Improvement of Reverse Osmosis

Through Pretreatment

by

J. R. Davis

M. K. Stenstrom

J. W. McCutchan

Water Resources Program

School of Engineering and Applied Science

UCLA

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ABSTRACT

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

The work described herein was initiated in April of 1976 at the Las Gallinas Valley Sanitary District, north of San Rafael, in Marin County, California. This site was selected due to the interest of two local agencies, the Marin Municipal Water District, and the Las Gallinas Valley Sanitary District. It was also selected due to the anticipated need for improved water resources in the county. Marin County was one of the most severely affected areas by the drought of 1976-77.

A 10 GPM (0.63 l/sec) reverse osmosis unit, using one-inch diameter, tube-style, cellulose acetate membranes was assembled at the site. The unit was initially operated using trickling filter effluent. The unit provided satisfactory effluent quality, but flux decline, due to membrane scaling and fouling, was excessive.

To improve production rates, a variety of types of treatment techniques were used to pretreat the trickling filter effluent prior to reverse osmosis treatment. Rapid sand filtration, chemical precipitation, flocculation, and clarification were evaluated. Average recovery rates were increased from a low of 25% using trickling filter effluent directly, to over 60%, using ferric chloride coagulation with sedimotion and filtration.

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Three years of operating data are presented in this report, along with detailed descriptions of flux maintenance and cleaning techniques. A mathematical analysis and model of the process are presented which can be used to determine the economically optimal design for municipal wastewater reclamation.

Chapter 1: INTRODUCTION

1.1 The Need for Wastewater Recycle in California

Water is a scarce and valuable commodity in California due to the arid nature of the region and the uneven distributions in water with respect to both time and space. The northern one-third of the state receives the bulk of the rainfall while most of the water demand is from the heavily populated semi-arid south. Moreover, a great majority of the state's rainfall comes during the winter months in contrast to the heavy summer demands.

Thirdly, the yearly rainfall is highly unpredictable with drought years of less than half the average rainfall interspersed with deluges of double to triple average values. This heterogenous distribution of water resources has resulted in the development of the world's largest system of reservoirs and canals to store and transport water to meet the state's temporal and spatial water demands. In spite of this highly developed water resources management system, severe water shortages are predicted before the year 2000, unless changes are made in the water-use policy of the state's agricultural economy.

To understand the severity of the state's water problem it is useful to review California water resources and water demand. Asano, Ghirelli, and Wassermann (1979) have compiled some useful statistics. The total water demand in California in 1975, an average rainfall year, was 35.0 million acre-ft.(11.4 trillion gallons), which was partially supplied by a groundwater overdraft of 1.8 million acre-ft. In the drought year of 1977 the overdraft increased to 6.2 million acre-ft. and by the year 2000 it is anticipated to reach this level during average rainfall years. A continued groundwater overdraft of this magnitude would rapidly diminish the state's groundwater resources and could not be maintained without drastic and unacceptable changes in the environment.

This increasing pressure on limited water supplies has led to a re-evaluation of excessive water use and wastewater disposal methods. As the gap between supply and demand closes, communities foreseeing water shortages are searching for ways of conserving and reusing water instead of discarding it into a heavily regulated environment.

Municipal wastewater recycle can lead to improvements in the environment and could significantly augment the state water supplies. In 1975 approximately 3.1 million acre-ft. (1 trillion gallons) of municipal wastewater was produced, of which 68%, or 2.1 million acre-ft.(684 billion gallons) were discharged to saline water (Asano, et. al., 1979; Calif. Dept. of Water Res., 1974). Moreover, the quantity of wastewater discharged to saline water is anticipated to increase to 3.7 million acre-ft. by the year 2000. It has been estimated that 2.5 million acre-ft. of this wastewater could be reclaimed and used for beneficial purposes (Calif. Dept. of Water Res., 1973; Calif. Dept. of Health Serv., 1979). This could reduce the projected year 2000 shortfall by approximately 40%. Therefore, it is apparent that wastewater recycling is an excellent source of additional water to decrease the impact of the predicted water shortage.

1.2 Current Reclamation Efforts

Though there are limitations and uncertainties pertaining to the direct reuse of treated wastewater for potable supplies, many communities are already using recycled wastewater for landscape and crop irrigation, industrial applications, and groundwater recharge (OSW and Bureau of Rec., 1972; Argo and Moutes, 1979, Flour, 1978).

Presently there are several full scale wastewater recycling projects in California. Prominent among these are the surface spreading operations by the County Sanitation Districts of Los Angeles (Asano, et. al., 1979), and the direct injection program at Water Factory 21 in Orange County (Argo and Moutes, 1979). During 1978 these projects and several smaller ones recycled a total of 0.184 million acre-ft. of treated wastewater.

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

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

1.3 Project Objectives

There have been numerous research projects investigating desalting capabilities of reverse osmosis. The California Department of Water Resources has recently sponsored investigations of desalting of irrigation drainage water (Antoniuk and McCutchan, 1973; Speight and McCutchan, 1979), for reuse and desalting ground water for potable supplies (Johnson and Loeb, 1969; Johnson, McCutchan, and Bennion, 1969), and application of reverse osmosis for municipal wastewater recycle (Wojcik, Lopez, and McCutchan, 1980). Los Angeles County investigated the application of the reverse osmosis process at their Pomona Wastewater Plant (Chen and Miele, 1971, 1972) and a 5 mgd spiral wound system was installed in Orange County, California (Argo and Moutes, 1979).

The investigation reported herein is an example of the use of simplest technology to achieve recycling objectives. A three-year study of a tubular reverse osmosis pilot plant is presented, with associated plant experience and operating data. The pilot plant is a portion of an ongoing study of reverse osmosis operation by the California Department of Water Resources.

A description of the unit is given in Table 1. A preliminary report of the system performance has been given by Wojcik, Lopez, and McCutchan (1979). Cooper and Richard (1978) and Cooper, Richard, Scarpace, and Straube (1977) have provided the results from studies on the unit's effectiveness in removing bacteria and viruses.

Table 1: RO Unit Specifications

Membrane Configuration	Tubular
Internal Diameter	.88 in.
Material	Cellulose Acetate
Annealing Temperature	88-90^{o} C
Number of Tubes	160
Operating Pressure	600 psi
Feed Rate	6.4 GPM

This study sought to evaluate the pretreatment requirements and the performance with trickling filter effluent on the feed. The RO pilot system was initially operated with only cartridge filtration of the feed water. From this system the pretreatment has evolved into the current system which includes chemical clarification, filtration, pH adjustment, and chlorination. During this development daily records of the flow and salinities of the feed, product and brine flows have provided documentation of improvements. These daily records, special short term tests, and cost data from previous DWR operation (Wojeck et. al., 1980), Oak Ridge National Laboratory (1980) and the EPA (1979) have provided the basis for a system's analysis to determine the economically optimal water reclamation plant configuration.

Chapter 2: THE REVERSE OSMOSIS CONFIGURATIONS AND OPERATING CHARACTERISTICS

2.1 The Status of Reverse Osmosis as a Desalination Alternative

The development of reverse osmosis as a process for desalinizing water was begun in the early 1950's. The early membranes, many of them organic substances (i.e., animal bladders) and metallic compounds, were not able to obtain economical fluxes while providing sufficient salt rejection ability. It wasn't until the development of the asymetric cellulose acetate membranes by Loeb and Sourirajian (1960), and the subsequent improvements by Manjikian, Loeb, and McCutchan (1965), that reverse osmosis became a viable desalination alternative. From these beginnings the process has become a major contributor to saline water and wastewater reclamation efforts.

Currently reverse osmosis is the most economical process for reducing the dissolved solids in brackish waters and removing most bacteria, heavy metals and viruses remaining after conventional wastewater treatment (Glueckstern, 1979;Channabasappa, 1977; OSW and Bureau of Rec.,1972, Goel and McCutchan, 1977). The value of reverse osmosis is evidenced by the increase in the number and capacity of reverse osmosis (RO) systems. Between 1971 and 1976 the number of membrane desalting plants (RO and electrodialysis) with capacity greater than 25,000 gallons per day (GPD) rose from 94 to 689 plants, and the total installed capacity of membrane desalting plants increased from 23 MGD in 1971 to 218 MGD in 1976. Eighty-five percent of these new membrane plants are reverse osmosis facilities (El-Ramby and Congdon, 1977). Fluor Corp. (1978) projected that the demand for output from membrane plants desalting brackish water will be over 4 BGD in 1985, and by the year 2000, almost 18 BGD.

2.2 THE THREE COMMON MEMBRANE CONFIGURATIONS

From the original plate and frame design used by the early researchers the membranes and their supports have evolved into three commonly used configurations: spiral wound, hollow fibre, and tubular. The choice of configuration is dependent upon the type of the water being treated. The spiral wound and the hollow fibre require water which is free from particulates (see Table 2).

The spiral wound membrane configuration employs membranes which form an envelope around a porous backing. The membranes are sealed on three edges with the open edge being glued to a plastic, perforated collection tube. This envelope is then wrapped around the tube and placed in a tubular pressure vessel. The feed water flows into the vessel and between the wrappings of the membranes. The product (desalted) water passes through the membranes and into the plastic collection tube which drains to collection tanks. The membranes normally come in three foot modules containing about 230 square feet of membrane area. The modules are connected in a combination serial and parallel arrangement to increase the recovery rate (OSW and Bureau of Rec., 1972).

The hollow fine fibre configuration developed by DuPont and Dow Chemical Corporations utilizes fibres with diameters of 25 to 250 microns (roughly comparable to the thickness of a human hair). The fibres have wall thicknessses of only 5 to 50 microns and are composed of unsupported membrane material. These hollow fibres are placed in a pressure vessel with the feed flowing external to the fibres. The product passes into the fibres and flows countercurrent on the inside to collection tubes. The small diameter of the fibres allows for a high packing density and prevents the fibres from collapsing. Descriptions of the spiral wound and hollow fibre configurations are given in the Desalting Handbook for Planners (OSW and Burreau of Rec, 1972).

Tubular membranes were utilized in this study. This configuration (Fig. 1) has the cellulose acetate membrane annealed to the inside of a porous membrane support which also functions as the pressure vessel. The 0.88 inch diameter, 10 foot long tubes are less prone to clogging than the spiral wound or the hollow fibre configuration and provide a configuration which is compatible with mechanical cleaning. Serial arrangement of the membranes allows for adjusting the recovery ratio (product flow/feed flow). The disadvantages of the tubular membranes are the large number of tubes required and the low packing density (ft^2/ft^3) relative to the other two configurations. A comparison of the three module configurations is given in Table 2, and a detailed description of the tubular configuration is given by Loeb (1965).

2.3 Membrane Flux

The flux through the membrane is related to the initial permeability and the rate of the flux decline. The initial permeability and

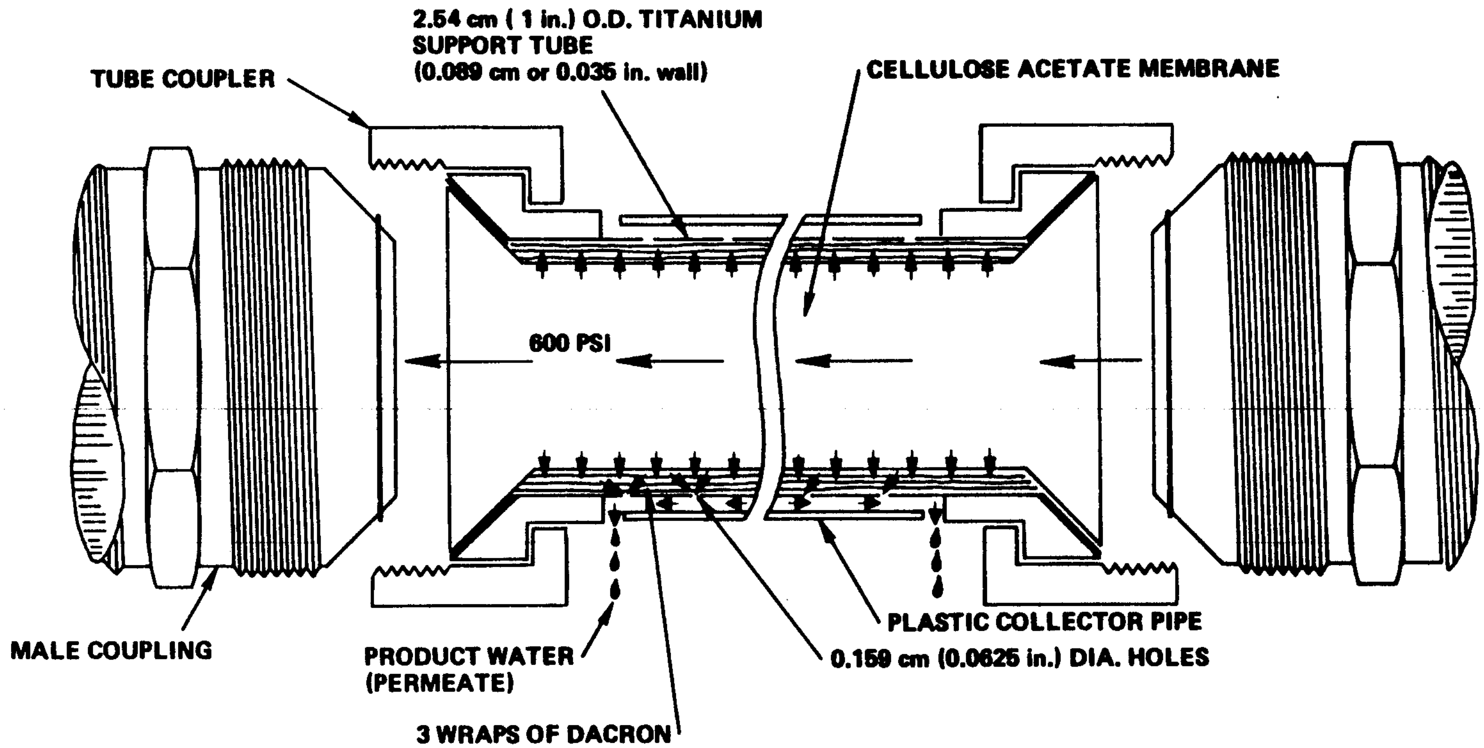


FIGURE 1: REVERSE OSMOSIS TUBULAR SECTION

Table 2: Comparison of Reverse Osmosis Module Configurations
(After: OSW and Bureau of Rec., 1972)

	Spiral Wound	Tubular	Hollow Fine Fibre
Membrane Surface Area per Volume, ft^2/ft^3	100 - 300	40 - 100	5,000 - 10,000
Product Water Flux, gfd	8 - 25	8 - 25	0.1 - 2
Typical Module Factors			
Brine Velocity, ft/sec.	0.7	1.5	0.04
Brine Channel Diameter, in.	0.005	0.5	0.004
Method of Membrane Replacement	As a membrane module assembly - on site	As tubes, on site	As entire pressure module on site, module returned to factory
Membrane Replacement Labor	Medium	High	Medium, requires equipment
High Pressure Limitation	Membrane compaction	Membrane compact.	Fibre collapse
Pressure Drop, Product Water Side	Medium	Low	High
Pressure Drop, Feed to Brine Exit	Medium	High	Low
Concentration Polarization Problem	Medium	High	Low
Membrane Cleaning - Mechanical	No	Yes	No
- Chemical	Yes - pH and solvent limited	Yes - pH and solvent limited	Yes - less restricted
Permissible Feed Ranges, ph	5.5 - 7.5	5.5 - 7.5	2 - 10
Permissible Temperature, ^oF	100	100	100

salt rejection capabilities of the cellulose acetate membranes are determined by the manufacturing process and the operating pressure.

Utilizing the Manjikian (1965) formulation for the membranes, the permeability can be changed by varying the annealing temperature. A higher annealling temperature yields a membrane with smaller pores. These smaller pores result in lower water permeability and higher salt rejection ability. Johnson, et.al., (1969) found empirically that a one degree centigrade increase in the annealing temperature decreases the permeation by about 1.5%. There is also a corresponding decrease in salt flow through the membrane. The optimal combination of membrane flux and salt rejection ability results with annealing temperatures between 86° and 93° Centigrade (Goel and McCutchan, 1977). Within normal temperature ranges an increase in operating temperature results in an increase in the permeability.

The initial flux across a membrane is also a function of the permeability and the change in pressure across the membrane. The simplest mathematical formulation to describe flux decline was presented by Rosenfeld and Loeb (1967) with flux as a function of the membrane permeability coefficient (A) and the differences in the osmotic and mechanical pressures across the membrane:

$$F_1 = A\ (\Delta P - \Delta\pi) \qquad (1)$$

in which ΔP is the pressure difference across the membrane, and $\Delta\pi$ is the osmotic pressure difference across the membrane.

The salt flux is given by:

$$F_2 = B\Delta C \qquad (2)$$

in which B is the membrane permeability coefficient for salt, and ΔC

the concentration difference across the membrane.

These equations were utilized in the work by Goel and McCutchan (1971, 1977) and Antoniuk and McCutchan (1973). Though not as theoretically comprehensive as those derived by Merten (1966), Speigler and Kedem (1966), or Sourirajan and Ohya (1977), they are adequate for low permeability membranes (Goel and McCutchan, 1977). The relationship of the intrinsic flux to annealing temperature and operating pressure are shown in Figures 2 and 3 (pages 13 - 14).

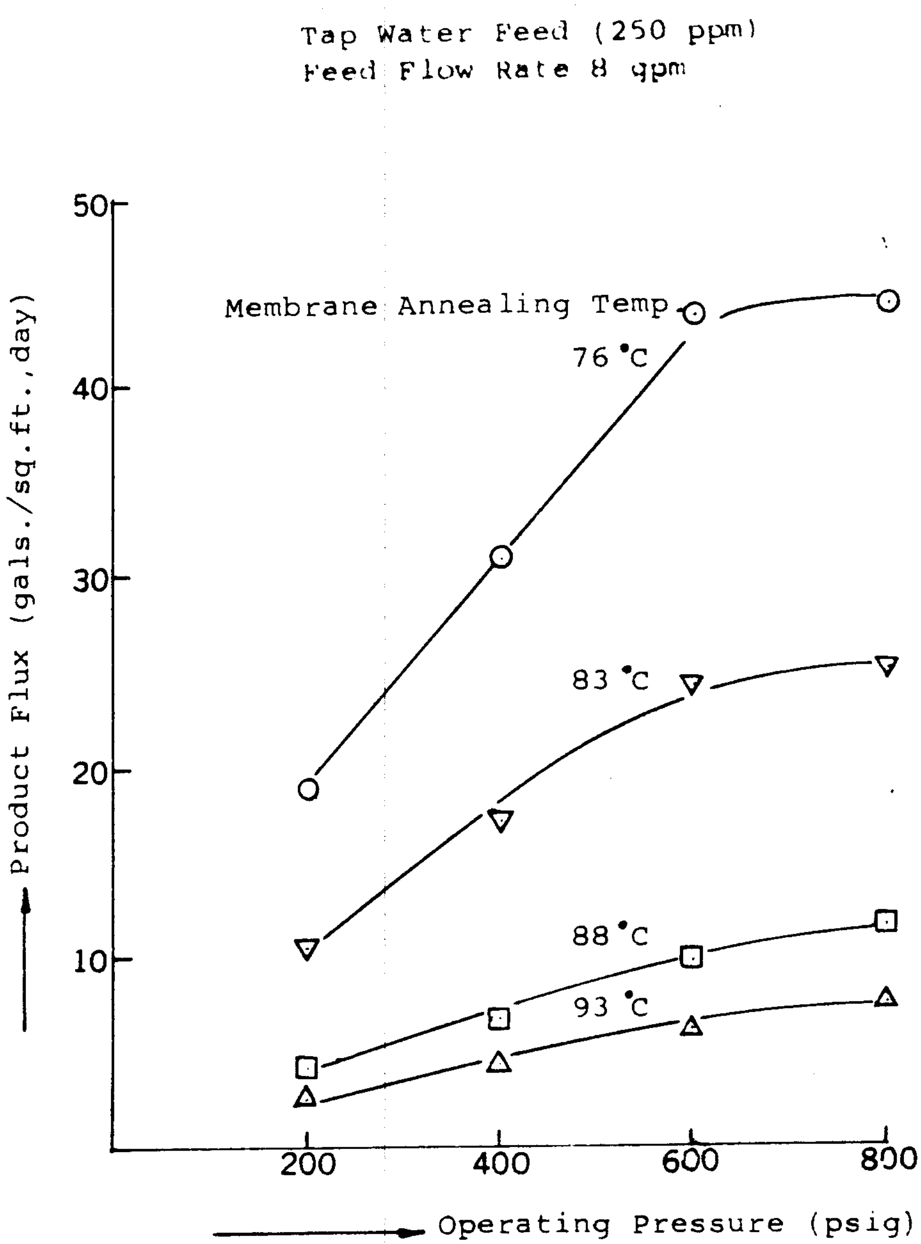


Figure 2; Product Flux versus Operating Pressure for Different Annealing Temperatures
(from Goel and McCutchan, 1977)

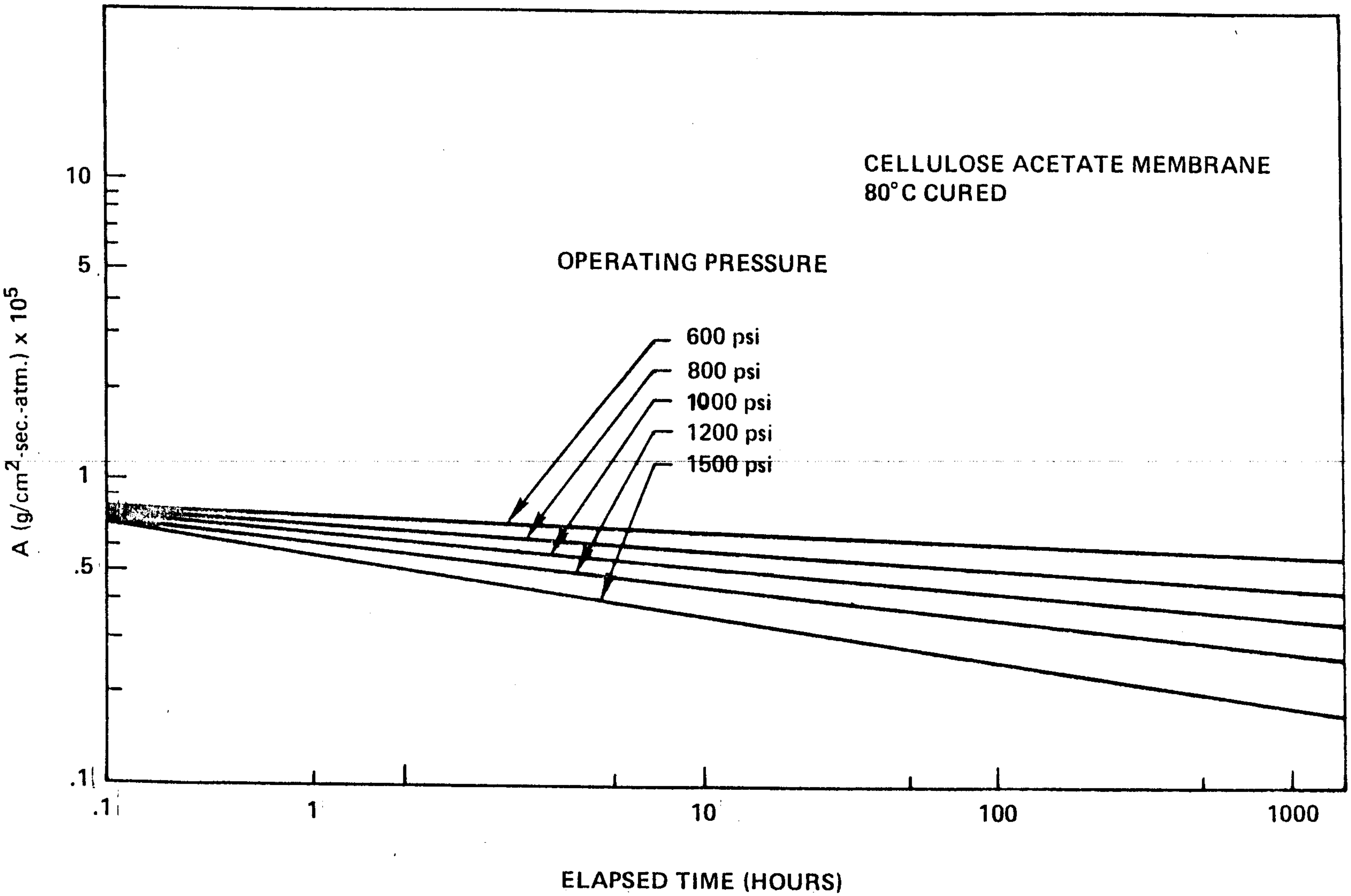


Figure 3 : Membrane Permeability Coefficient vs. Elapsed Time for Five Operating Pressures. (from Goel and McCutchan, 1977; based upon data of Merten, et. al.,1967)

Chapter 3: FLUX DECLINE PARAMETERS

3.1 Introduction

Reverse osmosis systems normally experience a severe drop in the product water flux during operation, especially during the first 20 hours after installation or cleaning. This is followed by a more gradual decline for the next 200 → 300 hours. Often at this point a severe drop in the flux may again occur (Thomas, et. al.,1973). These flux declines result in systems operating well below the intrinsic rate. The causes of these flux changes are related to three general processes; hydrolysis, compaction, and fouling. Establishing the causes, and designing systems which minimize the rate and effects of flux decline are major goals of reverse osmosis researchers.

3.2 Hydrolysis and Compaction

Hydrolysis of cellulose acetate membranes constitutes permanent damage to the membranes and results in an increase in both the water and salt flux through the membrane. The rate of hydrolysis is sensitive to the water chemistry, especially the chlorine residual. The chlorine reacts with the cellulose acetate causing hydrolysis, and since chlorine is normally used for control of organics care must be excercised in maintaining the residual of less than 1.0 ppm

(Channabasappa, 1977; Winfield, 1979a).

The rate of hydrolysis is also dependent upon the pH of the feedwater. The cellulose acetate membranes deteriorate rapidly when the pH is extremely acidic or basic. For maximum membrane life the pH needs to be maintained between 5.0 and 8.0 (Richard and Cooper, 1975; Winfield, 1979b).

Compaction occurs when membranes are subjected to high pressures. The structure is compressed resulting in a loss of product water flux. This compaction of the membranes results in a decrease in flux, but little change in the salt rejecting ability of the membranes. Though compaction is normally not as significant as fouling in reducing membrane flux, it is common to all membranes. The significance of compaction will vary with membrane formulation, operating pressure, and annealing temperature (Feuerstein and Bursztynsky, 1971; Kimura and Nakao, 1975; Pusch and Mossa, 1978; Goel and McCutchan, 1977). A general rule is that compaction effects increase with increases in the initial flux. The effects of compaction are usually significant only during the first 24 hours after the unit has been pressurized (Belfort, Babriella, and Marx, 1976).

3.3 Fouling of the membranes

Fouling occurs when substances or organisms in the feed wastewater become attached to the membrane surface or clog the membrane pores. The result is a decrease in the flux and usually an increase in salt rejection. Fouling of reverse osmosis membranes is the most significant cause of flux decline.

Reverse osmosis was orginally designed for use in seawater desalination where precipitation of salts onto the membranes was the determinant of maximum recoveries. The dissolved salts in the seawater quickly reach saturation as recovery increases and precipitate onto the membranes. With municipal wastewater as feed, much higher recovery ratios are possible without salt precipitation. As the recovery rates increase,the levels of organic and colloidal contaminants increase and become more significant.

