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Demineralization of Carbon-Treated Secondary Effluent by Spiral-Wound Reverse Osmosis Process

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DEMINERALIZATION OF CARBON-TREATED SECONDARY EFFLUENT BY SPIRAL-WOUND REVERSE OSMOSIS PROCESS

by

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FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimony to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution and it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems for the hazardous water pollutant discharges from municipal and community sources, for the preservation and treatment of public drinking water supplies, and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research; a most vital communications link between the researcher and the user community.

One of the goals of wastewater treatment is renovation of wastewater so that it can be reused. It is expected that partial demineralization of conventionally treated wastewater will be required if the wastewater is reused for any purpose which requires high quality water. Among the techniques for demineralization that which is newest but shows the most potential is reverse osmosis. In this process water is forced through a membrane which can reject salts. The permeability of these membranes is low so high pressure is required to achieve an economical production rate. Special configuration of the membrane and its support system are required to withstand the high pressure and maintain a high ratio of membrane surface to system volume. In the studies reported in here a reverse osmosis system using a spiral membrane-support configuration was tested for its efficacy in demineralization of secondary effluent. Included in the study was an evaluation of pretreatment of the reverse osmosis feed with activated carbon to reduce membrane fouling.

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ABSTRACT

A 56.8 cu m/day (15,000 gallons/day) spiral-wound reverse osmosis pilot plant, manufactured by the Gulf Environmental Systems Company, San Diego, California, was operated at the Pomona Advanced Wastewater Treatment Research Facility on the carbon-treated secondary effluent. The specific objectives for this study were (a) to establish the effective membrane life for wastewater demineralization with carbon adsorption pretreatment; (b) to determine the reliability of the process performance; and (c) to derive a realistic process cost estimate.

The study was first conducted on a constant feed pressure basis, and then it was run on a constant product water flux rate basis. During the first phase of the study, pH adjustment was not practiced for the weekly enzyme-detergent membrane cleaning procedures. However, this was practiced in the second phase of the study. The results from both phases of studies substantiated the fact that the membrane effective life was only about one year in demineralizing the carbon-treated secondary effluent.

A cost estimate for a 37,850 cu m/day (10 MGD) reverse osmosis plant indicated that for membranes with only one-year life the process cost was about 14.9¢/1,000 liters (57.4¢/1,000 gallons). However, the cost could be substantially reduced to 10.7¢/1,000 liters (41.3¢/1,000 gallons) for membranes with two-year life. Both cost estimates did not include the costs for carbon adsorption pretreatment and brine disposal. These cost estimates were based on August, 1973 material and construction costs.

This report was submitted by County Sanitation Districts of Los Angeles County in fulfillment of Contract No. 14-12-150 under the partial sponsorship of the Municipal Environmental Research Laboratory, Office of Research and Development, U.S. Environmental Protection Agency. Work was completed as of January 13, 1972.

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ACKNOWLEDGEMENTS

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The authors are deeply grateful to Dr. James E. Cruver of Gulf Environmental Systems Company, San Diego, California, for his advice and cooperation in this effort.

Mr. James Gratteau and Mr. Harold H. Takenaka, former project engineers at Pomona Advanced Wastewater Treatment Research Facility, were instrumental in initiating the pilot plant study.

The efforts of the laboratory and the pilot plant operating personnel of the Pomona Advanced Wastewater Treatment Research Facility are also gratefully acknowledged.

SECTION 1

INTRODUCTION

All uses of water serve to increase the mineral and organic contents of the water. The organic impurities are normally removed by the biological oxidation and activated-carbon adsorption processes, while the inorganic minerals are effectively removed by the demineralization processes, such as ion exchange, electrodialysis, and reverse osmosis. Therefore, the wastewater demineralization is an indispensable part of the total effort to achieve the following environmental goals:

A. To conserve the natural water qualities of the receiving water systems; and

B. To treat the wastewater to meet the quality requirements for various water reuses.

County Sanitation Districts of Los Angeles County in conjunction with the U.S. Environmental Protection Agency initiated a series of wastewater demineralization studies in 1967. Three demineralization processes--reverse osmosis, electrodialysis, and ion exchange were extensively studied at Pomona Advanced Wastewater Treatment Research Facility. Since reverse osmosis process was still at the development stage, several operating parameters had to be established in the beginning of the pilot plant study. The process was first applied directly to the secondary effluent without proper membrane cleaning procedures. This direct application was quickly proved to be a failure by the rapid decline in the system performance.

On June 16, 1969, a new experimental run with a 56.8 cu m/day (15,000 gallons/day) spiral-wound reverse osmosis pilot plant, manufactured by the Gulf Environmental Systems Company, was initiated with a carbon adsorption pretreatment on the secondary effluent. The objectives of this study were: (a) to evaluate the effect of the carbon adsorption pretreatment on the system performance; (b) to obtain data on the system reliability; (c) to establish the effective membrane life; and (d) to derive a realistic process cost estimate.

The study was divided into two phases. The first study was conducted with a constant operating pressure, while the second phase was conducted with a constant product water flux rate. After the initial 9,475 hours of on-stream operations in the first phase of the study, the pilot plant operation was temporarily suspended on August 16, 1970, as a result of the serious membrane deterioration. This was revealed by the substantial reduction in both product water flux rate and salt rejection. All the membrane modules were subsequently removed from the pilot plant system and sent to the Gulf Environmental Systems Company for membrane evaluation to determine the causes of membrane deterioration.

Based on the membrane evaluation results, the pilot plant operation was resumed on December 21, 1970 for the second phase of the study. New sets of operating conditions and membrane loading arrangement were employed in this second study. Only three of the original twenty-seven membrane modules were kept in the system for this new study, while fifteen of the other twenty-four modules were replaced with the new production membrane modules. The remaining nine modules were replaced with the partially used modules from a similar system being concurrently operated at Pomona Research Facility. All the used modules were still in good performance condition. This second part of the study was finally terminated on January 13, 1972, after a total of 7,803 hours of on-stream operation.

SECTION 2

CONCLUSIONS

The principal conclusions drawn from this pilot plant study are outlined as follows:

A. The Gulf Environmental System Company's spiral-wound reverse osmosis system was capable of achieving a 95 percent salt rejection, a 566 l/sq m/day (13.9 gal/sq ft/day) product water flux rate, and an 80 percent water recovery under a constant operating feed pressure of 32.1 Kg/sq cm (465 psi) in its initial stage of operation.

B. A regular membrane cleaning operation, including a weekly enzyme-detergent (BIZ) or sodium perborate cleaning and a daily air-tap water flushing, was essential even with a carbon adsorption pretreatment in controlling the product water flux decline, which resulted from membrane fouling.

C. A minimum brine flow at approximately 11.3 l/min (3 gpm) was helpful in minimizing the product water flux decline.

D. Both modes of operations, constant operating feed pressure, as in the first phase of the study, and constant product water flux rate, as in the second phase of study, showed similar performance and product water quality.

E. The water quality data prior to the deterioration of the membrane modules indicated that on the average the product water had:

a. Less than 3 percent of the feed phosphate content;

b. Less than 7 percent of the feed total chemical oxygen demand (TCOD) content;

c. Less than 1 percent of the feed sulfate content;

d. Less than 3 percent of the feed calcium content;

e. Less than 11 percent of the feed ammonia nitrogen content;

f. Less than 8 percent of the feed total dissolved solids (TDS) content; and

g. Less than 5 percent of the feed turbidity.

F. The cause of membrane deterioration was partially attributed to the hydrolysis of the membrane which was caused by the exposure to the high pH of the enzyme-detergent cleaning solution during the first phase of the study.

G. The results from both modes of pilot plant operations indicated that the effective membrane life was only one operation year based on initial performance parameters.

H. The process cost estimate for a 37,850 cu m/day (10 MGD) reverse osmosis plant is about 14.9¢/1,000 liters (57.4¢/1,000 gallons). However, if the membrane life could be improved from one year to two years, then the cost would be reduced to 10.7¢/1,000 liters (41.3¢/1,000 gallons). Both cost estimates do not include the costs for carbon adsorption pretreatment and brine disposal.

SECTION 3

RECOMMENDATIONS

The short membrane life as concluded from the study on wastewater demineralization is rather discouraging. An optimum membrane life was shown to be about three years for a practical and economical application of the reverse osmosis process to the wastewater demineralization(1). Therefore, it is recommended that further studies be pursued primarily in the areas of membrane improvement. Other parameters such as pretreatment methods, membrane cleaning techniques and frequency, feed pressure, brine recirculation, membrane module configuration, and brine velocity should also be thoroughly evaluated and investigated.

SECTION 4

PILOT PLANT DESCRIPTION

The 56.8 cu m/day (15,000 gallons/day) reverse osmosis pilot plant consisted of 9 steel pressure vessels. Each vessel measured 3.05 m (10 ft) in length and 10 cm (4 in) in diameter. Three ROGA spiral-wound membrane modules, manufactured by the Gulf Environmental Systems Company, were installed in each of the steel pressure vessels. Each membrane module was approximately 10 cm (4 in) in diameter, 0.91 m (3 ft) long, and contained 4.6 sq m (50 sq ft) of modified cellulose acetate membrane. The total membrane area in the pilot plant system was about 125 sq m (1,350 sq ft).

Figure 1 shows a schematic flow diagram of the spiral-wound reverse osmosis pilot plant. The carbon-treated secondary effluent was chlorinated to a 1 to 2 mg/l chlorine residual and acidified to a pH close to 5 using sulfuric acid before it was fed to the membrane system. The pilot plant system was in a 3-2-2-1-1 array to maintain sufficient brine velocities in the downstream modules. Some necessary provisions for a daily air-tap water flushing, a weekly enzyme-detergent cleaning cycle, and a chlorinated tap water flushing during downtimes were made. A flexible metal hose was installed between the main feed pump and the lead modules to prevent the fatigue failure of the piping in the system, which otherwise would be caused by the serious vibration of the feed pump.

Sufficient sample valves were installed on the pilot plant system, so that samples from the raw feed (carbon treated secondary effluent) blended feed (mixture of carbon-treated secondary effluent, sulfuric acid and chlorine solution), brine, and product streams could be taken regularly. Instrumentation was included to measure the temperature and the pressure of the blended feed, brine and product streams. A proportional chemical feed pump was used to add sulfuric acid to the feed stream for pH control. The pump rate was regulated by a pH controller. Chlorine was added to feed stream through a gas chlorinator.

The carbon-treated secondary effluent was obtained from the concurrent activated carbon adsorption pilot plant study at Pomona Research Facility. The carbon pilot plant was a four-stage downflow pressure system. Each stage contained about 3,020 Kg (6,650 lb) of Calgon Filtrasorb-400 granular activated carbon in a 1.83 m (6 ft) diameter steel column. The depth of the carbon bed was about 3.04 m (10 ft). Table 1 shows some of the physical characteristics of the activated carbon used

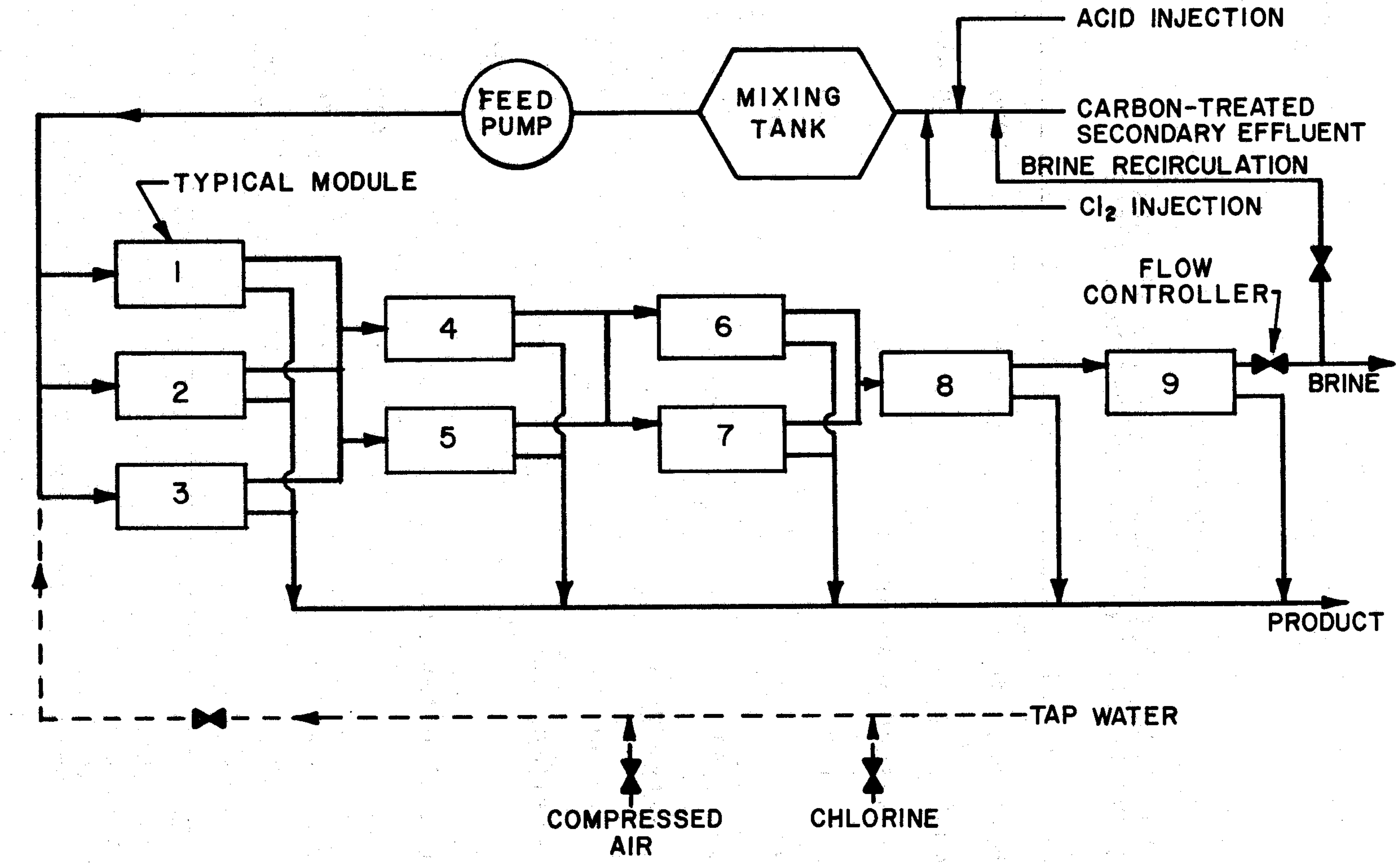


Figure 1. Schematic flow diagram of the reverse osmosis pilot plant.

TABLE 1

PHYSICAL CHARACTERISTICS OF THE GRANULAR ACTIVATED CARBON IN THE PRETREATMENT SYSTEM

Surface Area, m^2/g (BET)	:	1000
Apparent Density, g/ml	:	0.44
Density, backwashed & drained,	:	
lb/cu ft	:	25
Kg/cu m	:	401
Real Density, g/ml	:	2.1
Particle Density, g/ml	:	1.3
Effective Size, mm	:	0.55
Uniformity Coefficient	:	1.9
Pore Volume	:	0.94
Mean Particle Diameter, mm	:	0.9
Iodine No.	:	1000
Abrasion No. minimum	:	75
Ash, %	:	8.5

in the study. The empty-bed detention time for each stage of treatment was about 10 minutes. Therefore, a total of 40 minutes contact time was used in the study.

As shown in Figure 2, the carbon pilot plant included the carbon regeneration system. The carbon from the lead column was normally regenerated whenever the total chemical oxygen demand (TCOD) of the carbon plant effluent reached a level of approximately 10 mg/l. The lead carbon column was backwashed daily with a maximum backwash rate of 6.8 lps/sq m (10 gpm/sq ft). The results of the operation and performance of the four-stage carbon adsorption pilot plant were presented elsewhere(2).

