05/82	675.	1800.	35.	4.76	11.19	0.601
03/08/82	700.	1400.	20.	4.88	9.59	0.502
*03/08/82	700.	2000.	44.	4.81	11.86	0.630
03/10/82	725.	2700.	32.	4.92	8.61	0.447
+03/10/82	850.	2000.	62.	4.16	12.52	0.770
03/15/82	775.	1800.	35.	4.99	9.20	0.471
*03/15/82	700.	2000.	45.	4.88	11.23	0.589
03/17/82	650.	2100.	20.	4.87	8.94	0.464
03/18/82	490.	1050.	19.	4.88	8.84	0.463
+03/18/82	550.	1400.	33.	4.91	11.35	0.591

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

Appendix 2. Optimal Reclamation Plant Design from Phase 1.

THE LIMITING PARAMETER IS TDS FOR B(1)= 0.150

THE FOLLOWING WATER QUALITY RESULTS

	FILTER	RO	REQUIRED	BLENDED
TDS	1200.	250.0	500.0	500.0
TOC	18.	1.0	15.0	5.5
NTU	3.	0.5	2.0	1.2
TSS	4.	0.0	5.0	1.1

THE RATIO OF BLENDED RO PRODUCT WATER TO TOTAL PRODUCT FLOW = 0.737

ASSUMPTIONS

LABOR RATE = $ 12.00 PER HOUR

ELECTRICAL RATE = $ 0.05 PER KWH

INTEREST RATE = 8.00%

LIFE OF PROJECT = 20 YEARS

INFLATION RATIO = 1.000

PROJECT YEAR = 1979

COSTS PER KGALS FOR VARYING VALUES OF B AND THE CLEANING INTERVAL

	CLEANING INTERVAL(HOURS)									
B	8	16	24	32	40	48	56	64	72	80
0.15	1.60	1.57	1.59	1.61	1.64	1.67	1.69	1.72	1.74	1.76

Appendix 2. Continued

THE OPTIMAL DESIGN CONSISTS OF A CLARIFICATION/FILTRATION PRETREATMENT WITH FERRIC CHLORIDE ADDED AT 50. PPM

OPTIMAL SYSTEM SPECIFICATIONS

FILTER

INFLUENT FLOW:	1.83	MGD
LOADING RATE:	5.00	MGD
FILTER AREA:	253.	SQUARE FEET
DIAMETER:	18.	FEET
BACKWASH VELOCITY:	2.	FEET/MIN

CLARIFIER

INFLUENT FLOW:	1.83	MGD
LOADING RATE:	1000.00	MGD
CLARIFIER AREA:	1822.	SQUARE FEET
DIAMETER:	48.16	FEET
COAGULANT:	FERRIC CHLORIDE	

REVERSE OSMOSIS

INFLUENT FLOW:	1.56	MGD
PRODUCT FLOW:	0.74	MGD
PERCENT RECOVERY:	47.31	%
FLUX DECLINE INDEX (B):	0.15	
AVERAGE FLUX:	11.71	GPD/FT2
NUMBER OF MEMBRANES:	27359	
TOTAL AREA:	62927.18	SQUARE FEET
TIME REQUIRED FOR CLEANING:	2.0	HOURS
CLEANING INTERVAL:	16.0	HOURS
SULFURIC ACID INJECTED:	8.19	PPM
CHLORINE INJECTED:	2.00	PPM

Appendix 2. Continued

ENERGY AND LABOR ANALYSIS

	BUILDING ENERGY KWH/YR	COST($/YR)	PROCESS ENERGY KWH/YR	COST($/YR)	LABOR HRS/YR	$/YR
FILTER UNIT	41176.	2059.	62520.	3126.	1711.	20533.
SURFACE WASH	0.	0.	3904.	195.	72.	865.
MEDIA	0.	0.	0.	0.	0.	0.
BACKWASH	0.	0.	6066.	303.	197.	2367.
CLARIFIER UNIT	0.	0.	3844.	192.	173.	2076.
COAGULANT	2262.	113.	4900.	245.	308.	3701.
REVERSE OSMOSIS	30819.	1541.	3690893.	184545.	2001.	24008.
SULFURIC ACID	722.	36.	1630.	82.	80.	955.
CHLORINE	3175.	159.	396.	20.	348.	4171.
CLEANING	0.	0.	658.	33.	172.	2061.

UNIT COSTS

	TOTAL CAPITAL($/YR)	ANNUAL CAPITAL	O & M ($/YR)	WATER COSTS($/KGAL
VESSELS	308269.	31398.	26706.	0.17
SURFACE WASH	34017.	3465.	1214.	0.01
MEDIA	13204.	1345.	0.	0.00
BACKWASH	65133.	6634.	3425.	0.03
CLARIFIER	114782.	11691.	2662.	0.04
COAGULANTS	26155.	2664.	4228.	0.02
RO UNIT	1130297.	115124.	318825.	1.24
H2SO4	8979.	915.	1185.	0.01
CHLORINATION	6162.	628.	4684.	0.02
CLEANING	23102.	2353.	6178.	0.03