Belfort, et. al.(1976) divided the foulants into three categories:

1) dissolved inorganics - precipitates ($CaSO_4$, $CaCO_3$, $Mg(OH)_2$, $Fe(OH)_3$)
2) dissolved organics - humics, biological slimes and dissolved macromolecules, and
3) colloidal material.

3.3.1 Dissolved inorganics (salts)

Precipitation problems are common with seawater desalination,and for brackish waters where the solubility limits of various salts (especially $CaSO_4$) are exceeded due to brine concentration and the effects of concentration polarization.

To establish the maximum recovery and optimize the operating parameters the ratio of the concentration at the membrane wall to that of the bulk solution must by computed. This ratio is called the concentration polarization ratio. The concentration polarization ratio is an indicator of the tendency for salts to accumulate at the membrane surfaces.

Rosenfeld and Loeb (1967) used the equations of Brian (1965) to develop the following relationship between the concentration of a solution constituent at the membrane wall (C_w) and the bulk solution (C_b). The concentration polarization ratio is defined as follows:

$$\frac{C_w}{C_b} = \frac{1}{D_r} + \left(1 - \frac{1}{D_r} \right) \exp\left(\frac{F_1 N_{sc}{}^{0.67}}{V_b j_d} \right) \tag{3}$$

in which D_r is the ratio of the bulk brine concentration (C_d) to the product concentration (C_p), F_1 is the product flux, N_{sc} is the dimensionless Schmidt number for salt diffusion (kinematic viscosity/ diffusivity coefficient), U_b is the bulk brine velocity, j_d is the Chilton-Coburn mass transfer factor which is equivalent to $.023NRe^{-0.17}$ ($Re = U_b d/\mu$).

3.3.2 Dissolved Organics

Though removal of organics is one of the functions of RO, the dissolved organics can cause severe damage to the cellulose acetate membranes if they are not controlled. The organics in the wastewater will not only attach to the membranes and decrease permeability, but will attack the cellulose acetate resulting in physical deterioration of the membrane. To mitigate the effects of these dissolved organics, the water can be treated for their removal and/or, as is more frequently done, the influent stream can be chlorinated.

Unfortunately, as mentioned earlier, chlorine will cause membrane hydrolysis . Richard and Cooper (1975) studied the effects of chlorination on the microfauna of the membranes and concluded that a chlorine residual of 0.1 - 0.2 mg/l was effective at preventing bacterial attack on the membranes while not causing detectable damage to the membranes. A residual chlorine concentration of 1 ppm was set the maximum by Goel and McCutchan (1977). Winfield (1979a) was able to find a high correlation between the dissolved organics (as measured by UV absorbance at 270°) and the flux decline. Winfield's work illustrates the importance of dissolved organics and the need for some form of disinfection in the pretreatment of the RO feed.

3.3.3 Colloidal Material

Colloidal material is present in all wastewaters even after conventional treatment and are readily removed by the reverse osmosis process. As the product water flows through the tubes the colloidal concentrations increase due to increased brine concentration and the concentration polarization. As these colloidal particles and the salts become concentrated the negative charges surrounding the colloidal particles are compressed due to the increased salts concentration. The colloidal material may then aggregate and become attached to the membranes, resulting in a loss of flux (Brunelle, 1980).

Bevege, et. al. (1973) did research in which they were able to correlate the precipitation on the membranes with the streaming potential of the brine. Jackson and Landolt (1972), however, found that the fouling effects of $Fe(OH)_3$ were minimized at the isoelectric point, apparently because of the increased shear force upon the larger floc produced. This apparent conflict may be due to differences in feedwater characteristics and variation in operating parameters. It is indicative of the complex relationships which exist in reverse osmosis operations.

When turbidity is used as an indicator of the suspended solids concentration, a correlation with the flux decline has been noted by several authors. Feuerstein and Burstynski (1971) found the flux and turbidity could be related by:

$$A = 0.708 A_o T_f^{-0.379} \tag{4}$$

where A is the expected product water flux coefficient. A_o is the flux coefficient for a pure saline solution of comparable osmotic pressure and T_f is the turbidity (NTU) of the brine. Belfort (1976) also experimentally observed the dependence of flux upon turbidity as did Cruver and Nusbaum (1974).

3.4 The Effect of Velocity on Fouling

The concentration polarization relationship indicates the role of velocity in preventing severe flux decline. As the velocity of the brine increases, the ratio, C_w/C_b, approaches unity. The velocity appears to be one of the most critical of design parameters (Boari, et. al., 1978).

The importance of velocity in controlling the flux decline in RO modules was investigated by Thomas, et.al. (1973). In his study of the flux declines encountered with tubular membranes treating primary sewage, a "critical velocity" could be determined below which flux declines were more severe. This critical velocity is dependent upon the particle size and the axial velocity. When velocities greater than the critical velocity were used, the flux decline parameter

($b = \Delta\log$ flux/ $\Delta\log$ time) was generally between 0.045 and .14 while at velocities below this critical velocity the value ranged from 0.3 to .016. For primary sewage a marked improvement in performance was found with velocity greater than

$$V_c = (1.3 \pm 0.6)*(S_I) \quad (5)$$

when S_I is the initial flux of the membrane. The relationship of velocity and flux decline was much weaker during the first 20 hours after cleaning when rapid flux declines are common.

The relationship of velocity and flux decline were noted by Belfort, et.al. (1976) using artificially prepared wastewaters, and further supported by Boari, et.al. (1978), Sach and Zisner (1977), Jackson and Landolt (1972).

Kimura, et.al. (1975) were able to show that the velocity of the brine is more crucial than the annealing temperature of the membranes for long run times. The flux decline curves of membranes annealed at varying temperature will merge. When they tested membranes with differing initial fluxes, they found that the fluxes nearly converged after 150-200 hours operation. The final flux varied with the velocity. Nusbaum (1972), Sach and Zisner (1977), Boari, et.al. 1978), and Jackson and Landolt (1972) have also verified the dependency of flux upon velocity.

3.5 The Effectiveness of Chemical and Mechanical Cleaning

Maintenance of adequate flux rates through the membranes requires periodic cleaning. Various techniques have been developed for removing the scaling and fouling which have been deposited upon the membranes. Several of these have shown limited success such as Aerojets (1969) trials with continuous addition of Calgon and Biz. More successful techniques have used flushes with fresh water, (Welchsler, 1976), depressurization with detergent flushes (Boen and Johannsen, 1974), and periodic spongeball cleaning (Johnson, et.al., 1969, Yanagi and Mori, 1980, and Sach and Zisner, 1977).

Johnson, et.al. (1969) developed the technique of cleaning by depressured flushing with citric acid followed by flushing with oversized foam balls (spongeballs). The method was discovered when a membrane became unattached during their citric acid flush and traveled through the membrane tubes resulting in an increase in the membrane flux. It was found that when periodic spongeball cleaning was employed the average membrane life was increased due to reduced fouling. Kini-E and McCutchan (1966) were able to restore 80% of the original membrane flux by citric acid flushing and spongeball cleaning while Sach and Zisner (1977) successfully used the technique on non-cellulosic membranes.

The net benefit of cleaning is dependent upon the rate of flux decline, required down time, and the expense of labor and chemicals. To decrease downtime Yanagi and Mori (1980) treated municipal waste by spongeball cleaning every half hour without shutting the unit down. The oversize spongeballs, put through without depressurization with the influent stream, maintained the flux at 24 to 28 gfd. The technique, however, appeared to result in a decrease in membrane life. The tests were insufficient for quantifying the costs.

Finding the spongeball cleaning frequency which provides for the optimal trade-off between flux and membrane life will require long term experimentation, and is a subject of future research.

Chapter 4: PRETREATMENT

4.1 Economic Benefits of Pretreatment

All reverse osmosis feedwater receives some form of preparatory treatment before entering the unit. The level of preparation varies from screening of particulates to elaborate tertiary treatment schemes.

Increased pretreatment results in benefits due to reduced RO capital investment and savings in the operation and maintenance cost of the reverse osmosis plant. Pretreatment of the reverse osmosis feedwater results in the following benefits:

1. The reduced organic concentrations result in a reduced fouling rate and therefore, maintenance of higher flux and lower energy usage;
2. The membranes will last longer due to the lessened bacterial attachment and lessened abrasion by particulates;
3. Removal or control of precipitating salts will decrease the scaling and increase the potential recovery rate.

4.1.2 Increased Flux and Decreased Energy Usage

The increase in flux due to the decrease in fouling not only reduces membrane area requirements, but also results in energy savings. Due to the high pressure involved, the energy costs for pumping water through a reverse osmosis system tends to be a major portion of the operating expenses of the plant (up to 50%). The amount of energy required to produce a quantity of water is dependent upon the volume of water which must be pressurized. If the flux is increased at a given pressure the recovery ratio and therefore the energy needed to produce the product is decreased. This relationship is illustrated by data collected by Wojcik, et.al. (1978) at the Las Gallinas Pilot Plant. The total system energy usage went from 12 to 28 Kwh/Kgal as the recovery rate decreased from 50 to 30 percent.

The relationship between energy savings and flux becomes more difficult to determine if the number of membranes can be varied, i.e., at the system design state. An increase in membrane area will result in an increase in the production rate and head loss. The advantages of each, increased flux and greater area, will depend upon the various parameters, so an optimization of benefits would need to be performed to minimize the costs. As stated previously, this study will be limited to the relationships determined at the Las Gallinas Pilot Plant.

The savings related to membrane replacement costs are difficult to quantify as accurately, though they represent a significant portion of the O & M costs (20-40%) (OSW and Bureau of Rec., 1972; Argo and Moutes, 1979). Water Factory 21, which has intensive pretreatment facilities, achieves an average membrane life of approximately 36 months (Argo and Moutes, 1979). It appears that actual membrane life will vary with the wastewater quality; therefore, a conservative estimate for filtered feed water would be about 24 months (OSW and Bureau of Rec., 1972; Loeb, 1964; Wojcik, et.al., 1979).

Additional savings due to pretreatment will accrue as the result of reduction in the chemical requirements for adjusting the pH of the feed water. Acidic coagulants reduce the sulfuric acid required by reducing the pH and decreasing the buffering capacity.

4.2 Pretreatment Methods

Several researchers have attempted to operate the tubular reverse osmosis systems using primary treated wastewater for the feed. The RO system has demonstrated that it is capable of treating solids laden wastewater, butother problems result such as more severe flux declines, shorter life of membranes and pumping equipment, and higher critical velocities (Feuerstein and Bursztynsky, 1971; Chen and Miele, 1972; Thomas, et.al. 1973; Sach and Zisner, 1977). Thus, though the tubular unit is capable of treating primary wastewater, it is normally economically advantageous to include additional feedwater treatment (due to the relatively high costs of RO operation).

Secondary treatment reduces the organic and solids loading to the unit. The result is a lowering of the critical velocity (Thomas, et.al., 1973) or an increase in the average flux (Feuerstein and Bursztynsky, 1971). Some research has been done on the effects of different types of secondary treatment processes (Anderson and Mills, 1977; Boen and Johannsen, 1974; Feuerstein and Bursztynsky, 1970). Where salt precipitation is a problem lime-soda precipitation and/or ion exchange will be advantageous, but for municipal wastes colloidal and organic material often cause more severe problems, and their reduction is more crucial.

The simplest form of pretreatment is the use of cartridge filters preceding the RO module to remove particles which may damage the membranes. However, sensitivity of flux to cartridge filtration with pore sizes between 5 and 50 microns appears to be small. Johnson and Loeb (1966) and Winfield (1979) found little correlation between the flux decline when the pore size was varied within this range. The benefit of cartridge filtration appeared to be limited to the protection of equipment from large particulates.

More effective removal of colloidal material can be achieved by using conventional tertiary processes such as, coagulation-settling-filtration, coagulation-filtration, activated carbon treatments or a combination of these processes.

Granular activated carbon adsorption (GAC) preceded by filtration has proven to be the more effective pretreatment for increasing flux through the membranes, but is is also relatively expensive. The GAC pretreatment is being utilized at Water Factory 21 in Orange County, California, where the high quality GAC effluent can be blended with RO product before being injected into the groundwater aquifer (Argo & Moutes, 1979).

Though sand filtration has been found only marginally useful in reducing the flux decline, alum coagulation followed by sand filtration has also been found to be effective at decreasing the flux decline (Feuerstein and Bursztynsky, 1971; Boen and Johannsen; Aerojet, 1969). Nusbaum (1972) found that the mixed media filtration could replace the GAC as a pretreatment for secondary effluent with little loss of flux, and Channabasappa (1975) recommended coagulation/settling followed by sand filtration in order to decrease the loss of flux. The findings do favor some form of pretreatment and it is the purpose of this research to compare pretreatment alternatives, and determine their relative benefit.

4.3 Blending of Pretreated Water With The RO Product Water

For many applications of recycled water the quality of the reverse osmosis product is better than that demanded by users. The blending of secondary effluent which has been coagulated and filtered with the RO product can produce water of sufficient quality at a significantly reduced price, due to the lower costs of operating the coagulation-filtration process. The economics of this blending are dependent upon the use of product water and the TDS of the feedwater. For example, if the TDS of RO feedwater (filter effluent) is 3000 ppm, there is little opportunity for savings; a blend of 1:1 will have a TDS of about 1600-1700, but if the influent has a low TDS (i.e., 1000-1500) the higher blending ratios can decrease water costs by 10% to 40% depending upon the product requirements. This use of blending not only reduces costs, but allows some flexibility in meeting varying water quality requirements.

4.4 Health Benefits of Pretreatment

While this report concentrates upon the direct benefits of pretreatment in terms of capital and operating reductions which are related to the production of the reverse osmosis effluent, there are other important benefits of pretreatment which will have varying value depending upon product water use. Many of these benefits are difficult to evaluate but will become increasingly important in the future.

Health effects will become increasingly important as reclaimed wastewaters are used to augment the potable supplies. When wastewater is chlorinated, the chlorine reacts with the organic constituents in the water to form low molecular weight carcinogenic substances. Reverse osmosis is relatively ineffective at removing these by-products from the influent stream (Cooper and Richard, 1978); thus, any reduction in the concentration of organic compounds in the feed water should result in a decrease in the level of carcinogens in the product. The addition of coagulation/filtration or granular activated carbon adsorption effectively lowers the organic content and thereby the production of the organic by-products of chlorination.

Evaluation of this benefit would be dependent upon the use of the water; since the length of time before use, the amount of mixing with other waters, and possible filtering in groundwater basins are all factors in determining the health risks. A discussion of the potential health effects of these by-products is given by Page, et.al., (1979).

Chapter 5: LAS GALLINAS VALLEY CASE STUDY

5.1 Introduction

The Las Gallinas Pilot Plant Study is the latest of a series of UCLA research on the operations of tubular cellular acetate membranes. In this chapter studies involving brackish water desalting will be discussed briefly before a detailed analysis of the Las Gallinas Operations.

5.2 Coalinga, California 1965-1969

The development of high flux cellulose acetate membranes which retained a high salt rejection capacity by Loeb and Sourirajan (1960), and then improved by Manjikian, et. al. (1965), encouraged the UCLA researchers to test the membranes in a pilot scale desalination effort at Coalinga, California. The Coalinga groundwater supply contains 2500 ppm of salt. The reverse osmosis system was installed to provide desalted water to augment the city's potable supply. During the three and a half years of operation the most serious problem encountered was the decline in flux resulting from fouling of the reverse osmosis membranes. Subsequent analysis of membrane foulants showed $Fe(OH)_3$ to be the major constituent in the fouling material. The attempts at reducing this loss of flux met with some success.

The addition of cobalt catalized sodium sulfite, which prevented the oxidation of the iron by removing the oxygen from the feed, improved performance. Chlorine injection, which oxidizes iron but inhibits the growth of bacteria upon the membranes resulted in less flux decline, (Loeb and Johnson, 1966).

5.3 Firebaugh, California 1971-1979

The Interagency Wastewater Treatment Center at Firebaugh, California, has been concerned with the salinity increases in groundwater and soil due to irrigation water which permeates through the top soil, accumulating dissolved solids and then depositing these salts at a less permeable layer below. The resulting build-up of salts in the subsoil, if unattenuated, endangers agricultural production.

The Firebaugh, California, study was initiated by the California Department of Water Resources and UCLA in 1971 to ascertain the effectiveness of reverse osmosis in reducing the TDS of irrigation drainage for reuse.

The system contained 180 membranes and operated at 400 psi with an influent flow rate of approximately 6 gpm. The unit experienced severe fouling problems due to membrane scaling and the average recovery was only 33%. Addition of sodium hexametaphosphate (SHMP) improved performance, but scaling was still a problem.

Besides the low recovery the membranes also experienced physical deterioration. Deterioration of the membranes noted after 25 days of operation was attributed to bacterial growth in membrane areas which were not subject to a significant flux (i.e., at the flange ends where the couplings prevented flux), and chlorine addition to 1 ppm residual was initiated (Antoniuk and McCutchan, 1973). Still after 99 days of operation the tests were temporarily discontinued because of the excessive membrane deterioration; whether this was caused by the initial bacteria growth in areas protected from chlorine disinfection (Antoniuk and McCutchan, 1973) or by the attack upon the membranes by the chlorine as conjectured by Loeb and Johnson (1966) is not evident. The deterioration in evidence before chlorination favors a bacterial hypothesis.

5.4 Pilot Plant Description

In 1975 the UCLA Engineering Systems Department, with the sponsorship and support of the California Department of Water Resources, and in cooperation with the Las Gallinas Valley Sanitary District and Marin Municipal Water District, started the operation of a 160 tube reverse osmosis pilot plant at the Las Gallinas facility in Marin County, California. The purpose of the project was to investigate the application of reverse osmosis in reclamation of municipal wastewater.

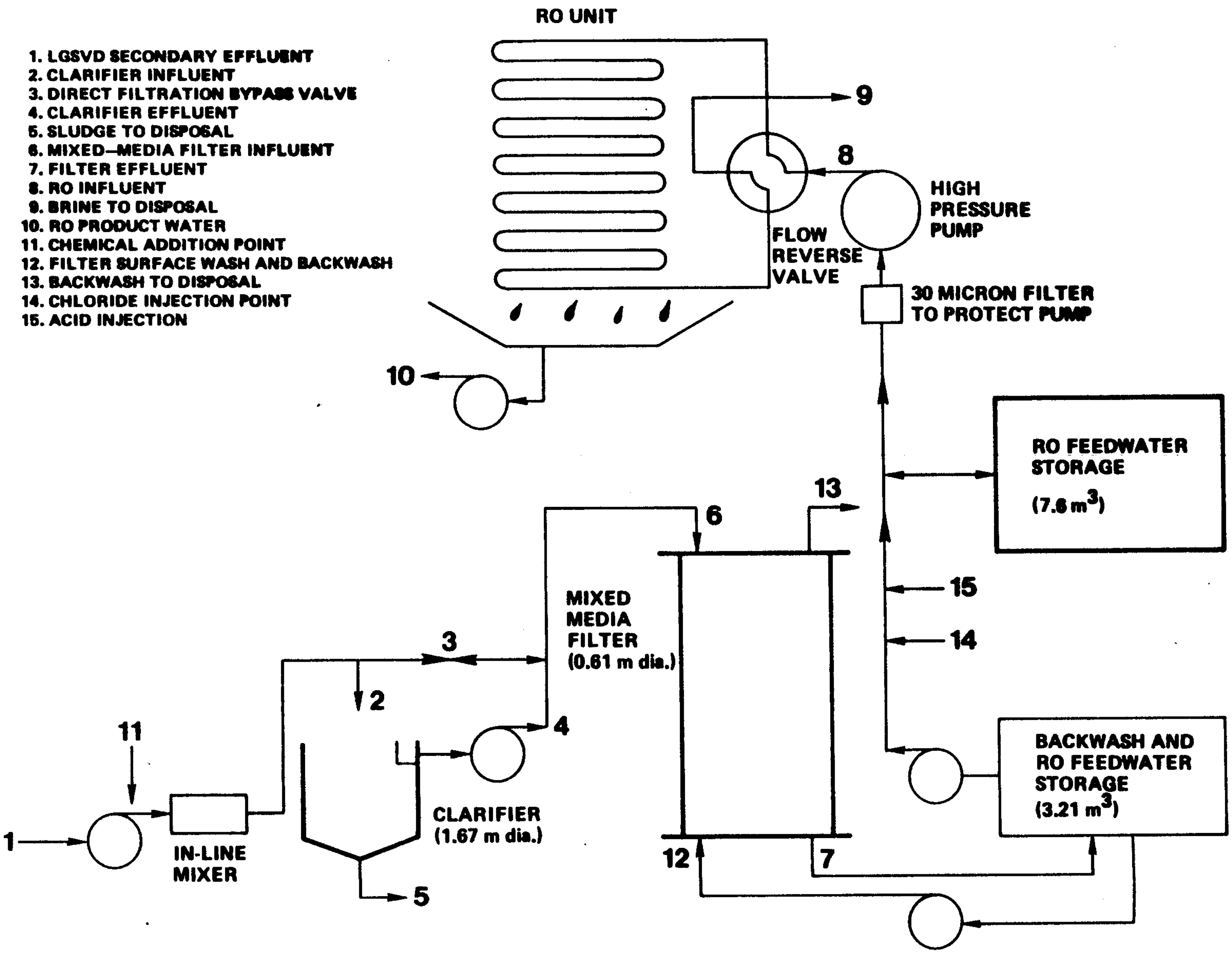
The pilot plant is located at the Las Gallinas Valley Sanitary district north of San Rafael, in Marin County, California. The sanitary district operates a secondary treatment facility using two stage trickling filters and clarification. The total plant flow consists almost entirely of municipal wastewater, augmented with heavy rainfall runoff from November to March. The flow rate ranges from 1.5 MGD ($.065 m^3/sec$) to as high as 10 MGD ($.44 M^3/sec$) depending upon the season and weather, with the trickling filters loaded at approximately 11 MGD/acre ($1.17 \times 10^{-4} m^3/m^2$-sec) and 84 $lbBOD_5$/1000

ft^3 (1.35 kg BOD_5/m^3) of media.

The RO pilot plant is composed of 160, ten foot (3.05 m) long tubes, with an internal membrane diameter of 0.88 inches (2.23 cm). Each membrane has an effective area approximately 2.24 ft^2 (2.08 m^2). Cellulose acetate membranes are used exclusively and are made by pilot plant operators using Eastman cellulose acetate formulation E-400-25, with membrane curing temperatures of 88-90^oC. The total membrane area of the unit is 358 ft^2(33.3 m^2). Figure 1 is a schematic diagram of a typical membrane; the membrane casting procedure and construction techniques are no different than those used in earlier studies at U.C.L.A. (Johnson, et.al., 1969; McCutchan and Goel, 1974; Goel and McCutchan, 1977, Speight and McCutchan, 1979) and are similar to the original membranes reported by Manjikian, et.al., (1965). The membranes are cast at the Department of Water Resources Firebaugh Facility.

The membranes are all arranged in a series, single stage configuration. A triplex positive displacement pump is used to pump feed water and can be operated over a range of 200 to 800 PSIG with flow rates ranging from 2 to 12 GPM (0.13 to 0.75 l/sec). The work reported herein was all performed at 600 PSIG (40.8 atmospheres) and the feedwater flow rate was normally 6.4 GPM (.404 l/sec) providing an inlet Reynolds number of approximately 23,000 and an inlet velocity of approximately 3.4 ft/s(1.05 m/s). The unit was also equipped with automatic flow reversal valves and spongeball cleaning facilities. A schematic is shown in Figure 4.

Figure 4: Plant Schematic



5.4.1 Chronology of RO Plant Operation

The RO pilot plant was originally placed in operation in April of 1976 treating trickling filter effluent which was filtered through a 30" diameter multi-media filter. This filter, besides providing feed for the RO unit, was operated by the Marin Municipal Water District (MMWD) to provide water for recycle. Subsequently MMWD installed a second filter to increase their recycling effort. The RO unit operated on filtered trickling filter effluent from the MMWD filters until May of 1979 when a smaller 24" diameter mixed-media filter was installed and dedicated to pretreatment of RO feedwater. No coagulating chemicals were used until May of 1979. The reclaimed water from the MMWD filters was used from 1976 through the drought of 1977 and continues until the present; the Marin Municipal Water District should be consulted for further information regarding their work.

This initial period from April 1976 to June 1979 was devoted to the development of membrane cleaning techniques and endurance testing of the RO membranes and equipment. The original cleaning technique was restricted to spongeball cleaning without chemical cleaning agents. During cleaning the unit was always depressurized and flushed with tap water or RO product water (later containing cleaning chemicals). Beginning in April 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 spongeball cleaning, was developed. Table 3 is a summary of the final cleaning procedure.

In March, 1978, chlorination of RO feedwater 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 operation.

5.4.2 Pretreatment with Coagulation-Clarification-Filtration

The current phase of UCLA participation at Las Gallinas began in December of 1978 with Zeta Potential determinations and jar tests to determine the efficiency of various coagulants. Tests were conducted using alum, ferric chloride, and various organic polymers. From these initial tests the alum, $FeCl_3$ and Nalco 7134 cationic polymer were selected for system testing.

From June 1979 until July 1979 feedwater was pretreated using direct filtration with cationic organic polymer (Nalco 7134). In July, 1979 a 5.5 ft. (1.7 m) diameter clarifier was installed and inorganic coagulants were used. The mixed-media 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 sec) 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 sec) rate independently of the RO feed rate and excess water was discharged with the Las Gallinas Valley Sanitary Districts Effluent. The filter media used was a commercially available media (Neptune Microfloc) consisting of 1.0 to 1.2 mm size

Table 3: Final Weekly Cleaning Procedure

Operation	Procedure
Citric Acid Flush	0.55 lbs (250 grams) of citric acid is added to 50 gallons (190 liters) of tap 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 lbs (500 grams) of a commercially available detergent (Biz) is added to 50 gallons (190 liters) of tap water at ambient temperature and circulated through the RO unit as before for one hour.
Spongeball Cleaning	After completion of chemical cleaning, ten 1 1/2" (3.8 cm) spongeballs are introduced into the RO feed at approximately one minute intervals, and are allowed to pass through the unit at approximately 2.7 ft/sec. (.52 m/sec).
Spongeball Cleaning	After approximately 70 hours of operation the unit is depressurized and the spongeball cleaning is repeated.

distributions of coal, a 0.42 to 0.55 mm size distribution of silica sand, and 0.2-0.3 mm size distribution of garnet sand.

The clarifier was operated at 10 GPM (.63 l/sec) also, 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 entire pilot plant, with the exception of cleaning chemical makeup and data collection, operated on an unattended, automatic basis. Operational attention was restricted to a daily check and daily data collection. Cleanings were normally performed on the remaining two days (usually Monday and Thursday). The total amount of operator time averaged about two hours per day.

Table 4 summarizes the various changes in operation of the unit. Figure 4 (page 34) is a schematic flow diagram for the entire reverse osmosis plant with pretreatment facilities.