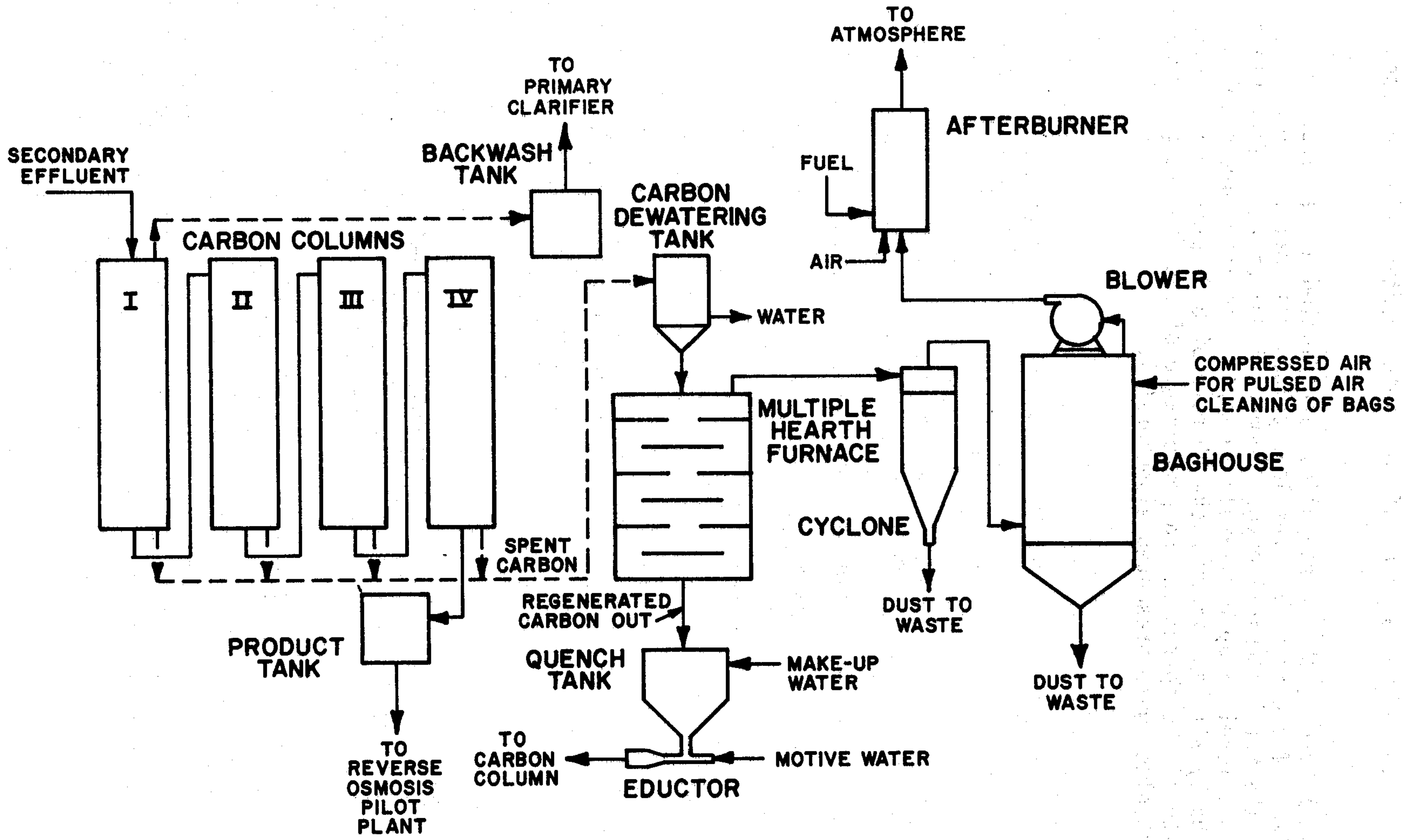


Figure 2. Schematic diagram of the activated carbon pretreatment system.

SECTION 5

PILOT PLANT OPERATION

OPERATING CONDITIONS

Phase I: Constant Feed Pressure Operation

The pilot plant was operated under a constant feed pressure of 32.1 Kg/sq cm (465 psi) during the first phase study. The other initial operating conditions for the pilot plant are summarized as follows:

A. Feed Water: Carbon-treated secondary effluent chlorinated to a chlorine residual of 1 to 2 mg/l.

B. Feed pH: Controlled to 5 using sulfuric acid.

C. Feed Flow: 61.3 lpm (16.2 gpm).

D. Product Flow: 50 lpm (13.2 gpm).

E. Brine Flow: 11.3 lpm (3 gpm).

F. Product Water Flux Rate: 566 l/sq m/day (13.9 gal/sq ft/day) at 25°C.

G. Water Recovery: 81.5 percent.

H. Salt Rejection: 95 percent.

A daily air-tap water flushing and a weekly enzyme-detergent cleaning cycle were conducted to maintain the product water flux rate during the first phase study.

Phase II: Constant Product Flux Rate Operation

During the second phase of the pilot plant study, the twenty-seven membrane modules in the system were made up of 12 used and 15 new production modules. The module loading arrangement for the system is shown in Table 2. A summary of the operating history of the modules is presented in Table 3.

The system for the second phase of the study was also operated on the carbon-treated secondary effluent with pre-chlorination to provide 1 to 2

TABLE 2

MODULE LOADING ARRANGEMENT AT START-UP OF CONSTANT PRODUCT FLUX RATE OPERATION

Pressure Vessel	Module Loading Arrangement
1	Used modules from other system
2	New production modules
3	Used modules from first part of study
4	Used modules from other system
5	New production modules
6	New production modules
7	New production modules
8	Used modules from other system
9	New production modules

Note: The sequence of the pressure vessels is shown in Figure 1.

TABLE 3

HISTORY OF MEMBRANE MODULES USED FOR CONSTANT PRODUCT FLUX RATE OPERATION

Pressure Vessel No. 1

12-21-70	:	Commenced study with 3 used modules which had 1,335 hours of operating time.
9-23-71	:	At 5,678 hours of unit operations, module #1, with 7,013 hours total operating time, was removed and replaced with a used module, which had 1,933 hours of operating time.
1-13-72	:	Study terminated at 7,803 hours of unit operation.

Module #1	- 7,013 hours of operation;
Module #1R	- 4,058 hours of operation;
Module #2	- 9,138 hours of operation;
Module #3	- 9,138 hours of operation.

Pressure Vessel No. 2

12-21-70	:	Commenced study with 3 new production modules.
9-23-71	:	At 5,678 hours of unit operations, module #1 was removed and replaced with a used module which had 1,933 hours of operating time.
1-13-72	:	Study terminated at 7,803 hours of unit operation.

Module #1	- 5,678 hours of operation;
Module #1R	- 4,058 hours of operation;
Module #2	- 7,803 hours of operation;
Module #3	- 7,803 hours of operation.

(continued)

TABLE 3 (Continued)

Pressure Vessel No. 3	
12-21-70	: Commenced study with 3 used modules which had 9,475 hours of operating time.
7-28-71	: At 4,414 hours of unit operations, all 3 modules were removed and replaced with 3 used modules which had 1,933 hours of operating time.
9-23-71	: At 5,678 hours of unit operation, module #1R was removed and replaced with a used module which had 1,933 hours of operating time.
1-13-72	: Study terminated at 7,803 hours of unit operations. Module #1, 2, & 3 - 13,899 hours of operation; Module #1R - 3,197 hours of operation; Module #1RR - 4,058 hours of operation; Module #2R and 3R - 5,322 hours of operation.

Pressure Veesel No. 4	
12-21-70	: Commenced study with 3 used modules which had 1,335 hours of operating time.
1-13-72	: Study terminated at 7,803 hours of unit operations. Module #1 - 9,138 hours of operation; Module #2 - 9,138 hours of operation; Module #3 - 9,138 hours of operation.

(Continued)

TABLE 3 (Continued)

Pressure Vessel
No. 5

12-21-70	:	Commenced study with 3 new production modules.
1-13-72	:	Study terminated at 7,803 hours of unit operation. Module #1 - 7,803 hours of operation; Module #2 - 7,803 hours of operation; Module #3 - 7,803 hours of operation.

Pressure Vessel
No. 6

Same as Pressure Vessel No. 5

Pressure Vessel
No. 7

Same as Pressure Vessel No. 5

Pressure Vessel
No. 8

Same as Pressure Vessel No. 4

Pressure Vessel
No. 9

Same as Pressure Vessel No. 5

Note: The sequence of the pressure vessels is shown in Figure 1.

mg/l of total residual chlorine. A summary of the operating conditions at the start-up and 100 hours later, when the water recovery was increased from 75 percent to 80 percent by increasing the brine recirculation from 9.5 lpm (2.5 gpm) to 12.5 lpm (3.3 gpm), is shown in Table 4. An attempt was made during the second phase of the study to operate the system at a constant 407 l/sq m/day (10 gal/sq ft/day) apparent product flux rate, and 80 percent water recovery by varying the feed pressure.

The membrane cleaning procedures adopted for both phases of the study were very similar; however, the pH of the cleaning solution was adjusted from 10.0 to 7.5 with sulfuric acid during the second phase of study. In addition to the enzyme-detergent (Biz) solution, a 2 percent sodium perborate solution was also tested in this study.

The procedures used in the air-tap water flushing and the enzyme-detergent (or sodium perborate solution) cleaning cycle throughout the entire study are described in the following sections.

MEMBRANE CLEANING

Enzyme-Detergent (or Sodium Perborate) Cleaning Procedure

The enzyme-detergent cleaning solution was made up by adding 2.84 Kg (100 oz) of a commercial enzyme-detergent, BIZ, into 379 l (100 gal) of tap water, while the sodium perborate cleaning solution was made up of 2 percent sodium perborate and 0.15 percent Triton X-100 non-ionic detergent with 1 percent (based on detergent weight) carboxy methyl cellulose (CMC) soil suspending agent. The enzyme-detergent (or sodium perborate) cleaning was conducted once a week.

During the cleaning cycle, the system (pressure vessels 1 to 9) was first filled with either enzyme-detergent or sodium perborate cleaning solution using the main feed pump. The pressure vessels which were not being flushed remained soaking in the cleaning solution, while others were being flushed according to the following sequence. Here the term flush refers to cycling the cleaning solution through the pressure vessels and membrane modules.

A. Pressure vessels 1 to 3 were flushed for 10 minutes at a feed pressure of 5.5 Kg/sq cm (80 psi), and at a flow rate of about 22.7 to 30.3 lpm (6 to 8 gpm) per pressure vessel.

B. Pressure vessels 4 to 5 were flushed for 20 minutes at a feed pressure of 5.5 Kg/sq cm (80 psi), and at a flow rate of about 22.7 to 30.3 lpm (6 to 8 gpm) per pressure vessel.

C. Pressure vessels 6 to 9 were flushed for 20 minutes at a feed pressure of 5.5 Kg/sq cm (80 psi), and at a flow rate of about 22.7 to 30.3 lpm (6 to 8 gpm) per pressure vessel.

TABLE 4. OPERATING CONDITIONS AT START-UP AND 100 HOURS LATER OF CONSTANT PRODUCT FLUX RATE OPERATION

Parameter	Start-Up	100 Hours
Feed Pressure		
Kg/sq cm	24.8	24.8
psi	360	360
Raw Feed Flow Rate		
lpm	47.3	44.3
gpm	12.5	11.7
Product Flow Rate		
lpm	35.6	35.6
gpm	9.4	9.4
Waste Brine Flow Rate		
lpm	11.7	8.7
gpm	3.1	2.3
Brine Recirculation Rate		
lpm	9.5	12.5
gpm	2.5	3.3
Water Recovery, %	75	80
Salt Rejection, %	95.5	95.5

During the cleaning cycle, the cleaning solution was recycled for the specified time period in the first set of pressure vessels, and then the same cleaning solution was applied to the next set of pressure vessels until the sequence was completed. After each cleaning solution flushing, an air-tap water flushing was also conducted to rinse the membrane modules.

Air-Tap Water Flushing Procedure

The air-tap water flushing was conducted once every day either as a main cleaning process in non-chemical solution flushing days or as a rinse process in chemical solution flushing days. For the air-tap water flushing, the same sequence of application to the pressure vessels was used as for the chemical solution flushing. This consisted of flushing each pressure vessel with tap water for two minutes and then with a mixture of air and tap water for another three minutes. The air-tap water mixture was, however, not recycled, it went directly to waste.

Acid Flush Procedure

This particular acid flushing was employed whenever the decline of the product water flux was due to the loss of pH control in the system. The procedure consisted of depressurizing the system and flushing with an acidified water (maintaining pH between 2 and 3) for thirty minutes. The acid flushing was then followed by a cleaning chemical solution and air-tap water flushing to provide maximum cleaning of the membrane modules.

SECTION 6

RESULTS AND DISCUSSIONS

CONSTANT FEED PRESSURE OPERATION

Product Water Flux Rate

The variation of the product water flux rate during the initial 800 hours of on-stream operation under a constant feed pressure of 32.1 Kg/sq cm (465 psi) is shown in Figure 3. As indicated in Figure 3, the product water flux rate decreased rapidly from 566 l/sq m/day (13.9 gal/sq ft/day) to 391 l/sq m/day (9.6 gal/sq ft/day) during the first 200 hours of on-stream operation. The primary cause for this rapid decrease in flux rate was possibly due to the high membrane compaction during the initial hours of operation. An indication that the decrease in flux rate for the initial period of operation was due to membrane compaction and not organic fouling was the fact that the enzyme-detergent flushing of the unit at 50 hours of operation failed to restore the product water flux rate. After 200 hours of operation, the weekly enzyme-detergent flushing procedure was found successful in removing the fouling materials and in controlling the decline of the product water flux rate.

At 250 hours of operation, the system operation was temporarily suspended due to the loss of carbon effluent feed which was caused by a power failure in the carbon pretreatment system. Chlorinated tap water with approximately 1 mg/l chlorine residual was run through the unit for about 65 hours until the carbon effluent feed was restored. When the system was placed back onstream, an increase in product water flux rate occurred. The increase was attributed to the flushing action resulting from 65 hours of chlorinated tap water feed.