COAGULATION/CLARIFICATION/FILTRATION $/KGAL= 0.28
REVERSE OSMOSIS SYSTEM $/KGAL= 1.29
TOTAL SYSTEM $/KGAL= 1.57

THE PLANT PRODUCES 1 MGD PER DAY OF THE SPECIFIED QUALITY WATER
THE TOTAL ANNUAL COSTS = $ 533288. COST/KGAL = $1.57

Appendix 3. Computer Model Program Listing.

```
C                               MAIN
C        THIS PROGRAM IS DESIGNED TO GIVE THE OPTIMAL CONFIGURATION
C     FOR A REVERSE OSMOSIS SYSTEM GIVEN THE ALTERNATIVES OF DIRECT
C     FILTRATION OR COAGULATION/FILTRATION WITH VARIOUS COAGULANT AIDS.
C     VARIOUS VALUES OF THE FLUX DECLINE INDEX (B) WHICH CORRESPOND WITH
C     DIFFERING LEVELS OF TREATMENT-(DIRECT FILTRATION AND-COAGULATION/
C     FILTRATION WITH DIFFERENT COAGULANTS) HAVE BEEN DETERMINED FROM
C     FIVE YEARS OF OPERATING AT LAS GALINAS, CALIFORNIA.  THESE ARE
C     TESTED ALONG WITH DIFFERING CLEANING FREQUENCIES TO DETERMINE THE
C     LEAST COST ALTERNATIVE.
C
C
C                                 SUBROUTINES
C     BLEND -  CALCULATES THE 'RATIO' OF THE RO PRODUCT TO THE THE 1MGD
C              PRODUCT FLOW REQUIRED TO PRODUCE THE INPUT WATER QUALITY
C              DEFINED BY FOUR(CAN BE USED UP TO 10) QUALITY PARAMETERS.
C     FLUX  -  DETERMINES THE AVERAGE FLUX PER SQUARE FOOT/DAY (AFD) FROM
C              THE CLEANING FREQUENCY AND RECOVERY AFTER CLEANING.
C     CLEAN -  CALCULATES THE COST OF CLEANING FROM THE FEED RATE AND THE
C              DOSES OF THE H2SO4 AND CHLORINEC
C     SIZE  -  PRODUCES THE SIZES OF THE UNITS REQUIRED TO PRODUCE 1MGD
C              FROM THE GIVEN AVERAGE FLUX AND BLENDING RATIO.
C     COST  -  THE CENTRAL ROUTINE FOR CALCULATING THE COSTS OF THE VARIOUS
C              UNIT PROCESSES USING THE DATA FROM THE EPA COST ESTIMATING
C              TECHNIQUES, 1979.  THE ANNUAL COSTS ARE CALCULATED USING
C              8% FOR 20 YEARS FOR THE BASE YEAR 1979, AND ARE UPDATED
C              FOR SUCCESSIVE YEARS.
C
C     READER   ALL THE INPUT DATA EXCEPT ADDITIONAL YEARS
C              FOR MULTIPLE YEAR ANALYSIS.
C
C
C     PROCESS  CALLS THE VARIOUS PROCESS ROUTINES
C
C
C
C                                  VARIABLES
C      THE CAPITAL COSTS ARE NONDIMENSIONED VARIABLES STARTING WITH CC_
C      EXCEPT IN THE OUTPUT SUBROUTINE WHERE THE UNIT CAPITAL COSTS ARE
C      REPRESENTED BY ARRAY BN(I), AND ANNUAL CAPITAL COSTS BY BNN(I)
C      ENERGY CONSUMPTION IS REPRESENTED BY ARRAY E(I) AND THE ENERGY
C      COSTS ARE GIVEN IN ARRAY EC(I), WHERE PKWH IS THE PRICE/KWH
C      LABOR - GIVEN IN ARRAY AL(I) IN HRS/YR AND LABOR COSTS BY ALC(I)
C      MAINTENANCE MATERIAL COST - ARE CONTAINED IN ARRAY AM(I) IN $/YR.
C
C
       DIMENSION BQ(100),BX(100),C(100),Z(102)
       COMMON /RO/ CT,FLUXI(10),REC(10),AFD,AF,MA,RECOV,RFEED,CF,RF,
      1MK,NCSTR,NHRS
       COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,FF,V
       COMMON /CLAR/ CRATE,CIN,CSIZE
       COMMON /UNIT1/ BB(20),TABLE(10,500),K,KM,TCOST,DOSE,ADOSE,CDOSE
       COMMON /UNIT2/ CCPU,CCFS,CCFM,CCFB,CCC,CCA,CCP,CCB,CCD,CCL,BNN(10)
      1,PARMAX(2,10),PARIN(10,10),PAROUT(10,10),IG(10,3),IC(10)
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      COMMON /UNIT3/ E(20),AM(20),AL(20),EC(20),ALC(10),SUML,ESUM,ECSUM,
     1SUMCL,SUMM,PKWH,HOURLY,AOPT(2,10),BOPT(2,10),AI,FINFL,IYP
      COMMON /OPTIM/ CFOPT,TMAX,TMAX1,IWRITE,KBOPT,IOPT
      REAL MA
      DATA A/1000000./,CFE/8./
C NHRS = NUMBER OF CLEANING FREQUENCIES(CF) TO TO BE USED
C MK = NUMBER OF VALUES OF FLUX DECLINE INDEX TO BE USED
C CT = TIME REQUIRED FOR CLEANING
C FLUXI & REC = THE FLOW AND THE RECOVERY AFTER CLEANING
C FRATE AND CRATE ARE THE LOADING RATES FOR THE FILTER AND CLARIFIER
C V = UPFLOW VELOCITY (FT/MIN) FOR THE FILTER BACKWASH
C SET IWRITE = 0 FOR TERSE OUTPUT
      IWRITE = 1