5.4.3 The Twenty-Four Hour Tests

In order to evaluate the effectiveness of various coagulants in preventing flux decline twenty-four hour tests with close monitoring of the flux and TDS were conducted. The tests were conducted during three periods during which 3-4 tests were run on consecutive days. This served to lessen the variability due to changing influent composition and temperature variations. The general procedure follows:

1) To prepare for the test, injection of the coagulant to be tested was begun at the influent pump discharge. The chemical clarification - filtration system was operated independently

Table 4: Chronological Summary of Pilot Plant Operation

DATE	HOUR		
4/27/76	0	Pilot plant started up on trickling filter effluent after multi-media filtration.	Weekly spongeball clean ing without cleaning chemicals.
4/18/77	8,500	Cleaning prodedure changed by the addition of two hour Biz detergent flush.	Various concentrations of Biz (up to 2.1 g/l) were used for flushing.
6/20/77	10,000	Citric acid substituted for Biz detergent.	Various concentrations (.04-.53g/l) were used.
9/26/77	12,400	Returned to Biz detergent.	Concentrations between 1.05 and 1.32 g/l were used.
1/1/78	14,700	Final cleaning procedure developed, using one hour citric acid flush, followed by one hour Biz detergent flush, followed by sponge ball cleaning.	0.66 g/l citric acid concentration and 1.32 g/l Biz used for flush.
3/23/78	16,700	Began chlorination of multi-media filter effluent.	Chlorine residual ranged from .5 to 6.0 mg.1 aver aging about 2.5 mg/l.
5/15/78	18,000	Influent pH control initiated by addition of sulfuric acid.	Set acid injection for ph = 5.5.
8/1/78	19,800	Automatic sponge ball clean-ing started using reverse flow to initiate sponge ball cleaning.	Cleaning frequency set to 6 hours.
6/1/79	27,100	Mixed media filter with cation ic polymer operation begun.	Cationic polymer dosage optimized by Zeta Potent ial measurements.
7/6/79	28,000	Clarifier installed. Pilot plant operated with coagula-tion, clarification and filtration until shut down.	Optimal concentrations of alum, $FeCl_3$ and cationic polymers tested during this period.
1/7/80	32,400	Unit shut down.	

of the reverse osmosis unit to allow the pretreatment system to reach steady-state for the coagulant. During this process the water remaining in the 2000 gallon storage tank provided feed for the RO unit.

2) The multi-media filter was backwashed.

3) The RO unit was shut down and chemically cleaned with a one hour citric acid flush, followed by a one-hour Biz flush, followed by ten oversize spongeballs introduced at one-minute intervals.

4) During the cleaning the feedwater storage tank was drained, flushed with water from the pretreatment system and filled. Normally enough water to begin the test was available when the cleaning was completed.

5) The unit was started and adjusted to a feed rate of 6.4 GPM and to 600 psi operating pressure. Initial data collection was begun thirty minutes after start-up.

6) The brine and product flows were determined by timing 30 - 60 seconds of flow into .264 gallon (1000 ml) graduated cylinders and recording the results in milliters per minute and gallons per minute. The feed flow was calculated by summing these two flows. The total dissolved solids (TDS) was measured with a TDS meter and recorded. Also recorded were the filter effluent turbidity, power usage, operating pressure, and pH. A sample data collection sheet is enclosed in Appendix C.

7) The measurements were repeated at hourly intervals for the first several hours and then repeated the next morning.

Chapter 6: EXPERIMENTAL RESULTS

6.1.1 Flux and the Effects of Cleaning

The early results with the unit were disappointing in that very low recovery rates were obtained. The recovery averaged about 25% with fluxes in the range of 4.5 to 5.0 gal/ft^2 day (GSFD) (7.6-8.5 1/m^2 hr). The spongeball cleaning was effective at first in that fluxes were increased from approximately 5.0 GSFD (8.5 1/m^2 hr) before cleaning to approximately 9 to 10 GSFD (15 to 17 1/m^2 hr) after cleaning. After approximately 8000 hours operation the flux before cleaning decreased to approximately 3.5 GSFD (6 1/m^2 hr) while the flux after cleaning could only be restored to about 4.2 to 4.5 GSFD (7.1-7.6 1/m^2 hr). This rapid deterioration was due to the accumulation of insoluble salts on the membrane surface, which could not be scrubbed from the surface by the spongeballs.

The use of the enzyme detergent partially restored the membrane fluxes, but results were still disappointing. Starting in April of 1977 the fluxes after detergent and spongeball cleaning gradually increased from 4-4.5 GSFD (6.8-7.6 1/m^2 hr) to a maximum of about 5 GSFD (8.5 1/m^2 hr). On June 20, 1977 the first citric acid cleaning was performed, which restored membrane flux to 12.5 GSFD (21.2 1/m^2 hr). This flux after cleaning was maintained until the end of September when flushing only with the enzyme detergent was resumed. The flux after cleaning gradually declined and by December

1977 had declined to the previous levels of 4 to 4.5 GSFD (6.8-7.6 $1/m^2$ hr). Beginning in March of 1978 the final cleaning procedures shown in Table 3 were consistently used and flux after cleaning again stabilized at about 12.5 GSFD (21.2 $1.m^2$ hr). The results of the improvements in cleaning technique can be seen in Figure 5 which shows the before and after cleaning fluxes for the entire period of the investigation. The increases in flux due to improvement in cleaning are obvious. The increases in fluxes after January, 1978 are attributed to improved pretreatment, rather than improved membrane cleaning.

6.2.2 <u>Flux Decline and Effects of Pretreatment</u>

Improvements made in recovery and flux maintenance after January 1978 are largely due to improvements in RO feed water quality. Chlorination of RO feed water was begun in March 1978 and feedwater pH control (pH controlled to approximately 5.5) was begun in May, 1978, and resulted in small increases in the before cleaning fluxes. Before cleaning fluxes increased from approximately 4 GSFD (6.8 $1/m^2$ hr) to approximately 6 GSFD (10.2 $1/m^2$ hr). The installation of the auto-spongeball cleaning devices in August of 1978 coincides with increases in before cleaning fluxes to as high as 8 GSFD (13.6 $1/m^2$ hr). Unfortunately the high before cleaning fluxes fell back to the 5-7 GSFD (8.5-11.8 $1/m^2$ hr) range during the period from October, 1978 to May, 1979. At present there is no explanation of this decreasing trend.

The use of chemical coagulation and clarification had very

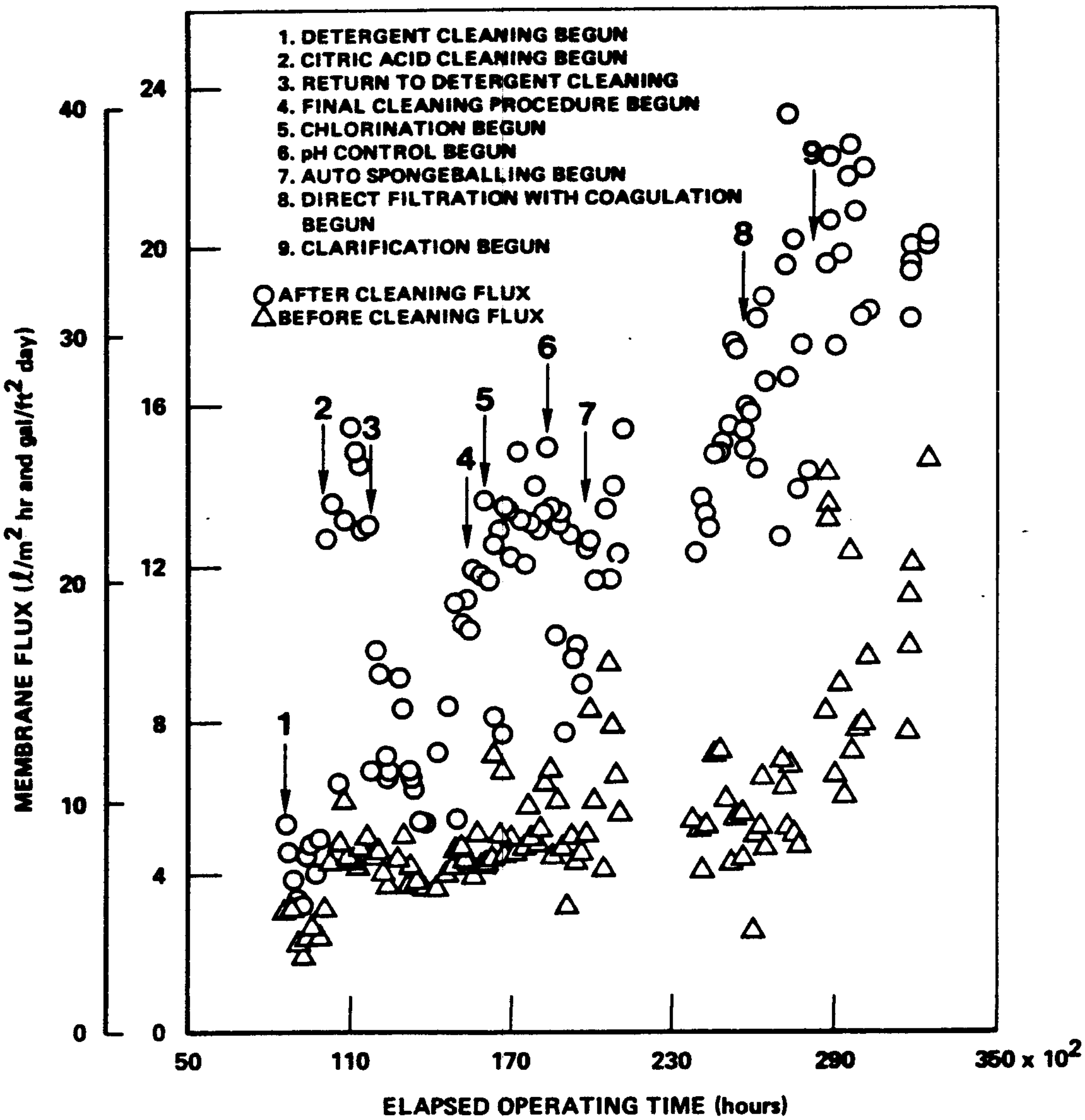


Figure 5: Membrane Fluxes Before and After Cleaning vs. Time

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 (23.7 to 42.4 l/m^2 hr) respectively, during the final phases of the study when the inorganic coagulants were used.

During the last months of the study a series of 24 hour flux decline tests were made using various concentrations of ferric chloride, alum, and organic coagulants. Flux decline curves for representative 24 hour tests for each coagulant, and uncoagulated, filtered trickling filter effluent are shown in Figure 6. The flux decline coefficients (slope of a log-log plot of flux and time) are shown in Table 5 and are compared with decline coefficients calculated by Thomas, et. al. (1973). The flux decline coefficients show that the ferric chloride coagulant produced feed water with the least tendency to foul the membranes. The effectiveness of the organic coagulant using direct filtration and alum with clarification were about equal. The uncoagulated water, as expected, produced the largest fouling rate.

6.1.3 Flux Variation of Membranes

Figures 7 and 8 show the results of flux measurements for individual membranes. In Figure 7 the upper set of data points represents flow measurements taken one-half hour after the unit was started after cleaning. It reflects the variability in the initial

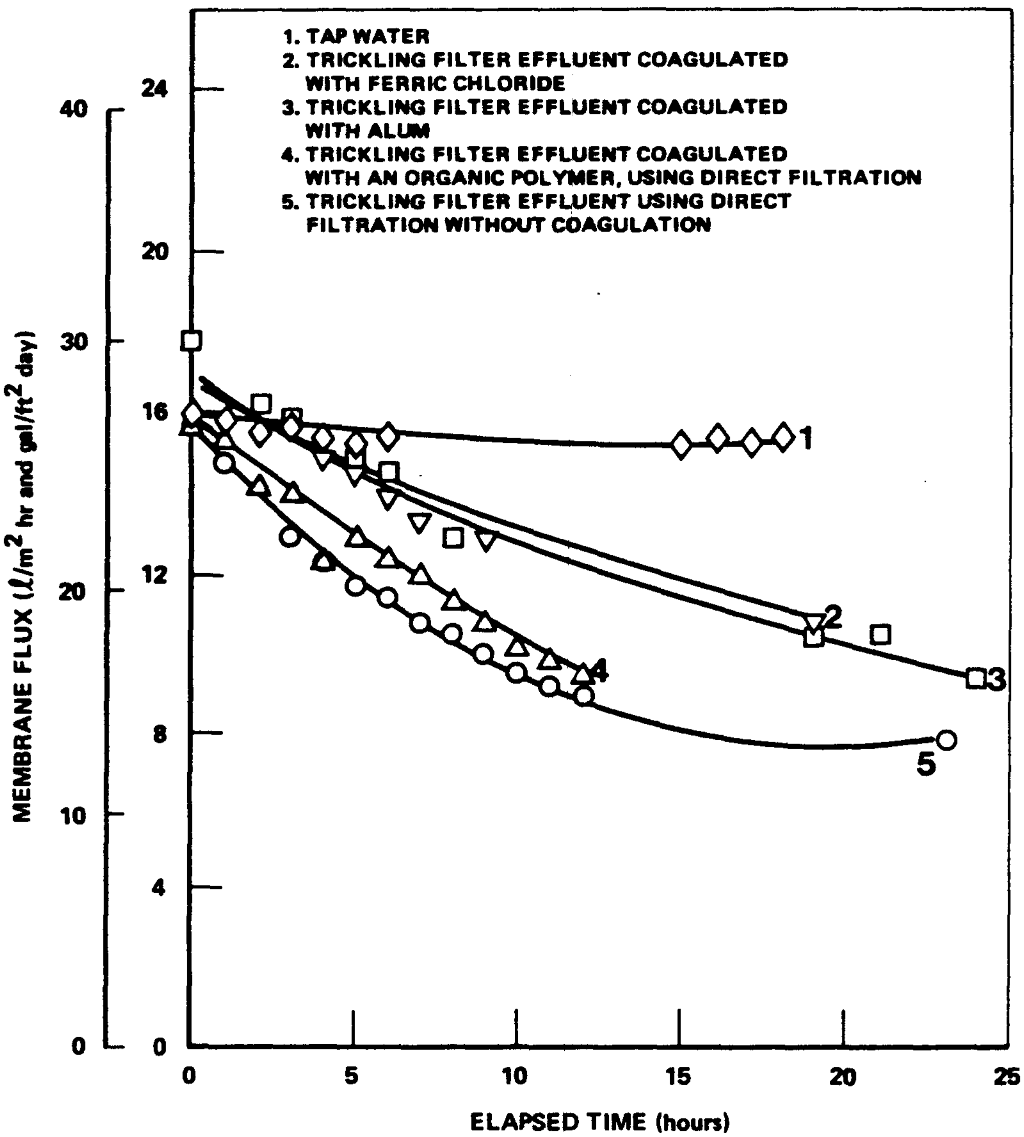


Figure 6: Flux Decline From 24 Hour Tests

Table 5: Flux Decline Coefficients for Various Types of Feed Waters

Flux Decline Coefficient	Feed Water Type	Reference
0.243	Trickling Effluent with Dual Media Filtration	This Study
0.202	Trickling Filter Effluent with alum coagulation, clarification and mixed-media filtration	This Study
0.204	Trickling Filter Effluent with Organic Polymer coagulation and direct mixed-media filtration	This Study
0.146	Trickling Filter Effluent with Ferric Chloride coagulation, clarification, and mixed-media filtration	This Study
0.0136	Tap Water (TDS 100)	This Study
0.9	Raw Wastewater	Calculated by Thomas et. al. (1973) from the data of Feuerstein and Bursztynsky (1970).
0.56	Primary Effluent	Calculated by Thomas et. al., as previously.
0.35	Secondary Effluent	Calculated by Thomas et. al., as previously.
0.14	Carbon-Treated Secondary Effluent	Calculated by Thomas et. al., as previously.

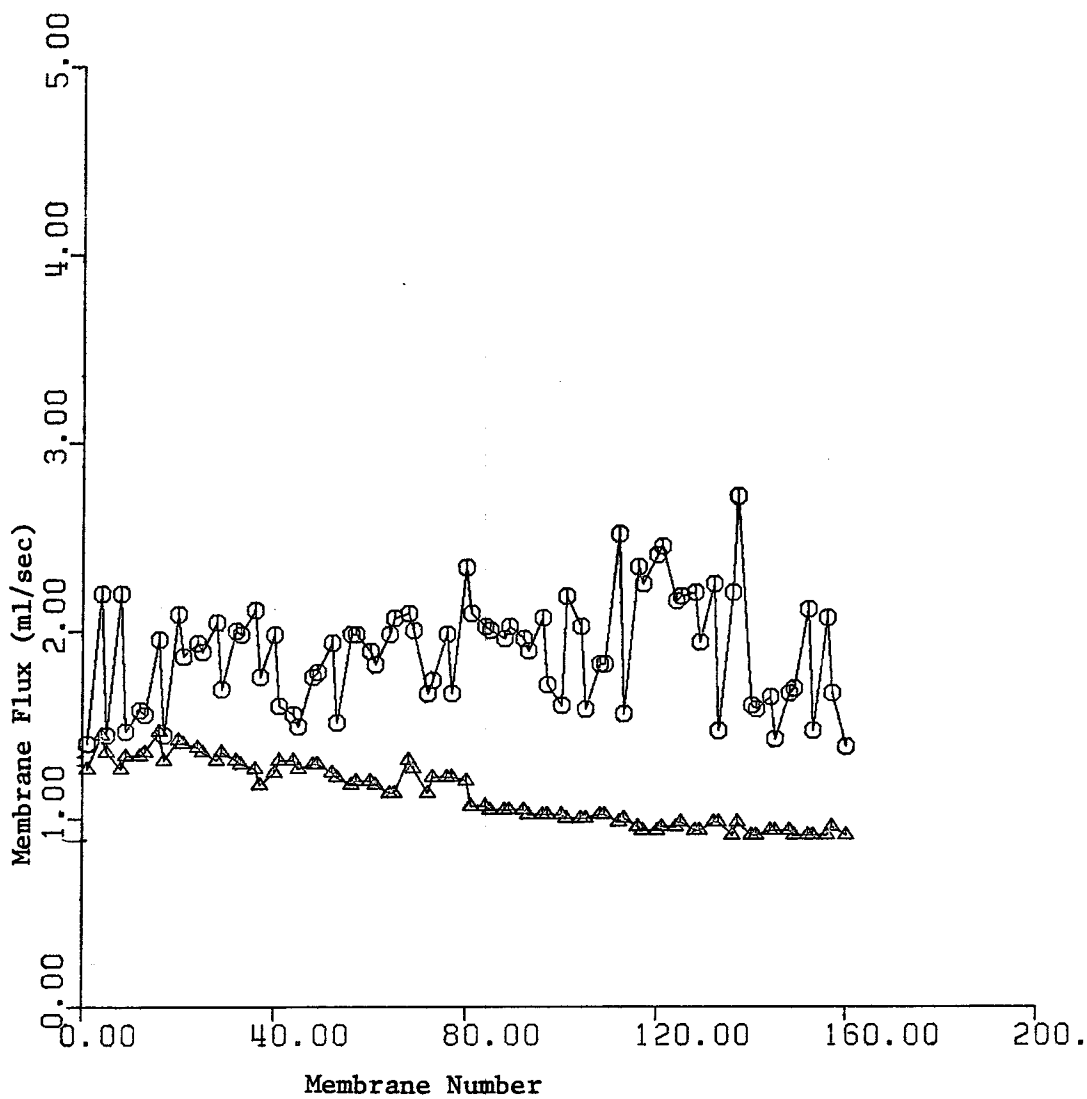


Figure 7: Membrane Flux versus the Membrane Number

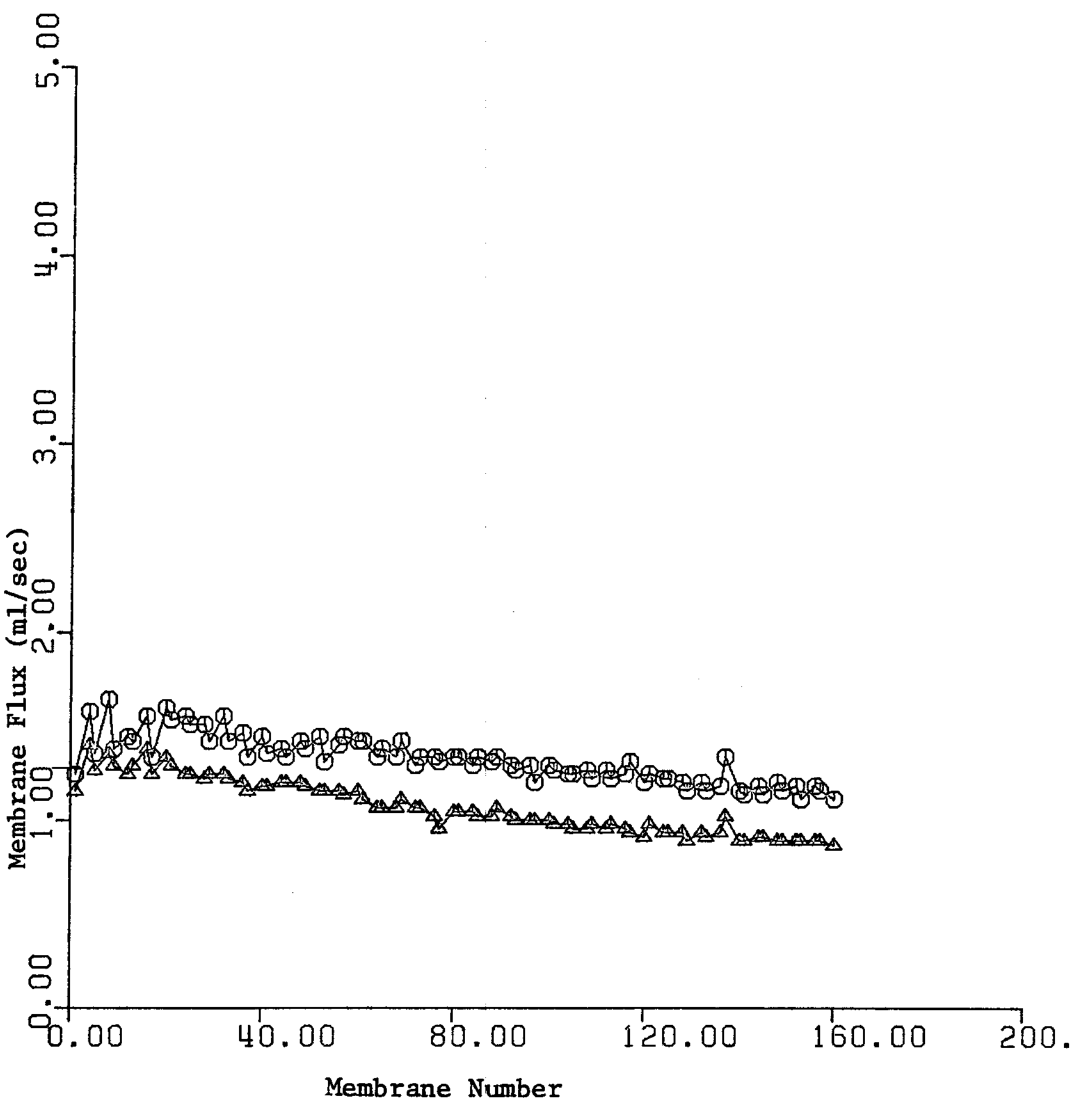


Figure 8: Membrane Flux versus Membrane Number

fluxes of the individual membranes. The lower set of data represent the fluxes an hour later. The individual membranes fluxes appear to be strongly influenced by the hydraulic characteristics of the system. This is also shown by the similar measurements shown in Figure 8. The two sets of data for Figure 8 were collected after the unit had operated for several hours and then approximately nine hours later. Though the data is insufficient to support quantitative conclusions, it appears to support the work of Kimura and Nakao (1975), who found the velocity to be more of a flux determinant than the initial membrane flux.

Also, because the data shows no sharp changes in flux, it appears that the salts are not becoming saturated and precipitating upon the membranes. This was supported by a test which showed sodium hexametaphosphate to be ineffective at reducing the flux decline.

6.2 RO Product Water Quality

RO product water quality was routinely measured for TDS and turbidity and measured quarterly for a broad spectrum of contaminants, including trace metals, boron, total organic carbon (TOC) and nutrients. Table 6 shows two such analysis of RO feedwater and product water quality. The levels of contaminants shown in the feedwater on September 17 are typical of the feedwater produced by direct filtration with an organic coagulant. The best quality feedwater was not analyzed for all the contaminants shown in Table 6, but for only a small subset of contaminants, including TOC, total suspended solids, and turbidity, which averaged approximately 15 mg/1, 4 mg/1 and 3.0

Table 6: RO Feed Water and Product Water Quality on March 19, 1979 and September 17,1979*

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

*All units are in mg/l. Values represent averages of measured water quality before and after chemical cleaning. Analysis performed by DWR.

+ as $CaCO_3$

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

NTU, respectively.

The product water quality shown in Table 6 is excellent for most reclaimed water applications and in fact is much higher than needed for many applications. The only contaminants which might be of concern for non-potable recycle uses are boron and sodium. The boron might be of concern when irrigating boron sensitive plants. The sodium concentration is not prohibitively high except that the balance between it and the divalent cations such as calcium and magnesium is poor. If the RO product water were used as the only source of irrigation water for sensitive crops, it might be advantageous to adjust the sodium to calcium ratio by adding calcium to the product water.

6.2.2 Effects of Alum and $FeCl_3$ on the Water Quality During 24 Hour Tests

The effluent from the pretreatment system was analyzed during the initial twenty four hour tests with $FeCl_3$ and Alum for TSS, turbidity, COD, and TOC. The results are shown in Figures 9 through 16 (pages 52-59). Both coagulants were effective at improving the quality of the filter effluent. The plots reflect the improved water quality which would allow the filtered water to be used directly in some recycling applications, and to be blended with RO product water where there are more stringent requirements.