As indicated in Figure 3, there were two sharp drops in product water flux rate at 430 hours and 550 hours of operation. These drops were caused by the problems with the acid feed system. The acid pump air-locked after 430 hours of operation and it resulted in a loss of feed pH control for approximately 12 hours. At 550 hours, an electrical failure in the pH monitoring system resulted in a partial loss of pH control over a 3 day weekend. As soon as each malfunction in the acid feed system was noted, the system was taken offstream and corrective measures were taken to restore the product water flux rate before it was placed back onstream. In both cases, the acid flush cleaning procedure as described in previous section was applied successfully to the system to restore the flux rate. The incidents fully

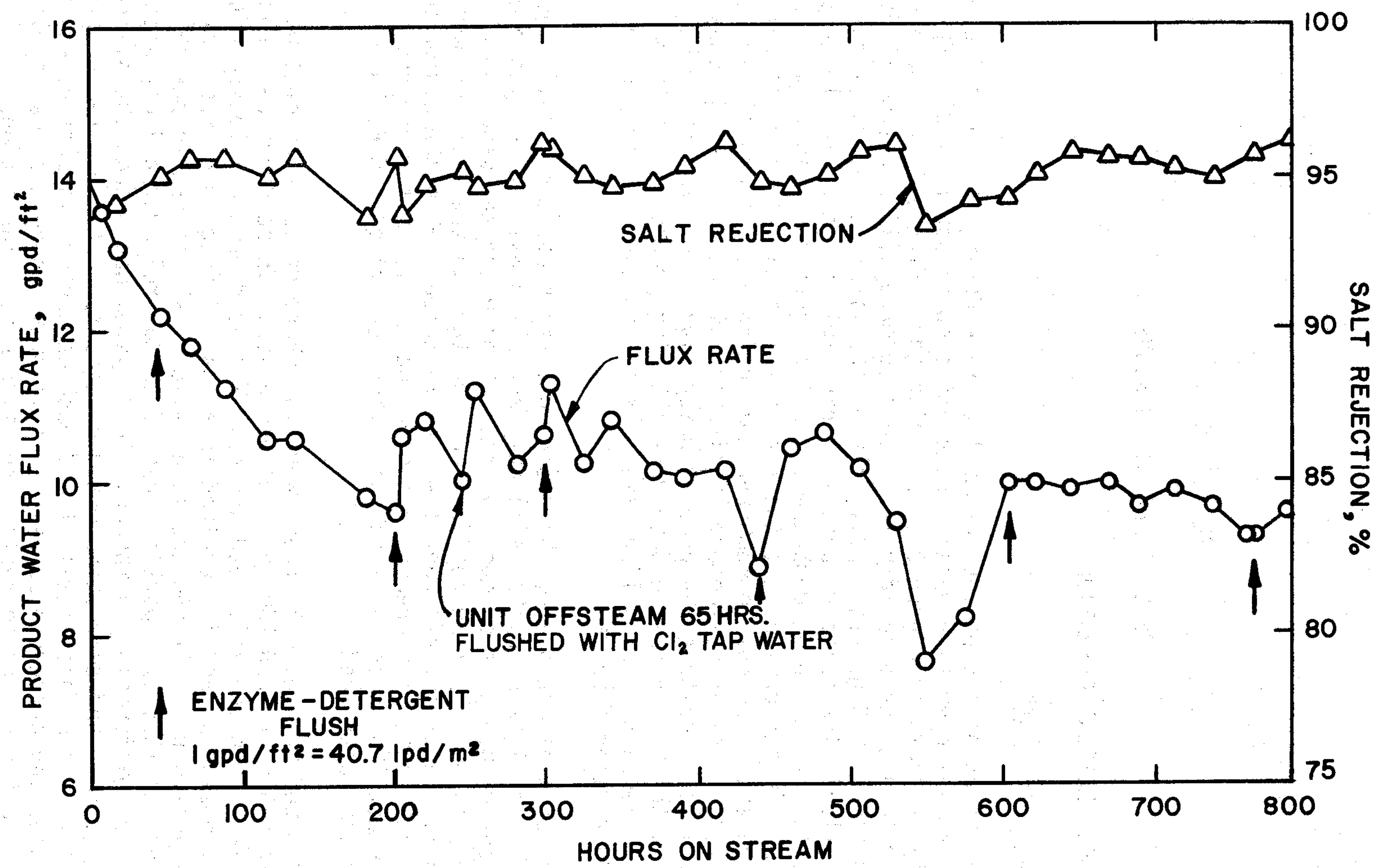


Figure 3. Variation of flux rate and salt rejection during the initial 800 hours of operation with constant feed pressure of 32.1 Kg/sq cm (465 psi).

substantiated the fact that the pH control was very essential in preventing the membrane fouling caused by the precipitation of calcium salts.

Figure 4 summarizes the variations in the product water flux rate and in the total COD of the feed water from July 1, 1969 to February 28, 1970. This period corresponded to the operating times from 200 hours to 5,600 hours. The solid circles on the product water flux curve of Figure 4 indicate the applications of enzyme-detergent cleaning to the system. The solid triangles on the total feed COD curve indicate times when methanol was present in the feed water. This methanol leakage occurred on several different occasions during a denitrification study being conducted concurrently in the carbon adsorption system and resulted in abnormally high COD values. Some special notes are shown under each curve of Figure 4 to explain the deviations from the normal operation.

The effectiveness of the enzyme-detergent flushing in restoring the product water flux rate is illustrated by the distance between the two solid circles. The two solid circles are, respectively, the product water flux rates before and after the enzyme-detergent cleaning cycle.

The decline of the product water flux rate during the entire first phase of this pilot plant study is shown in Figure 5. The product water flux rate decreased from 566 l/sq m/day (13.9 gal/sq ft/day) at time zero to 350 l/sq m/day (8.6 gal/sq ft/day) at 6,000 hours of operation. However, the product water flux rate between the period of 6,000 hours to 9,475 hours (end of the first phase of study) was found to increase from 350 l/sq m/day (8.6 gal/sq ft/day) to 374 l/sq m/day (9.2 gal/sq ft/day). This increase in flux rate corresponded with a decrease in the overall salt rejection.

The flux decline slope was determined several times during the study. The initial slope, determined after 1,500 hours of operation, was - 0.09. At 6,000 hours, the flux decline slope changed to - 0.07. After 6,000 hours, the product water flux rate began to increase due to the deterioration of the membrane. This caused a reversal of the flux decline slope. Finally, at 9,475 hours, the flux decline slope was about - 0.055. The most meaningful flux decline slope would be that calculated for the first 6,000 hours of operation, that is - 0.07.

Salt Rejection

The salt rejection variations from July 1, 1969 to April 30, 1970 are shown in Figure 6. This period corresponds to the operation times from 200 hours to 6,950 hours. The salt rejection was found to decrease slightly when the concentration of the nitrate ion in the feed water increased due to either the nitrification of the Pomona activated sludge plant, which supplied the secondary effluent to the carbon pretreatment system, or the addition of sodium nitrate to the feed of the carbon adsorption system during the denitrification study. The reason for this decrease in salt rejection was that the nitrate ion was not rejected as well as other ions

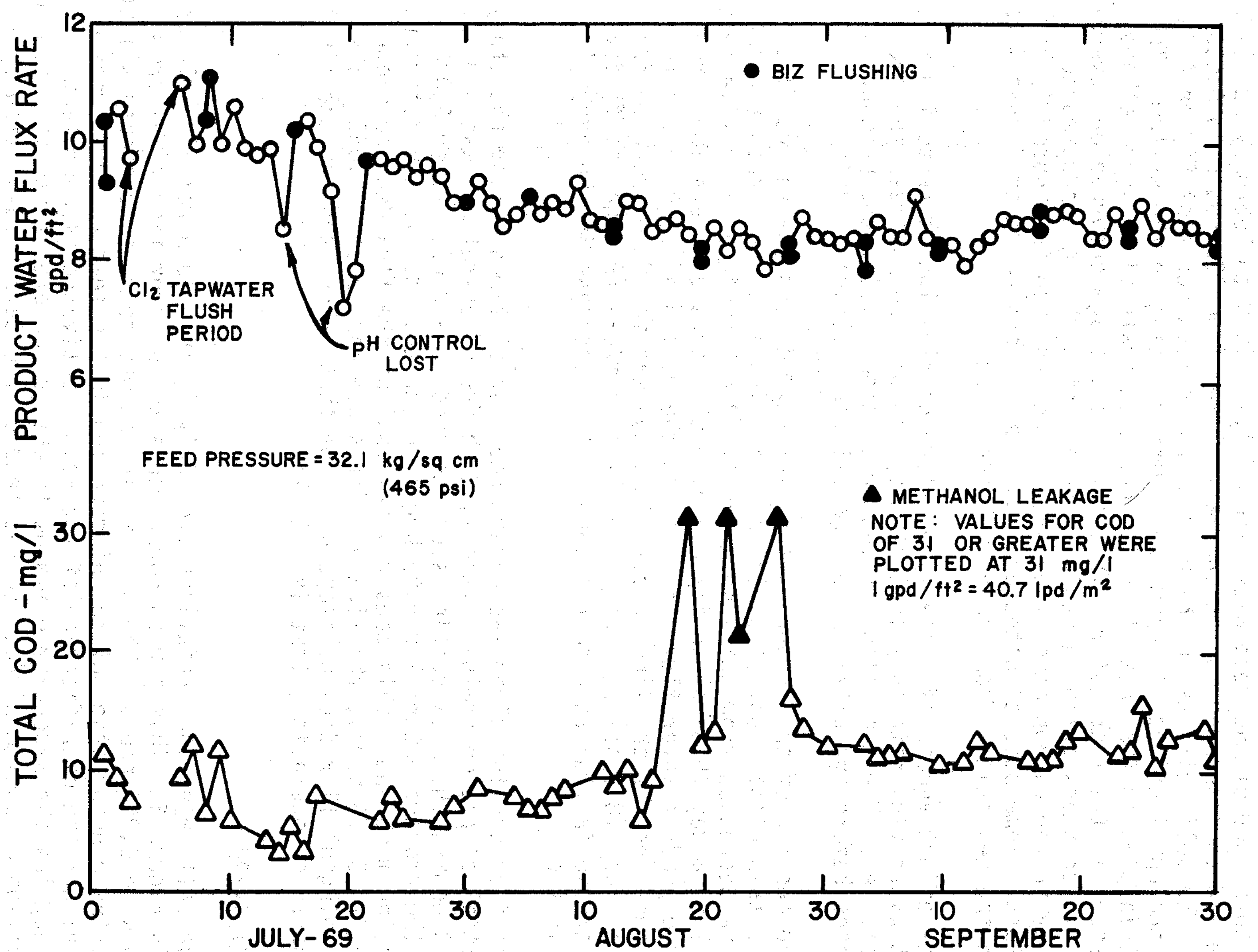


Figure 4. Total feed COD and product water flux rate vs. operation time. (constant feed pressure operation)

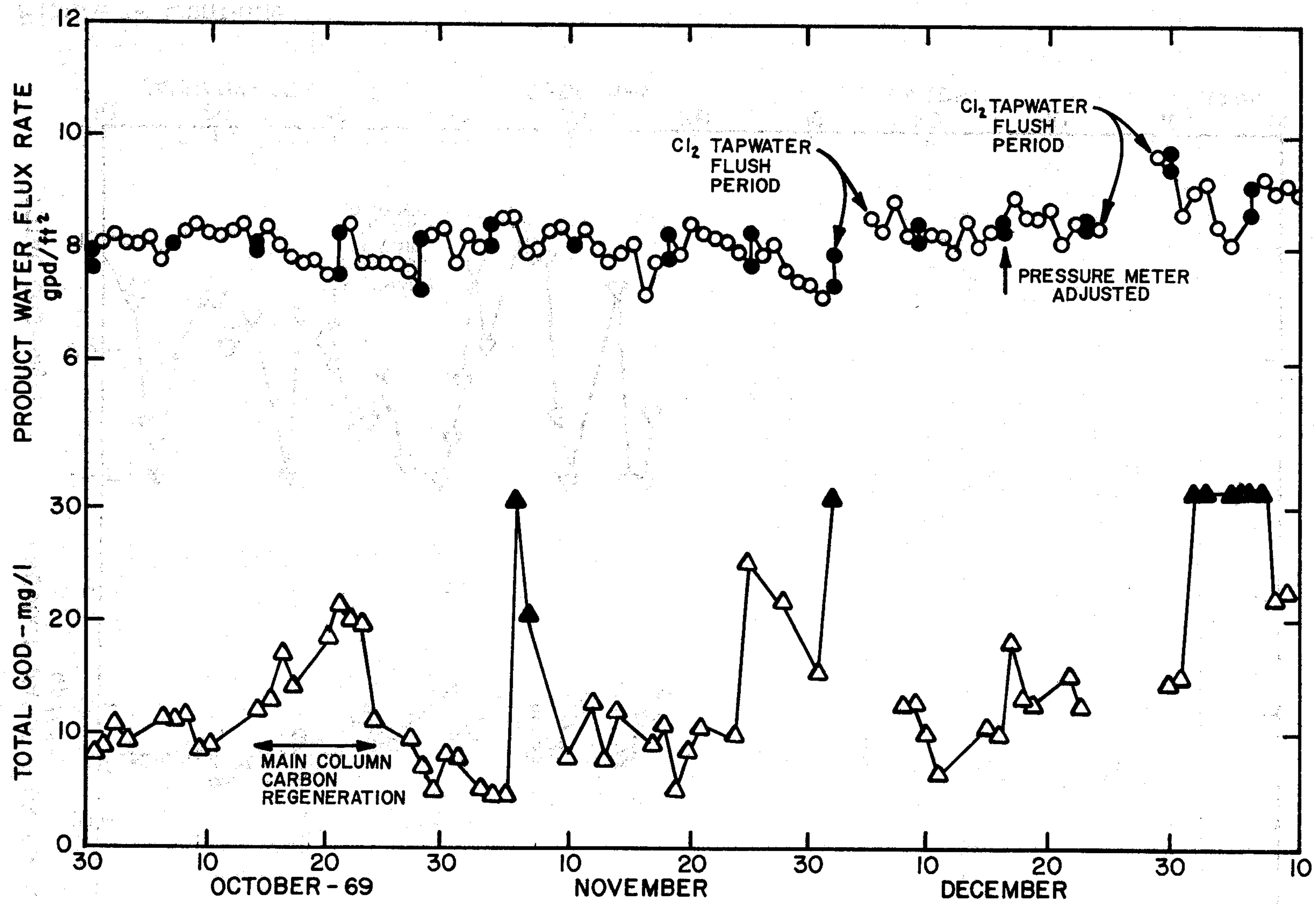
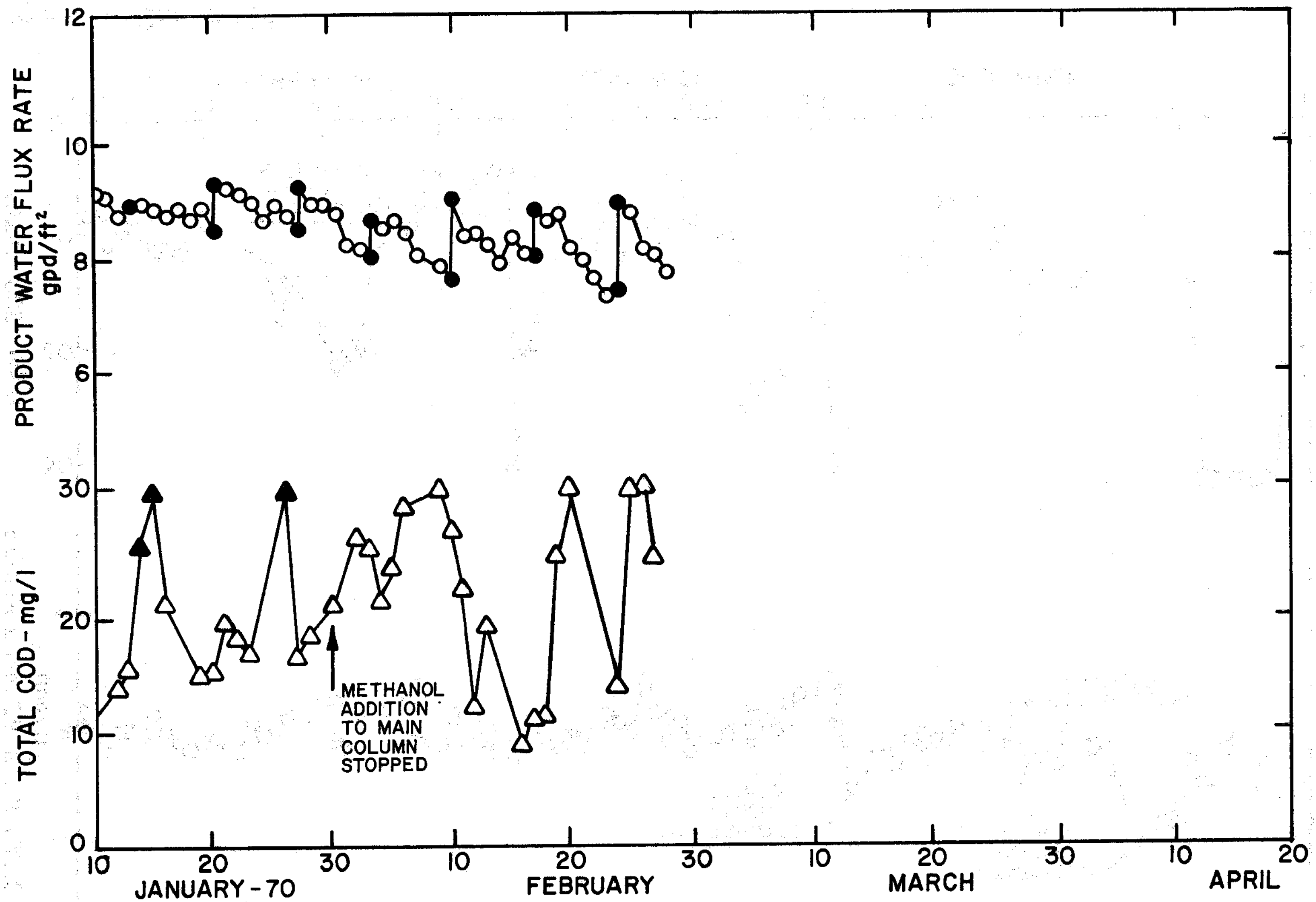