C.. CALL SUBROUTINE READER WHICH READS IN ALL THE INPUT DATA
      CALL READER
C.. SET THE COUNTERS FOR OPTIMIZATION
C.. TMAX IS SET TO A LARGE VALUE IS COMPARED TO SYSTEM COST TO RETAIN
C   THE LOWEST OR OPTIMAL COST
C.. IOPT IS SET TO 1 DURING THE LOOPS BUT IS SETTO 2 FOR THE FINAL
C   SUMMARY
1     IOPT = 1
      TMAX = 9999999.
      CF = CFE
      DO 50 KM = 1,MK
      CALL BLEND
      TMAX1 = 99999999.
      DO 40 J = 1,NHRS
C
      CALL PROCES
      TABLE(KM,J) = TCOST
      C(J) = CF
      CF = CF + CFE
   40 CONTINUE
      IF(IWRITE.EQ.0) GO TO 45
      CF=CFOPT
      CALL OUTPUT(CFE)
   45 CF = CFE
   50 CONTINUE
      IF(IWRITE.NE.0) GO TO 100
C  COMPUTE OPTIMAL VALUES
      IOPT=2
      KM = KBOPT
      CF = CFOPT
      CALL PROCES
      CALL OUTPUT(CFE)
100   READ(5,1000,END=110) IYP
1000  FORMAT(I4)
      CALL UPYEAR
      GOTO 1
110   STOP
      END
```

```
      BLOCK DATA
      COMMON /RO/ CT,FLUXI(10),REC(10),AFD,AF,MA,RECOV,RFEED,CF,RF,
     1MK,NCSTR,NHRS
      COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,FF,V --
      COMMON /CLAR/ CRATE,CIN,CSIZE
      COMMON /UNIT1/ BB(20),TABLE(10,500),K,KM,TCOST,DOSE,ADOSE,CDOSE
      COMMON /UNIT2/ CCFU,CCFS,CCFM,CCFB,CCC,CCA,CCR,CCB,CCD,CCL,BNN(10)
     1,PARMAX(2,10),PARIN(10,10),PAROUT(10,10),IG(10,3),IC(10)
      COMMON /UNIT3/ E(20),AM(20),AL(20),EC(20),ALC(10),SUML,ESUM,ECSUM,
     1SUMCL,SUMM,PKWH,HOURLY,AOPT(2,10),BOPT(2,10),AI,FINFL,IYP
      COMMON /OPTIM/ CFOPT,TMAX,TMAX1,IWRITE,KBOPT,IOPT
      REAL MA
      DATA NHRS/10/,CT/2./,FRATE/5./,CRATE/1000./,V/2./
      END


      SUBROUTINE PROCES
C..THIS SUBROUTINE CALLS THE CLEAN, FLUX, SIZE, AND COST
C  ROUTINES AND INITIALIZES THE ARRAYS PRIOR TO CALLING
      DIMENSION ANN(10)
      COMMON /UNIT2/ CCFU,CCFS,CCFM,CCFB,CCC,CCA,CCR,CCB,CCD,CCL,BNN(10)
     1,PARMAX(2,10),PARIN(10,10),PAROUT(10,10),IG(10,3),IC(10)
      COMMON /UNIT3/ E(20),AM(20),AL(20),EC(20),ALC(10),SUML,ESUM,ECSUM,
     1SUMCL,SUMM,PKWH,HOURLY,AOPT(2,10),BOPT(2,10),AI,FINFL,IYP
      EQUIVALENCE (CCFU,ANN(1))
C  INITIALIZATION
      DO 10 I = 1,20
      E(I) = 0.
      AM(I) = 0.
10    AL(I) = 0.
      DO 20 I=1,10
20    ANN(I)=0.
      CALL FLUX
      CALL SIZE
      CALL CLEAN
      CALL COST
      RETURN
      END
```

```
      SUBROUTINE CLEAN
C
C                        SUBROUTINE CLEAN
C
      COMMON /RO/ CT,FLUXI(10),REC(10),AFD,AF,MA,RECOV,RFEED,CF,RF,
     1MK,NCSTR,NHRS
      COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,FF,V
      COMMON /CLAR/ CRATE,CIN,CSIZE
      COMMON /UNIT1/ BB(20),TABLE(10,500),K,KM,TCOST,DOSE,ADOSE,CDOSE
      COMMON /UNIT2/ CCFU,CCFS,CCFM,CCFB,CCC,CCA,CCR,CCB,CCD,CCL,BNN(10)
     1,PARMAX(2,10),PARIN(10,10),PAROUT(10,10),IG(10,3),IC(10)
      COMMON /UNIT3/ E(20),AM(20),AL(20),EC(20),ALC(10),SUML,ESUM,ECSUM,
     1SUMCL,SUMM,PKWH,HOURLY,AOPT(2,10),BOPT(2,10),AI,FINFL,IYP
      COMMON /OPTIM/ CFOPT,TMAX,TMAX1,IWRITE,KBOPT,IOPT
C
C        CFACT = THE SIZE OF REQUIRED TANKS FOR WASH WATER
C        ACETIC = ACETIC ACID DOSE/MEMBRANE PER CLEANING
C        BIZ = BIZ DOSE/MEMBRANE PER CLEANING
C        QM = COST OF CLEANING CHEMICALS/YEAR
C        CT/CF = CLEANING TIME/CLEANING FREQUENCY
C        UPTIME = PROPORTION OF TIME SYSTEM IS OPERATING
C
      REAL MA
      DATA UPTIME/0.8/,PA/0.05/,PB/0.02/,ACETIC/.00344/,BIZ/0.00689/
      TM = RFEED/(AFD*2.3)
      QM = (ACETIC*PA + BIZ*PB) *TM *CF/24.*340.
      Q = 0.03125 * TM * CT/((CF +CT)*UPTIME)
      CFACT = Q * CT * 60.
      CCL = 10. **(0.2758 *ALOG10(CFACT) + 3.31)
      E(19) = 0.0
      E(20) = 10. **(1.00 * ALOG10(Q) + 1.077)
      AL(10)= 10. **(0.062 *ALOG10(Q) + 2.127)
      AM(10)= 10. **(0.281 *ALOG10(Q) + 2.157)     + QM
      FIN = FIN + CFACT
      RETURN
      END
```