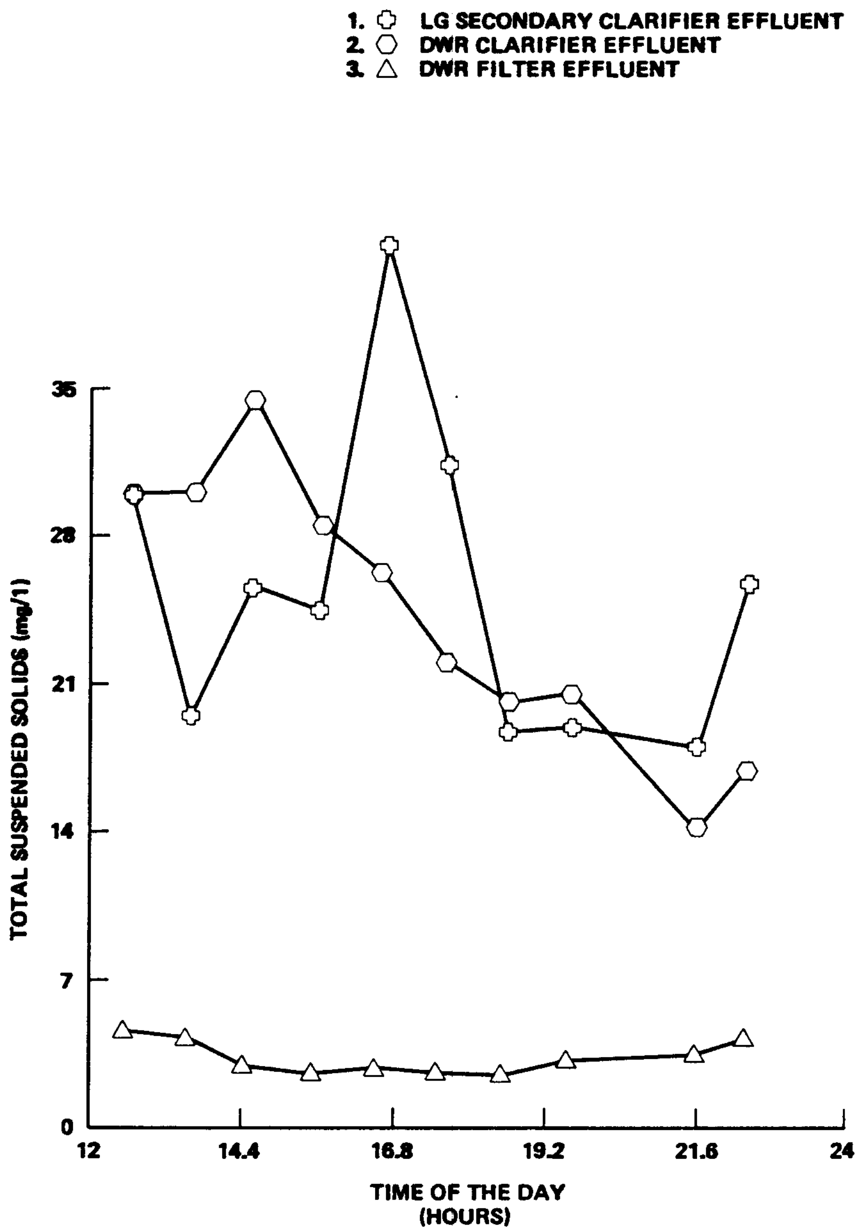


Figure 9: Total Suspended Solids versus Time Using Ferric Chloride

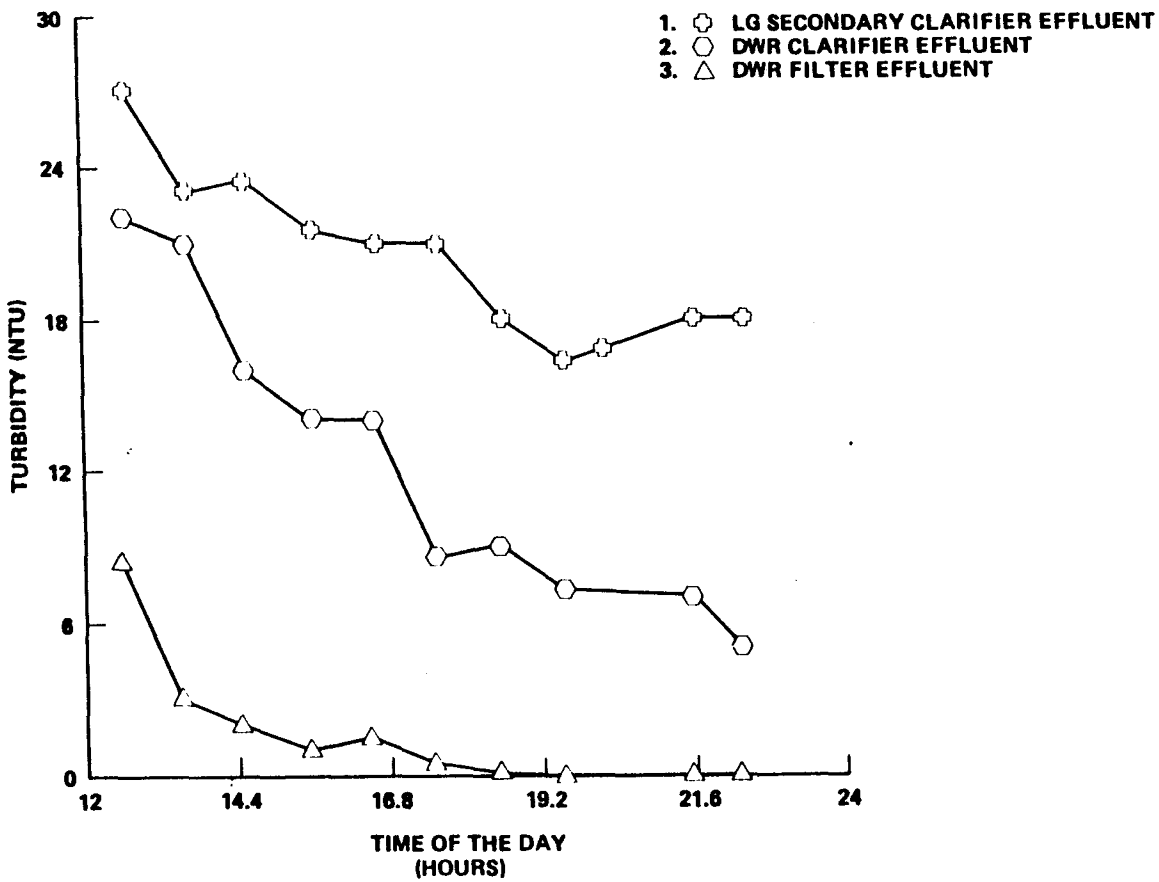


Figure 10: Turbidities versus Time Using Ferric Chloride

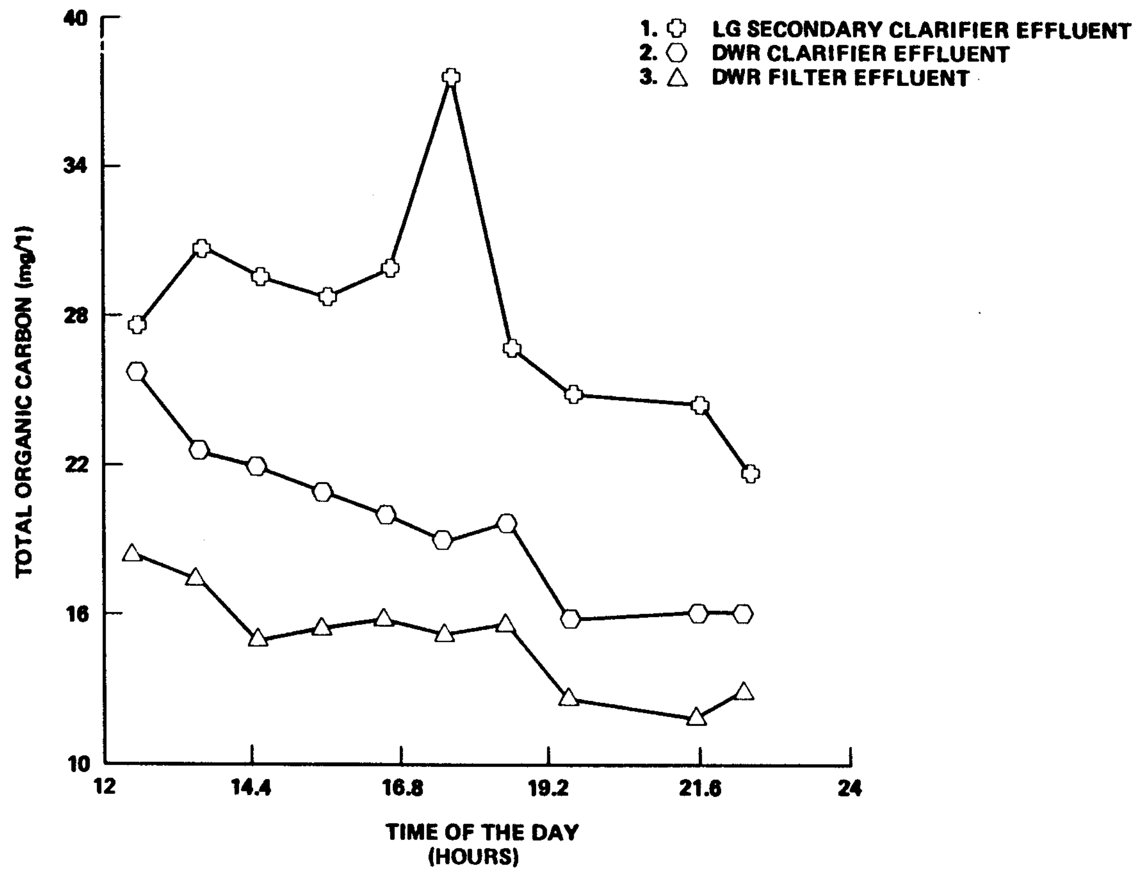


Figure 11: TOC versus Time Using Ferric Chloride

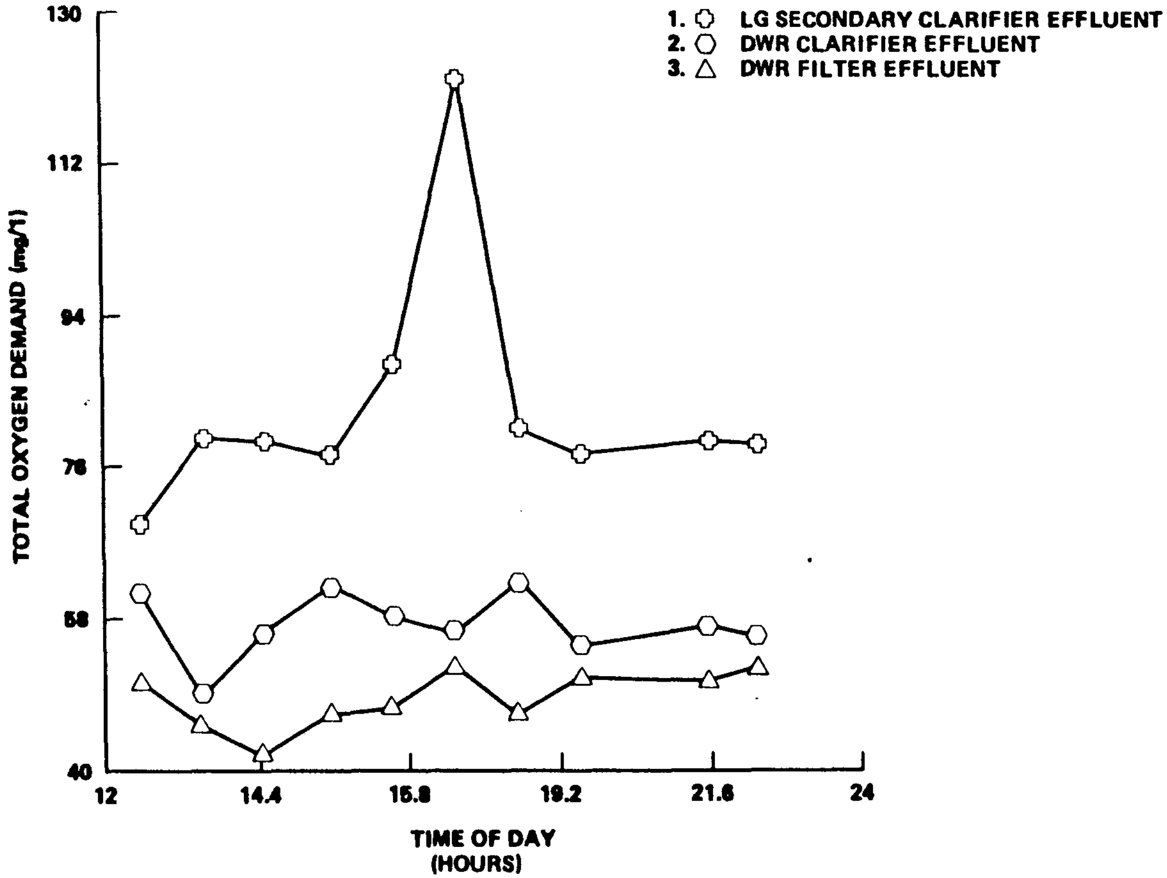


Figure 12: TOD versus Time Using Ferric Chloride

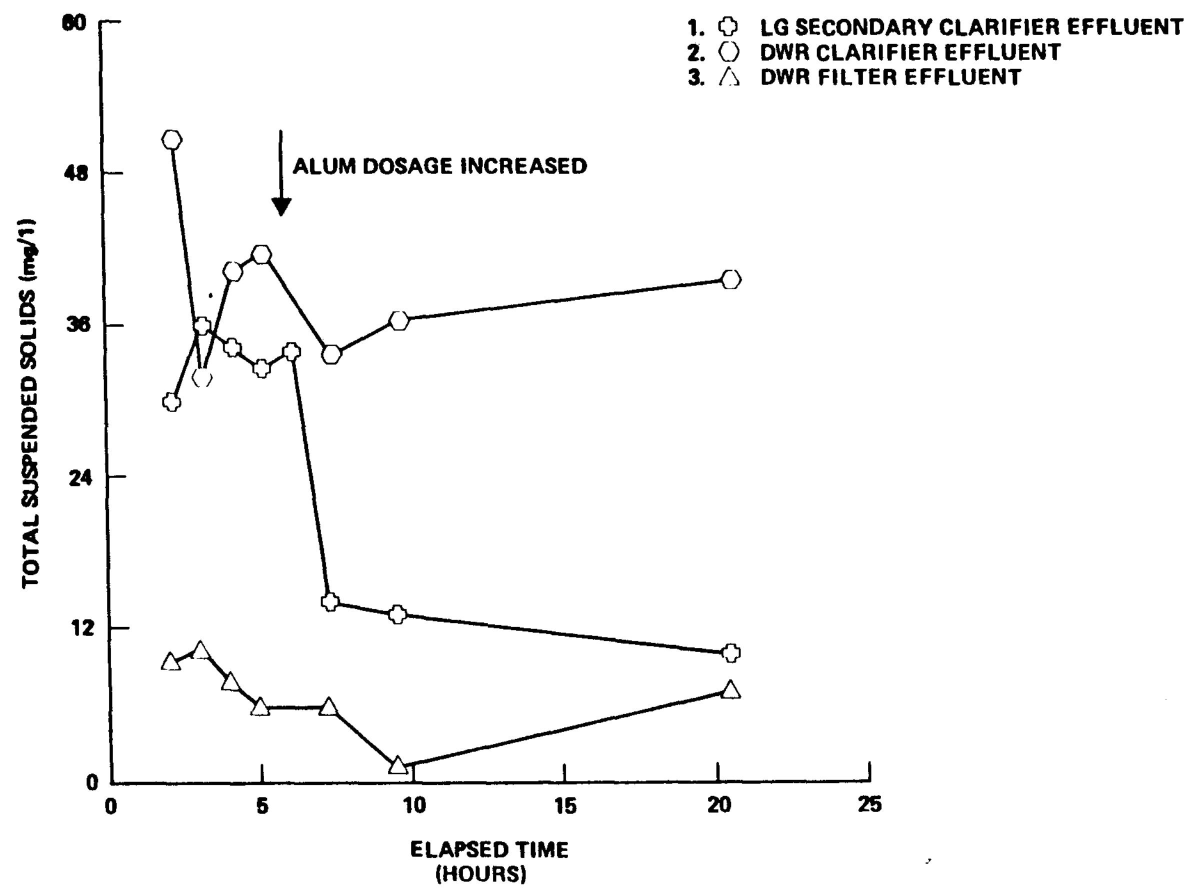


Figure 13: TSS versus Time Using Alum

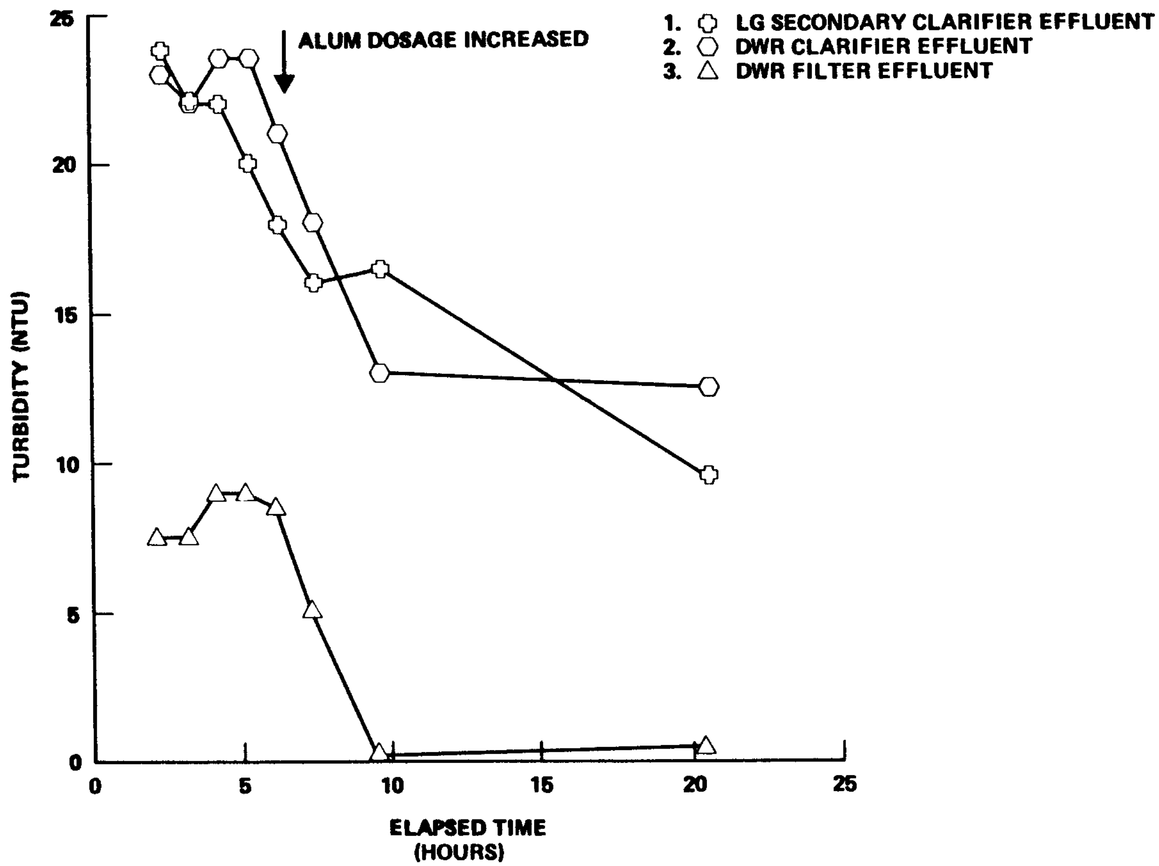


Figure 14: Turbidity versus Time Using Alum

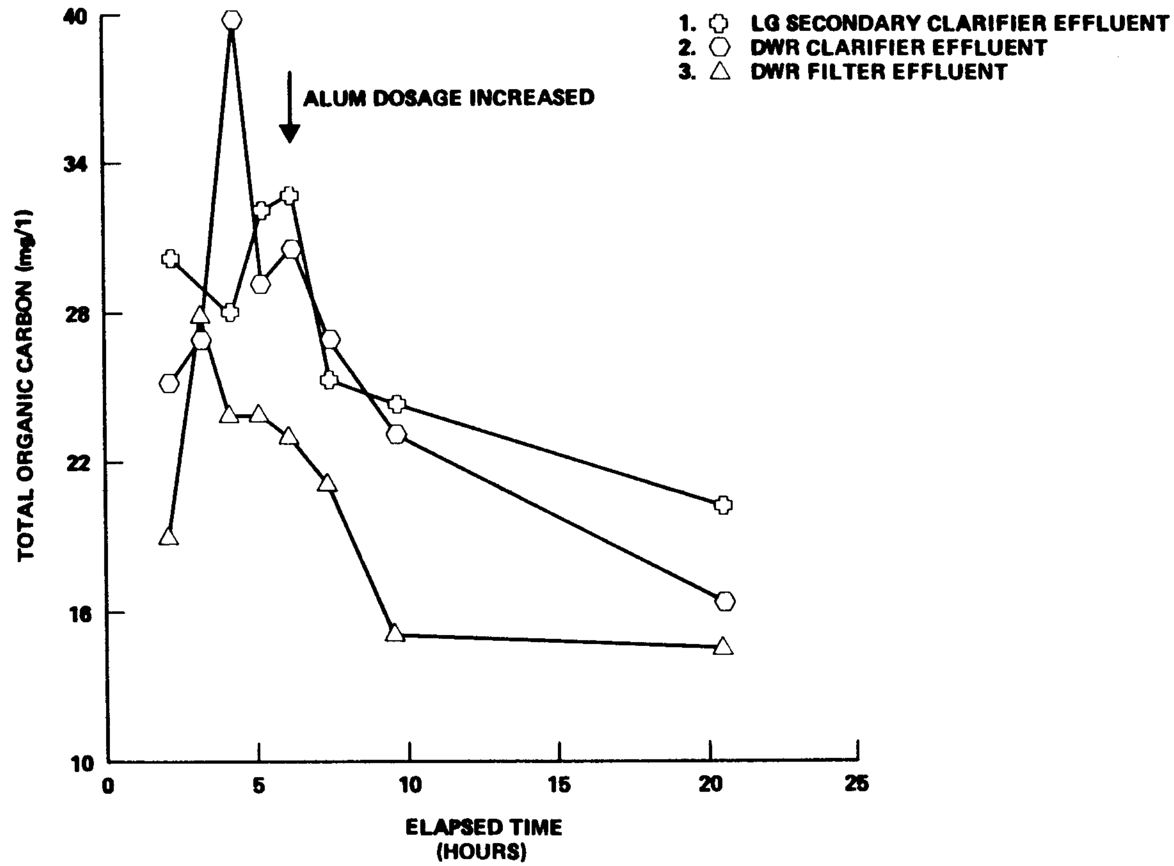


Figure 15: TOC versus Time Using Alum

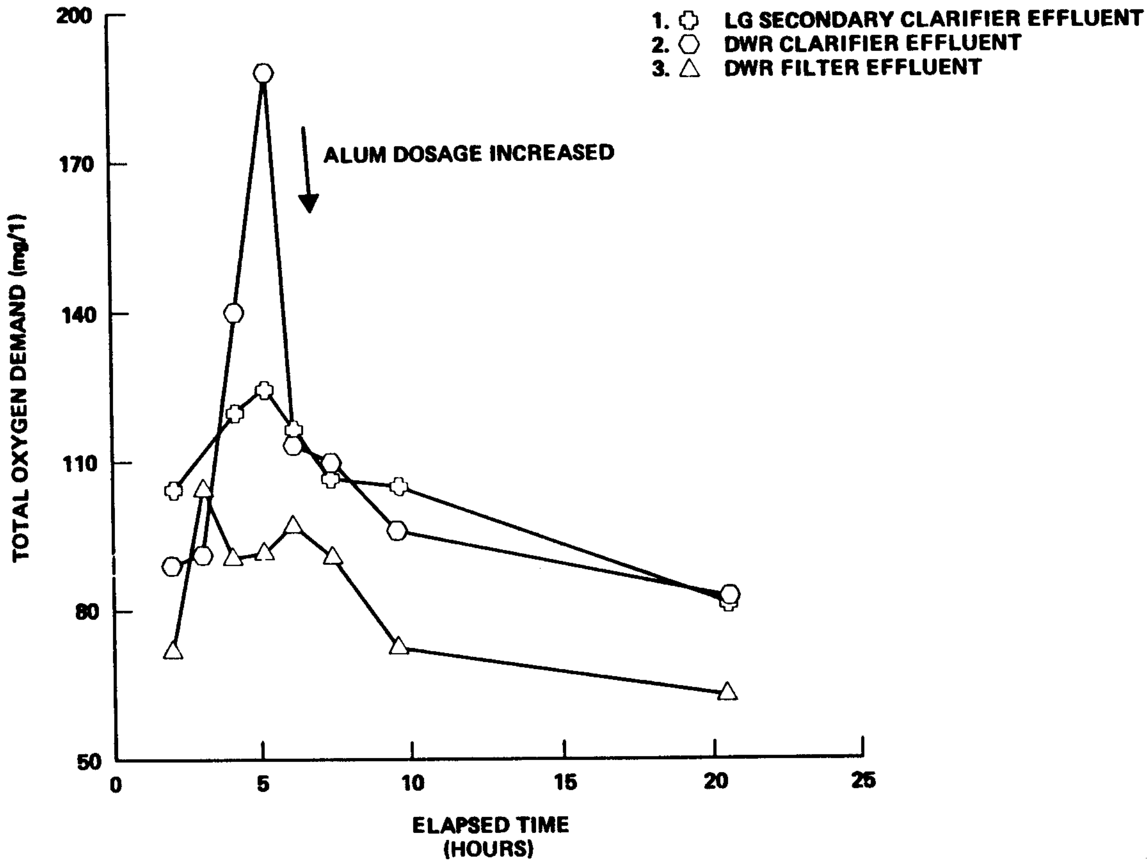


Figure 16: TOD versus Time Using Alum

6.3 Membrane Life During the Study

It has been reported by Kimura and Nakao (1975) that frequent spongeball cleaning of membranes may tend to reduce membrane life. In the present study there is no way to determine the decrease in life expectancy due to cleaning; however, it is interesting to note the membrane replacement statistics. From the start-up of the pilot plant until November, 1979, 363 membranes were replaced, not including membranes which failed within 30 days after insertion. Membranes which failed in less than 30 days were judged to be defective due to the casting or insertion technique. The average membrane life was 420 days with a standard deviation of 262 days. Membranes were replaced when product TDS was above 500 mg/l or when after cleaning flux declined to less than 5.0 GSFD (8.5 l/m^3 hr). The majority of membranes were replaced due to high product TDS concentration resulting from membrane rupture.

It appears that the major source of membrane failure is due to corrosion of end fittings and not due to hydrolysis or other types of membrane failure. The end fittings of the membranes were brass and it appears that an electrochemical corrosion reaction gradually reduced the bevel of the couplings, which eventually allowed the high pressure flow to reach the dacron backing material, producing very rapid failure. An alternate type of coupling using a PVC liner was manufactured to alleviate the corrosion problems, and where these were installed there was essentially no corrosion, and reduced membrane failure.

Chapter 7: DISCUSSION OF EXPERIMENTAL RESULTS

7.1 The Effectiveness of the Cleaning Procedures

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

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

7.2 Chlorination and pH Control

The use of chlorination and pH control appears to have a modest beneficial effect on membrane flux decline. This result is similar to the results of Winfield (1979a, 1979b) with hollow fiber membranes, Hanagi and Mori (1980) with tube membranes, and Cruver and Nusbaum (1974) and Nusbaum, Cruver, and Sleigh (1972) with spiral wound membranes. The improved flux appears to be related to inhibition of bacterial growth on the membranes, and to reduced pH where many of the scaling salts are more soluble. The fate of the chlorination byproducts in the pilot plant, and the effect of chlorine on the membranes have been investigated by Cooper, et. al., (1977, 1980).

7.3 The Benefit of the Pretreatment System

Pretreatment with coagulation, clarification, and filtration had the most dramatic effect on flux maintenance. Fluxes before cleaning increased from 8 GSFD (13.6 l/m^2 hr) to values ranging as high as 12 to 14 GSFD (20 to 24 l/m^2 hr). This flux increase is attributed to removal of the submicron colloidal particles, as evidenced by relatively ineffective results of filtration without coagulation. The amount of carry over of potentially fouling colloidal and suspended material to the RO unit, as measured during cleaning, declined dramatically with enhanced pretreatment.

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

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

Chapter 8: SYSTEMS APPROACH

8.1 Introduction

Several researchers have developed parameter optimization routines for tubular membranes. Hatfield (1967) developed a nonlinear programming model for desalination of brackish waters to maximize flux. Later Hatfield and Graves (1970) presented an improved version of the model. Fan, et. al. (1969) and McCutchan and Goel (1974) presented an optimization which included multi-staging of the units.

Goel and McCutchan (1977) then presented a model for optimizing a single pass tubular system. Antoniuk and McCutchan (1973) attempted to predict system performance for a unit operating on irrigation field water at Firebaugh, California. This same facility was also utilized by Speight and McCutchan (1979) to develop a mathematical model for economic optimization of a system.