Figure 4. Continued



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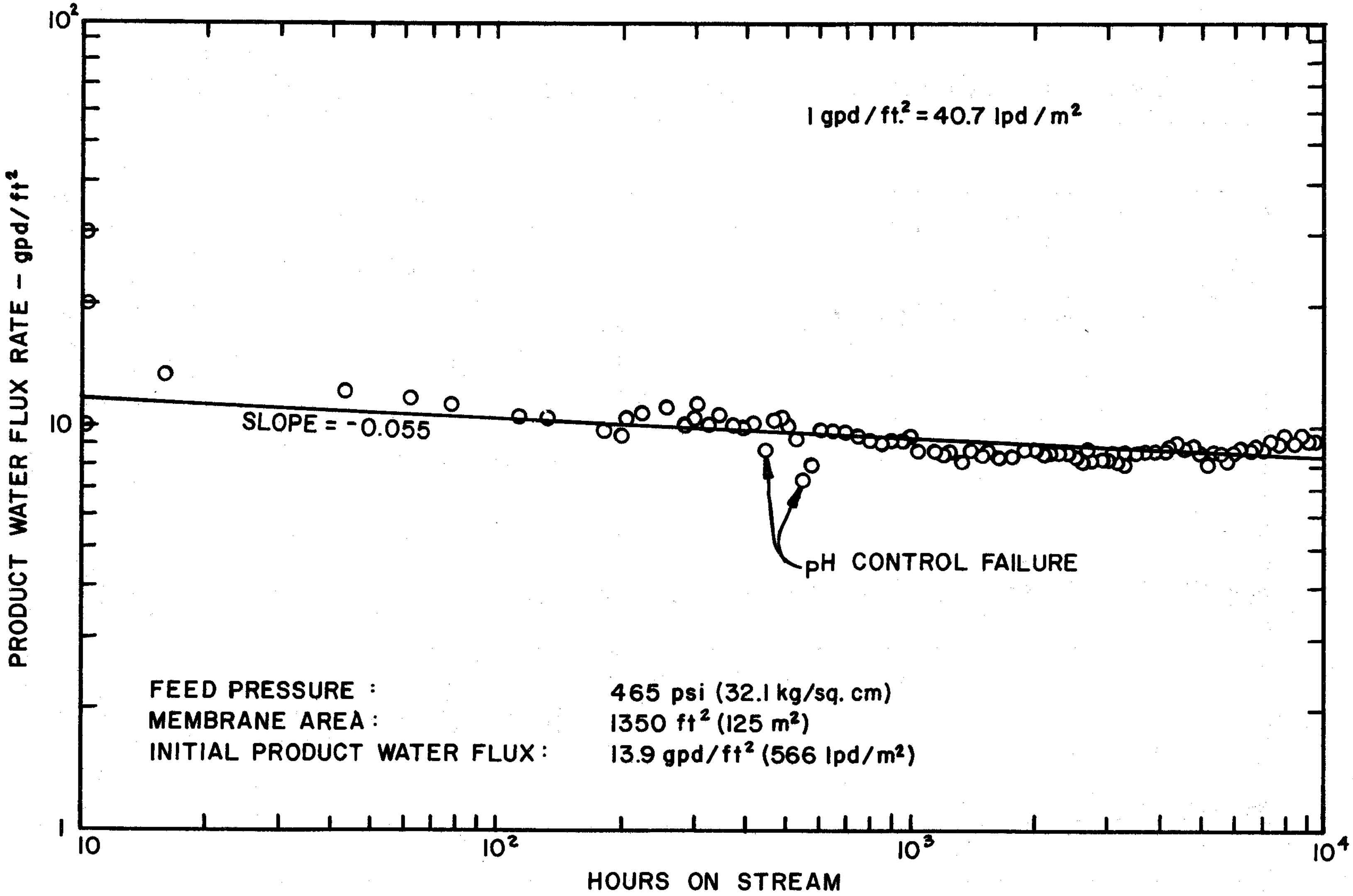

Figure 5. Decline rate of product water flux under constant feed pressure operation.

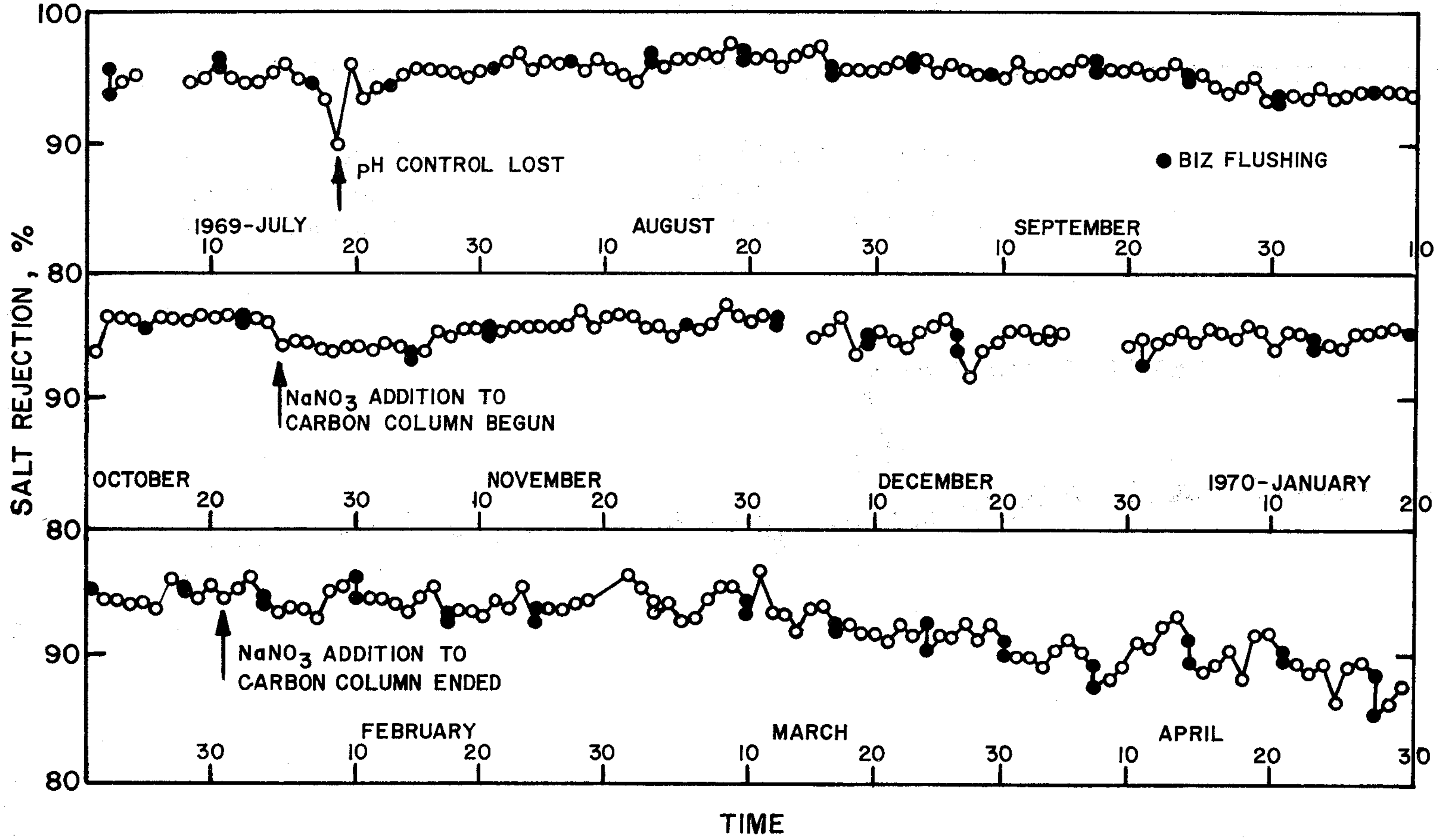

Figure 6. Variation of rejection vs. operation time under constant feed pressure operation.

in the feed. Thus, the increased concentration of the nitrate ion in the feed caused the overall rejection to decrease slightly.

During March of 1970, at approximately 6,000 hours of operation, the overall salt rejection started to decrease slowly. This trend continued throughout the remainder of the study. On August 16, 1970, when the system was taken offstream, the overall salt rejection decreased to 77 percent after a total of 9,475 hours of on-stream operation.

Figures 7 through 15 summarize the salt rejection variations from June 18, 1969 to August 14, 1970 for each of the nine pressure vessels. As indicated in these figures, the greatest decline in salt rejection occurred in pressure vessels 1, 2, and 3. The initial and final salt rejection values for these pressure vessels averaged about 93 percent and 45 percent, respectively. The salt rejection for the pressure vessels 4 through 7 decreased from 93 percent to 80 percent, and for the pressure vessels 8 and 9 from 94 percent to 90 percent.

Since each pressure vessel contained three spiral-wound modules in series, the salt rejection calculated for each pressure vessel represented the overall performance of the three modules. In order to determine which modules in each pressure vessel were responsible for the decline in salt rejection, a conductivity probe was used to make conductivity measurements of the entire system. The results of these measurements are summarized in Table 5. All modules in pressure vessels 1, 2, and 3 showed a deterioration in their ability to reject salts. The No. 2 module in the pressure vessel 4, No. 2 and 3 modules in the pressure vessel 5, and the No. 2 module in the pressure vessel 6 also showed a decline in salt rejection.

Water Quality

The chemical analyses conducted on the feed water, product water, and the brine waste during the first phase of the pilot plant study are summarized in Table 6. The percent rejections for the various ions are calculated using the blended feed and product values only. As indicated in the table, the overall rejection of the inorganic ions, as measured by the TDS reduction, was about 91 percent. The system demonstrated excellent rejection of calcium, magnesium, sulfate, and phosphate ions, while it seemed very poor in the rejection of potassium and nitrate ions.

CONSTANT PRODUCT FLUX RATE OPERATION

Feed Pressure

During the first phase of study, the feed pressure for the system operation was maintained constant at about 32.1 Kg/sq cm (465 psi). However, the feed pressure was varied during the second phase of the pilot plant study to maintain a constant product water flux rate of 407 l/sq m/day (10 gal/sq ft/day).

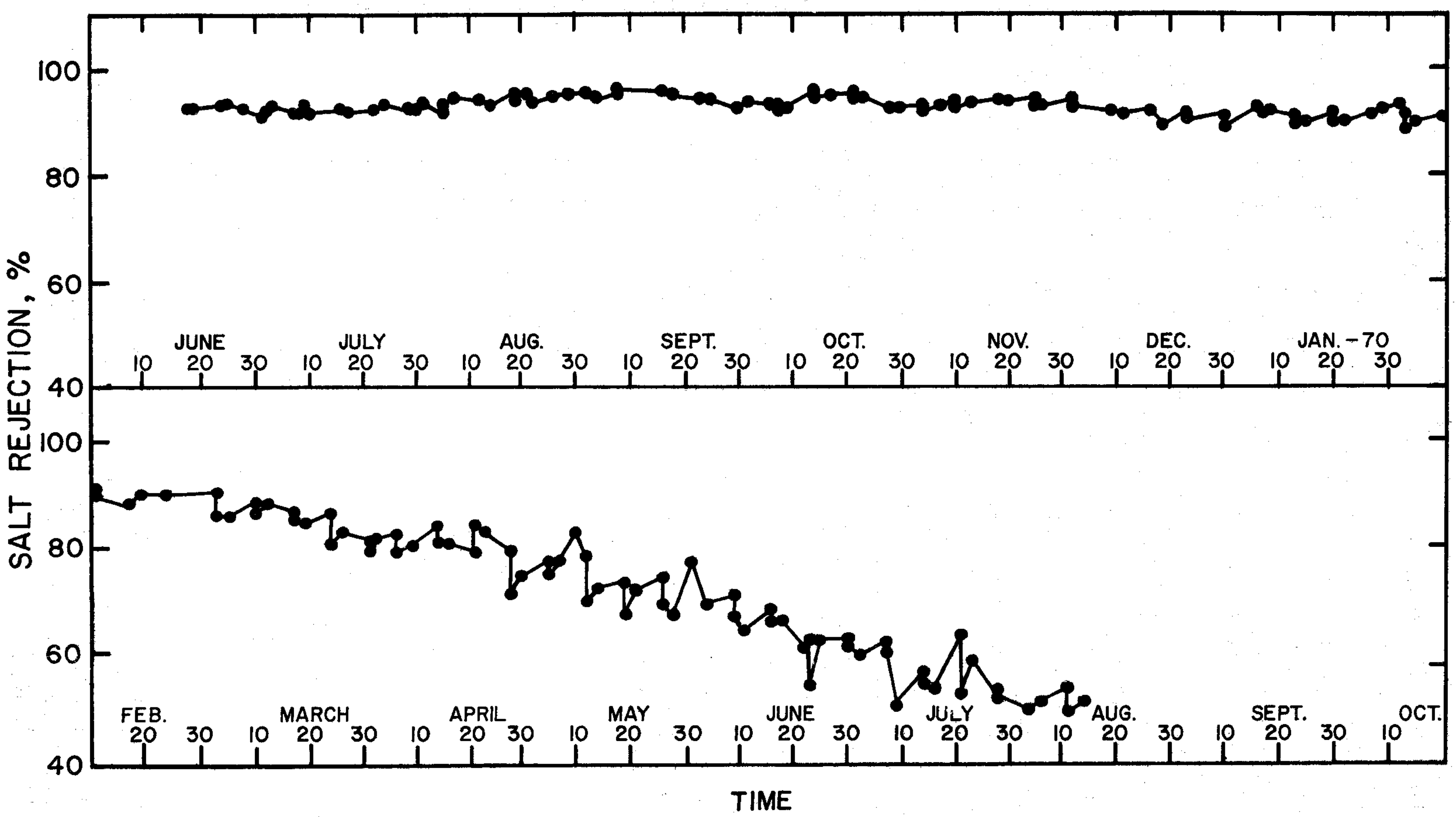

Figure 7. Salt rejection vs. operation time in pressure vessel no. 1.