Appendix 3. Continued

```
      SUBROUTINE FLUX
C
C                         SUBROUTINE FLUX
C
C FSUM IS THE TOTAL FLOW BETWEEN CLEANINGS
C CF = THE CLEANING FREQUENCY
C CT = CLEANING TIME
C FLUXI = FLUX AFTER CLEANING
C REC = THE RECOVERY AFTER CLEANING
C FSUM/TOTAL TIME = AVERAGE FLUX
      DIMENSION F(500),T(500),Z(102),AAF(500)
      COMMON /RO/ CT,FLUXI(10),REC(10),AFD,AF,MA,RECOV,RFEED,CF,RF,
     1MK,NCSTR,NHRS
      COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,FF,V
      COMMON /CLAR/ CRATE,CIN,CSIZE
      COMMON /UNIT1/ BB(20),TABLE(10,500),K,KM,TCOST,DOSE,ADOSE,CDOSE
      COMMON /UNIT2/ CCFU,CCFS,CCFM,CCFB,CCC,CCA,CCR,CCB,CCD,CCL,BNN(10)
     1,PARMAX(2,10),PARIN(10,10),PAROUT(10,10),IG(10,3),IC(10)
      REAL MA
      FMAX = 0.0
      FSUM = 0.
      KK = CF+0.5
      F(1) = FLUXI(KM)
      N = KK +1
      DO 10 I = 2,N
      F(I) = FLUXI(KM) * (1./FLOAT(I)**BB(KM))
      FSUM = FSUM +(F(I) + F(I-1))/2.
   10 CONTINUE
      AF = FSUM/(KK + CT)
      RECOV = AF * REC(KM)/FLUXI(KM)
      AFD = AF
      CF = FLOAT (KK)
      RETURN
      END
```

```
      SUBROUTINE BLEND                                          --
C
C                         SUBROUTINE BLEND
C
C
      COMMON /RO/ CT,FLUXI(10),REC(10),AFD,AF,MA,RECOV,RFEED,CF,RF,
     1MK,NCSTR,NHRS
      COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,FF,V
      COMMON /CLAR/ CRATE,CIN,CSIZE
      COMMON /UNIT1/ BB(20),TABLE(10,500),K,KM,TCOST,DOSE,ADOSE,CDOSE
      COMMON /UNIT2/ CCFU,CCFS,CCFM,CCFB,CCC,CCA,CCR,CCB,CCD,CCL,BNN(10)
     1,PARMAX(2,10),PARIN(10,10),PAROUT(10,10),IG(10,3),IC(10)
      COMMON /OPTIM/ CFOPT,TMAX,IMAX1,IWRITE,KBOPT,IOPT
      REAL MA
      RATIOM = .0
      KK = 1
C  LOOP FOR CALCULATING THE BLENDING RATIO. FIRST FIND THE CONTROLLING
C  WATER QUALITY PARAMETER
      DO 10 K=1,NCSTR
C.. CHECK TO SEE IF BLENDING IS REQUIRED. BYPASS RATIO CALCULATIONS
C   IF BLENDING IS NOT REQUIRED.
      IF(PARIN(KM,K).LE.PARMAX(1,K)) GOTO 10
      RATIO = (PARMAX(1,K) - PARIN(KM,K))/(PAROUT(KM,K) - PARIN(KM,K))
      IF(RATIO.LT.RATIOM) GO TO 10
      RATIOM = RATIO
      KK = K
10    CONTINUE
C PRINT-OUT OF BLENDED RATIOS
      RATIO = RATIOM
      DO  20 I = 1,NCSTR
20    PARMAX(2,I) = (RATIO*PAROUT(KM,I)) + ((1.-RATIO) * PARIN(KM,I))
      WRITE(6,1000) IC(KK),KM,BB(KM),(IC(K),PARIN(KM,K),PAROUT(KM,K)
     1,(PARMAX(I,K),I=1,2),K=1,NCSTR)
1000  FORMAT(///,T20,'THE LIMITING PARAMETER IS ',A3,' FOR B(',I1,')=',
     1F6.3,///,T20,'THE FOLLOWING WATER QUALITY RESULTS',/,T20,
     2'FILTER',T30,'RO ',T40,'REQUIRED',T50,'BLENDED',
     34(/,T10,A3,T19,F8.1,T26,F8.1,T39,F8.1,T50,F8.1))
      WRITE(6,1010) RATIO
1010  FORMAT(//,' THE RATIO OF BLENDED RO PRODUCT WATER TO TOTAL',
     1' PRODUCT FLOW = ',F5.3)
      RETURN
      END
```

```
      SUBROUTINE READER
C.. THIS SUBROUTINE READS THE INPUT DATA THEN WRITES OUT THE STARTING
C.. VALUES OF B, INITIAL FLUXES, AND TREATMENT OBJECTIVES
      COMMON /RO/ CT,FLUXI(10),REC(10),AFD,AF,MA,RECOV,RFEED,CF,RF,
     1MK,NCSTR,NHRS
      COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,FF,V
      COMMON /CLAR/ CRATE,CIN,CSIZE
      COMMON /UNIT1/ BB(20),TABLE(10,500),K,KM,TCOST,DOSE,ADOSE,CDOSE
      COMMON /UNIT2/ CCFU,CCFS,CCFM,CCFB,CCC,CCA,CCR,CCB,CCD,CCL,BNN(10)
     1,PARMAX(2,10),PARIN(10,10),PAROUT(10,10),IG(10,3),IC(10)
      COMMON /UNIT3/ E(20),AM(20),AL(20),EC(20),ALC(10),SUML,FSUM,ECSUM,
     1SUMCL,SUMM,PKWH,HOURLY,AOPT(2,10),BOPT(2,10),AI,FINFL,IYP
      COMMON /OPTIM/ CFOPT,TMAX,TMAX1,IWRITE,KBOPT,IOPT
      DIMENSION AIA(5),IYEAR(5),FINFLA(5),PKWHA(5)
      REAL MA,LABORC(5)
      DATA IYEARB/1979/
C.. READ THE NUMBER OF FLUX DECLINE ALTERNATIVES
      READ(5,1000) MK
1000  FORMAT(I2)
C.. WRITE OUT THE NUMBER OF ALTERNATIVES
      WRITE(6,1010) MK
1010  FORMAT('1 THE NUMBER OF ALTERNATIVES IS ',I2)
C.. READ THE NUMBER OF FINISH WATER QUALITY CONSTRAINTS
      READ(5,1000) NCSTR
C.. WRITE OUT THE NUMBER OF CONSTRAINTS
      WRITE(6,1020) NCSTR
1020  FORMAT('0THE NUMBER OF WATER QUALITY PARAMETERS IS  ',I2)
C.. READ AND WRITE THE PARAMETER AND ITS MAXIMUM VALUE
      WRITE(6,1030)
1030  FORMAT(T21,'WATER QUALITY OBJECTIVES',//,' PARAMETER',
     1T15,'MAXIMUM VALUE')
      DO 10 I=1,NCSTR
      READ(5,1040) IC(I),PARMAX(1,I)
1040  FORMAT(A4,6X,F10.0)
10    WRITE(6,1050) IC(I),PARMAX(1,I)
1050  FORMAT(T6,A4,T18,F10.2)
C.. READ THE INPUT DATA FOR EACH ALTERNATIVE
      LOOP=0
20    LOOP=LOOP+1
      IF(LOOP-MK) 30,30,60
C.. READ THE B VALUES, INITIAL FLUXES AND RECOVERIES, AND WATER QUALITY
C   OBJECTIVES.