8.1 Calculations and Assumptions for Calculating the Cost of 1 MGD Facility

Using the data collected in this study, and current cost data compiled by Oak Ridge National Laboratory (1980) and the EPA (1979), an economic simulation model has been developed. The model is based upon using the RO product water in conjunction with a specific quantity of RO feedwater, 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 feedwater to meet specified water quality standards, such as TDS, TOC, turbidity. Pretreatment level and cleaning frequency are considered as variables while RO operating pressure, membrane characteristics, and velocity are considered constant.

8.3 Basis for Cost Estimates

The least cost system of pretreatment was found by simulating the RO unit with various pretreatment systems. The basis of comparison for the simulation were the flux decline indices for the various pretreatment operations.

The flux decline indices ($\dot{B}$) were calculated for each of the pretreatment alternatives from the tests described in section 5.4.3. From this value the optimal cleaning frequency for the Las Gallinas system was determined with the least cost pretreatment system for producing 1 MGD of a given quality product water.

The costs for the system were obtained from EPA documents (1979), Oak Ridge National Laboratory (1980) and local estimates of power and labor costs. The costs of the pretreatment systems were functions assumed to be log-linear with parameters calculated from the appropriate reference. The assumption of log-linearity corresponds well with the functions in the range of 1-10 MGD. The equations for the cost functions are given in the following form:

$$\text{Log}_{10}(\text{Cost Variable}) = a * \text{Log}_{10}(\text{Size Variable}) + b. \qquad (6)$$

The functions were calculated for each of 5 categories specified in the EPA report (1979) which allows variation of the costs for labor and energy. The size variables used to represent each process are given in Table 7, the values of a and b are given in Table 8, and Table 9 shows a sample calculation.

Detailed cost estimates for reverse osmosis plants are given in Desalting Seawater and Brackish Water: Cost Update, 1979 (ORNL, 1979) as well as the EPA document, which contained the pretreatment figures. The Oak Ridge cost figures were slightly less for total costs than the costs estimated in the EPA document. Assumption of an 8% discount rate for a 20-year amoritization were used in calculations. Labor costs were placed at $12.00/hr. instead of the ten dollar figure of the EPA (1979). The EPA energy consumption estimates were used in lieu of those measured in the pilot plant, due to the relative inefficiencies of pilot plant operation, and the cost of energy was assumed to be 5¢/Kwh.

It was assumed that the flux decline relationships and the relationship of the cleaning techniques are transferable to a 1 MGD facility. It is also assumed that though the EPA costs (1979), which are based upon spiral wound and hollow fibre installations, will not necessarily equal the tubular costs, they will vary in a similar fashion with recovery rate and total capacity, thus insuring the same relationship for comparative costs.

Brine disposal costs have not been included and no function was included for variation of membrane life with changes in cleaning frequency, because there is insufficient data from which to derive a relationship. Determination of such a membrane life function would probably require years of operation with the specific system. It would appear that more frequent spongeballing would reduce membrane life, (Belfort,1976) but there is some evidence which indicates that

Table 7: Size Variables and Design Basis

FILTER	SIZE VARIABLE	DESIGN BASIS	RATE
Vessels, Tanks, Etc.	Filter Area	Loading Rate	5 GPM/Ft^2
Surface Wash		"	"
Filter Media	"	"	"
Backwash	GPM	Upflow Velocity	2 ft/min
CLARIFIER			
Vessel	Clarifier Area	Loading Rate	1000 GPD/Ft^2
Polymer	lbs/Day	Dose	5 ppm
$FeCl_3$	lbs/Hr	"	50 ppm
Alum	lbs/Hr	"	60 ppm
REVERSE OSMOSIS			
Vessels	Feed Flow	Flux Decline Index (B)	.13 -.23
Chlorine	lbs/Day	Dose	2 ppm-9 ppm
H_2SO_4	GPD	Dose	15 ppm

TABLE 8: COST COEFFICIENTS

COST SECTION	TOTAL CAPITAL ($)		ENERGY				OPERATION AND MAINTENANCE			
			Lights, Heating & Cooling		Process		Materials ($/yr)		Labor (Person Hrs/yr)	
FILTER	a	b	a	b	a	b	a	b	a	b
Tanks & Vessels	.32	4.72	210,00		.968	2.47	.785	2.47	.301	2.51
Surface Wash	.24	3.96	NA	NA	.888	1.45	.139	2.00	.486	.69
Filter Media	.65	2.55	NA	NA	NA	NA	NA	NA	NA	NA
Backwash	.37	3.49	NA	NA	1.00	1.38	.281	2.24	.062	2.146
CLARIFIER										
Vessel	.322	4.01	NA	NA	.172	3.024	.640	1.57	.154	1.736
Polymer	20,200[1]		8210[1]		17300[1]		270[1]		198[1]	
$FeCl_3$[2]	.278	4.00	.574	3.20	4900[1]		.067	2.186	.062	.067
Alum	.232	4.08	.574	3.216	4900[1]		200[1]		0.62	3.97
REVERSE OSMOSIS										
System	.848	5.89	.901	5.023	.962	6.382	.188	3.265	.886	4.988
H_2SO_4	.1186	3.82	3680[1]		1630[1]		.330	1.53	.222	1.56
Chlorine	.3625	3.752	.517	3.448	.173	2.58	.1066	2.53	.177	2.570
Cleaning	.276	3.31	NA	NA	1.0	1.077	.062	2.127	.281	2.157

1 - Value is approximately constant for considered values

2 - Based on costs for $FeSO_4$

Table 9 : Sample Calculation
Determination of Filter Vessel Capital Costs

Equation 5 shows the relationship between the cost variable and the size variable;

$$\mathrm{Log}_{10}(\text{Cost Variable}) = a * \mathrm{Log}_{10}(\text{Size Variable}) + b$$

Table 7 shows that the size variable for filter vessels is the filter area, and the loading rate is 5 GPM. If the feed flow is 1.7 mgd;

$$1.7\text{x } 10^6 \text{ gals/day} = 1.7\text{x } 10^6/1440 \text{ gals/min}$$
$$= 1180 \text{ GPM.}$$

Therefore, the filter area at 5 GPM/ft^2 is given by

$$\text{Area} = 1180 \text{ GPM}/(5 \text{ GPM}/ft^2) = 236 \text{ ft}^2.$$

Table 8 shows the values of the coefficients;

$$a = 0.32 \quad \text{and} \quad b = 4.72,$$

where the cost variable is in units of dollars[1].

Substitution into equation 5 yields the following;

$$\mathrm{Log}_{10}(\text{Cost Variable}) = .32 * \mathrm{Log}_{10}(236) + 4.72.$$

$$\mathrm{Log}_{10}(\text{Cost Variable}) = 5.478$$

Cost Variable = 301531, which is equivalent to

$ 301,531.

[1]Had the cost variable been in units of hrs/yr or Kwh/yr, the final step would be to multiply the cost variable by the $/hr for the labor or the $/Kwh for energy.

cleaner membranes have longer effective lives (Kuiper, Van Hezel, and Bom, 1974). Membrane life could have a significant effect upon the total water cost, but since the cleaning frequencies for the various systems differ only slightly, the relationship between costs would be similar.

8.4 The Optimal System

Table 10 shows a hypothetical analysis for a plant producing 1 MGD of reclaimed water. The water quality required for recycle is a TDS of less than 750 mg/l, turbidity of less than 2.0 NTU, TSS less than 5 mg/l and TOC less than 5.0 mg/l. Using the values obtained in the experimental study (RO product water: TDS - 250 mg/l, turbidity - 0.1, TOC - 1.0, and TSS = 0, with feedwater quality; TDS - 1200 mg/l, turbidity = 3.0, TOC = 18, and TSS = 4.0), a blending ratio of 0.76 MGD RO product flow to 0.24 MGD of feed blending flow is required. The size and cost of each pretreatment system, with associated flux decline coefficients, are calculated along with the optimal cleaning frequency. The optimal cleaning frequency affects RO plant size since the cleaning requires the RO unit to be down for 2.0 hours.

The optimal reclamation plant size and cost breakdown are also shown in Table 10. For this hypothetical case ferric chloride with clarification/filtration was found to be optimal and the optimal cleaning frequency was found to be once per 16 hours. The total system cost was found to be $1.55/1000 gal of reclaimed water produced. This cost is relatively insensitive to cleaning frequency over the range of once per 12 hours to once per 26 hours.

In all hypothetical cases considered thus far, the most advanced pretreatment system of these evaluated experimentally has been optimal. For this case the cost varies from a low of \$1.55/1000 gal to a high of \$2.23/1000 gal. The results of this program are considered to be tentative.

Table 10: Optimal Reclamation Plant Design

The limiting parameter for the desired water quality is the TOC and the blending ratio is .76 MGD (.033 m^3/sec) of RO product water per 1 MGD (.044 m^3/sec) of production.

The following water quality results;

	Filter	RO	Required	Blended
TDS	1200.0	250.0	750.0	473.5
NTU	3.0	0.1	2.0	0.8
TOC	18.0	1.0	5.0	5.0
TSS	4.0	0.0	5.0	0.9

Optimal System's Specifications

Filter:			
Influent Flow	:	1.37 MGD	(0.071 m^3/sec)
Loading Rate	:	5.00 GPM/ft^2	(3.4 1/m^2sec)
Filter Area	:	222 ft^2	(20.6 m^2)
Diameter	:	17. ft	(5.2 m)
Backwash Velocity	:	2.0 ft/min	(0.010 m/sec)
Clarifier:			
Influent Flow	:	1.63 MGD	(0.071 m^3/sec)
Loading Rate	:	1000 gpd/ft^2	(40.7 m^3/m^2day)
Clarifier Area	:	1602 ft^2	(148. m^2)
Diameter	:	45. ft	(13.8 m)
Coagulant	:	Ferric Chloride	
Reverse Osmosis:			
Influent Flow	:	1.37 MGD	(0.059 m^3/sec)
Product Flow	:	0.76 MGD	(0.033 m^3/sec)
Percent Recovery	:	56 %	
Flux Decline Index (B)	:	0.15	
Average Flux	:	13.58 GPD/ft^2	(23.0 1/m^2/sec)
Number of Membranes	:	24484	
Total Membrane Area	:	56314 ft^2	(5231 m^2)
Time Req'd for Cleaning	:	2.0 hrs	
Optimal Cleaning Interval	:	16.0 hrs	
Sulfuric Acid Injected	:	8.19 ppm	
Chlorine Injected	:	2.00 ppm	

Assumptions

Labor	=	\$12.00 /hr
Electrical Rate	=	\$ 0.05 /Kwh
Interest Rate	=	8 %
Life of Project	=	20 years

Table 10 Optimal Reclamation Plant Design (Continued)

Unit Costs

	Total Capital ($)	Total Capital ($/yr)	O & M ($/yr)	Water Costs ($/kgal)	Water Costs ($/m^3)
Vessels	295873.	30135.	26436.	0.17	0.045
Surface Wash	32986.	3360.	1199.	0.01	0.003
Media	12144.	1237.	0.	0.004	0.001
Backwash	62114.	6326.	3408.	0.03	0.008
Clarifier	110138.	11218.	2645.	0.04	0.018
Coagulants	25312.	2578.	4219.	0.02	0.005
RO Unit	1011413.	103015.	315734.	1.23	0.325
H_2SO_4	8841.	900.	1181.	0.01	0.003
Chlorination	5904.	601.	4668.	0.02	0.005
Cleaning	32496.	3310.	6038.	0.03	0.008

Coagulation/Clarification/Filtration	0.27 $/Kgal	(0.07 $/m^3)
Reverse Osmosis System	1.28 $/Kgal	(0.34 $/m^3)
Total System	1.55 $/Kgal	(0.41 $/m^3)

Energy and Labor Analysis

	Building Energy		Process Energy			
	Kwh/yr	Cost($/yr)	Kwh/yr	Cost($/yr)	Hrs/yr	$/yr
Filter Unit	41176.	2059.	55220.	2761.	1646	19756.
Surface Wash	0.	0.	3484.	174.	68	813.
Media	0.	0.	0.	0.	0	0.
Backwash	0.	0.	5336.	267.	196	2384.
Clarifier Unit	0.	o.	3760.	188.	170	2036.
Coagulant	2114.	106.	4900.	245.	306	3674.
RO Unit	27387.	1369.	3253701.	162685.	1952	23424.
Sulfuric Acid	722.	36.	1630.	82.	76	914.
Chlorine	2988.	149.	388.	19.	343	4119.
Cleaning	0.	0.	2267.	113.	185	2226.

Chapter 9: CONCLUSIONS AND FUTURE RESEARCH

An experimental and economic analysis of a 10 GPM pilot RO plant has been presented. From the experimental study the following conclusions are made:

1. Cleaning using flushes of citric acid, followed by enzyme detergent and spongeball cleaning are effective at maintaining membrane flux to essentially the initial flux levels. The citric acid is the major cleaning agent. Enzyme detergent and/or spongeball cleaning without citric acid are relatively ineffective.

2. Automatic spongeball cleaning technique appears to have promise for maintaining membrane flux between chemical cleanings. Further testing is required.

3. The major factor contributing to membrane degradation for the type of membrane construction shown in Figure 1 is corrosion, resulting in an average membrane life of 420 days (10,000 hours). This will be tested in future studys which will use the PVC liners at the couplings.

4. For all the conditions tested and analyzed, the greatest level of pretreatment provided the most economical operation.

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

5. For the optimal conditions investigated, reclaimed water containing approximately 500 mg/l TDS can be produced for approximately \$0.42/m^3 (\$1.60/1000 gal). The costs estimated are considered tentative.

6. The pretreatment system had a significant effect on energy consumption. The optimal energy requirements were approximately 2.6 KWhr/m^3 (10 KWhr/1000 gal).

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

It is anticipated that the existing pilot plant will be operated for at least two more years. During this time several major modifications are planned to further observe the units performance. To test the RO system on activated sludge effluent, installation of a pilot activated sludge plant is planned for the fall of 1980.

The system will be operated with increasingly consistent pretreatment and cleaning procedures, and thus a more accurate appraisal of membrane life will be possible. Investigation into velocity relationships, (which have largely been ignored in this study) are planned and possibly a test of a parallel leg configuration. Additionally, several novel flux control techniques, such as those proposed by Belfort and Marx (1979) are being considered.

APPENDIX A - REFERENCES

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APPENDIX B - FORTRAN PROGRAM

```
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  THREE 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           7% FOR 20 YEARS.
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 PP(20),BQ(100),BX(100)
      COMMON /RO/ CT,F1,REC,AFD,AF,MA,RECOV,RFEED,CF,RF,MK
      COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,FF,V
      COMMON /C/ CRATE,CIN,CSIZE
      COMMON /UNIT1/ B,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)
      COMMON /UNIT3/ E(20),AM(20),AL(20),EC(20),ALC(10),SUML,ESUM,ECSUM,
     1SUMCL,SUMM,PKWH,HOURLY,AOPT(10),BOPT(10),AI
      COMMON /OPTIM/ KBOPT,CFOPT,TMAX
      EQUIVALENCE (CCF,PP(1))
      REAL MA
      DATA JOKER/1/,A/1000000./,INDEX/0/,P/1./
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 F1 & REC = THE FLOW AND THE RECOVERY AFTER CLEANING
C FRATE AND CRATE ARE THE LOADING RATES FOR THE FILTER AND CLARIFIER
```

```
      TMAX = 99999999.
      NHRS = 10
      MK = 4
      FEED = 6.2
      CT = 2.
      F1 = 5.2
      REC = F1/FEED
      FRATE = 5.
      CRATE = 1000.
      V = 2.0
      CF = 4.0
      BB(1) = .1456
      BB(2) = .2060
      BB(3) = .2020
      BB(4) = .2425
C
    1 DO 50 KM = 1,4
      DO 40 J = 1,NHRS
C
C  INITIALIZATION
      DO 10 I = 1,20
      PP(I) = 0.0
      E(I) = 0.
      AM(I) = 0.
      AL(I) = 0.
   10 CONTINUE
C
      B = BB(KM)
      CALL FLUX(B)
      IF(P.GT.10.) GO TO 30
   20 CALL BLEND
      P = 20.
   30 CALL SIZE
      CALL CLEAN
      KGB = KM
      CALL COST(KGB)
      TABLE(KM,J) = TCOST
      CF = CF + 4.
   40 CONTINUE
      CF = 4.
   50 CONTINUE
C  INITIALIZATION
      DO 55 I = 1,20
      PP(I) = 0.0
      E(I) = 0.
      AM(I) = 0.
      AL(I) = 0.
   55 CONTINUE
C  COMPUTE OPTIMAL VALUES
      KGB = KBOPT
      KM = KGB
      B = BB(KM)
      CF = CFOPT
```

```
      CALL FLUX(B)
      CALL SIZE
      CALL CLEAN
      CALL COST(KGB)
   60 CALL OUTPUT(NHRS)
  100 STOP
      END
      SUBROUTINE CLEAN
C
C                          SUBROUTINE CLEAN
C
      COMMON /RO/ CT,F1,REC,AFD,AF,MA,RECOV,RFEED,CF,RF,MK
      COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,FF,V
      COMMON /C/ CRATE,CIN,CSIZE
      COMMON /UNIT1/ B,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)
      COMMON /UNIT3/ E(20),AM(20),AL(20),EC(20),ALC(10),SUML,ESUM,ECSUM,
     1SUMCL,SUMM,PKWH,HOURLY,AOPT(10),BOPT(10),AI
      COMMON /OPTIM/ KBOPT,CFOPT,TMAX
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
      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
      SUBROUTINE FLUX(B)
C
C                     SUBROUTINE FLUX
C
C FSUM IS THE TOTAL FLOW BETWEEN CLEANINGS
C CF = THE CLEANING FREQUENCY
C CT = CLEANING TIME
C F1 = 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,F1,REC,AFD,AF,MA,RECOV,RFEED,CF,RF,MK
      COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,FF,V
```

```
      COMMON /C/ CRATE,CIN,CSIZE
      COMMON /UNIT2/ CCFU,CCFS,CCFM,CCFB,CCC,CCA,CCR,CCB,CCD,CCL
      REAL MA
      FMAX = 0.0
      FSUM = 0
      KK = CF
      F(1) = F1
      N = KK +1
      DO 10 I = 2,N
      T(I) = FLOAT(I)
      F(I) = F1 * (1/T(I)**B)
      FSUM = FSUM +(F(I) + F(I-1))/2.
   10 CONTINUE
      AF = FSUM/(KK + CT)
      RECOV = AF * REC/F1
      AFD = AF * 1440. /368.
      CF = FLOAT (KK)
      RETURN
      END
      SUBROUTINE BLEND
C
C                        SUBROUTINE BLEND
C
      DIMENSION PAR(2,20),FPAR(20),ROPAR(20),CRIT(20)
      COMMON /RO/ CT,F1,REC,AFD,AF,MA,RECOV,RFEED,CF,RF,MK
      COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,FF,V
      COMMON /C/ CRATE,CIN,CSIZE
      COMMON /UNIT2/ CCFU,CCFS,CCFM,CCFB,CCC,CCA,CCR,CCB,CCD,CCL,BNN(10)
      REAL MA
      N = 4
      XX = 0.0
C  LOOP FOR CALCULATING THE BLENDING RATIO
      DO 1 K= 1,N
      IF(K.GT.N) GO TO 1
      READ(5,10) CRIT(K),PAR(1,K),FPAR(K),ROPAR(K)
   10 FORMAT(A3,3F8.1)
      IF(FPAR(K).LE.PAR(1,K)) GO TO 20
      IF(ROPAR(K).GT.PAR(1,K)) GO TO 30
      RATIO = (PAR(1,K) - FPAR(K))/(ROPAR(K) - FPAR(K))
      IF(RATIO.LT.XX) GO TO 1
      XX = RATIO
      KK = K
      GO TO 1
   20 RATIO = 0
      KK = 1
      GO TO 1
   30 WRITE(6,40) CRIT(KK),CRIT(KK),PAR(1,K)
   40 FORMAT(//////,T20,'THE RO PRODUCT WATER HAS A ',A3,' GREATER THAN
     CTHAT REQUIRED',/,'THEREFORE WATER WITH ',A3,' LESS THAN',F8.1,'
     CCANNOT BE PRODUCED WITH THIS SYSTEM')
      STOP
    1 CONTINUE
```

```
      RATIO = XX
      DO  2 I = 1,N
      PAR(2,I) = (RATIO*ROPAR(I)) + ((1-RATIO) * FPAR(I))
    2 CONTINUE
      WRITE(6,50) CRIT(KK),(CRIT(K),FPAR(K),ROPAR(K),
     1(PAR(I,K),I=1,2),K=1,4)
   50 FORMAT(///,T20,'THE LIMITING PARAMETER IS ',A3,/,
     1//,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))
      RETURN
      END
      SUBROUTINE SIZE
C
C                        SUBROUTINE SIZE
C
C     CALCULATES THE SIZES OF THE VARIOUS COMPONENTS OF THE SYSTEM
      COMMON /RO/ CT,F1,REC,AFD,AF,MA,RECOV,RFEED,CF,RF,MK
      COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,FF,V
      COMMON /C/ CRATE,CIN,CSIZE
      COMMON /UNIT2/ CCFU,CCFS,CCFM,CCFB,CCC,CCA,CCR,CCB,CCD,CCL,BNN(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(KGB)
C
C                        SUBROUTINE COST
C
      COMMON /RO/ CT,F1,REC,AFD,AF,MA,RECOV,RFEED,CF,RF,MK
      COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,FF,V
      COMMON /C/ CRATE,CIN,CSIZE
      COMMON /UNIT1/ B,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)
      COMMON /UNIT3/ E(20),AM(20),AL(20),EC(20),ALC(10),SUML,ESUM,ECSUM,
     1SUMCL,SUMM,PKWH,HOURLY,AOPT(10),BOPT(10),AI
      COMMON /OPTIM/ KBOPT,CFOPT,TMAX
```

```
      DIMENSION ANN(10)
      EQUIVALENCE (CCFU,ANN(1))
      REAL MA
      DATA N/20/,CFT/20./
      AI = 0.08
      PKWH = .05
      HOURLY = 12.
      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***********************************************************
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 = 10.**ECOST
        E(1)   = RFT * 210000.
        E(2)   = 10. **(0.968 * ALOG10(FSIZE) + 2.47)
       AM(1)   = 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 = 10. ** ECOST
        E(3)   = 0.0
        E(4)   = 10. **(0.8877* ALOG10(FSIZE) + 1.4585)
       AM(2)   = 10. **(0.139 * ALOG10(FSIZE) + 2.0)
       AL(2)   = 10. **(0.486 * ALOG10(FSIZE) + 0.69)
C
C                               MEDIA
C---------------
      CCFM = 10. **(0.652 * ALOG10(FSIZE) + 2.554)
        E(5) = 0.
        E(6) = 0.
       AM(3) = 0.
       AL(3) = 0.
C
C                               BACKWASH
C-----------------
      Q = FSIZE *  V * 7.48
      CCFB = 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)   = 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
      IF(KGB.NE.4) GO TO 10
      K = 1
      GO TO 40
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-------------
   10 ECOST = 0.322 * ALOG10(CSIZE) + 4.01
      CCC = 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) = 10. **(0.640 * ALOG10(CSIZE) + 0.574)
C
C                           COAGULANTS
C              BRANCH TO THE PROPER CHEMICAL FEED
C
      GO TO (20,25,30),KGB
C
C                     NALCO 7134 @ 10PPM
C------------------
   25 CCA = 20200
      DOSE =  5
      PRICE = 1.50
      AMT = WF * DOSE * PRICE*340.
        E(11)= RFT * 8210.
        E(12)= 17300.
       AL(6) = 198.
       AM(6) = 270. + AMT
      K = 2
      GO TO 40
C
C                     FECL3 @ 50 PPM
C------------------
   20 DOSE = 50
      AMT = WF * DOSE/24.
      PRICE = .10
      CCA = 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 = 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) = 200. + PRICE * AMT
      K = 4
C***************************************************************
C*                   REVERSE OSMOSIS                           *
C***************************************************************
C CCR = CAPITAL COSTS
C
   40 CCR = 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) = 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 = 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) = 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 = 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)  = 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 + CCR + 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
C                SELECT LEAST TOTAL ANNUAL COST
C-------------------
      IF(TCOST.LT.TMAX) GO TO 80
      RETURN
   80 KBOPT = KGB
      CFOPT = CF
      TMAX = TCOST
      DO 90 I = 1,10
      BOPT(I) = EC(2*I-1) +EC(2*I) + ALC(I) + AM(I)
      AOPT(I) =(BNN(I)+BOPT(I))/340000
   90 CONTINUE
      RETURN
      END
      SUBROUTINE OUTPUT(NHRS)
C
```

```
C               SUBROUTINE OUTPUT
C
      COMMON /RO/ CT,F1,REC,AFD,AF,MA,RECOV,RFEED,CF,RF,MK
      COMMON /FILTER/ FRATE,FOUT,FIN,FSIZE,RATIO,FF,V
      COMMON /C/ CRATE,CIN,CSIZE
      COMMON /UNIT1/ B,BB(20),TABLE(10,500),K,KM,TCOST,DOSE,ADOSE,CDOSE
      COMMON /UNIT2/ BN(10),BNN(10)
      COMMON /UNIT3/ E(20),AM(20),AL(20),EC(20),ALC(10),SUML,ESUM,ECSUM,
     1SUMCL,SUMM,PKWH,HOURLY,AOPT(10),BOPT(10),AI
      DIMENSION T(4,4),AA(4,3),G(10,3)
      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
      READ(5,5) ((T(I,J),I=1,4),J=1,4)
    5 FORMAT(4(1X,4A4))
C----------------------ASSUMPTIONS-------------------------------
      WRITE(6,220) HOURLY,PKWH,AI
  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',//)
C*******************************************************************
C         TABLE OF ANNUAL COSTS AND FLUX DECLINE INDEX
C
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 I = 1,MK
      DO 30 J = 1,NHRS
      TABLE(I,J) = TABLE(I,J)/340000.
   30 CONTINUE
      WRITE(6,40) (I,I=4,40,4)
   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(I),(TABLE(I,J),J=1,10),I=1,MK)
   50 FORMAT(/,T5,F5.2,T10,10F6.2)
```

```
C*********************************************************************
C*********          OPTIMAL SYSTEM          ********************
C*********                                  *********************
C**************************************************************************
C---BRANCH TO OPTIMAL DESIGN -----
C
      GO TO (60,80,100,120),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 RECOV = RECOV * 100.
      NM = IFIX(RFEED/(AFD*2.3))
      RFEED = RFEED/(10.**6)
C------------REVERSE OSMOSIS SPECIFICATIONS---------------------
      WRITE(6,180) RFEED,RATIO,RECOV,B,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 INCEX (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  $/YR')
C-----------TOTAL AND ANNUAL COSTS OF UNIT PROCESSES-------------------
C
      WRITE(6,210) (E(2*J-1),EC(2*J-1),E(2*J),EC(2*J),AL(J)
     1,ALC(J),J=1,10)
  210 FORMAT(/,T7,'FILTER UNIT',6F9.0,//,T6,'SURFACE WASH',6F9.0,
     1//,T12,'MEDIA',6F9.0,//,T9,' BACKWASH',6F9.0,//,
     2T4,'CLARIFIER UNIT',6F9.0,//,T8,' COAGULANT',6F9.0,//,
     3T3,'REVERSE OSMOSIS',6F9.0,//,
     4T5,'SULFURIC ACID',6F9.0,//,T9,' CHLORINE',6F9.0,//,
     5T10,'CLEANING',6F9.0)
      READ(5,213) ((G(J,I),I=1,3), J=1,10)
  213 FORMAT(10(3A4,/))
      WRITE(6,214)
      WRITE(6,215) ((G(J,I),I=1,3),BN(J),BNN(J),BOPT(J),AOPT(J), J=1,10)
  214 FORMAT(//,T35,'UNIT COSTS',//,T15,'TOTAL CAPITAL($/YR)',T37,'ANNUA
     1L CAPITAL',T54,'O & M ($/YR)',T67,'WATER COSTS($/KGAL)')
  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(I)
```

```
      GO TO 217
  216 RCAP = RCAP + AOPT(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---------------------------
      CKG = TCOST/340000.
      WRITE(6,230) TCOST,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
C     SUBROUTINE JDPLOT(JD,XVALS,YVALS,X1,Y1,KM)
C     COMMON /SCLR/FX,DX,FY,DY,ITESTX,ITESTY
C     DIMENSION XVALS(100),YVALS(100),Z(105)
C     DX = X1
C     FX = 0
C     FY = 0
C     DY = Y1
C     ITESTX = 1
C     ITESTY = 1
C     IF(KM.EQ.1) CALL CPLOT(3,JD,-1,XVALS,YVALS,1,1,Z)
C     CALL CPLOT(3,JD,1,XVALS,YVALS,1,1,Z)
C     IF(KM.EQ.4)   CALL CPLOT(3,JD,5,XVALS,YVALS,1,1,Z)
C     RETURN
C     END
*RUN
```