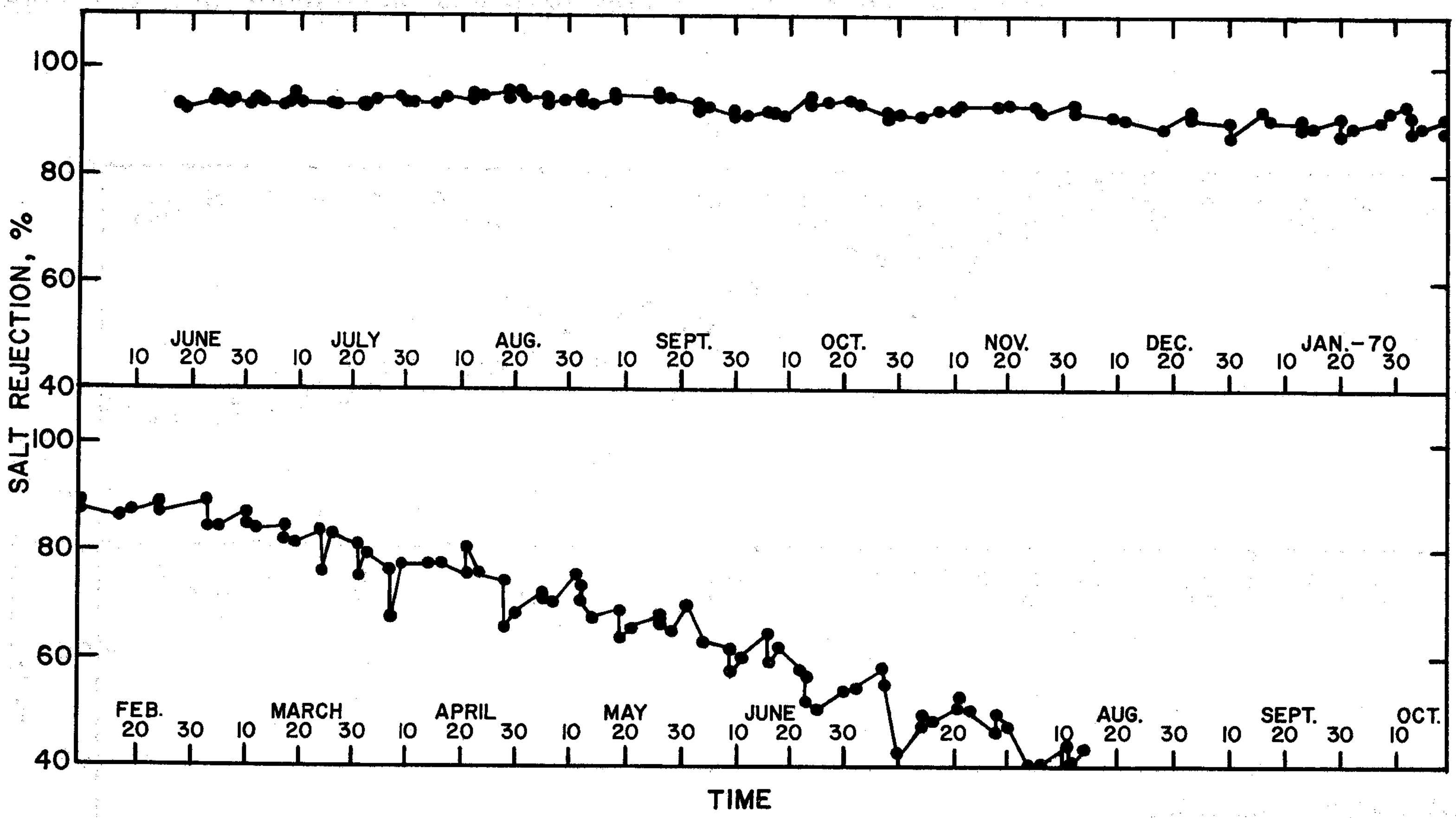

Figure 8. Salt rejection vs. operation time in pressure vessel no. 2.

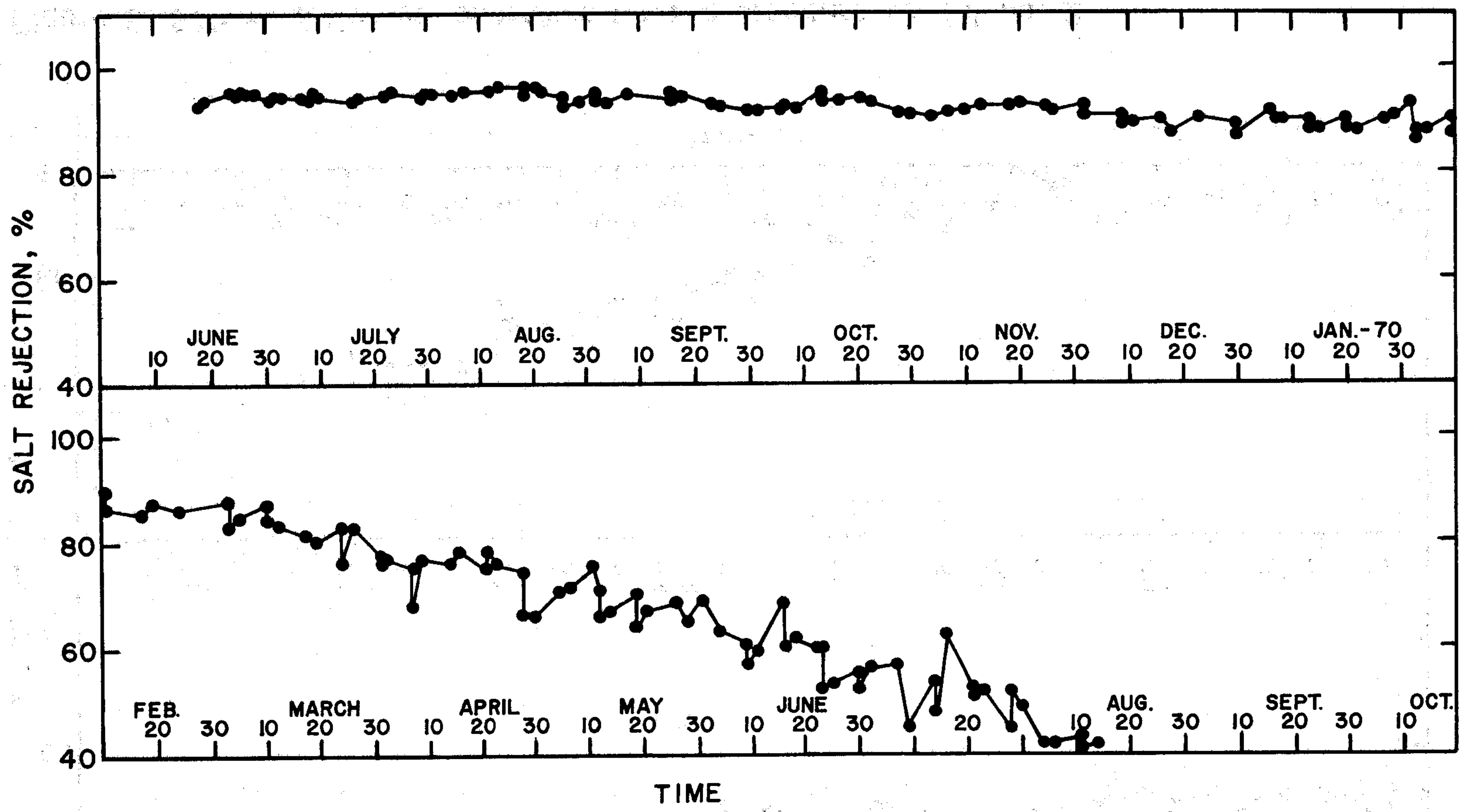

Figure 9. Salt rejection vs. operation time in pressure vessel no. 3.

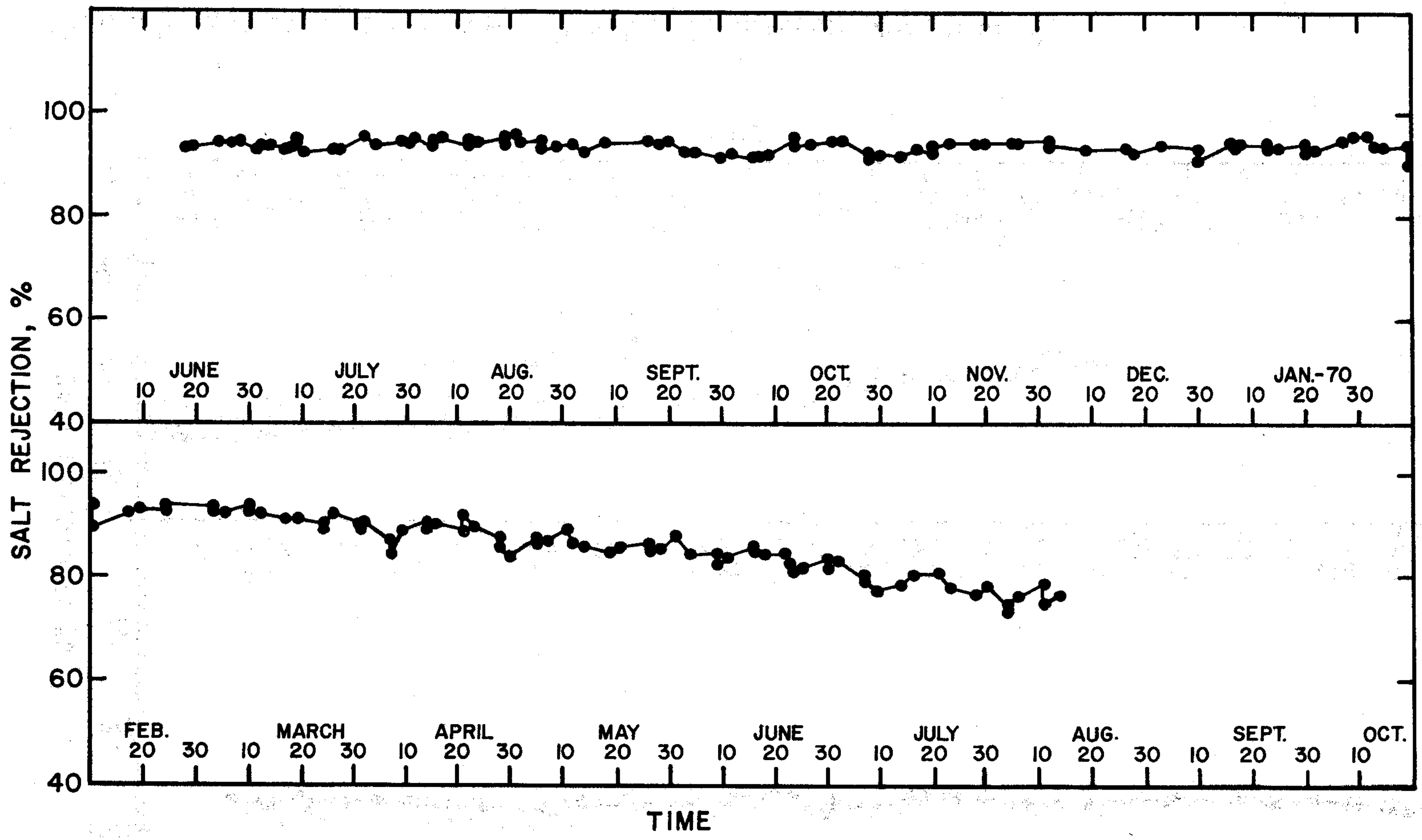

Figure 10. Salt rejection vs. operation time in pressure vessel no. 4.

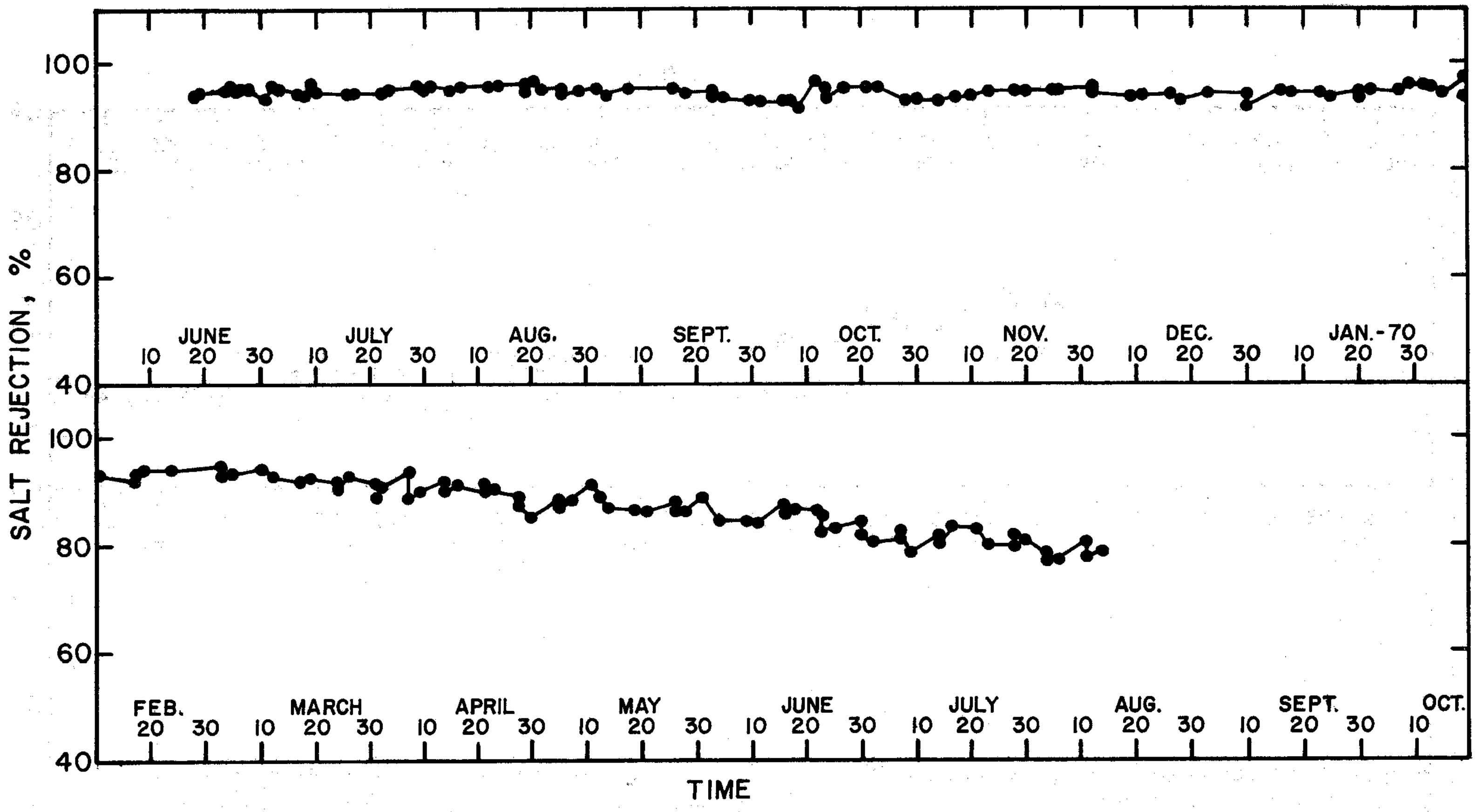

Figure 11. Salt rejection vs. operation time in pressure vessel no. 5.

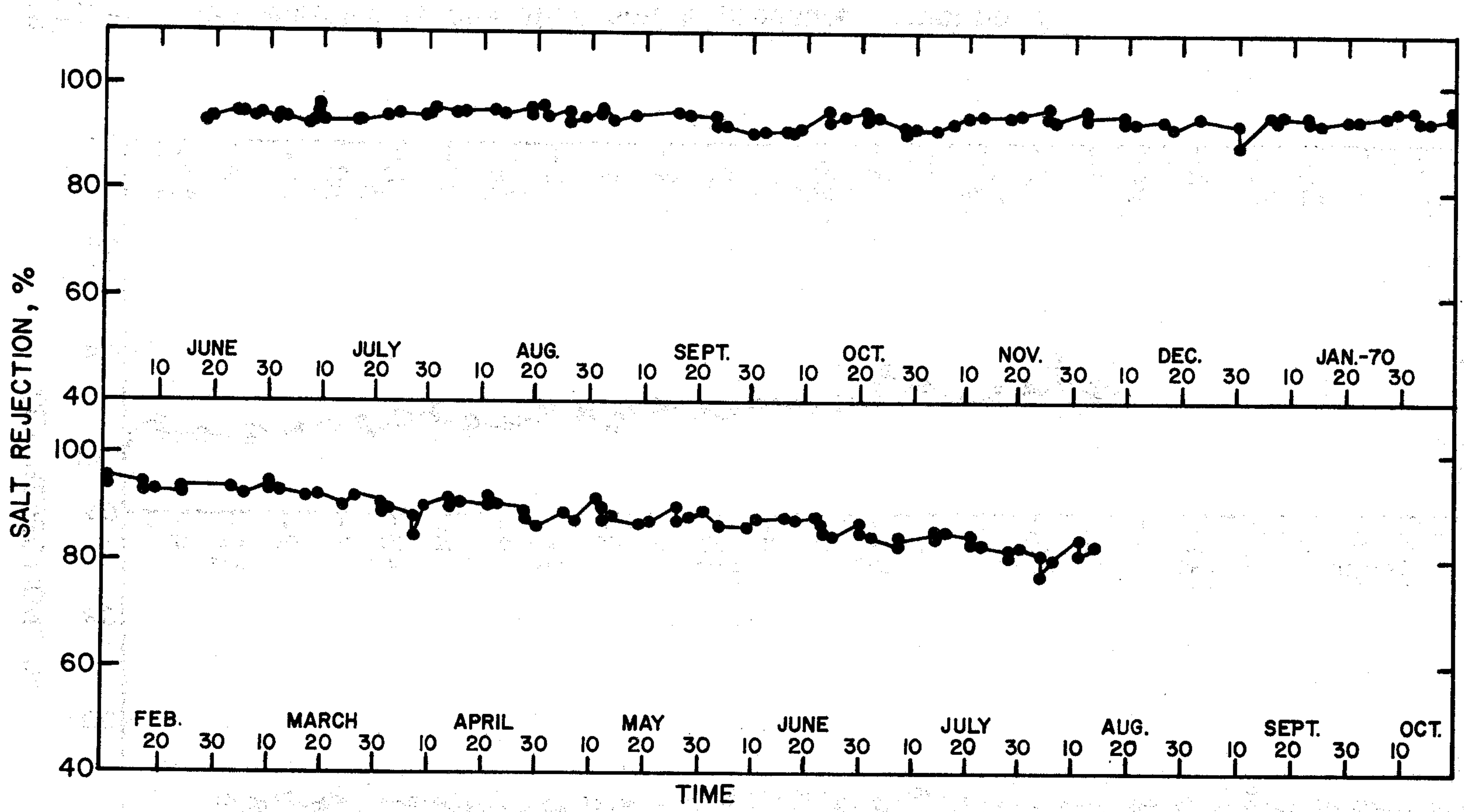

Figure 12. Salt rejection vs. operation time in pressure vessel no. 6.