C.. CHECK TO SEE IF THE REQUIRED VALUES ARE SUPPLIED VALUES ARE
C   CONSISTENT WITH THE MAXIMUM VALUES
30    READ(5,1060) BB(LOOP),REC(LOOP),FLUXI(LOOP)
1060  FORMAT(8F10.0)
      READ(5,1060) (PARIN(LOOP,J),J=1,NCSTR)
      READ(5,1060)  (PAROUT(LOOP,J),J=1,NCSTR)
C.. CHECK TO SEE IF THE RO PRODUCT WATER QUALITY EXCEED THE REQUIRED
      DO 40 J=1,NCSTR
      IF(PAROUT(LOOP,J).LE.PARMAX(1,J)) GOTO 50
40    CONTINUE
C.. AN UNFEASIBLE ALTERNATIVE HAS BEEN SELECTED.  WRITE OUT AN ERROR
C.. MESSAGE AND STOP
```

```
      WRITE(6,1080) J,IC(J),PARMAX(1,J),PAROUT(LOOP,J)
1080  FORMAT(' AN UNFEASIBLE ALTERNATIVE HAS BEEN SPECIFIED FOR ',
     1'OPTION NO. ',I2,//,' FOR PARAMETER ',A4,' AN OUTPUT VALUE OF '
     2,F10.2,' IS REQUIRED, BUT THE RO PRODUCT WATER IS ONLY',F10.2)
      STOP
C.. END THE LOOP
50    GOTO 20
C.. WRITE OUT THE INPUT DATA
60    WRITE(6,1100) (IC(I),I=1,NCSTR)
1100  FORMAT(///,' NO.',T5,'B VALUE',' INITIAL',2X,'INITIAL',10(4X,A4))
      WRITE(6,1110)
1110  FORMAT(1X,T13,'RECOVERY  FLUX')
      DO 80 I=1,MK
      WRITE(6,1120) I,BB(I),REC(I),FLUXI(I),(PARIN(I,J),J=1,NCSTR)
1120  FORMAT(1X,I2,F7.3,F8.3,1X,F7.2,1X,10(1X,F7.2))
80    WRITE(6,1130) (PAROUT(I,J),J=1,NCSTR)
1130  FORMAT(' OUTPUT VALUES',T28,10(1X,F7.2))
C.. READ IN THE HEADINGS AND TITLES USED IN THE OUTPUT ROUTINES
      READ(5,1150) ((IG(J,I),I=1,3), J=1,10)
1150  FORMAT(9(3A4,/),3A4)
C.. READ THE INFLATION, HOURLY COSTS, POWER COSTS, AND
C   ELECTRICITY COST FOR EACH YEAR.
      READ(5,1000) NYEARS
      IYEARM=IYEARB+NYEARS-1
      DO 90 I=1,NYEARS
      READ(5,1160) IYEAR(I),FINFLA(I),LABORC(I),AIA(I),PKWHA(I)
1160  FORMAT(I4,6X,4F10.0)
90    IYEAR(I)=(IYEAR(I)-IYEARB)+1
C.. READ THE FIRST STUDY YEAR
      READ(5,1170) IYP
1170  FORMAT(I4)
      ENTRY UPYEAR
      IF ((IYP.GE.IYEARB).AND. (IYP.LE.IYEARM)) GOTO 100
      WRITE(6,1180) IYEARB,IYP,IYEARM
1180  FORMAT(////,' THE YEAR SPECIFIED IS OUTSIDE THE POSSIBLE RANGE',
     1//,' THE YEAR MUST BE BETWEEN ',I4,' AND ',I4,//,1X,I4,' WAS ',
     2 'SPECIFIED')
      STOP
100   IY=(IYP-IYEARB)+1
      AI=AIA(IY)
      HOURLY=LABORC(IY)
      PKWH=PKWHA(IY)
      FINFL=FINFLA(IY)
      RETURN
      END
```

```
      SUBROUTINE SIZE
C
C                           SUBROUTINE SIZE
C
C     CALCULATES THE SIZES OF THE VARIOUS COMPONENTS OF THE SYSTEM
      COMMON /RO/ CT,FLUXI(10),REC(10),AFD,AF,MA,RECOV,RFEED,CF,RF,
     1MK,NCSTR,NHRS
      COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,FF,V
      COMMON /CLAR/ CRATE,CIN,CSIZE
      COMMON /UNIT2/ CCFU,CCFS,CCFM,CCFB,CCC,CCA,CCR,CCB,CCD,CCL,BNN(10)
     1,PARMAX(2,10),PARIN(10,10),PAROUT(10,10),IG(10,3),IC(10)
      REAL MA
      A = 10.0**6
C BN = NUMBER OF BACKWASHES PER DAY
      BN = 2
C MEMBRANE AREA          AFD = AVERAGE FLUX/FT2-DAY    MA
      MA = RATIO * A/AFD
C FEED FLOW RATE TO RO                                  RFEED
      RFEED = RATIO * A/RECOV
C FILTER OUTPUT                                          FOUT
      FOUT = RFEED + (1-RATIO) *A
C FILTER INPUT = FOUT + BACKWASH WATER (2%)              FIN
      FIN = FOUT
C FILTER SURFACE AREA AT 5 GPM/FT2                      FSIZE
      FSIZE = FIN /(FRATE * 1440.)
      FIN = FIN + BN*V*FSIZE
C CLARIFIER AT 1000 GPD/FT2                             CSIZE
      CSIZE = FIN/1000.
      RETURN
      END
```

```
      SUBROUTINE COST
C
C                             SUBROUTINE COST
C
      COMMON /RO/ CT,FLUXI(10),REC(10),AFD,AF,MA,RECOV,RFEED,CF,RF,
     1MK,NCSTR,NHRS
      COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,FF,V
      COMMON /CLAR/ CRATE,CIN,CSIZE
      COMMON /UNIT1/ BB(20),TABLE(10,500),K,KM,TCOST,DOSE,ADOSE,CDOSE
      COMMON /UNIT2/ CCFU,CCFS,CCFM,CCFB,CCC,CCA,CCR,CCB,CCD,CCL,BNN(10)
     1,PARMAX(2,10),PARIN(10,10),PAROUT(10,10),IG(10,3),IC(10)
      COMMON /UNIT3/ E(20),AM(20),AL(20),EC(20),ALC(10),SUML,ESUM,ECSUM,
     1SUMCL,SUMM,PKWH,HOURLY,AOPT(2,10),BOPT(2,10),AI,FINFL,IYP
      COMMON /OPTIM/ CFOPT,TMAX,TMAX1,IWRITE,KBOPT,IOPT
      DIMENSION ANN(10)
      EQUIVALENCE (CCFU,ANN(1))
      REAL MA
      DATA N/20/,CFT/20./
      RFT = CFT/102.
      A = 10.**6
C FF = MGD TO FILTER,  WF = MILLIONS OF GALLONS TO FILTER, RF=MGD TO RO
      FF = FIN/A
      WF = FF * 8.34
      RF = RFEED/A
C.. INSERT TWO INFLATION FACTORS, ONE FOR MAINTENANCE AND ONE
C   FOR LABOR, TO BE USED LATER BUT FOR THE PRESENT SET TO
C   THE SINGLE FACTOR, FINFL
      FINFLM=FINFL
      FINFLC=FINFL
C***********************************************************
C*                            FILTER                       *
C***********************************************************