APPENDIX C - DAILY RECORDS FOR LAS GALLINAS FACILITY

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
04/27/76				10.00	16.51	0.42
04/28/76				0.00	0.00	
04/29/76				0.00	12.2	
04/30/76				0.00	10.96	
05/01/76				0.00	10.76	
05/02/76				0.00	10.33	
05/03/76				0.00	10.29	
05/04/76				0.00	10.02	
05/05/76				0.00	9.63	
05/06/76				0.00	8.18	
05/07/76				0.00	8.06	
05/08/76				0.00	7.55	
05/10/76				0.00	5.99	
05/11/76				0.00	5.56	
05/12/76				0.00	12.72	
05/13/76				0.00	11.19	
05/14/76				0.00	0.00	
06/30/76				0.00	15.30	
07/01/76				0.00	11.11	
07/02/76				0.00	10.21	
07/03/76				0.00	11.19	
07/06/76				0.00	8.18	
07/07/76				0.00	7.63	
07/08/76				0.00	6.89	
07/09/76				0.00	11.50	
07/10/76				0.00	8.57	
07/12/76				0.00	8.30	
07/13/76				0.00	0.00	
07/14/76				0.00	9.63	
07/15/76				0.00	8.22	
07/16/76				0.00	0.00	
07/21/76				0.00	0.00	
07/23/76				0.00	0.00	
07/24/76				0.00	7.43	
07/26/76				0.00	7.12	
07/30/76				0.00	0.00	
07/31/76				0.00	5.56	
08/02/76				0.00	4.85	
08/03/76				0.00	4.50	
08/04/76				0.00	9.78	

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT FLUX (GSF)	RECOVERY
08/05/76				0.00	16.33	
08/25/76				0.00	12.40	
08/26/76				0.00	0.00	
08/27/76				0.00	10.85	
08/28/76				0.00	10.44	
08/29/76				0.00	0.00	
09/28/76				0.00	19.44	
09/29/76				0.00	13.54	
09/30/76				0.00	17.41	
10/01/76				0.00	11.78	
10/02/76				0.00	10.25	
10/03/76				0.00	9.82	
10/04/76				0.00	15.50	
10/05/76				0.00	16.63	
10/06/76				0.00	11.39	
10/07/76				0.00	13.70	
10/08/76				0.00	9.51	
10/11/76				0.00	11.94	
10/11/76				0.00	14.21	
10/12/76				0.00	10.02	
10/13/76				0.00	10.64	
10/14/76				0.00	9.20	
10/15/76				0.00	8.33	
10/15/76				0.00	9.23	
10/16/76				0.00	7.54	
10/17/76				0.00	5.97	
10/18/76				0.00	4.70	
10/18/76				0.00	11.19	
10/19/76				0.00	6.79	
10/20/76				0.00	6.41	
10/21/76				0.00	6.18	
10/22/76				0.00	5.83	
10/23/76				0.00	6.51	
10/24/76				0.00	5.35	
10/25/76				0.00	4.93	
10/25/76				0.00	9.04	
10/26/76				0.00	7.12	
10/27/76				0.00	6.81	
10/28/76				0.00	6.22	
10/29/76				0.00	5.68	

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
10/30/76				0.00	6.22	
10/31/76				0.00	5.87	
11/01/76				0.00	5.36	
+11/01/76	1150	1350	25.0	6.09	9.16	0.38
11/02/76	1050	1450	87.5	6.07	7.16	0.30
11/03/76	1125	1550	92.5	6.04	6.89	0.29
11/04/76	1100	1200	95.0	5.97	6.14	0.26
11/05/76	1000	1400	85.0	5.87	6.42	0.28
11/06/76	1325	2300	27.5	5.62	6.10	0.28
11/07/76	1250	1650	05.0	5.97	5.99	0.26
11/08/76	1175	1550	95.0	5.77	5.32	0.24
+11/08/76	1275	1850	20.0	6.11	7.36	0.31
11/09/76	1150	1600	85.0	6.20	6.93	0.29
11/10/76	1350	1900	05.0	5.95	6.73	0.29
11/11/76	1100	1550	90.0	6.09	6.18	0.26
11/15/76	1200	1600	12.5	6.33	8.45	0.34
11/16/76	1075	1550	26.0	0.00	0.00	
11/16/76	1075	1550	26.0	6.24	6.50	0.27
+11/16/76	1100	1600	90.0	6.53	7.51	0.29
11/17/76	950	1350	75.0	6.50	6.89	0.27
11/18/76	1050	1450	90.0	6.14	6.57	0.27
11/19/76	900	1100	82.5	5.91	6.18	0.27
11/20/76	1300	1800	55.0	6.03	6.14	0.26
11/21/76	1500	1900	62.5	5.95	5.71	0.25
11/22/76	1375	1750	32.5	5.84	5.48	0.24
11/23/76	1200	2300	92.5	5.30	4.85	0.23
11/24/76	1800	2250	80.0	5.44	4.58	0.22
+11/30/76	950	1300	00.0	5.64	6.57	0.30
12/01/76	900	1250	70.0	5.35	5.56	0.27
12/02/76	1050	1400	95.0	6.24	5.40	0.22
12/03/76	1100	1600	97.5	5.63	4.89	0.22
12/04/76	1150	1960	10.0	5.83	5.52	0.24
12/05/76	1000	1300	82.5	5.76	5.17	0.23
+12/06/76	1250	1800	32.5	5.96	6.89	0.30
12/06/76	1150	1525	95.0	6.76	5.83	0.22
12/07/76	1000	1300	12.5	6.14	5.32	0.22
12/08/76	1300	1700	07.5	5.86	5.36	0.23
12/09/76	1300	1350	12.5	5.84	5.32	0.23
12/10/76	1250	1500	12.5	7.03	4.70	0.17
12/11/76	1150	1450	17.5	5.93	4.81	0.21

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA (1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT FLUX (GSF)	RECOVERY
12/12/76	1200	1700	10.0	6.28	4.85	0.20
12/13/76	1100	1400	05.0	5.44	4.30	0.20
+12/13/76	1150	1450	60.0	5.63	4.58	0.21
12/14/76	1000	1700	95.0	5.43	4.93	0.23
12/15/76				0.00	0.00	
12/16/76	1050	1100	72.5	7.91	5.01	0.16
12/17/76	1000	1200	67.5	6.29	5.09	0.21
12/18/76	1800	1400	10.5	5.73	4.77	0.21
12/19/76				0.00	0.00	
12/20/76	1450	2100	35.0	5.43	3.99	0.19
12/21/76	1700	2300	55.0	5.62	3.80	0.17
12/22/76	1900	2200	97.5	8.17	3.40	0.11
01/24/77	1150	1450	62.5	10.28	8.80	0.22
01/25/77	1150	1450		8.36	7.26	0.22
01/26/77	1050	1300	02.5	8.25	6.81	0.21
01/26/77	1300	1550	54.0	8.41	7.10	0.22
01/27/77				8.50	6.42	0.19
01/28/77	1150	1350	35.0	8.67	5.40	0.16
01/29/77	1325	1500	65.0	8.56	4.77	0.14
01/30/77	1350	1500	85.0	7.66	3.95	0.13
01/31/77	1175	1400	52.0	7.81	3.48	0.11
+01/31/77	1400	1750	46.5	8.51	7.28	0.22
02/01/77	1300	1750	40.0	8.71	5.71	0.17
02/02/77	1500	1800	60.0	8.28	5.40	0.17
02/03/77	1600	1850	60.0	8.37	5.20	0.16
02/04/77	1600	1900	70.0	7.91	4.93	0.16
02/05/77	1650	1950	77.5	7.51	4.81	0.16
02/06/77	1650	2300	95.0	7.40	4.77	0.16
02/07/77	1550	1850	75.0	7.29	4.27	0.15
+02/07/77		2000	92.5	7.18	6.56	0.23
02/08/77	1600	2350	97.5	7.21	4.97	0.18
02/09/77	1450	1800	87.5	7.51	4.66	0.16
02/10/77	1550	1800	87.5	7.45	4.46	0.15
02/11/77	1600	2150	75.0	6.79	4.46	0.17
02/14/77	1500	1900	90.0	7.44	7.83	0.27
02/15/77	1500	2200	77.0	7.29	5.48	0.19
02/15/77	1600			0.00	0.00	
03/14/77	1475			0.00	0.00	
+03/14/77	1500	2050	90.0	8.59	7.65	0.23
03/15/77	1450	1725	38.0	8.56	5.81	0.17

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

	DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
	03/16/77	1050	1400	17.5	8.52	6.81	0.20
	03/17/77	1525	1925	62.5	8.64	6.07	0.18
	03/17/77	1600	1950	65.0	8.66	6.35	0.19
	03/18/77	1625	1900	67.5	8.88	5.63	0.16
	03/19/77	1700	1975	95.0	8.61	5.71	0.17
	03/20/77	1700	1950	85.0	8.70	4.81	0.14
	03/21/77	1750	2000	65.0	8.82	5.48	0.16
	03/22/77	1625	1850	60.0	8.65	4.29	0.13
+	03/22/77	1650	2050	45.0	7.89	6.89	0.22
	03/23/77	1650	1950	70.0	8.06	5.24	0.17
	03/24/77	1600	1900	50.0	8.03	4.70	0.15
	03/25/77	1500	1750	50.0	7.92	4.89	0.16
	03/29/77	1550	2200	22.5	5.66	6.07	0.27
	03/30/77	1475	1825	67.5	5.61	4.13	0.19
	03/30/77	1550	2075	75.0	4.22	4.44	0.27
	03/31/77	1500	2000	05.0	4.19	3.56	0.22
	04/01/77	1500	2150	20.0	4.25	3.29	0.20
	04/02/77	1600	1900	17.5	4.77	3.37	0.18
	04/03/77	1600	7150		4.94	3.33	0.17
	04/04/77	1450	1800	97.5	4.67	2.89	0.16
+	04/04/77	1600	2200	91.5	4.77	5.32	0.29
	04/05/77	1600	2000	70.0	4.65	3.87	0.21
	04/06/77	1650	2300	80.0	5.33	3.51	0.17
	04/07/77	1750	2250	97.5	5.47	3.26	0.15
	04/08/77	1850	2150	97.5	5.31	3.02	0.15
	04/09/77	1700	2250	50.0	4.77	3.15	0.17
	04/10/77	1600	2200	87.5	5.25	2.99	0.15
	04/11/77	1550	1900	65.5	4.48	2.79	0.16
+	04/11/77	1600	2050	87.5	4.20	4.23	0.26
	04/12/77	1500	1900	62.5	4.15	3.68	0.23
	04/13/77	1500	1600	65.0	4.11	3.23	0.20
	04/14/77	1550	1900	07.5	4.21	3.37	0.20
	04/15/77	1500	1500	37.5	3.26	2.95	0.23
	04/16/77	1525	1700	65.0	4.16	2.93	0.18
	04/17/77	1300	1800	40.0	4.34	3.13	0.18
	04/18/77	1475	1750	17.5	4.41	3.09	0.18
*	04/18/77	1450	2200	01.0	4.19	5.29	0.32
	04/19/77	1500	1750	15.0	4.20	4.59	0.28
	04/20/77	1500	1650	10.0	4.26	3.56	0.21
	04/21/77	1480	1875	92.5	4.15	3.35	0.21

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT FLUX (GSF)	RECOVERY
+04/21/77	1450	2100	20.0	4.53	4.54	0.26
04/22/77	1200	1500	50.0	4.63	3.85	0.21
04/23/77	1250	1700	65.0	4.47	3.85	0.22
04/24/77	1300	1550	57.5	4.30	3.31	0.20
04/25/77	1200	1480	48.5	4.21	3.15	0.19
*04/25/77	1180	1975	43.5	4.59	4.60	0.26
04/26/77	1500	1500	62.5	4.70	3.41	0.19
04/27/77	1475	1900	97.5	3.93	2.79	0.18
04/28/77	1425	1625	79.0	3.92	2.65	0.17
+04/28/77	1425	1775	94.0	4.41	3.54	0.20
04/29/77	1500	1700	90.0	4.34	2.97	0.18
04/30/77	1425	1650	87.5	4.13	2.61	0.16
05/01/77	1450	1450	55.0	2.55	2.32	0.23
05/02/77				0.00	0.00	
*05/02/77	1500	1975	02.5	3.91	3.87	0.25
05/03/77	1500	1800	92.5	3.89	2.61	0.17
05/04/77	1300	1900	92.5	4.21	2.32	0.14
05/05/77				0.00	0.00	
05/05/77	1675	1900	27.5	5.00	2.32	0.12
05/07/77				0.00	0.00	
05/08/77	1900	2200	40.0	4.16	2.58	0.16
05/09/77	1700	1950	18.5	4.20	2.34	0.14
*05/09/77	1680	2125	52.5	4.67	3.37	0.18
05/10/77	1650	1900	90.0	4.59	2.78	0.15
05/10/77	1625	1950	51.5	4.46	2.99	0.17
05/11/77	1300	1200	57.5	4.54	2.40	0.14
05/12/77	1300	1490	76.0	4.43	2.40	0.14
+05/12/77	1380	1550	64.0	4.43	3.23	0.19
05/13/77	1650	1920	83.0	4.59	2.65	0.15
05/13/77	1600	1750	15.0	4.61	3.00	0.17
05/14/77	1700	1850	77.5	4.69	2.43	0.13
05/15/77	1700	2000	10.0	4.12	2.48	0.15
05/16/77	1550	1750	81.5	4.42	1.94	0.11
*05/16/77	1550	1800	26.5	4.66	3.26	0.18
05/17/77	1500	1750	48.0	4.59	2.58	0.14
05/17/77	1580	1730	91.0	4.71	3.73	0.20
05/18/77	1500	1800	30.0	4.77	2.74	0.15
05/19/77	1550	1850	27.5	4.81	2.79	0.15
+05/19/77	1550	2025	31.0	4.80	4.31	0.23
05/20/77	1580	1900	10.0	4.83	3.40	0.18

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
05/20/77				4.52	3.51	0.20
05/20/77	1650	2100	22.5	4.56	4.10	0.23
05/21/77	1400	2050	20.0	4.61	3.15	0.17
05/22/77	1500	1400	85.0	4.54	2.68	0.15
05/23/77	1500	1800	20.0	3.55	2.43	0.18
*05/23/77	1450	1980	06.0	4.44	4.51	0.26
05/24/77				0.00	0.00	
05/24/77	1450	1920	58.5	4.86	4.25	0.22
05/25/77	1550	1700	10.0	4.92	3.12	0.16
05/26/77	1380	1600	07.5	4.85	2.90	0.15
+05/26/77	1400	1720	44.0	4.99	3.85	0.20
05/27/77	1450	1710	95.0	4.90	3.33	0.17
05/28/77	1450	1650	02.5	5.05	2.82	0.14
05/29/77	1450	1950	10.0	5.03	3.13	0.16
05/30/77	1550	1800	17.5	5.30	2.74	0.13
*05/30/77	1750	2420	66.5	3.97	4.78	0.31
05/31/77	1730	2350	61.5	3.68	2.90	0.20
05/31/77				3.84	3.05	0.20
05/31/77				3.99	3.13	0.20
05/31/77	1330	1600	25.0	4.74	3.24	0.17
06/01/77	2000	2300	07.5	5.73	2.79	0.12
06/02/77	2325	2650	68.0	4.74	2.44	0.13
+06/02/77	2730	3300	81.5	4.39	3.04	0.18
*06/06/77	2350	3280	00.0	3.53	4.05	0.29
06/07/77	2350	2770	62.5	4.71	3.14	0.17
06/08/77	2300	2300	20.0	4.87	2.53	0.13
06/09/77	1950	2200	20.0	3.61	2.14	0.15
+06/09/77	2000	2250	92.5	4.02	2.92	0.19
06/10/77	1900	2250	81.5	4.11	2.77	0.17
06/10/77				4.05	2.58	0.16
06/10/77				4.13	2.71	0.17
06/10/77	1950	2320	85.0	4.48	2.81	0.16
06/11/77	1800	2350	80.0	4.54	2.58	0.15
06/12/77	1750	2000	55.0	4.58	2.50	0.14
06/13/77	1790	2050	46.0	4.53	2.43	0.14
*06/13/77	1770	2450	65.0	4.55	4.91	0.28
06/14/77	1700	2000	32.5	4.46	3.51	0.20
06/15/77	1700	2100	45.0	3.48	3.31	0.24
06/16/77	1650	2050	47.5	3.74	2.95	0.20
+06/16/77	1800	2330	20.5	4.56	4.40	0.25

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT FLUX (GSF)	RECOVERY
06/17/77	1750	2200	42.5	4.14	3.48	0.21
06/17/77				4.02	3.22	0.21
06/17/77				4.04	3.53	0.22
06/17/77	1800	2450	10.0	4.03	3.90	0.25
06/18/77	1800	2300	65.0	4.05	3.77	0.24
06/19/77	1800	2200	62.5	3.83	3.10	0.21
06/20/77	1700	2100	55.0	4.03	3.15	0.20
*06/20/77	1650	4000	85.5	4.27	12.48	0.75
06/21/77	1550	3000	27.5	4.58	6.55	0.37
06/21/77	1630	2700	09.0	4.57	7.75	0.43
06/22/77	1525	1550	90.0	4.55	5.22	0.29
06/22/77	1870	2700	45.0	4.61	6.73	0.37
06/23/77	1500	1900	95.0	4.96	5.04	0.26
+06/23/77	1630	2550	92.5	4.63	6.77	0.37
06/24/77	1600	2250	00.0	4.62	5.68	0.31
06/24/77				4.65	5.81	0.32
06/24/77				4.54	5.98	0.34
06/24/77	1600	2450	95.0	4.46	6.40	0.37
06/25/77	1300	2200	07.5	4.62	5.67	0.31
06/26/77	1750	2400	30.0	4.61	5.01	0.28
06/27/77	1800	2320	50.0	4.48	4.30	0.25
*06/27/77	1950	9999	64.0	4.67	13.41	0.73
06/28/77	1780	3380	45.0	4.36	8.06	0.47
06/28/77				0.00	0.00	
07/05/77	2950	9999	56.5	4.10	7.98	0.50
07/06/77	2570	3600	66.5	4.07	5.06	0.32
07/07/77	2270	3200	39.0	4.06	4.34	0.27
+07/07/77	2300	4450	05.0	4.07	7.96	0.50
07/08/77	2100	3050	05.0	4.05	5.26	0.33
07/09/77	1950	2700	62.5	4.50	5.01	0.28
07/10/77	1950	2500	77.5	4.53	4.38	0.25
07/11/77	1800	2430	65.5	4.44	4.77	0.27
*07/12/77	1800	3900	75.0	4.41	6.33	0.37
07/13/77	1500	2800	00.0	4.14	6.89	0.43
07/14/77	1900	2100	10.0	4.23	6.02	0.36
07/14/77	2100	2550	52.5	4.15	5.69	0.35
+07/14/77	2200	3600	17.5	4.54	7.43	0.42
07/15/77	2000	3200	80.0	5.19	6.87	0.34
07/16/77	1900	2850	80.0	3.96	5.27	0.34
07/17/77	1800	5000		4.03	7.49	0.48

+ - SPONGEBALL CLEANING
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DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT FLUX (GSF)	RECOVERY
07/18/77	1850	2950	24.0	4.02	5.90	0.38
*07/18/77	1800	4900	77.5	4.34	12.91	0.76
07/19/77	1750	2350	70.0	4.45	6.66	0.38
07/20/77	1700	2900	45.0	3.08	4.81	0.40
07/21/77	1600	2300	30.0	3.97	6.13	0.39
+07/21/77	1600	1600	24.5	3.95	9.09	0.59
07/22/77	1500	2300	35.0	4.28	6.51	0.39
07/23/77	1800	2100	95.0	4.16	5.32	0.33
07/24/77	1400	1600	95.0	4.24	4.42	0.27
07/25/77	1430	2500	40.5	4.42	4.49	0.26
*07/25/77	1430	3300	26.5	4.20	15.29	0.93
07/26/77	1550	2500	45.0	4.35	6.89	0.41
07/27/77	1500	2500	95.0	3.66	5.01	0.35
07/28/77	2000	2000	25.0	3.73	5.27	0.36
07/29/77	2100	3200	75.0	3.64	4.91	0.34
+07/29/77	1950	2900	20.0	4.94	13.48	0.70
07/30/77	2400	2500	55.0	4.08	6.46	0.40
07/31/77	2100	2500	75.0	4.42	5.22	0.30
08/01/77	3300	4000	10.0	3.40	4.35	0.33
*08/01/77	3620	4700	50.0	4.47	14.68	0.84
08/02/77	3180	3780	15.0	4.47	6.32	0.36
08/02/77				3.79	7.40	0.50
08/03/77	1800	3100	80.0	3.50	5.37	0.39
08/04/77	2180	3250	82.5	3.46	4.25	0.31
+08/04/77	2350	3470	25.0	4.45	5.92	0.34
08/05/77				4.66	5.73	0.31
08/06/77				4.86	5.06	0.27
08/07/77				4.55	5.11	0.29
08/08/77	1650	2046	43.5	3.96	4.26	0.27
*08/08/77	1650	2475	06.0	4.30	14.36	0.85
08/09/77				4.46	6.82	0.39
08/10/77				4.32	5.94	0.35
08/11/77	1530	2000	07.0	4.19	5.35	0.33
+08/11/77	1584	2640	33.0	4.13	7.06	0.44
08/12/77				4.29	5.83	0.35
08/13/77				3.83	4.86	0.32
08/14/77				3.71	5.01	0.35
08/15/77	1848	2442	96.0	3.65	4.71	0.33
*08/15/77	1980	3300	01.5	4.35	12.71	0.75
08/15/77	1980	9999	15.0	4.53	12.71	0.72

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT FLUX (GSF)	RECOVERY
08/16/77				4.51	6.30	0.36
08/17/77				4.49	5.58	0.32
08/18/77				4.47	5.19	0.30
+08/18/77				4.33	7.03	0.41
08/19/77				4.13	6.33	0.39
08/20/77				3.85	5.40	0.36
08/21/77				3.95	5.53	0.36
08/22/77	1550	2150	00.0	4.22	4.93	0.30
*08/22/77	1650	2850	52.5	4.17	12.81	0.78
08/23/77	1800	2750	55.0	4.30	6.77	0.40
08/24/77	2050	2650	85.0	4.27	6.07	0.36
08/25/77	2050	2850	10.0	4.29	5.95	0.35
+08/25/77	2150	2900	75.0	4.33	6.65	0.39
08/26/77	2150	3100	07.5	4.28	6.22	0.37
08/27/77	2050	3650	95.0	4.59	7.28	0.41
08/28/77	2000	2750	65.0	4.30	5.75	0.34
08/29/77				0.00	0.00	
*08/29/77	2150	3950	10.0	3.62	6.64	0.47
08/30/77	2100	2500	00.0	2.96	5.78	0.50
08/31/77	1900	1750	52.5	3.47	4.91	0.36
09/01/77	2400	3600	50.0	3.94	4.99	0.32
+09/01/77	2300	2800	50.0	4.02	6.35	0.40
09/02/77	1900	2200	05.0	4.36	5.06	0.30
09/03/77	1750	2400	65.0	4.12	5.42	0.34
09/04/77	1950	2650	01.5	4.21	5.22	0.32
09/04/77	1750	2270	62.5	4.32	4.46	0.26
*09/05/77	1700	2680	82.5	4.53	9.67	0.55
09/06/77	1400	1500	05.0	4.63	5.53	0.30
09/07/77	1500	2000	02.5	3.79	5.47	0.37
09/08/77	1200	1600	45.0	4.03	5.22	0.33
+09/08/77	1400	1000	00.0	3.77	6.41	0.43
09/09/77	1200	1200	50.0	3.75	4.96	0.34
09/10/77	1400	1400	15.0	3.07	3.96	0.33
09/11/77	1200	1500	80.0	3.35	4.45	0.34
09/12/77	1100	2200	95.0	3.96	4.65	0.30
*09/12/77	1800	2500	65.0	3.56	9.09	0.65
09/13/77	1600	1400	60.0	3.51	5.06	0.37
09/14/77	1800	2300	20.0	4.00	4.80	0.31
09/15/77	1750	2400	95.0	3.91	4.55	0.30
+09/15/77	1750	2500	72.5	3.96	5.99	0.39

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
09/16/77	1600	2100	05.5	4.23	4.99	0.30
09/17/77	1700	2300	45.0	4.05	4.46	0.28
09/18/77	1750	3050	40.0	4.05	4.46	0.28
09/19/77	1820	2320	10.0	3.86	4.05	0.27
*09/22/77	1900	4000	25.0	3.14	7.08	0.58
09/23/77	1900	2100	85.0	3.11	4.65	0.38
09/24/77	1700	3100	20.0	3.51	4.65	0.34
09/25/77	1700	1600	15.0	3.13	5.94	0.49
09/26/77	2030	2750	00.0	3.07	3.74	0.31
*09/26/77				4.05	6.46	0.41
*09/27/77	2000	3400	92.5	3.97	6.66	0.43
09/28/77	2100	2600	75.0	3.86	5.19	0.34
09/29/77	2100	9999		2.94	8.11	0.70
09/30/77				0.00	0.00	
10/01/77	1900	2400	55.0	3.38	5.58	0.42
10/02/77	1900	2700	75.0	3.39	4.80	0.36
10/03/77				0.00	0.00	
10/04/77	1650	2550	95.0	4.05	8.42	0.53
10/04/77	1750	2925	47.5	4.37	7.70	0.45
10/05/77	1650	2200	05.0	4.30	4.65	0.28
10/06/77	1600	2100	05.0	4.34	4.60	0.27
+10/06/77	1700	2100	60.0	4.04	7.13	0.45
10/07/77	1500	1800	15.0	3.99	5.27	0.34
10/08/77				0.00	0.00	
10/09/77	1600	2000	80.0	4.01	3.93	0.25
10/10/77				0.00	0.00	
10/11/77	1480	1950	65.0	3.96	4.43	0.29
10/11/77				4.47	5.04	0.29
*10/11/77	1500	2080	39.0	4.54	8.97	0.51
10/12/77	1550	2000	20.0	4.67	6.51	0.36
10/13/77	1700	2400	00.0	4.66	5.83	0.32
+10/13/77	1600	3100	20.0	4.33	9.71	0.57
10/14/77	1900	3400	90.0	4.00	6.25	0.40
10/15/77	2100	2700	75.0	4.12	4.96	0.31
10/16/77				0.00	0.00	
10/17/77	1800	2200	25.0	4.20	5.89	0.36
10/18/77	2300	2800	85.0	3.78	5.02	0.34
*10/18/77	2050	4150	35.0	3.06	8.21	0.69
10/19/77	2000	3400	75.0	3.06	4.85	0.40
10/20/77	1800	2650	30.0	3.09	4.50	0.37