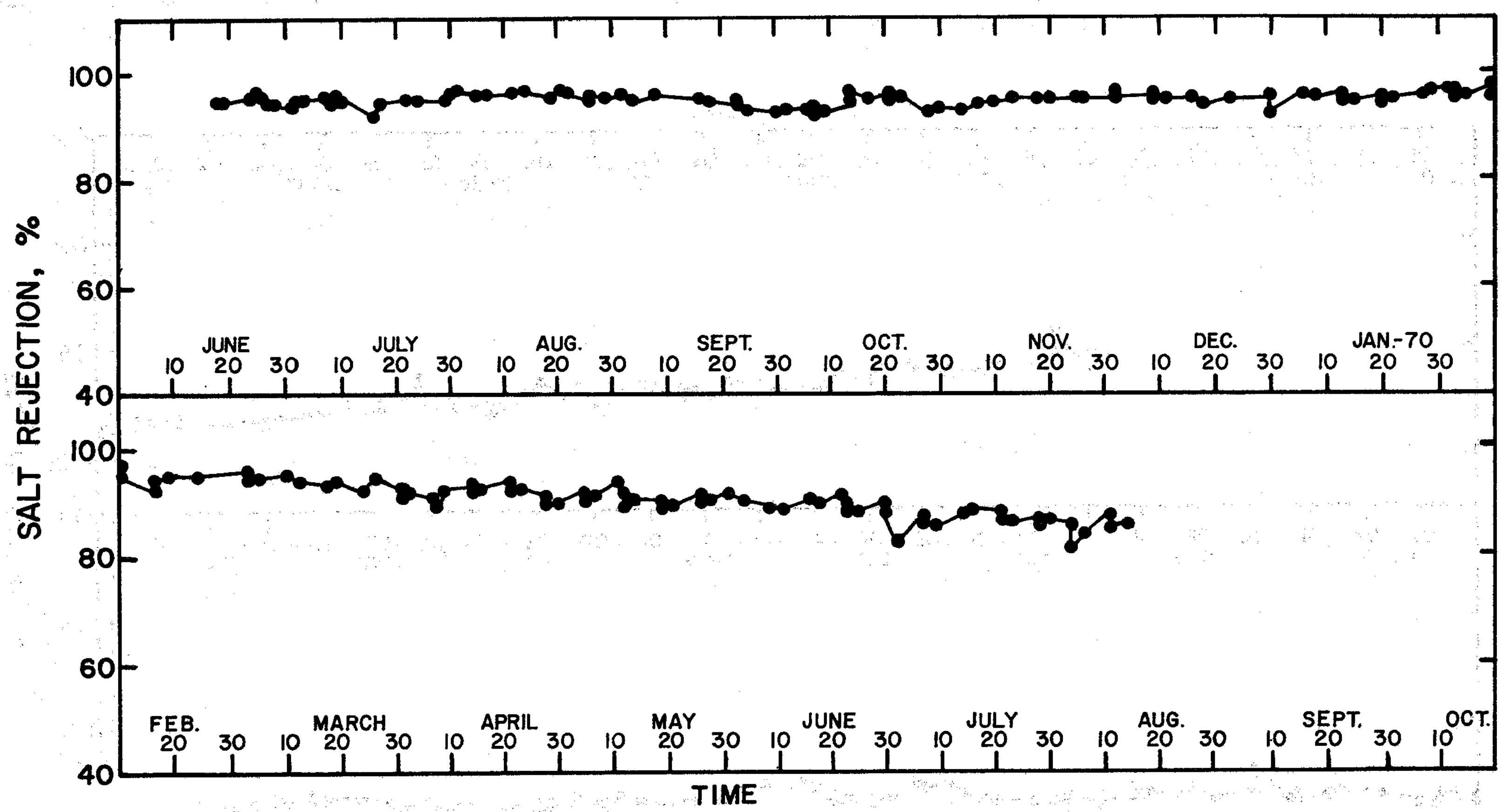

Figure 13. Salt rejection vs. operation time in pressure vessel no. 7.

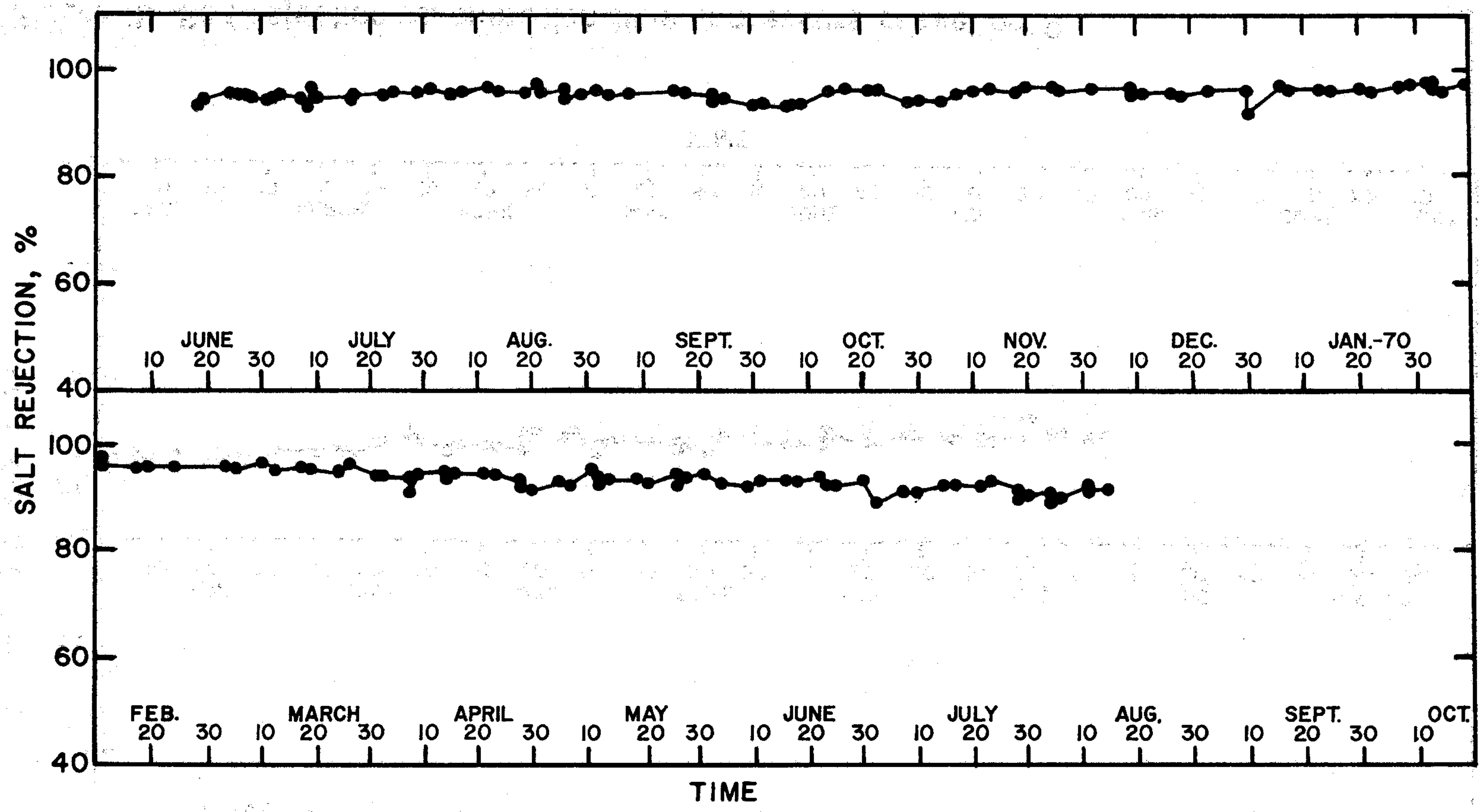

Figure 14. Salt rejection vs. operation time in pressure vessel no. 8.

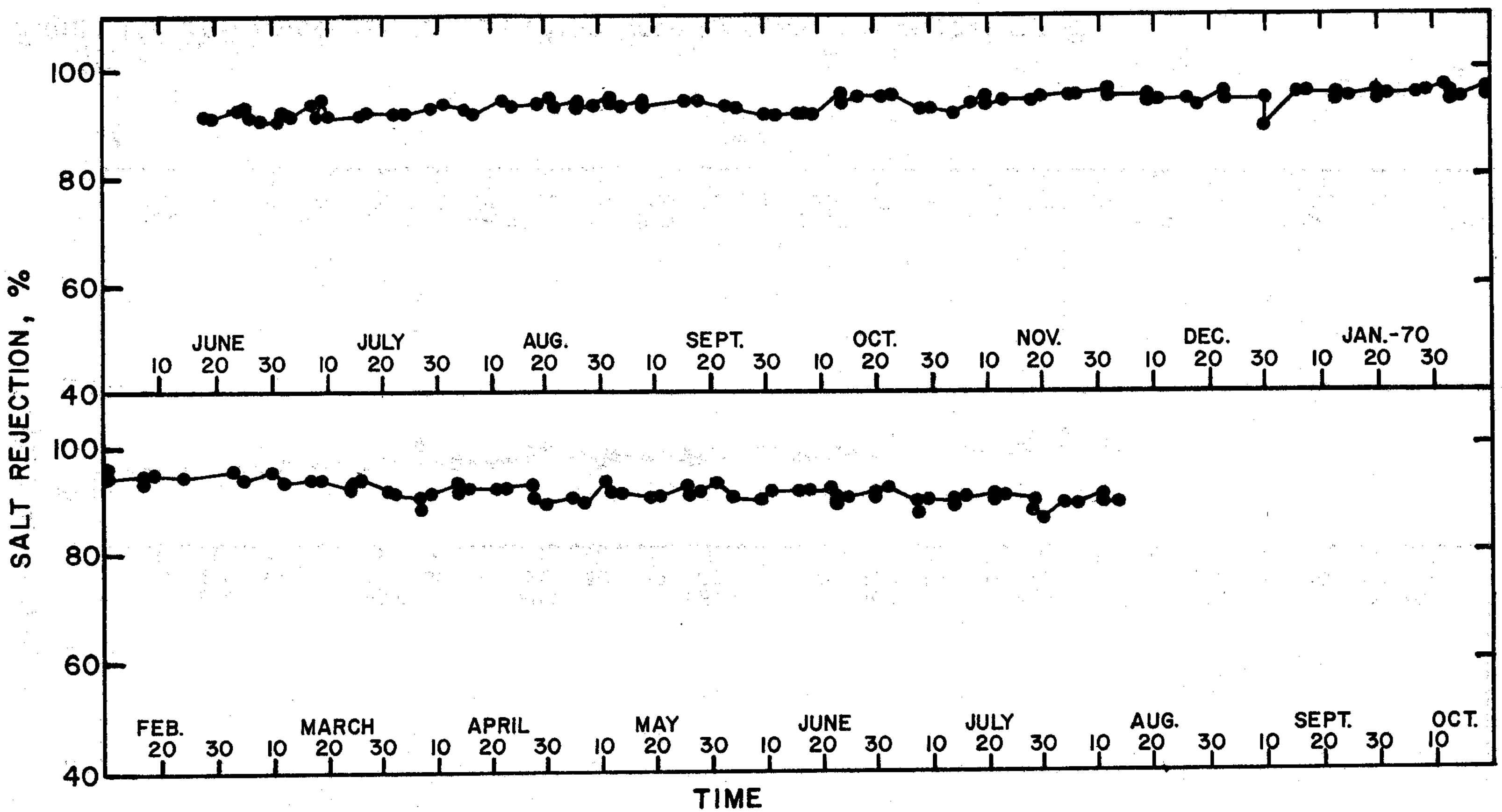

Figure 15. Salt rejection vs. operation time in pressure vessel no. 9.

TABLE 5. INDIVIDUAL MODULE SALT REJECTION TESTS CONDUCTED AT THE END OF CONSTANT FEED PRESSURE OPERATION STUDY

Pressure Vessel	Module #1				Module #2				Module #3			
1	Off	Off	Off	Off	Off	Off	Off	Off	Off	Off	Off	Off
2	Off	Off	Off	Off	Off	Off	Off	Off	Off	Off	Off	Off
3	Off	Off	Off	Off	Off	Off	Off	Off	Off	Off	Off	Off
4	7.8	8.2	7.7	7.6	6.9	Off	Off	7.4	6.3	8.6	8.7	10.0
5	5.9	6.2	5.9	5.7	5.5	5.8	6.9	Off	Off	8.0	9.6	9.0
6	6.7	7.9	8.2	5.4	7.5	Off	Off	Off	9.8	10.0	6.6	7.4
7	6.4	5.2	6.9	6.8	4.1	3.9	4.0	4.4	7.5	8.3	7.8	8.0
8	4.9	3.8	3.7	4.3	5.2	5.4	5.8	5.9	6.5	6.8	3.6	4.1
9	6.5	7.0	5.9	5.1	6.2	6.0	7.2	6.7	6.7	6.8	6.7	6.4

Notes:
1. Measurements taken at one foot (30.5 cm) intervals using an Industrial Instruments Model RA 4-WA-S4-Kf.
2. Readings should be multiplied by 30 (cell constant) to get conductivity, μmhos/cm.
3. Off reading was off scale.

TABLE 6. SUMMARY OF WATER QUALITY ANALYSES FOR THE PERIOD OF ZERO TO 9,475 HOURS OF CONSTANT FEED PRESSURE OPERATION

Parameter	Blended Feed	Product	Brine	Rejection %
Sodium, mg/l Na	129	15.0	452	88.5
Potassium, mg/l K	16.5	4.7	49.1	71.5
Calcium, mg/l Ca	40.8	1.5	132	96.5
Magnesium, mg/l Mg	24.5	0.8	98.5	97.0
Chloride, mg/l Cl	95	14.1	326	85.0
Sulfate, mg/l SO_4	318	3.0	1310	99.0
Phosphate, mg/l PO_4-P	10.4	0.15	38.8	98.5
Ammonia, mg/l NH_3-N	13.9	1.6	44.9	88.5
Nitrate, mg/l NO_3-N	7.7	3.5	16.6	54.5
Turbidity, JTU	1.0	0.1	2.9	90.0
Total COD, mg/l	10.1	1.0	32.7	90.0
TDS, mg/l	744	67	2800	91.0

Notes: 1. Analyses were run on once-a-week grab samples taken at 8:00 A.M.