C CALCULATE THE COST OF  FILTRATION
C CCF = CAPITAL COST FOR FILTER (EPA,1979)
C
C                            FILTER UNIT
C -----------
      ECOST = 0.32 * ALOG10(FSIZE) + 4.72
      CCFU = FINFLC * 10.**ECOST
        E(1)    = RFT * 210000.
        E(2)    = 10.  **(0.968 * ALOG10(FSIZE) + 2.47)
       AM(1)    = FINFLM * 10. **(0.785 * ALOG10(FSIZE) + 1.427)
       AL(1)    = 10.  **(0.301 * ALOG10(FSIZE) + 2.51)
C
C                             SURFACE WASH
C--------------
      ECOST = 0.24 * ALOG10(FSIZE) + 3.955
      CCFS = FINFLC * 10. ** ECOST
        E(3)    = 0.0
        E(4)    = 10.  **(0.8877* ALOG10(FSIZE) + 1.4585)
       AM(2)    = FINFLM * 10. **(0.139 * ALOG10(FSIZE) + 2.0)
       AL(2)    = 10.  **(0.486 * ALOG10(FSIZE) + 0.69)
C
C                                    MEDIA
```

```
C---------------
      CCFM = FINFLC * 10. **(0.652 * ALOG10(FSIZE) + 2.554)
        E(5) = 0.
        E(6) = 0.
       AM(3) = FINFLM * 0.
       AL(3) = 0.
C
C                                  BACKWASH
C-----------------
      Q = FSIZE *  V * 7.48
      CCFB = FINFLC * 10. **(0.37 * ALOG10(Q) + 3.49)
        E(7)    = 0.0
        E(8)    = 10. **(1.00 * ALOG10(FSIZE) + 1.38)
       AL(4)    = 10. **(0.062 * ALOG10(FSIZE) + 2.146)
       AM(4)    = FINFLM * 10. **(0.281 * ALOG10(FSIZE) + 2.24)
C
C                    TOTAL CAPITAL COST FOR FILTER (CCF)
C-------------------
      CCF = CCFU + CCFS + CCFM + CCFB
C IF PRETREATMENT INCLUDES MORE THAN FILTRATION GO TO 10
C*************************************************************
C*                            CLARIFIER                      *
C*************************************************************
C OVERFLOW = 1000. GPD/FT2
C CCC = CLARIFIER CAPITAL COSTS
C CCA = CAPITAL COSTS FOR CHEMICAL FEED SYSTEMS
C
C                   CLARIFIER UNIT
C-------------
C.. BY PASS THE COST CALCULATIONS FOR THE OPTIONS WITHOUT
C   COAGULATION KM= 4 OR 6
      IF ((KM.EQ.4).OR.(KM.EQ.6)) GOTO 15
   10 ECOST = 0.322 * ALOG10(CSIZE) + 4.01
      CCC = FINFLC * 10. ** ECOST
       E(9) = 0.0
       E(10)= 10. **(0.172 * ALOG10(CSIZE) + 3.024)
      AL(5) = 10. **(0.154 * ALOG10(CSIZE) + 1.736)
      AM(5) = FINFLM * 10. **(0.640 * ALOG10(CSIZE) + 0.574)
C
C                             COAGULANTS
C            BRANCH TO THE PROPER CHEMICAL FEED
C
15    GO TO (20,25,30,35,20,35,30,35),KM
C
C                        NALCO 7134 @ 10PPM
C-------------------
   25 CCA = FINFLC * 20200.
      DOSE =  5
      PRICE = 1.50
      AMT = WF * DOSE * PRICE*340.
        E(11)= RFT * 8210.
        E(12)= 17300.
       AL(6) = 198.
       AM(6) = FINFLM * 270. + AMT
```

```
      K = 2
      GO TO 40
C
C                         FECL3 @ 50 PPM
C-------------------
   20 DOSE = 50
      AMT = WF * DOSE/24.
      PRICE = .10
      CCA = FINFLC * 10. **(0.278 * ALOG10(AMT) + 4.00)
       E(11) = RFT * (10. **(.574 * ALOG10(AMT) + 3.20))
       E(12) = 4900.
       AL(6) = 10. **(0.062 * ALOG10(AMT) + 2.396)
       AM(6) = 10. **(0.067 * ALOG10(AMT) + 2.186) + PRICE*AMT
      K = 3
      GO TO 40
C
C                         ALUM @ 60 PPM
C-------------------
   30 DOSE = 60
      CONC1 = .5
      PRICE = .05
      AMT = WF * DOSE/24.
      AMTC = AMT/CONC1
      CCA = FINFLC * 10. ** (0.232 * ALOG10(AMTC) + 4.08)
       E(11) = RFT * (10. **(0.574 * ALOG10(AMT) + 3.216))
       E(12) = 4900.
       AL(6) = 10. **(0.062 * ALOG10(AMT) + 3.97)
       AM(6) = FINFLM * (200. + PRICE * AMT)
      K = 4
      GO TO 40
   35 K=1
C                         NO CHEMICALS
C   CHEMICAL COSTS HAVE BEEN INITIALIZED TO ZERO IN MAIN ROUTINE
C
C*********************************************************************
C*                        REVERSE OSMOSIS                           *
C*********************************************************************
C CCR = CAPITAL COSTS
C
   40 CCR = FINFLC * 10. **(0.848 * ALOG10(RF) + 5.89)
      E(13) = RFT * (10. **(0.901 * ALOG10(RF) + 5.023))
      E(14) = 10. **(0.962 * ALOG10(RF) + 6.382)
      AL(7) = 10. **(0.188 * ALOG10(RF) + 3.265)
      AM(7) = FINFLM * 10. **(0.886 * ALOG10(RF) + 4.988)
C
C                    PRETREATMENT CHEMICALS
C
C                      SULFURIC ACID
C-------------------
      CONC = 0.96
      PRIC1 = 7.40
      ADOSE = 15.0/1.8318
      AMT = ADOSE * RF/CONC
      CCB = FINFLC * 10. **(0.1186 * ALOG10(AMT) + 3.82)
```

```
         E(15)  = RFT * 3680.
         E(16)  = 1630.
         AL(8)  = 10.  **(0.330 * ALOG10(AMT) + 1.53)
         AM(8)  = FINFLM*(10.  **(0.222 * ALOG10(AMT)+1.56)+PRIC1*AMT)
C