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT FLUX (GSF)	RECOVERY
+10/20/77	1400	1900	15.0	2.63	6.71	0.65
10/21/77	1700	1800	10.0	2.57	4.23	0.42
10/22/77	1650	2500	57.5	3.31	4.31	0.33
10/23/77	1550	2200	27.5	3.31	4.10	0.32
10/24/77	1500	1100	45.0	3.35	4.03	0.31
10/25/77	1500	2100	22.5	2.99	3.72	0.32
*10/25/77	1550	2300	79.0	4.28	6.66	0.40
10/26/77	1600	2200	20.0	4.30	5.22	0.31
10/27/77	1600	1700	15.0	3.91	4.55	0.30
+10/27/77	1600	2100	75.0	3.80	5.47	0.37
10/28/77	1700	3000	55.0	3.70	4.86	0.34
10/29/77	1900	2600	05.0	3.86	4.99	0.33
10/30/77	1500	2500	87.5	3.85	4.96	0.33
10/31/77	1500	2000	45.0	3.72	4.23	0.29
*10/31/77	1600	2300	50.0	3.80	6.51	0.44
11/01/77	1575	1675	30.0	3.89	5.04	0.33
11/01/77	1600	1850	60.0	4.27	4.87	0.29
11/02/77	1700	2200	95.0	4.25	5.47	0.33
11/03/77	1550	1800	40.0	3.81	5.22	0.35
+11/03/77	1200	1800	95.0	3.82	6.26	0.42
11/04/77	1400	2100	95.0	3.80	4.75	0.32
11/05/77	1400	2700	00.0	3.12	4.55	0.37
11/06/77	1400	1900	95.0	3.14	4.13	0.34
11/07/77	1400	1950	85.0	3.02	4.07	0.34
*11/07/77	1400	2130	62.5	4.35	6.20	0.36
11/08/77	1600	2600	45.0	4.30	5.17	0.31
11/09/77	1500	2300	55.0	4.08	5.01	0.31
11/10/77	1550	1600	35.0	3.80	4.86	0.33
11/11/77	1600	2300	10.0	3.91	4.34	0.28
11/12/77	2300	2700	50.0	3.91	4.34	0.28
11/13/77	2900	3700	00.0	3.89	4.28	0.28
11/14/77	2950	3650	85.0	3.85	3.79	0.25
*11/14/77	2950	4300	65.0	3.80	5.37	0.36
11/15/77	2600	3800	25.0	2.34	4.49	0.49
11/16/77	2200	2900	00.0	3.87	4.50	0.30
11/17/77	1800	2500	85.0	4.04	4.45	0.28
11/18/77	1700	1800	00.0	3.84	4.18	0.28
+11/18/77	1700	2500	75.0	3.84	4.78	0.32
11/19/77	1650	2150	37.5	3.76	4.10	0.28
11/20/77	1550	1850	52.5	3.75	3.87	0.26

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT FLUX (GSF)	RECOVERY
11/21/77	1575	2000	12.5	3.71	3.69	0.25
*11/21/77	1575	2350	80.0	3.97	5.33	0.34
11/28/77	2050	3300	05.0	4.32	7.59	0.45
11/29/77	1900	2500	90.0	4.37	4.70	0.27
11/30/77	1800	2200	07.5	4.40	5.42	0.32
12/01/77	1800	2400	75.0	3.49	3.93	0.29
12/02/77	1700	3000	60.0	3.52	4.80	0.35
12/02/77				0.00	0.00	
+12/03/77	1600	3000	75.0	4.15	8.47	0.52
12/04/77	1500	2000	25.0	3.73	5.32	0.36
12/05/77	1575	2350	34.0	4.29	6.16	0.37
12/06/77	1500	2200	05.0	4.37	4.81	0.28
12/07/77				0.00	0.00	
12/08/77	1400	2000	85.0	4.30	5.27	0.31
+12/08/77	1600	3200	65.0	4.28	7.33	0.44
12/09/77	2100	2500	70.0	0.00	0.00	
12/10/77	2750	3600	52.5	3.68	4.16	0.29
12/11/77	3200	4200	15.0	3.72	3.82	0.26
12/12/77	3450	4450	62.5	3.67	3.67	0.26
*12/12/77	3550	9999	75.0	4.20	7.11	0.43
12/13/77	3800	4850	55.0	4.38	5.27	0.31
12/14/77	3000	3700	85.0	4.29	5.01	0.30
12/15/77	1800	2300	60.0	3.81	5.01	0.34
12/16/77	1950	2100	25.0	3.88	4.75	0.31
+12/16/77	1400		25.0	3.66	5.83	0.41
12/17/77	1300	1500	80.0	3.56	5.17	0.37
12/18/77	1500	1700	60.0	3.49	4.93	0.36
12/19/77	1700	2500	60.0	4.05	4.50	0.28
+12/19/77	1800	2400	20.0	3.66	6.35	0.44
12/20/77	1800	2900	07.5	3.83	4.86	0.32
12/21/77	1500	2800	97.5	3.64	4.75	0.33
12/22/77	1700	2500	65.0	3.80	4.75	0.32
+12/22/77	1750	2620	32.5	3.81	6.25	0.42
12/23/77	1200	2000	92.5	3.70	5.78	0.40
12/24/77	1600	2000	45.0	3.07	5.22	0.43
12/25/77	1900	2600	55.0	3.47	4.70	0.35
12/26/77	1900	2200	30.0	3.47	4.36	0.32
12/27/77	1900	2600	71.5	3.46	4.02	0.30
*12/27/77	1800	2850	46.5	3.59	8.28	0.59
12/28/77	1600	2600	22.5	3.72	6.30	0.43

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
12/29/77	1400	2000	90.0	3.81	5.63	0.38
+12/29/77	1350	1800	35.0	3.77	8.26	0.56
12/30/77	1400	2300	97.5	3.80	6.41	0.43
12/31/77	1400	1500	25.0	3.81	5.42	0.36
01/01/78	1000	1900	70.0	3.95	4.70	0.30
01/02/78	1350	1500	92.5	3.89	4.28	0.28
01/03/78	1200	1620	56.5	3.78	4.26	0.29
*01/03/78	1200	2700	31.5	4.12	10.85	0.67
01/04/78	1300	2100	85.0	4.10	6.66	0.41
01/05/78	1400	1600	95.0	3.30	5.27	0.41
01/06/78	950	1500	07.5	3.30	4.80	0.37
+01/06/78	1000	1550	70.0	3.85	9.09	0.60
01/07/78	1350	2800	20.0	3.80	6.61	0.44
01/08/78	1600	2450	82.5	3.74	5.35	0.36
01/09/78	1380	1420	02.5	3.74	4.64	0.32
*01/09/78	1100	2875	11.0	2.16	5.46	0.64
01/10/78	1200	2100	05.0	4.36	7.65	0.45
01/11/78	1400	2350	60.0	4.36	6.20	0.36
01/12/78	1500	1600	35.0	4.20	5.27	0.32
+01/12/78		2200	60.0	3.88	10.53	0.69
01/13/78	1300	1500	30.0	3.87	5.63	0.37
01/14/78	700	800	10.0	3.81	5.22	0.35
01/15/78	700	1000	40.0	3.68	4.80	0.33
01/16/78	750	1000	45.0	3.66	4.70	0.33
*01/16/78	750	2200	92.5	3.70	10.31	0.71
01/17/78	700	1500	50.0	3.69	8.37	0.58
01/18/78	800	900	60.0	3.72	6.87	0.47
01/19/78	700	1050	10.0	3.86	5.91	0.39
+01/19/78	700	950	30.0	3.34	10.17	0.78
01/20/78	875	1400	65.0	3.27	7.28	0.57
01/21/78	900	1400	47.5	3.23	6.05	0.48
01/22/78	950	1200	00.0	3.24	4.99	0.39
01/23/78	950	2100	80.0	3.19	4.31	0.35
*01/23/78	1000	1600	72.5	4.34	10.95	0.64
01/24/78	975	1500	50.0	4.40	7.20	0.42
01/25/78	1000	1100	90.0	3.22	4.96	0.39
01/26/78	1200	1500	15.0	3.25	4.65	0.37
+01/27/78		2200	85.0	3.52	10.64	0.77
01/28/78	1100	1850	22.5	3.52	5.94	0.43
01/29/78	1050	1350	10.0	3.54	5.04	0.36

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
01/30/78	950	1350	75.0	3.50	4.31	0.31
*01/30/78	1000	2700	35.0	3.56	10.17	0.73
01/31/78	1100	2400	95.0	3.60	6.15	0.44
02/01/78	975	1000	15.0	3.23	4.99	0.39
02/02/78	1100	1600	40.0	3.12	4.68	0.38
+02/02/78	1100	1225	70.0	3.19	7.75	0.62
02/03/78	1100	1600	77.5	3.27	5.58	0.44
02/04/78	1300	1950	85.0	2.85	4.44	0.40
02/05/78	1000	1200	80.0	2.93	4.23	0.37
02/06/78	950	1750	32.5	2.87	3.97	0.35
*02/06/78	1025	2800	85.0	3.99	11.67	0.75
02/07/78	900	1800	92.5	3.84	7.49	0.50
02/08/78	700	850	20.0	3.97	6.77	0.44
02/09/78	750	1200	35.0	3.56	5.78	0.41
+02/09/78	600	1175	30.0	3.60	9.81	0.70
02/10/78	800	1950	72.5	3.59	7.65	0.54
02/11/78	850	1400	50.0	3.93	6.59	0.43
02/12/78	800	1200	37.5	3.80	5.65	0.38
02/13/78	600	950	00.0	3.72	5.76	0.40
02/14/78	700	900	20.0	3.63	5.01	0.35
*02/14/78	700		60.0	3.25	11.57	0.91
02/15/78	800	2100	60.0	4.21	7.49	0.45
02/16/78	700	1000	32.5	3.93	6.10	0.40
02/17/78	800	900	39.0	3.91	5.58	0.36
+02/17/78	700	1100	90.0	3.62	8.57	0.61
02/18/78	1000	1550	25.0	3.51	6.61	0.48
02/19/78	700	900	70.0	3.51	5.42	0.39
02/20/78	750	850	05.0	3.54	4.65	0.34
02/21/78	800	1115	47.5	3.49	4.23	0.31
*02/22/78	700	850	65.0	3.80	13.43	0.90
02/23/78		2100	00.0	3.79	6.35	0.43
02/24/78		1000	85.0	3.71	5.42	0.37
+02/24/78	900	1300	30.0	3.62	10.95	0.77
02/25/78	950	2200	90.0	3.63	5.94	0.42
02/26/78	900	1700	85.0	3.72	5.06	0.35
02/27/78	980	1380	75.0	3.64	4.34	0.30
02/27/78	1000	1175	75.0	3.63	4.39	0.31
*02/27/78	980	2900	42.5	4.01	11.43	0.73
02/28/78	1050	1800	06.0	3.18	5.67	0.46
03/01/78	1000	1800	60.0	3.19	4.75	0.38

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
03/02/78	1200	1600	40.0	3.21	4.39	0.35
+03/02/78		1350	08.0	3.22	7.23	0.57
03/03/78	900	1900	30.0	3.21	6.15	0.49
03/04/78	800	1800	75.0	3.17	5.27	0.42
03/05/78	700	900	68.0	3.23	5.01	0.40
03/06/78	800	1200	75.0	3.18	4.39	0.35
*03/06/78	850		35.0	3.15	12.35	1.00
03/07/78	1000	2400	10.0	3.06	6.40	0.53
03/07/78				3.49	7.05	0.52
03/07/78	1000	1200	00.0	3.37	8.01	0.61
03/08/78	1200	2200	85.0	2.82	5.94	0.54
03/09/78	800	1000	20.0	2.82	5.58	0.50
+03/09/78	800	1000	85.0	3.25	8.89	0.70
03/10/78	900	2575	82.0	3.26	8.99	0.70
03/13/78	1000	2900	65.0	4.41	12.19	0.71
*03/13/78	1000	3450	70.0	4.22	12.69	0.77
03/13/78	1000	2150	32.0	4.19	10.73	0.65
03/14/78	925	1500	68.0	4.30	7.13	0.42
03/15/78	950	1250	80.0	4.28	5.99	0.36
03/16/78	975	1400	82.0	4.34	5.84	0.34
+03/16/78	925	1200	25.0	4.03	11.67	0.74
03/17/78	900	1500	20.0	3.75	9.40	0.64
03/18/78	1000	1500	50.0	3.80	6.40	0.43
03/19/78	900	1200	90.0	3.75	5.58	0.38
03/20/78	700	1000	28.0	4.09	4.96	0.31
03/20/78	900	1350	22.0	3.76	5.01	0.34
*03/20/78	975	1950	18.0	3.63	13.33	0.94
03/20/78	1000	2150	58.0	3.26	10.76	0.84
03/21/78	980	1700	90.0	3.64	6.61	0.46
03/21/78				3.83	6.82	0.46
03/21/78	1000	2000	72.0	3.82	7.57	0.51
03/22/78	1000	1425	15.0	3.81	6.04	0.40
03/23/78	990	1100	00.0	3.76	5.11	0.35
+03/23/78	975	2100	62.0	3.90	11.14	0.73
03/24/78	1000	2200	02.0	3.78	6.74	0.46
03/25/78	950	1500	90.0	3.74	5.50	0.38
03/26/78	950	1350	88.0	3.75	5.06	0.35
03/27/78	950	1450	95.0	3.76	4.52	0.31
*03/27/78	1000	1700	78.0	3.60	13.22	0.94
03/28/78	1050	1700	92.0	3.69	5.91	0.41

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
03/29/78	1250	2800	89.0	3.69	5.04	0.35
03/30/78	1300	2800	60.0	3.77	4.75	0.32
+03/30/78	1300	1700	00.0	4.13	11.21	0.69
03/31/78	1200	1900	35.0	4.06	6.16	0.39
03/31/78	1150	1700	26.0	4.08	5.64	0.35
04/01/78	1000	1950	92.0	4.15	5.54	0.34
04/02/78	1050	1450	00.0	4.10	5.40	0.34
04/03/78	1000	1400	82.0	4.14	4.93	0.30
*04/03/78	1000	2500	92.0	4.04	11.98	0.76
04/04/78	1050	1120	18.0	4.07	6.38	0.40
04/05/78	900	1400	95.0	4.08	5.53	0.35
04/06/78	1000	1300	00.0	4.10	4.91	0.31
+04/06/78	1000	1200	52.0	3.91	10.33	0.68
04/07/78	900	1800	40.0	3.97	6.64	0.43
04/08/78	1200	1150	15.0	4.55	5.83	0.33
04/09/78	1100	1950	25.0	4.14	5.27	0.32
04/10/78	1000	1500	05.0	3.85	4.55	0.30
*04/10/78	1050	3600	80.0	4.12	14.62	0.91
04/11/78	1050	1850	38.0	4.05	6.04	0.38
04/12/78	1000	1400	10.0	4.15	5.40	0.33
04/13/78	1000	1350	00.0	3.96	4.85	0.31
+04/13/78			55.0	4.09	11.57	0.72
04/14/78	1000	1700	20.0	4.09	6.20	0.39
04/15/78	1000	1400	95.0	3.89	5.11	0.34
04/16/78	900	1250	68.0	3.96	5.11	0.33
04/17/78				0.00	0.00	
*04/17/78	950	4200	90.0	3.62	12.91	0.91
04/18/78	950	1600	74.0	3.65	6.46	0.45
04/19/78	1000	1300	05.0	3.64	5.17	0.36
04/20/78	1000	1425	98.0	3.76	4.80	0.33
+04/20/78	1000	2000	50.0	3.99	11.05	0.71
04/21/78	1000	1500	95.0	3.84	6.56	0.44
04/22/78	1000	1400	25.0	3.83	5.53	0.37
04/23/78	800	1400	00.0	4.24	5.53	0.33
04/24/78	925	1250	20.0	3.91	4.68	0.31
*04/24/78	980	2700	35.0	4.20	11.83	0.72
04/25/78	975	1550	65.0	4.22	6.80	0.41
04/26/78	600	1100	35.0	4.37	6.56	0.38
04/27/78	750	1100	28.0	4.22	6.14	0.37
+04/27/78	800	1175	65.0	0.00	5.58	

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA (1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT FLUX (GSF)	RECOVERY
04/28/78	900	1195	22.0	3.58	7.49	0.54
04/29/78		1300	80.0	4.11	6.89	0.43
04/30/78			95.0	4.28	6.23	0.37
05/01/78	900	1200	55.0	4.48	5.73	0.33
*05/01/78	850	1650	20.0	4.22	12.86	0.78
05/02/78	950	1550	64.0	4.26	7.15	0.43
05/03/78	1100	1450	90.0	3.95	5.53	0.36
05/04/78	1000	1400	95.0	4.09	5.06	0.32
+05/04/78	1000	1300	65.0	4.07	11.88	0.75
05/05/78	1000	1000	90.0	4.15	6.92	0.43
05/06/78	1000	1650	05.0	4.34	5.86	0.35
05/07/78	900	1250	68.0	4.36	5.27	0.31
05/08/78	900	1225	72.0	4.26	4.91	0.29
*05/08/78	900	2150	62.0	4.24	13.79	0.83
05/09/78	950	1450	70.0	4.52	6.64	0.38
05/10/78	1000	1400	95.0	4.46	5.68	0.33
05/11/78	1000	1100	80.0	4.53	5.22	0.29
+05/11/78	1000	1400	28.0	4.09	8.78	0.55
05/12/78	975	1100	05.0	4.37	7.80	0.46
05/13/78	800	1300	15.0	4.83	6.10	0.32
05/14/78	1000	1700	95.0	4.28	5.58	0.33
05/15/78	930	1250	89.0	4.40	4.80	0.28
*05/15/78	950	2600	32.0	4.36	12.71	0.75
05/16/78	950	1700	82.0	4.32	7.28	0.43
05/17/78	950	1450	20.0	4.30	6.51	0.39
05/18/78	950	1300	72.0	4.32	5.17	0.31
+05/18/78	950	1825	92.0	4.31	9.23	0.55
05/19/78	875	1350	80.0	4.22	6.71	0.41
05/20/78	950	1350	90.0	4.30	5.48	0.33
05/21/78	950	1350	00.0	4.32	5.53	0.33
05/22/78	975	1175	00.0	4.74	5.17	0.28
*05/22/78	975	2000	50.0	3.91	13.12	0.86
05/23/78	1100	1825	94.0	4.42	6.88	0.40
05/24/78	1000	1150	90.0	5.06	5.94	0.30
05/25/78	1100	1300	60.0	5.06	5.37	0.27
+05/25/78	1500	2300	55.0	5.25	12.19	0.59
05/26/78	1150	1800	40.0	5.22	8.11	0.40
05/27/78	750	1600	15.0	5.37	5.73	0.27
05/28/78	900	1200	85.0	5.31	6.61	0.32
05/29/78			10.0	5.19	6.15	0.30

\+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
05/29/78	1025	1400	49.0	5.27	6.35	0.31
*05/29/78	1050	9999	35.0	3.94	14.77	0.96
05/30/78	1150	1900	38.0	5.61	8.52	0.39
05/31/78	2000	2800	10.0	5.53	6.56	0.30
06/01/78	1750	1800	35.0	5.56	5.63	0.26
+06/01/78	1600	3500	00.0	5.84	12.29	0.54
06/02/78	1200	1800	60.0	5.97	9.40	0.40
06/03/78	1100	1700	35.0	5.79	7.05	0.31
06/04/78	1100	1600	35.0	5.84	7.36	0.32
06/05/78	1050	1525	26.0	5.78	6.67	0.29
*06/05/78	1050	4050	39.0	4.27	13.25	0.79
06/06/78	1000	1600	50.0	4.31	5.60	0.33
06/06/78	1000	1600	50.0	4.29	7.08	0.42
06/07/78	1050	1150	50.0	4.42	6.25	0.36
06/08/78	2200	3000	00.0	4.06	3.92	0.25
+06/08/78	2100	2900	55.0	4.08	12.81	0.80
06/09/78	2000	3000	50.0	4.22	6.92	0.42
06/10/78	2050	3000	75.0	3.93	5.06	0.33
06/11/78	1600	2400	45.0	3.96	4.65	0.30
06/12/78	1500	2000	74.0	4.09	4.47	0.28
*06/12/78	1250	2000	05.0	4.81	10.02	0.53
06/13/78	950	1950	34.0	4.86	9.30	0.49
06/14/78	1700	2900	10.0	4.69	6.35	0.35
+06/15/78	950	2050	25.0	4.90	11.47	0.60
06/16/78	1700	2600	50.0	4.94	8.93	0.46
06/17/78	1600	2400	65.0	5.02	7.67	0.39
06/18/78	1650	3300	68.0	5.04	6.71	0.34
06/19/78	1425	1925	70.0	4.89	5.90	0.31
*06/19/78	1150	2250	50.0	5.97	13.12	0.56
06/19/78	1150	2250	50.0	4.75	12.81	0.69
06/20/78	1750	3800	95.0	4.70	8.47	0.46
06/21/78	2500	4400	75.0	4.77	7.07	0.38
06/22/78	2800	3500	25.0	4.74	5.73	0.31
+06/22/78	3500	9999	50.0	4.79	11.67	0.62
06/24/78	3900	6000	00.0	4.94	9.23	0.48
06/26/78	1050	1450	62.0	4.31	4.71	0.28
*06/26/78	1020	1600	62.0	4.98	7.59	0.39
06/27/78	1000	1200	40.0	5.00	5.27	0.27
06/28/78	1000	1300	42.0	4.87	5.32	0.28
06/29/78	3400	2100	28.0	3.89	4.44	0.29

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT FLUX (GSF)	RECOVERY
+06/29/78	2000	4750	38.0	4.17	12.81	0.78
06/30/78	1100	2000	45.0	4.61	9.56	0.53
07/01/78	850	1400	00.0	3.99	6.51	0.42
07/02/78	1500	2100	10.0	3.97	4.39	0.28
07/03/78	1725	2050	30.0	3.92	3.23	0.21
*07/03/78	1550	2750	50.0	4.30	12.60	0.75
07/04/78	3500	9999	00.0	4.09	8.57	0.54
07/05/78	9999	9999	75.0	3.38	10.12	0.77
07/06/78	1000	1700	55.0	3.38	6.71	0.51
+07/06/78	1700	3500	65.0	3.76	12.29	0.84
07/07/78	2000	2700	50.0	3.80	6.61	0.44
07/08/78				0.00	0.00	
07/09/78				0.00	0.00	
07/10/78	850	1200	76.0	4.30	4.99	0.30
*07/10/78	920	1780	95.0	4.27	9.49	0.57
07/11/78	1100	1900	10.0	3.82	6.79	0.45
07/12/78	725	1100	30.0	3.83	5.01	0.33
07/13/78	1400	2100	70.0	4.14	4.55	0.28
+07/13/78	1500	3000	08.0	3.85	6.71	0.45
07/14/78	725	1000	02.0	4.02	7.03	0.45
07/15/78	1200	1700	10.0	3.96	5.17	0.33
07/16/78	800	1100	85.0	3.79	3.87	0.26
07/17/78	1500	2000	88.0	3.83	4.34	0.29
*07/17/78	1450	2800	05.0	4.20	9.74	0.59
07/18/78	1000	1700	00.0	4.13	5.94	0.37
07/19/78	1100	1900	38.0	4.24	6.66	0.40
07/20/78	1200	1700	65.0	4.21	5.11	0.31
+07/20/78	1400	1850	75.0	4.02	7.26	0.46
07/21/78	1750	2500	72.0	4.05	6.46	0.41
07/22/78	1900	2700	35.0	4.19	5.45	0.33
07/23/78	1500	2100	02.0	4.15	4.86	0.30
07/24/78	1450	1950	62.0	4.09	4.57	0.29
*07/24/78	1400	2700	98.0	4.12	8.80	0.55
07/25/78	1200	1800	50.0	4.22	6.14	0.37
07/26/78	1200	1750	80.0	4.40	5.63	0.33
07/27/78	1000	1100	55.0	4.29	5.20	0.31
+07/27/78	1350	2150	50.0	4.28	9.82	0.59
07/28/78	950	1600	35.0	4.15	6.34	0.39
07/29/78	850	1250	38.0	4.18	5.40	0.33
07/30/78	950	1300	65.0	4.22	5.05	0.31

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT FLUX (GSF)	RECOVERY
07/31/78	1050	3300	00.0	1.32	5.01	0.97
*07/31/78	1250	3450	48.0	4.49	12.17	0.69
08/01/78	1025	1700	02.0	4.60	7.47	0.42
08/02/78	1100	1500	50.0	4.61	7.28	0.40
08/03/78	1050	1450	19.0	5.13	5.53	0.28
08/03/78				6.31	8.58	0.35
08/03/78	1300	2180	45.0	4.22	4.83	0.29
08/04/78	1100	2100	50.0	5.58	10.69	0.49
08/05/78	1100	1400	63.0	5.62	9.40	0.43
08/06/78	1100	1600	38.0	5.54	8.78	0.41
08/07/78	1000	1525	88.0	5.54	8.14	0.38
*08/07/78	1000	1600	38.0	5.40	12.44	0.59
08/09/78	1000	2150	38.0	5.55	12.56	0.58
08/09/78	975	1600	02.0	5.42	10.72	0.51
08/10/78	1050	1600	12.0	5.47	8.30	0.39
08/11/78	1000	1300	98.0	5.57	7.12	0.33
08/12/78	1200	2350	50.0	5.52	6.77	0.31
08/13/78	1100	2000	00.0	5.56	6.89	0.32
08/14/78	1050	1150	92.0	5.51	5.87	0.27
*08/14/78	1100	2300	30.0	5.57	11.39	0.52
08/15/78	1100	1350	10.0	5.76	9.09	0.40
08/16/78				5.49	11.07	0.52
08/17/78	1600	2600	82.0	5.33	8.68	0.42
08/18/78	1900	2100	35.0	5.61	9.30	0.42
08/19/78	1700	2650	88.0	5.24	7.29	0.36
08/20/78	1600	2300	92.0	5.53	7.49	0.35
08/21/78	1500	1900	38.0	5.44	9.28	0.44
08/22/78	1850	2000	75.0	5.54	7.44	0.34
08/23/78	1300	1800	00.0	5.04	8.57	0.43
08/24/78	1450	2100	30.0	4.66	6.66	0.37
08/25/78			50.0	4.42	8.63	0.50
08/26/78	1100	1950	15.0	4.62	8.99	0.50
08/27/78	1100	2000	25.0	4.34	7.90	0.46
08/28/78	825	1550	15.0	4.34	4.15	0.24
*08/28/78	800	1350	70.0	5.54	13.22	0.61
08/29/78	1000	1950	42.0	5.48	10.53	0.49
08/30/78	1000	1300	15.0	5.12	10.12	0.51
08/31/78	1100	1500	70.0	5.76	9.50	0.42
09/01/78	950	1350	95.0	5.65	9.50	0.43
09/02/78	1100	2150	07.0	5.45	10.89	0.51