2. Blended feed was a mixture of carbon-treated secondary effluent, sulfuric acid and chlorine solution.

3. Rejection (%) = 100X (Blended feed concentration - Product concentration)/(Blended feed concentration)

4. COD = Chemical oxygen demand.

5. TDS = Total dissolved solids.

A summary of the overall performance during the second phase of the pilot plant study is shown in Figure 16. As indicated in Figure 16, the initial feed pressure necessary to maintain this constant flux was about 24.8 Kg/sq cm (360 psi). The system remained at this feed pressure until 120 hours of operation when the water recovery was increased from 75 to 80 percent. After 150 hours of operation, the feed pressure required for the system to maintain the 407 l/sq m/day (10 gal/sq ft/day) product flux rate was found to fluctuate between 26.9 Kg/sq cm (390 psi) and 34.5 Kg/sq cm (500 psi). Further increase of the feed pressure was noted at about 2,400 hours of operation. This increase was believed to be a result of the insufficient velocity in the circulation of the cleaning solution and the water flush through the pressure vessels during the membrane cleaning cycle. At 2,830 hours of operation, the modules were cleaned twice a week instead of once a week. This new practice was continued for a three week period to thoroughly clean up the membrane surface. The cleaning solution flow rate through each pressure vessel was increased from 11.4 to 34.1 lpm (3 to 9 gpm). The water flushing flow rate was also increased from 11.4 to 26.5 lpm (3 to 7 gpm). After this flow rate adjustment, there was a decrease in the feed pressure. The pilot plant system was depressurized for approximately 84 hours after an enzyme-detergent cleaning at 3,553 hours of operation. This special depressurization treatment resulted in a 8.3 Kg/sq cm (120 psi) decrease in the feed pressure to maintain the constant 407 l/sq m/day (10 gal/sq ft/day) product water flux rate. At the end of the pilot plant study, the rapid decline of the salt rejection was accompanied with a low feed pressure, about 20.7 Kg/sq cm (300 psi). This behavior could be attributed to some membrane breakup developed in the system.

<u>Salt Rejection</u>

During the second phase of the pilot plant study, the product water flux rate was kept constant at 407 l/sq m/day (10 gal/sq ft/day) by varying the feed pressure. As indicated in Figure 16, under this mode of operation, the overall salt rejection was steadily maintained at 95 percent throughout the initial 3,600 hours of opeation. After 3,600 hours of operation, the salt rejection started to decline gradually. This decline was primarily attributed to the poor salt rejection of the modules in pressure vessel No. 2. At 4,414 hours of operation, these modules were replaced with three used modules which had 1,933 hours of operating time accumulated from other similar study. At the time of the module replacement, the salt rejection was about 60 percent for the original set of modules. The new set of modules substantially improved the salt rejection to 94 percent. However, the product water flux rate for the new set of modules in the pressure vessel No. 3 was only about 317 l/sq m/day (7.8 gal/sq ft/day), while the overall flux rate for the entire system was 407 l/sq m/day (10 gal/sq ft/day). The explanation was that the membranes might have been affected by the irreversible compaction and fouling. In addition, the three replacement used modules were operated under 38 Kg/sq cm (550 psi) feed pressure in a previous study, while they were operated under 27.6 Kg/sq cm (400 psi) in this study.

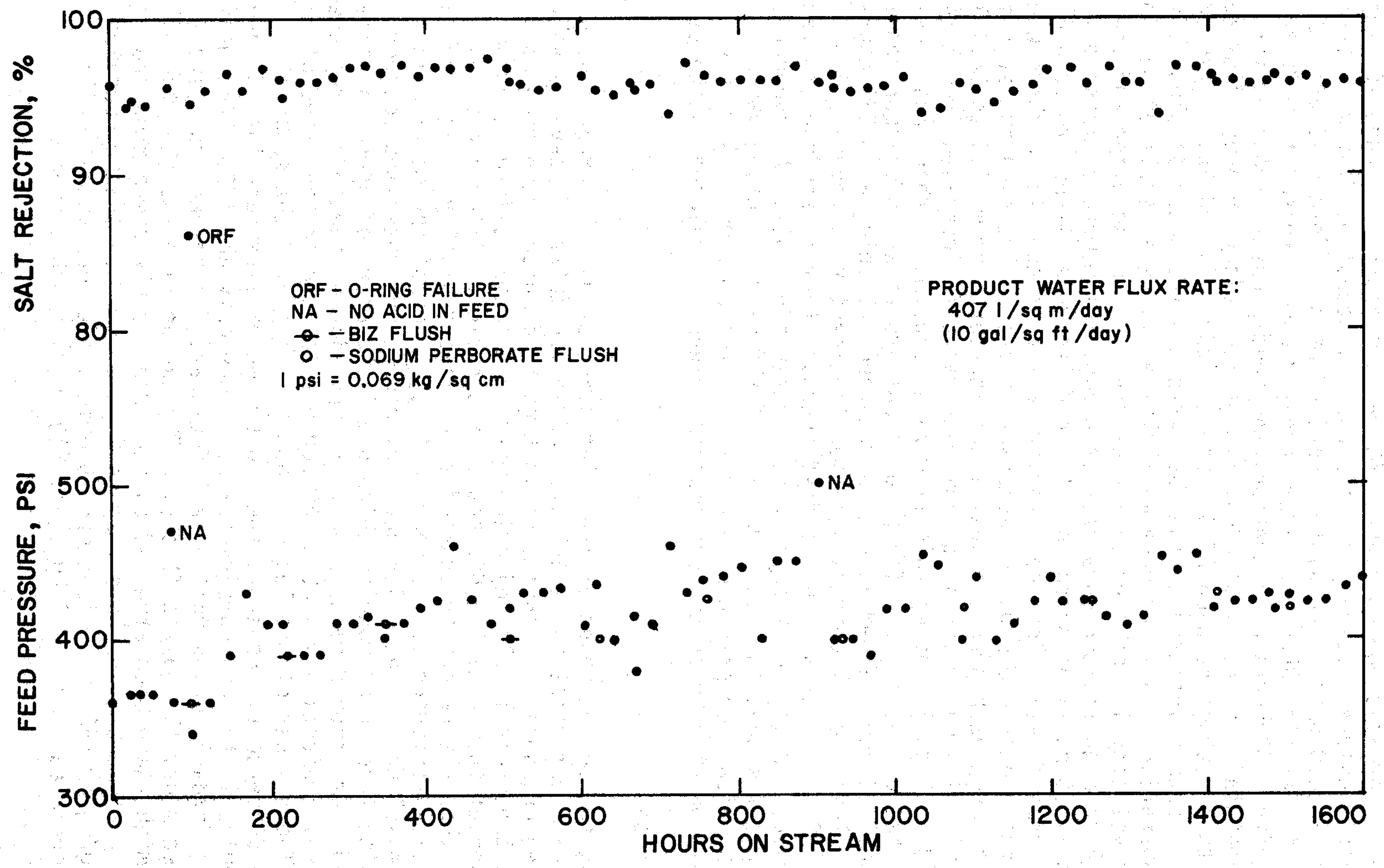

Figure 16. Salt rejection and feed pressure variation vs. operation time under constant flux rate operation.

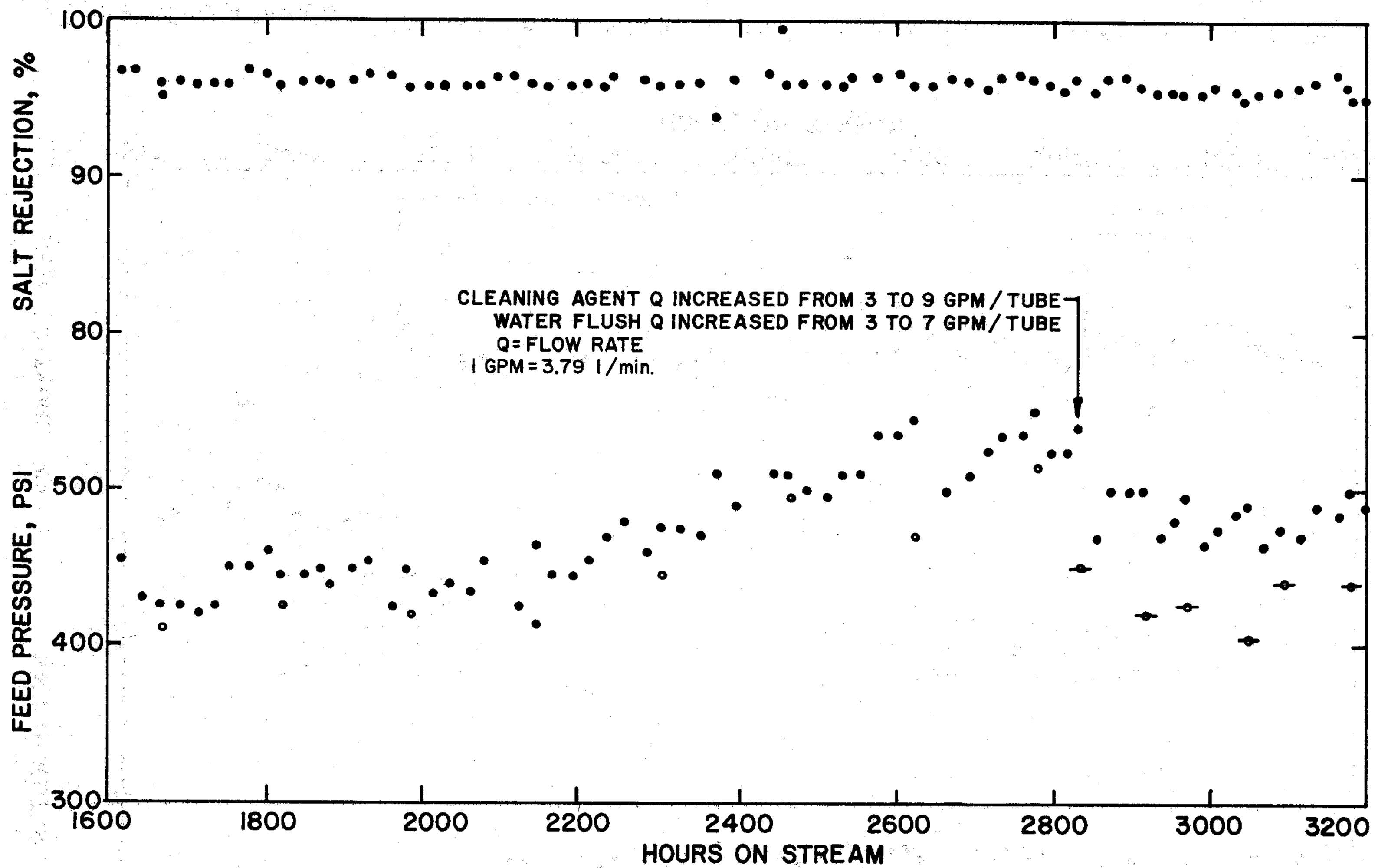

Figure 16. Continued.

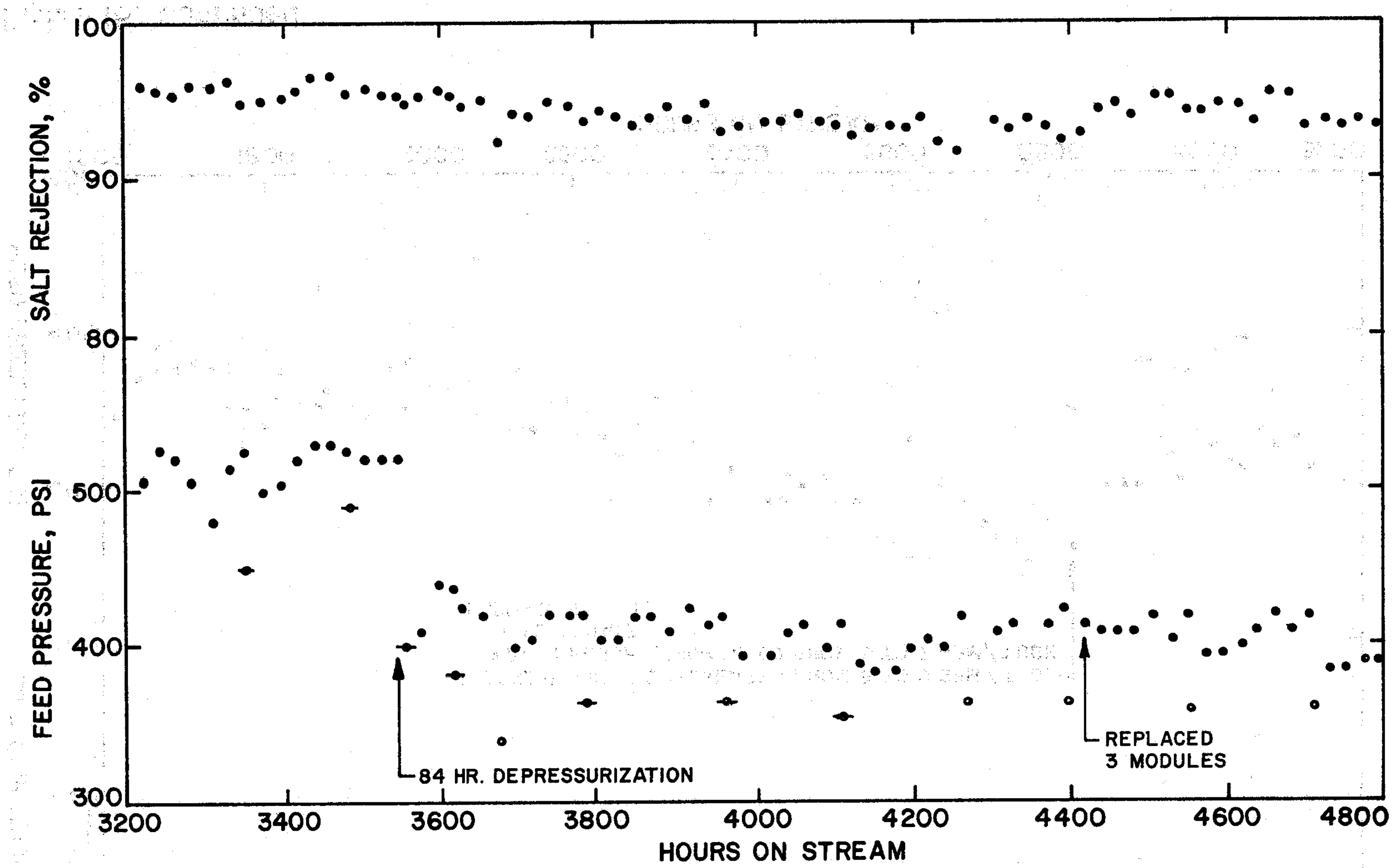

Figure 16. Continued.