C                        CHLORINE
C-------------------
      IF(K.GT.1) CDOSE = 2.
      IF(K.LE.1) CDOSE = 9.
      PRIC2 = 0.25
      AMT2 = WF * CDOSE/24.
      CCD = FINFLC * 10. **(0.3625 * ALOG10(AMT2) + 3.752)
       E(17)  = 10.  **(0.517 * ALOG10(AMT2) + 3.448)
       E(18)  = 10.  **(0.173 * ALOG10(AMT2) + 2.58)
      AL(9)   = 10.  **(0.1066 * ALOG10(AMT2) + 2.53)
      AM(9)   = FINFLM*(10.**(0.177*ALOG10(AMT2)+2.570)+PRIC2*AMT2)
C
C**************************************************************
C*************   TOTAL COSTS   ********************************
C**************************************************************
C
      SUML = 0.
      ESUM = 0.
      SUMM = 0.
      ECSUM = 0.
      SUMCL = 0.
      DO 50 I = 1,20
      EC(I)  = E(I)  * PKWH
      ESUM = ESUM + E(I)
      ECSUM = ECSUM + EC(I)
   50 CONTINUE
      DO 60 J = 1,10
      ALC(J)  = AL(J)  * HOURLY
      SUML = SUML + AL(J)
      SUMCL = SUMCL + ALC(J)
      SUMM = SUMM + AM(J)
   60 CONTINUE
C                   TOTAL CAPITAL(TCAP) AND O&M (TOM) COSTS
C-------------------
      TCAP = CCF + CCC + CCA + CCF + CCB + CCD  +CCL
      TOM =   SUMM + SUMCL + ECSUM
C
C                     ANNUAL CAPITAL COSTS
C-------------------
      AT = (1 + AI)**N
      F    = (AI * AT)/(AT - 1)
      DO 70 I = 1,10
      BNN(I)  = ANN(I)  * F
70    CONTINUE
      ACAP = TCAP * F
      TCOST = TOM + ACAP
      IF(TCOST.GT.TMAX1) GOTO 78
C.. THIS BRANCH IS TAKEN IF THE PRESENT CLEANING FREQUENCY GIVES
C   A LOWER TOTAL COST
```

```
      DO 75 I = 1,10
      BOPT(1,I) = EC(2*I-1) +EC(2*I) + ALC(I) + AM(I)
75    AOPT(1,I) =(BNN(I)+BOPT(1,I))/340000.
C                    SELECT LEAST TOTAL ANNUAL COST
C-------------------
C
      TMAX1=TCOST
78    IF(TCOST.LT.TMAX) GO TO 80
      RETURN
C.. THIS BRANCH IS TAKEN IF THE LOWEST COST FOR THE OPTIMAL CLEANING
C   FREQUENCY FOR THIS B VALUE IS THE LOWEST SO FAR IN THE ANALYSIS
C   TMAX IS THE GLOBAL OPTIMUM
80    KBOPT = KM
      CFOPT = CF
      TMAX = TCOST
      DO 90 I = 1,10
      BOPT(2,I) = EC(2*I-1) +EC(2*I) + ALC(I) + AM(I)
90    AOPT(2,I) =(BNN(I)+BOPT(2,I))/340000.
      RETURN
      END
```

```
      SUBROUTINE OUTPUT(CFE)
C
C                 SUBROUTINE OUTPUT
C
      COMMON /OPTIM/ CFOPT,TMAX,TMAX1,IWRITE,KBOPT,IOPT
      COMMON /RO/ CT,FLUXI(10),REC(10),AFD,AF,MA,RECOV,RFEED,CF,RF,
     1MK,NCSTR,NHRS
      COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,PF,V
      COMMON /CLAR/ CRATE,CIN,CSIZE
      COMMON /UNIT1/ BB(20),TABLE(10,500),K,KM,TCOST,DOSE,ADOSE,CDOSE
      COMMON /UNIT2/ BN(10),BNN(10)
     1,PARMAX(2,10),PARIN(10,10),PAROUT(10,10),IG(10,3),IC(10)
      COMMON /UNIT3/ E(20),AM(20),AL(20),EC(20),ALC(10),SUML,ESUM,ECSUM,
     1SUMCL,SUMM,PKWH,HOURLY,AOPT(2,10),BOPT(2,10),AI,FINFL,IYP
      DIMENSION AA(4,3)
      DATA AA/'    ','  NA','LCO ','7134',' FER','RIC ','CHLO','RIDE',
     1'    ','    ','    ','ALUM'/
      REAL MA
C  READ IN THE NAMES OF THE UNIT PROCESSES; COLS 2-17= PRESSURE FILTER
C  COLS 19-34= CLARIFIER, COLS 36-51= CHEMICAL FEED, COLS 53-68 =
C  REVERSE OSMOSIS
C-----------------------ASSUMPTIONS--------------------------------
      PAI=100*AI
      WRITE(6,220) HOURLY,PKWH,PAI,FINFL,IYP
  220 FORMAT(///,T40,'ASSUMPTIONS',/,T30,'LABOR RATE = $',F6.2,' PER HOU
     1R',//,T25,'ELECTRICAL RATE = $',F6.2,' PER KWH',//,T26,'INTEREST
     2RATE = ',F7.2,'%',//,T25,'LIFE OF PROJECT =      20 YEARS',//,
     3T25,'INFLATION RATIO =',F10.3,//,T28,'PROJECT YEAR = ',I9)
C*************************************************************************
C          TABLE OF ANNUAL COSTS AND FLUX DECLINE INDEX
C
C-----------------------TABLE HEADINGS----------------------
C     WRITE(6,10) (I,I=1,40,4)
C  10 FORMAT(/////,T20,'TOTAL ANNUAL COSTS FOR VARYING VALUES OF THE FLU
C    1X DECLINE INDEX (B) AND THE CLEANING INTERVAL',//,T50,
C    2'CLEANING INTERVAL(HOURS)',/,T3,'B',T5,10I10)
C
C-----------------------B VALUES AND TOTAL ANNUAL COSTS----------
C     WRITE(6,20) (BB(I),(TABLE(I,J),J=1,10),I=1,MK)
C  20 FORMAT(/,(F5.2,T10,10(F9.0,1X)))
C
C-----------------------B VALUES AND COSTS PER THOUSAND GALLONS-----
      DO 30 J = 1,NHRS
      TABLE(KM,J) = TABLE(KM,J)/340000.
   30 CONTINUE
      ICF=CFE+0.5
      ICFT=ICF*NHRS
      WRITE(6,40) (I,I=ICF,ICFT,ICF)
   40 FORMAT(////,T10,'COSTS PER KGALS FOR VARYING VALUES OF B AND THE C
     1LEANING INTERVAL',//,T30,'CLEANING INTERVAL(HOURS)',/T8,'B',T10,
     210I6)
      WRITE(6,50) BB(KM),(TABLE(KM,J),J=1,10)
   50 FORMAT(/,T5,F5.2,T10,10F6.2)
C*********************************************************************
```