+ - SPONGEBALL CLEANING
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DAILY FLUX DATA - LAS GALINAS, CALIFORNIA (1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT FLUX (GSF)	RECOVERY
09/03/78	1000	1200	90.0	5.45	10.19	0.48
09/04/78	1100	1900	55.0	5.39	10.53	0.50
09/05/78	800	1350	00.0	5.44	9.30	0.44
*09/06/78	950	1900	90.0	5.39	11.44	0.54
09/07/78	1100	1700	10.0	4.97	8.26	0.42
09/08/78	1100	1700	98.0	5.49	7.65	0.36
09/09/78	1200	1600	50.0	5.54	7.23	0.33
09/10/78	1000	1300	65.0	5.60	7.85	0.36
09/11/78	800	1300	95.0	5.49	7.75	0.36
*09/11/78	880	2000	98.0	5.33	13.74	0.66
09/12/78	2200	3400	95.0	5.83	10.02	0.44
09/13/78	1700	2800	95.0	5.37	7.40	0.35
09/14/78	2500	4000	00.0	5.38	12.17	0.58
09/15/78	2000	3000	00.0	5.34	8.30	0.40
09/16/78	1500	2000	85.0	5.39	8.06	0.38
09/17/78	1800	2150	50.0	5.37	7.08	0.34
09/18/78	1250	1700	10.0	5.30	6.50	0.31
*09/18/78	1250	2850	32.0	4.51	12.05	0.68
09/19/78	3600	4900	00.0	4.55	6.46	0.36
09/20/78	1000	1700	00.0	4.04	6.50	0.41
09/21/78	1100	1500	15.0	4.67	6.07	0.33
+09/21/78	1100	1400	90.0	4.53	11.54	0.65
09/21/78	1150	2400	70.0	5.24	12.29	0.60
09/22/78	900	1900	90.0	3.65	11.78	0.82
09/23/78	1000	1000	20.0	4.49	7.43	0.42
09/24/78	1000	1400	30.0	4.46	6.73	0.39
09/25/78	1000	1200	80.0	4.27	5.56	0.33
*09/25/78	925	1780	74.0	4.48	15.18	0.87
09/26/78	1400	2550	12.0	4.33	7.83	0.46
09/26/78	1150	2350	90.0	4.11	9.00	0.56
09/26/78	1050	2200	00.0	4.39	8.92	0.52
09/27/78	950	1550	70.0	4.34	6.97	0.41
09/27/78	950	1800	40.0	4.17	8.84	0.54
09/28/78	1000	1700	90.0	4.46	6.42	0.37
+09/28/78	950	1750	62.0	4.34	14.24	0.84
09/29/78	900	3900	35.0	4.39	9.31	0.54
09/30/78	900	1500	88.0	5.56	8.73	0.40
01/12/79	2000	3700	20.0	3.68	7.36	0.51
01/13/79	1500	2350	18.0	3.71	5.52	0.38
01/14/79	1200	1700	58.0	3.66	4.77	0.33

+ - SPONGEBALL CLEANING
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DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
01/15/79	850	1000	95.0	3.70	5.40	0.37
*01/15/79	1050	2850	20.0	4.77	12.09	0.65
01/16/79	1100	1300	45.0	4.72	8.45	0.46
01/17/79	1200	1600	75.0	5.01	6.18	0.32
+01/18/79	1200	5000	78.0	4.95	14.44	0.75
01/19/79	700	1600	30.0	4.96	7.83	0.40
01/20/79	1250	1800	05.0	5.14	6.81	0.34
01/21/79	1250	1750	90.0	5.02	5.48	0.28
01/22/79	1100	1600	85.0	4.91	5.17	0.27
*01/22/79	1250	9999	62.0	5.03	13.42	0.68
01/23/79	1500	2250	02.0	5.10	6.73	0.34
01/24/79	1500	2000	10.0	4.91	5.17	0.27
01/25/79	1400	1700	40.0	4.83	5.28	0.28
+01/25/79	1400	2400	30.0	4.64	13.42	0.74
01/26/79	1400	2900	25.0	4.15	5.28	0.33
01/27/79	1450	2000	35.0	4.06	4.73	0.30
01/28/79	1500	1850	50.0	4.19	4.42	0.27
01/29/79	1400	1600	50.0	4.28	4.15	0.25
*01/29/79	2200	2800	65.0	4.20	13.03	0.79
01/29/79	1600	3800	65.0	5.66	13.03	0.59
01/30/79	1500	2500	25.0	5.50	7.04	0.33
01/31/79	1500	2500	55.0	5.01	9.70	0.50
02/01/79	1600	2000	80.0	4.94	6.93	0.36
02/02/79	1300	2300	22.0	4.76	5.60	0.30
+02/02/79	1300	2100	00.0	4.88	11.66	0.61
02/03/79	1350	2700	30.0	4.70	6.61	0.36
02/04/79	1250	1800	20.0	4.77	5.56	0.30
02/05/79	1250	1500	05.0	4.65	5.17	0.28
*02/05/79	1400	4750	50.0	4.49	12.72	0.72
02/06/79	1300	2400	95.0	4.11	6.18	0.38
02/06/79	1500	2600	30.0	4.51	7.55	0.43
02/07/79	1400	2000	30.0	4.16	4.93	0.30
02/08/79	1200	2000	25.0	4.12	4.54	0.28
+02/08/79	1200	1800	90.0	4.51	13.73	0.78
02/09/79	1200	2000	45.0	4.51	6.93	0.39
02/10/79	1400	1950	55.0	4.77	6.07	0.32
02/11/79	1350	1300	85.0	4.97	5.40	0.28
02/12/79	1300	1350	45.0	4.29	4.81	0.29
02/13/79	1400	1900	65.0	4.89	4.15	0.22
*02/13/79	1400	2500	80.0	3.99	14.52	0.93

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DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
02/14/79	650	1200	70.0	5.05	9.27	0.47
02/14/79	750	1050	60.0	10.01	13.81	0.35
02/15/79	950	1300	70.0	9.74	9.74	0.26
02/15/79				5.83	10.60	0.46
+02/15/79	1000	3200	55.0	5.47	17.37	0.81
02/16/79	1000	1600	02.0	5.52	8.80	0.41
02/17/79	1000	1500	82.0	3.43	8.10	0.60
02/18/79	1000	1500	95.0	6.00	6.97	0.30
02/19/79	1000	1600	85.0	6.01	8.02	0.34
02/20/79	1000	1300	72.0	6.02	7.04	0.30
*02/20/79	950	2900	10.0	6.14	14.56	0.61
02/21/79	750	1300	62.0	6.20	9.94	0.41
+02/22/79	1450	4100	02.0	6.84	15.18	0.57
02/23/79	900	1100	55.0	7.02	10.96	0.40
02/24/79	875	1150	68.0	6.98	9.94	0.36
02/25/79	900	1100	82.0	5.84	8.18	0.36
02/26/79	950	1200	68.0	6.77	7.08	0.27
*02/26/79	950	9999	12.0	6.35	14.79	0.60
02/27/79	1000	1700	95.0	6.43	9.04	0.36
02/27/79				4.03	15.77	1.00
02/27/79				3.92	14.17	0.92
02/27/79	1000	4200	05.0	4.59	15.81	0.88
02/28/79	1000	1600	95.0	4.67	7.32	0.40
03/01/79	900	1250	52.0	4.70	5.79	0.31
+03/01/79	900	1400	08.0	5.52	14.28	0.66
03/02/79	900	1400	90.0	5.18	8.69	0.43
03/03/79	950	1500	82.0	5.20	7.32	0.36
03/04/79	850	2000	05.0	5.14	4.81	0.24
03/05/79	850	1100	80.0	5.56	5.87	0.27
*03/05/79	1750	3000	35.0	7.64	15.26	0.51
03/06/79	1575	2450	25.0	7.40	10.64	0.37
03/06/79				3.78	13.73	0.93
03/06/79	1450	9999	45.0	4.05	12.76	0.80
03/07/79	1700	2700	70.0	4.02	7.04	0.45
03/08/79	1600	2200	20.0	4.07	5.79	0.36
03/09/79	1600	2100	30.0	3.61	5.05	0.36
03/10/79	1500	2050	22.0	3.64	4.93	0.35
03/11/79	1500	2000	55.0	4.16	4.93	0.30
03/12/79	1650	2050	28.0	3.99	4.30	0.28
*03/12/79	800	1400	70.0	4.44	17.37	1.00

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DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
03/13/79	775	1200	86.0	4.37	7.00	0.41
03/13/79	1200	1400	60.0	4.05	13.23	0.83
03/14/79	875	1100	00.0	4.00	6.57	0.42
03/15/79	750	850	05.0	4.23	5.79	0.35
+03/15/79	750	1150	45.0	4.25	16.20	0.97
03/16/79	800	1400	00.0	4.14	7.32	0.45
03/17/79	700	1050	98.0	4.00	5.60	0.36
03/18/79	750	1000	02.0	4.18	5.44	0.33
03/19/79	800	1050	95.0	4.07	5.40	0.34
*03/19/79	700	2950	32.0	4.86	17.14	0.90
03/20/79	800	1250	92.0	4.70	8.06	0.44
03/21/79	800	1200	98.0	4.78	6.50	0.35
03/22/79	700	1200	92.0	4.80	6.18	0.33
+03/22/79	950	1900	45.0	5.09	16.00	0.80
03/23/79	950	1450	08.0	5.15	6.93	0.34
03/24/79	900	1200	22.0	4.93	5.87	0.30
03/25/79	900	1200	50.0	5.07	5.79	0.29
03/26/79	1000	1300	00.0	5.13	5.40	0.27
*03/26/79	950	1800	28.0	5.24	15.14	0.74
03/27/79	1050	1925	30.0	4.09	7.12	0.44
03/28/79	1000	1300	32.0	4.01	5.56	0.35
*03/28/79	1000	3550	30.0	3.96	14.63	0.94
03/29/79	1050	1600	07.0	4.06	7.83	0.49
03/29/79	1500	1900	68.0	3.73	10.37	0.71
03/29/79	1050	1500	50.0	4.25	9.59	0.58
03/30/79	1200	1950	20.0	3.83	6.65	0.44
03/31/79	1100	1650	02.0	3.86	5.56	0.37
04/01/79	1100	1600	25.0	3.88	5.13	0.34
04/02/79	1050	1500	25.0	3.93	4.42	0.29
*04/02/79	1050	3550	30.0	5.62	15.73	0.72
04/03/79	1100	1700	88.0	5.49	8.53	0.40
04/03/79	1500	1750	41.0	4.32	9.20	0.54
04/04/79	1200	1700	00.0	4.48	5.56	0.32
04/05/79	1000	1450	45.0	4.33	4.85	0.29
+04/05/79	1000	1500	38.0	4.49	15.10	0.86
04/06/79	1000	1400	40.0	4.17	6.18	0.38
04/07/79			18.0	4.51	5.67	0.32
04/08/79	1000	1500	12.0	4.23	5.40	0.33
04/09/79	900	1200	40.0	4.49	5.60	0.32
*04/09/79	1300	3150	78.0	4.51	15.53	0.88

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
04/10/79	1200	1900	90.0	4.78	7.04	0.38
04/10/79	1500	3450	52.0	5.38	16.40	0.78
04/11/79	1450	2200	00.0	5.05	8.80	0.45
04/11/79	1450	3850	48.0	3.61	10.21	0.72
04/12/79	1000	1500	00.0	3.54	5.40	0.39
04/13/79	1200	1800	55.0	2.99	4.15	0.35
04/14/79	1100	1500	22.0	3.38	3.99	0.30
04/15/79	1150	1550	60.0	3.12	4.46	0.37
04/16/79	1250	1100	75.0	3.24	2.54	0.20
*04/16/79	1300	9999	28.0	4.38	14.17	0.83
04/17/79	2300	3700	65.0	2.98	5.67	0.49
04/18/79	950	1250	08.0	3.71	5.05	0.35
*04/18/79	950	3500	70.0	5.08	18.00	0.91
04/19/79	950	1550	90.0	4.14	6.81	0.42
04/19/79	1500	1550	90.0	3.78	7.16	0.48
04/19/79	1350	2550	60.0	3.03	7.47	0.63
04/19/79	2200	2750	88.0	3.89	7.79	0.51
04/20/79	1300	1550	50.0	3.65	4.81	0.34
04/20/79	1050	1650	00.0	4.53	7.43	0.42
04/21/79	1000	1700	22.0	3.99	5.91	0.38
04/22/79	1000	1300	35.0	3.48	5.56	0.41
04/23/79	950	1500	08.0	3.00	4.34	0.37
04/24/79	900	1700	10.0	5.15	6.42	0.32
04/25/79	1000	1100	20.0	4.61	5.24	0.29
*04/25/79	1200	2000	58.0	5.48	18.51	0.86
04/26/79	1100	1600	25.0	3.73	7.67	0.53
04/27/79	1100	1500	45.0	5.39	7.04	0.33
04/28/79	1200	1700	12.0	5.45	7.16	0.34
04/29/79	1100	1600	98.0	5.43	6.89	0.32
04/30/79	1000	1500	85.0	5.39	6.42	0.30
*04/30/79	1100	3150	50.0	5.71	16.36	0.73
05/01/79	1050	1400	95.0	5.06	7.55	0.38
05/01/79	1350	4000	36.0	5.07	13.77	0.69
05/01/79	1800	9999	62.0	5.59	16.71	0.76
05/01/79	2600	5000	02.0	5.14	14.44	0.72
05/02/79	1300	1600	18.0	6.02	12.60	0.53
05/03/79	1200	1950	35.0	3.62	14.17	1.00
05/04/79	1200	3000	45.0	3.62	10.02	0.71
05/05/79	1100	1550	92.0	3.85	5.56	0.37
05/06/79	1000	1300	75.0	3.46	4.85	0.36

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
05/07/79	1000	1300	95.0	3.46	4.66	0.34
05/07/79				0.00	0.00	
*05/08/79	1050	4800	12.0	4.41	14.13	0.82
05/09/79	1300	2200	25.0	3.40	12.83	0.96
05/21/79	1000	2150	02.0	5.00	12.48	0.64
05/22/79	1100	3900	10.0	5.23	7.67	0.37
05/23/79	1000	1900	50.0	3.43	6.38	0.48
05/24/79	1025	1650	98.0	3.40	5.87	0.44
+05/27/79	1500	9999	00.0	4.22	13.81	0.84
05/28/79	800	1500	88.0	4.10	7.28	0.45
05/29/79	1350	1800	88.0	6.01	6.85	0.29
*05/29/79	1550	2400	00.0	5.01	19.33	0.99
05/30/79	875	2000	90.0	4.15	6.73	0.41
05/31/79	1950	2200	00.0	4.17	6.18	0.38
*05/31/79	1200	9999	80.0	5.91	23.09	1.00
06/01/79	1250	1900	68.0	5.05	9.74	0.49
06/01/79	2350	9999	00.0	4.29	15.89	0.95
06/02/79	2500	1900	75.0	4.72	12.17	0.66
06/03/79	2300	1600	00.0	3.54	8.30	0.60
06/04/79	1100	1700	98.0	3.35	9.63	0.73
06/05/79	1200	1850	05.0	3.05	5.95	0.50
06/06/79	1000	1350	10.0	3.54	5.17	0.37
*06/06/79	1050	9999	00.0	4.25	16.47	0.99
06/07/79	1250	1950	45.0	4.59	8.88	0.49
06/08/79	1400	3200	58.0	3.57	8.18	0.59
06/09/79	950	1500	00.0	5.51	8.53	0.40
06/10/79	1300	1900	52.0	5.54	8.06	0.37
06/11/79	1250	1700	35.0	5.38	6.73	0.32
*06/11/79	1400	9999	25.0	5.30	19.92	0.96
06/12/79	2000	2500	55.0	4.85	7.94	0.42
06/12/79	1550	3400	62.0	4.84	18.00	0.95
06/13/79	1500	2700	35.0	4.98	11.35	0.58
06/14/79	1450	1600	95.0	1.76	5.09	0.74
06/19/79	1100	9999	75.0	3.86	13.54	0.90
06/19/79	1800	4400	60.0	3.40	11.35	0.85
06/19/79	1150	2500	58.0	4.19	16.00	0.98
06/20/79	1100	1750	12.0	3.62	6.73	0.48
06/21/79	1200	1700	08.0	3.74	5.56	0.38
06/22/79	600	800	10.0	2.54	4.97	0.50
*06/24/79	1450	9999	50.0	4.80	17.30	0.92

+ - SPONGEBALL CLEANING

* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
06/24/79	1450	9999	75.0	3.70	13.66	0.94
06/25/79	1450	2850	28.0	3.51	7.24	0.53
06/25/79	2650	4050	50.0	4.25	10.96	0.66
06/26/79	1600	2000	25.0	4.07	7.04	0.44
06/27/79	1400	2200	12.0	2.99	5.36	0.46
06/28/79	1300	2050	58.0	3.02	4.70	0.40
06/28/79				0.00	0.00	
07/05/79				0.00	0.00	
*07/05/79	1050	9999	65.0	4.25	14.09	0.85
07/06/79	950	1550	08.0	4.33	7.47	0.44
07/06/79	2175	4000	55.0	4.36	8.69	0.51
08/01/79	1200		12.0	6.55	9.12	0.36
08/02/79	1200	2000	50.0	6.55	15.73	0.61
+08/02/79	1200	2900	30.0	4.78	18.59	0.99
08/03/79	1600	3250	08.0	4.63	14.09	0.78
*08/03/79	1325	4600	89.0	6.55	19.37	0.76
08/03/79	1400	4650	78.0	6.40	18.16	0.73
08/03/79	1425	4300	59.0	6.45	17.30	0.69
08/03/79	1450	3775	53.0	6.44	15.81	0.63
08/03/79	1450	2800	62.0	6.45	15.10	0.60
08/03/79	1450	2375	16.0	6.44	14.36	0.57
08/03/79	1900	3100	30.0	6.49	14.01	0.55
08/03/79	1400	2800	88.0	6.43	13.19	0.52
08/03/79	1375	2700	79.0	6.44	12.87	0.51
08/03/79	1375	2550	78.0	6.47	12.21	0.48
08/03/79	1375	2400	66.0	6.48	11.66	0.46
08/03/79	1350	2350	65.0	6.46	11.23	0.44
08/03/79	1325	2300	60.0	6.47	10.96	0.43
08/04/79	1400	1200	45.0	6.49	9.51	0.37
08/05/79	1300	1825	42.0	6.40	8.06	0.32
*08/05/79	1250	4150	50.0	6.54	19.37	0.76
08/05/79	1400	4400	29.0	6.67	18.94	0.73
08/05/79	1375	4450	75.0	6.41	17.53	0.70
08/05/79	1450	4180	38.0	6.43	17.22	0.68
08/05/79	1700	3625	64.0	6.39	15.10	0.60
08/05/79	1525	3925	22.0	6.40	15.85	0.63
08/05/79	1550	3700	22.0	6.45	15.18	0.60
08/05/79	1575	3650	29.0	6.41	14.75	0.59
08/05/79	1600	3450	26.0	6.40	13.85	0.55
08/05/79	1625	3300	12.0	6.42	13.27	0.53

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA (1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW (GPM)	PRODUCT FLUX (GSF)	RECOVERY
08/05/79	1625	3100	06.0	6.38	12.44	0.50
08/05/79	1575	2925	02.0	6.44	12.05	0.48
08/05/79	1575	2850	99.0	6.40	11.47	0.46
08/06/79	1575	3350	13.0	6.31	12.91	0.52
*08/07/79	1750	4600	62.0	6.54	20.43	0.80
08/07/79	1975	9999	58.0	6.65	20.19	0.78
08/07/79	2000	9999	56.0	6.61	19.56	0.76
08/07/79	1975	6300	25.0	6.61	19.17	0.74
08/07/79	1975	5750	00.0	6.57	18.39	0.72
08/07/79	1950	5500	68.0	6.62	18.00	0.69
08/07/79	1950	5200	50.0	6.64	17.06	0.66
08/07/79	1925	4900	24.0	6.66	16.40	0.63
08/07/79	1950	4450	10.0	6.65	15.81	0.61
08/08/79	1800	3475	46.0	6.63	13.23	0.51
08/08/79	1900	4600	78.0	4.50	13.30	0.76
08/08/79	1825	3450	59.0	6.62	12.87	0.50
08/08/79				6.57	13.93	0.54
08/08/79				6.53	14.87	0.58
*08/09/79	1975	8900	12.0	6.57	21.99	0.86
08/09/79	2450	6650	00.0	6.53	19.88	0.78
08/09/79	2000	6400	62.0	6.63	19.96	0.77
08/09/79	2075	6900	60.0	6.57	19.56	0.76
08/09/79	2100	6400	40.0	6.59	19.21	0.75
08/09/79	2100	5900	08.0	6.55	18.35	0.72
08/09/79	2100	5700	85.0	6.63	17.80	0.69
08/09/79	2050	5000	30.0	6.62	15.81	0.61
08/10/79	1925	3750	46.0	6.55	12.76	0.50
08/10/79	1950	3770	48.0	6.53	12.83	0.50
08/11/79	1925	3250	55.0	6.45	11.50	0.46
08/12/79	1600	2950	50.0	4.39	8.33	0.49
08/13/79	1500	2000	10.0	6.27	6.34	0.26
08/14/79				5.13	19.88	0.99
08/14/79	1250	1100	82.0	5.40	20.90	0.99
08/14/79	162	425	44.0	5.78	21.60	0.96
08/15/79	1500	2900	00.0	5.50	11.23	0.52
08/16/79	1450	2600	82.0	5.99	10.68	0.46
+08/16/79	1400	3700	22.0	6.90	17.30	0.64
08/17/79	1400	2300	70.0	6.50	10.96	0.43
08/18/79	1350	2000	55.0	6.44	8.57	0.34
08/19/79	1700	2800	60.0	4.31	8.61	0.51

+ - SPONGEBALL CLEANING
* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
08/20/79	1250	1600	40.0	6.31	6.50	0.26
*08/20/79	1300	4650	05.0	5.95	17.26	0.74
08/21/79	2000	3900	62.0	4.20	12.21	0.74
08/22/79	1400	2400	22.0	4.15	8.80	0.54
08/23/79	1400	1500	85.0	4.17	6.81	0.42
+08/23/79	1400	2000	10.0	6.21	19.45	0.80
08/24/79	1350	3600	90.0	4.25	11.35	0.68
08/25/79	1250	2300	95.0	4.26	8.61	0.52
08/26/79	1100	2900	90.0	6.38	16.51	0.66
08/27/79	1000	1600	70.0	4.25	8.77	0.53
*08/27/79	1050	4100	40.0	6.16	19.56	0.81
08/28/79	1250	2100	48.0	4.12	9.00	0.56
08/29/79	1200	2000	65.0	5.54	17.45	0.81
08/30/79	1250	1800	55.0	5.51	9.59	0.44
+08/30/79	1100	1500	12.0	5.15	19.13	0.95
08/31/79	1150	2050	70.0	4.93	10.21	0.53
09/01/79	850	1300	00.0	3.46	7.75	0.57
09/02/79	1050	1400	12.0	4.96	8.06	0.42
09/03/79	1000	1300	20.0	5.05	7.79	0.39
09/04/79	975	975	98.0	4.86	5.99	0.31
*09/04/79	900	1900	62.0	6.08	21.52	0.90
09/05/79		1900	20.0	5.83	12.09	0.53
09/06/79	1400	2500	55.0	5.36	9.82	0.47
09/06/79	1400	4400	25.0	5.86	17.14	0.75
09/07/79	1375	2650	60.0	6.70	14.01	0.53
09/08/79	2000	4000	70.0	6.75	14.20	0.54
09/09/79	1450	2800	02.0	6.71	13.15	0.50
09/10/79	1650	2950	04.0	6.57	12.09	0.47
*09/10/79	1750	8900	15.0	6.47	22.30	0.88
09/11/79	1650	3950	42.0	6.40	15.34	0.61
09/12/79	1700	8400	05.0	6.29	20.07	0.82
09/14/79	1600	3350	16.0	6.50	13.77	0.54
09/14/79	1625	6400	55.0	6.45	19.84	0.79
09/15/79	1500	1900	52.0	6.02	8.88	0.38
09/16/79	1500	2100	40.0	6.25	7.94	0.32
09/17/79	1100	1200		3.98	7.12	0.46
*09/17/79	1350	5000	90.0	6.74	20.66	0.78
09/18/79	1400	1500	40.0	6.17	9.90	0.41
09/19/79	1400	1500	30.0	6.92	8.88	0.33
09/20/79	1200	1600	40.0	6.23	7.87	0.32

+ - SPONGEBALL CLEANING

* - CHEMICAL AND SPONGEBALL CLEANING

DAILY FLUX DATA - LAS GALINAS, CALIFORNIA(1976-1980)

DATE	TDS FEED	TDS BRINE	TDS PRODUCT	FEED FLOW(GPM)	PRODUCT FLUX(GSF)	RECOVERY
+09/20/79	1200	3400	05.0	6.23	15.93	0.65
09/21/79	650	2400	05.0	6.17	9.08	0.38
09/22/79	1250	1850	45.0	6.41	8.57	0.34
09/23/79	1100	1600	42.0	6.39	8.26	0.33
09/24/79	1150	1650	16.0	5.95	7.67	0.33
*09/24/79	1175	4300	50.0	6.25	18.00	0.74
09/25/79	1300	2500	32.0	6.25	12.68	0.52
09/26/79	1250	2200	85.0	6.39	11.15	0.45
09/27/79	1200		55.0	6.13	11.58	0.48
+09/27/79	1200	2000	95.0	8.60	17.14	0.51
09/28/79	1200	1950	20.0	6.20	12.91	0.53
09/29/79	1200	1600	15.0	6.16	9.63	0.40
09/30/79	1200	1600	00.0	5.94	7.12	0.31
10/01/79	1200	1725	20.0	6.21	7.83	0.32
*10/01/79	1150	6000	48.0	6.11	21.76	0.91
10/02/79	1400	1500	40.0	6.17	11.54	0.48
10/03/79	1450	2450	80.0	6.27	11.11	0.45
10/04/79	1600	1450	58.0	6.34	9.94	0.40
+10/04/79	1600	1950	32.0	6.12	17.77	0.74
10/05/79	600	2800	75.0	6.28	13.85	0.56
10/06/79	1650	3100	42.0	6.37	12.95	0.52
10/06/79	1750	3100	30.0	6.30	10.72	0.43
10/08/79	1900	1600	15.0	6.00	9.43	0.40
*10/09/79	3000	9999	75.0	6.12	18.16	0.76
10/09/79	1600	6440	39.0	6.11	19.92	0.83
12/04/79	1200	1900	55.0	6.69	8.80	0.34
12/05/79	1600	2200	65.0	6.53	7.79	0.30
12/06/79	1500	2000	60.0	6.59	7.90	0.31
12/07/79	1400	2000	45.0	6.52	7.75	0.30
12/08/79	1400	1800	55.0	6.42	7.75	0.31
12/09/79	1150	1600	08.0	6.33	7.51	0.30
*12/09/79	1250	6250	00.0	4.98	17.96	0.92
12/10/79	1200	2200	52.0	5.06	9.78	0.49
12/10/79	2700	4400	00.0	4.08	15.46	0.97
12/10/79	1600	5800	80.0	4.45	16.00	0.92
12/10/79	1400	4650	65.0	4.40	16.00	0.93
12/10/79	1250	4650	38.0	5.40	17.45	0.83
12/11/79	1200	2100	72.0	5.01	9.70	0.50
*12/11/79	1200	3800	85.0	6.23	19.13	0.78
12/11/79				6.44	19.37	0.77

+ - SPONGEBALL CLEANING
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