Appendix 3. Continued

```
C*********             OPTIMAL SYSTEM          **********************
C*********                                     **********************
C***************************************************************************
C---BRANCH TO OPTIMAL DESIGN -----
C
      GO TO (60,80,100,120,100,60,120),K
C
   60 WRITE(6,70)
   70 FORMAT('1',/,T10,'THE OPTIMAL DESIGN REQUIRES FILTRATION ONLY')
      GO TO 140
C
   80 WRITE(6,90) DOSE
   90 FORMAT('1',//,T10,'THE OPTIMAL DESIGN CONSISTS OF A CLARIFICATION/
     CFILTRATION PRETREATMENT',/T20,'WITH NALCO 7134 POLMER ADDED AT ',F
     C4.0,' PPM')
      GO TO 140
C
  100 WRITE(6,110) DOSE
  110 FORMAT('1',//,T10,'THE OPTIMAL DESIGN CONSISTS OF A CLARIFICATION/
     1FILTRATION PRETREATMENT',/T20,'WITH FERRIC CHLORIDE ADDED AT ',F4.
     20,' PPM')
      GO TO 140
C
  120 WRITE(6,130) DOSE
  130 FORMAT('1',///,T10,'THE OPTIMAL DESIGN CONSISTS OF A CLARIFICATION
     C/FILTRATION PRETREATMENT',/T20,'WITH ALUM ADDED AT ',F4.0,' PPM')
C***************************************************************************
C                         FILTER
C
  140 D = (4.*FSIZE/3.1418)**0.5
C------------------------FILTER SPECIFICATIONS------------------
      WRITE(6,150) FF, FRATE, FSIZE,D,V
  150 FORMAT(///,T40,'OPTIMAL SYSTEM SPECIFICATIONS',//,T10,'FILTER',/,
     1T26,'INFLUENT FLOW:',F12.2,T54,'MGD',/
     2T27,'LOADING RATE:',F12.2,T54,'MGD',/
     3T28,'FILTER AREA:',F10.0,T54,'SQUARE FEET',/
     4T31,'DIAMETER:',F10.0,T54,'FEET',/
     5T22,'BACKWASH VELOCITY:',F10.0,T54,'FEET/MIN')
      IF(K.EQ.1) GO TO 170
C
C                         CLARIFIER
C
      D2 = (4.*CSIZE/3.1418)**0.5
      JJ = K-1
C---------------CLARIFIER SPECIFICATIONS-------------------------
      WRITE(6,160) FF,CRATE,CSIZE,D2,(AA(I,JJ),I = 1,4)
  160 FORMAT(//,T10,'CLARIFIER',/,
     1T26,'INFLUENT FLOW:',F12.2,T54,'MGD',/
     2T27,'LOADING RATE:',F12.2,T54,'MGD',/
     3T25,'CLARIFIER AREA:',F10.0,T54,'SQUARE FEET',/
     4T31,'DIAMETER:',F12.2,T54,'FEET',/,T30,'COAGULANT:',4A4)
C
C                   REVERSE OSMOSIS
C
```

```
 170 PRECOV = RECOV * 100.
     NM = IFIX(RECOV * RFEED/(AFD*2.3))
     FFEED = RFEED/(10.**6)
C------------REVERSE OSMOSIS SPECIFICATIONS--------------------
     WRITE(6,180) RFEED,RATIO,PRECOV,BB(KBOPT),AFD,NM,MA,CT
 180 FORMAT(//,T10,'REVERSE OSMOSIS ',//,
    1T26,'INFLUENT FLOW:',F12.2,T54,'MGD',/,T27,'PRODUCT FLOW:',F12.2,T
    254,'MGD',/,T23,'PERCENT RECOVERY:',F12.2,T54,'%',/,
    3T17,'FLUX DECLINE INDEX (B):',F12.2,/,T27,'AVERAGE FLUX:',F12.2,
    4T54,'GPD/FT2',/,T20,'NUMBER OF MEMBRANES:',I9,/,
    5T29,'TOTAL AREA:',F12.2,T54,'SQUARE FEET',/,
    6T13,'TIME REQUIRED FOR CLEANING:',F12.1,T54,'HOURS')
C------RO-CONTINUED
     WRITE(6,190) CF,ADOSE,CDOSE
 190 FORMAT(' ',T22,'CLEANING INTERVAL:',F12.1,T54,'HOURS',/,
    1T17,'SULFURIC ACID INJECTED:',F12.2,T54,'PPM',/,
    2T22,'CHLORINE INJECTED:',F12.2,T54,'PPM')
C***********************************************************************
C                        TOTAL COSTS
C***********************************************************************
C-------------HEADINGS FOR THE UNITS PROCESS COSTS----------------
     WRITE(6,200)
 200 FORMAT('1',///,T25,'ENERGY AND LABOR ANALYSIS',//,
    1T13,'BUILDING ENERGY',
    2T35,'PROCESS ENERGY',T55,'LABOR',/,
    3T12,'KWH/YR COST($/YR)',T32,'KWH/YR COST($/YR)',
    4T53,'HRS/YR  $/YR')
C-----------TOTAL AND ANNUAL COSTS OF UNIT PROCESSES------------------
C
     WRITE(6,210)  (E(2*J-1),EC(2*J-1),E(2*J),EC(2*J),AL(J)
    1,ALC(J),J=1,10)
 210 FORMAT(/,T7,'FILTER UNIT',6F9.0,//,T6,'SURFACE WASH',6F9.0,
    1//,T13,'MEDIA',6F9.0,//,T9,' BACKWASH',6F9.0,//,
    2T4,'CLARIFIER UNIT',6F9.0,//,T8,' COAGULANT',6F9.0,//,
    3T3,'REVERSE OSMOSIS',6F9.0,//,
    4T5,'SULFURIC ACID',6F9.0,//,T9,' CHLORINE',6F9.0,//,
    5T10,'CLEANING',6F9.0)
     WRITE(6,214)
 214 FORMAT(//,T35,'UNIT COSTS',//,T15,'TOTAL CAPITAL($/YR)',T37,'ANNUA
    1L CAPITAL',T54,'O & M  ($/YR)',T67,'WATER COSTS($/KGAL)')
     WRITE(6,215) ((IG(J,I),I=1,3),BN(J),BNN(J),BOPT(IOPT,J),
    1AOPT(IOPT,J),J=1,10)
 215 FORMAT(/,10(T5,3A4,T25,F8.0,T45,F7.0,T60,F7.0,T75,F5.2,/))
     CAP = 0.
     RCAP = 0.
     DO 217 I = 1,10
     IF(I.GE.7) GO TO 216
     CAP = CAP + AOPT(IOPT,I)
     GO TO 217
 216 RCAP = RCAP + AOPT(IOPT,I)
 217 CONTINUE
     ADD = CAP + RCAP
     WRITE(6,218) CAP,RCAP,ADD
 218 FORMAT(/,T20,'COAGULATION/CLARIFICATION/FILTRATION $/KGAL=',F5.2,
```

```
     1/,T10,'                      REVERSE OSMOSIS SYSTEM $/KGAL=',F5.2,
     2/,T10,'                                TOTAL SYSTEM $/KGAL=',F5.2)
C
C----------------TOTAL COST AND PRICE/KGAL---------------------------
      TMAX2=TMAX
      IF(IWRITE.NE.0) TMAX2=TMAX1
      CKG = TMAX2/340000.
      WRITE(6,230) TMAX2,CKG
  230 FORMAT(///,T15,' THE PLANT PRODUCES 1 MGD PER DAY OF THE SPECIFIED
     1 QUALITY WATER',/,T10,'THE TOTAL ANNUAL COSTS =  $',F9.0,10X,
     2'COST/KGAL = $',F4.2)
      RETURN
      END
```

```
      SUBROUTINE JDPLOT
      COMMON /SCLR/FX,DX,FY,DY,ITESTX,ITESTY
      DIMENSION XVALS(100),YVALS(100),Z(105)
      DO 1 I = 1,10
      XVALS(I) = I
      YVALS(I) = I
    1 CONTINUE
      DX = 8
      FX = 0
      FY = 1.0
      DY = .25
      ITESTX = 1
      ITESTY = 1
C     CALL CPLOTO(3,10,-2,XVALS,YVALS,1,1,Z)
      RETURN
      END
```