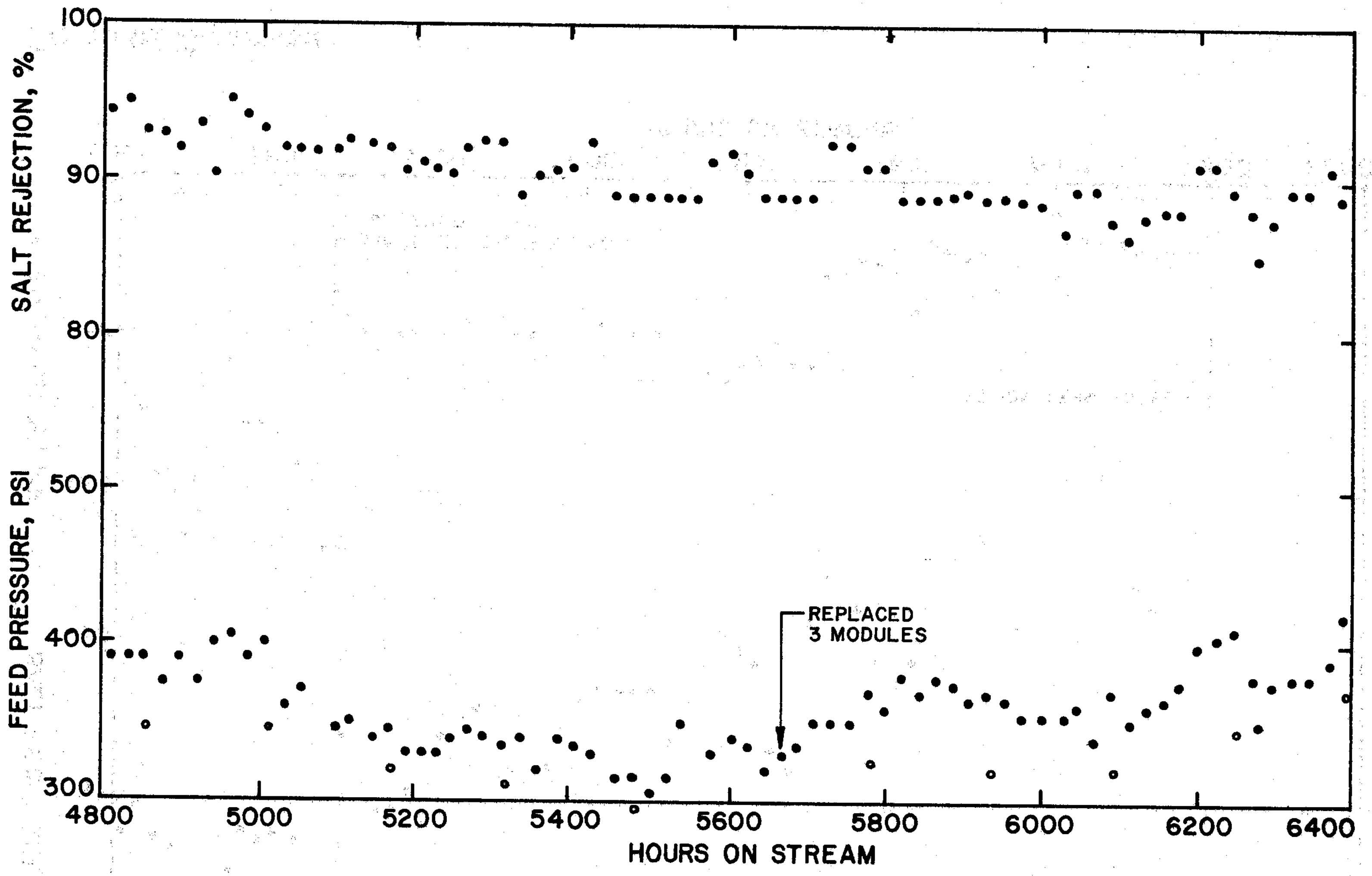

Figure 16. Continued.

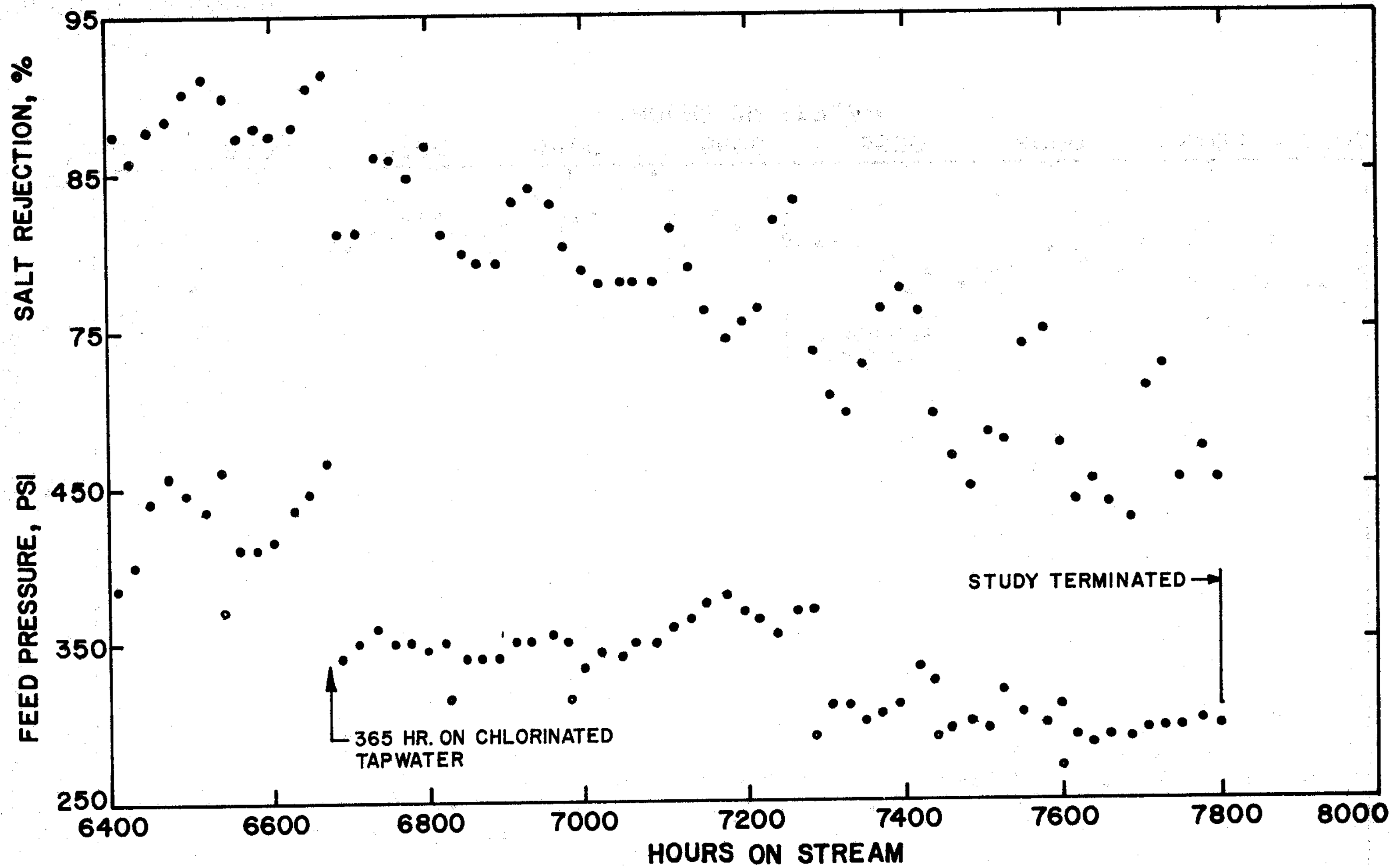

Figure 16. Continued.

At 6,677 hours of operation, the feed pump failed, which necessitated a preservation of the membrane modules by feeding the system with chlorinated tap water for a period of 365 hours until the pump was repaired. Upon resumption of the system operation, the salt rejection was found to have decreased from 91 percent to 81 percent. The occurrence of the substantial decrease in salt rejection efficiency at the same time that the system was inoperative and on chlorinated tap water could have been a coincidence. Within the next 1,100 hours of operation, the salt rejection rapidly declined to 65 percent at which time the entire study was terminated at 7,803 hours of operation.

Water Quality

The average water quality data for the system operations during the initial 6,700 hours are presented in Table 7. As shown in both Table 6 and Table 7, similar rejection efficiencies for various ions were achieved by the system operated with constant feed pressure in the first phase of the study and the system operated with constant product flux rate in the second phase of the study. However, the ion rejection by the system with constant product flux rate mode of operation greatly reduced after 6,700 hours of on-stream operation. Table 8 shows the summary of the water quality analyses from 6,700 to 7,803 hours of operation in the second phase of the pilot plant study.

MEMBRANE MODULE STABILITY

Under Constant Feed Pressure Operation

The performance of the membrane modules in each of the nine pressure vessels (three modules per vessel) from time zero to 4,800 hours of operation is summarized in Table 9. The following points of clarification may be necessary for interpreting the data presented in this table.

A. The difference between the initial (at time zero) and present (at 4,800 hours of operation) feed pressures in the downstream modules was the result of decreased pressure drop through the system. At time zero, the total feed (equivalent to product plus brine) was about 61.3 lpm (16.2 gpm), while at 4,800 hours the total feed was dropped to 45.4 lpm (12 gpm). Consequently, the pressure drop through the system also decreased from 5.9 Kg/sq cm (85 psi) at time zero to 3.1 Kg/sq cm (45 psi) at 4,800 hours. This caused the feed pressure in the downstream vessels to increase. The initial and present feed pressure of vessels No. 1, 2, and 3 remained the same because the feed pressure was a controlled operating parameter.

B. The small differences between the "A" (water permeability coefficient) values specified by the Gulf Environmental Systems Company (GESCO) and those calculated from the initial operating conditions could be explained as follows: The values of "A" specified by GESCO were taken from a test run at 41.4 Kg/sq cm (600 psi) and 2,000 mg/l sodium chloride feed solution. The "A" values calculated at time zero were based on a

TABLE 7. SUMMARY OF WATER QUALITY ANALYSES FOR THE PERIOD OF ZERO TO 6,700 HOURS OF CONSTANT PRODUCT FLUX RATE OPERATION

Parameter	Raw Feed mg/l	Blended Feed mg/l	Product mg/l	Brine mg/l	Rejection %
Na	109	168	23.8	390	86.9
K	13.3	20.2	2.5	40.2	87.1
Ca	58.5	97.3	2.4	226	97.5
Mg	12.6	56.9	0.47	49.0	99.2
Cl	104	173	31.1	520	82.1
SO_4	69	467	4.3	1093	99.1
PO_4-P	18.3	15.4	0.55	34.1	96.4
NH_3-N	14.4	24.3	2.5	52.1	89.6
NO_3-N	1.30	1.43	0.85	2.17	40.9
TCOD	8.1	12.7	0.43	26.5	96.6
DCOD	4.7	7.4	0.38	18.7	94.8
TTOC	2.4	3.2	0.47	6.3	85.1
DTOC	1.2	2.5	0.40	5.0	83.7
TDS	547	1084	77.7	1778	92.8
TURBIDITY, JTU	1.2	1.8	0	3.1	100

Notes: 1. Raw feed was carbon-treated secondary effluent.

2. Analyses were run on once-a-week grab samples taken at 8:00 A.M.

3. Difference between raw feed and blended feed was due to H_2SO_4 addition, chlorination and brine recirculation.

4. TOC - Total organic carbon.

TABLE 8, SUMMARY OF WATER QUALITY ANALYSES FOR THE PERIOD OF 6,700 TO 7,803 HOURS OF CONSTANT PRODUCT FLUX RATE OPERATION

Parameter	Raw Feed mg/l	Blended Feed mg/l	Product mg/l	Brine mg/l	Rejection %
Na	114	136	57	257	58.1
K	12.3	14.9	5.8	27.3	61.1
Ca	53.0	67.7	11.5	135.6	83.0
Mg	11.4	14.3	2.89	24.6	79.8
Cl	132	143	109	193	23.8
SO_4	58.6	324	52.8	760	83.7
PO_4-P	10.0	12.2	3.56	28.2	70.8
NH_3-N	17.0	18.3	7.88	36.7	56.9
NO_3-N	0.75	0.55	0.47	0.45	14.5
TCOD	5.0	6.0	2.4	138	60.0
DCOD	2.6	3.4	0.8	6.3	76.6
TDS	572	807	291	1491	63.9
TURBIDITY, JTU	1.4	2.3	0	3.6	100

Notes: 1. Raw feed was carbon-treated secondary effluent.

2. Analyses were run on once-a-week grab samples. taken at 8:00 A.M.

3. Difference between raw feed and blended feed was due to H_2SO_4 addition, chlorination, and brine recirculation.

TABLE 9

PERFORMANCE OF MODULES IN EACH PRESSURE VESSEL
FROM TIME ZERO TO 4,800 HOURS OF CONSTANT FEED PRESSURE OPERATION

Pressure Vessel	1	2	3	4	5	6	7	8	9
Initial Feed Pressure, psi	465	465	465	430	430	405	405	390	380
Present Feed Pressure, psi	465	465	465	450	450	440	440	428	420
Initial Flux Rate, gpd/ft^2	15.4	15.8	14.8	13.9	14.2	13.8	12.9	12.8	12.1
Present Flux Rate, gpd/ft^2	10.7	10.9	10.5	9.2	9.4	8.8	8.0	8.5	8.6
% Reduction in Flux Rate	30	31	29	34	34	36	38	34	29
Initial "A" x 10^5 Specified by GESCO	2.17	2.21	2.19	2.18	2.17	2.23	2.03	2.18	2.16
Initial "A" x 10^5 Calculated	2.29	2.36	2.20	2.25	2.29	2.36	2.20	2.28	2.22
Present "A" x 10^5 Calculated	1.61	1.64	1.56	1.43	1.46	1.40	1.26	1.38	1.43
% Reduction in "A" Value	30	31	29	36	36	41	43	39	36

(Continued)

TABLE 9 (continued)

Pressure Vessel	1	2	3	4	5	6	7	8	9
Initial Influent Flow, gpm	5.36	5.36	5.36	5.65	5.65	4.19	4.19	5.59	4.26
Initial Effluent Flow, gpm	3.77	3.77	3.77	4.19	4.19	2.80	2.80	4.26	3.00
Average Initial Flow, gpm	4.57	4.57	4.57	4.92	4.92	3.50	3.50	4.93	3.63
Present Influent Flow, gpm	3.94	3.94	3.94	4.24	4.24	3.27	3.27	4.79	3.90
Present Effluent Flow, gpm	2.83	2.83	2.83	3.27	3.27	2.40	2.40	3.90	3.00
Average Present Flow, gpm	3.39	3.39	3.39	3.76	3.76	2.84	2.84	4.35	3.45
Initial Salt Rejection Specified by GESCO	93.5	94.5	93.0	93.5	94.5	94.0	94.0	93.5	92.5
Initial Salt Rejection Calculated	93.0	93.0	93.0	93.5	94.0	93.5	94.5	93.5	91.5
Present Salt Rejection Calculated	90.0	89.0	88.0	93.0	95.0	94.0	96.0	96.5	96.0
Flux Decline Slope	-0.066	-0.077	-0.081	-0.077	-0.075	-0.104	-0.089	-0.077	-0.052

Notes: 1. "Initial" = at time zero; "Present" = at 4,800 hours of operation.
2. GESCO = Gulf Environmental Systems Company, San Diego, California.

feed containing approximately 700 mg/l TDS and a feed pressure varying from 26.2 Kg/sq cm (380 psi) to 32.1 Kg/sq cm (465 psi). The GESCO pointed out that at lower feed pressure, the value of "A" was higher than at higher feed pressure. This would explain why all "A" values calculated from the initial conditions were slightly higher than the values specified by the GESCO. A second explanation for the discrepancy in "A" values was that the osmotic pressure of the feed water was neglected in the calculations. This simplified the calculations and only introduced an error of 1 or 2 percent.

C. The influent, effluent, and average flows were calculated by assuming equal flow distribution in the parallel pressure vessels.

The following observations were made from the data presented in Table 9:

A. The modules in the pressure vessels 6 and 7, which experienced the highest reduction in flux rates, the highest reduction in "A" values, and the highest in flux decline slopes, showed the lowest average feed flows. Initially, the feed flow to these modules averaged 13.2 lpm (3.5 gpm); at 4,800 hours of operation, this declined to 10.6 lpm (2.8 gpm). The GESCO recommended that the minimum flow in each module should be between 11.4 lpm (3 gpm) and 15.1 lpm (4 gpm).

B. The modules in vessel 9 had the lowest flux decline slope. Since these modules received the poorest quality feed, one would expect the flux decline slope to be greater than that experienced in the preceding modules. This apparent discrepancy may be explained by noting the differences between initial (26.2 Kg/sq cm or 380 psi) and present (29 Kg/sq cm or 420 psi) feed pressures. The lower flux decline slope observed for the modules in the vessel 9 occurred because the operating pressure increased with time. This indicates a true picture of the effect of fouling on the entire system cannot be obtained by looking at the individual flux decline slopes. A better measure of fouling would be the decrease in "A" values between time zero and 4,800 hours. The percent reduction in "A" value for the modules in vessel 9 was greater than those experienced by the modules in vessels 1, 2, and 3. This indicates that the fouling in the downstream modules was more severe than in the upstream modules. The changes in the feed pressure and average flows during the study make it difficult to determine the effects of fouling through the entire system.

C. The salt rejection for the modules in the vessels 1, 2, and 3 decreased from the initial values. While they stayed the same in vessels 4 and 5, they increased in vessels 6 to 9. The modules in the vessels 1, 2, and 3 had exhibited an increase in salt rejection up to approximately 3,000 hours after which the salt rejection started to decrease slowly.

At the end of the first phase of the pilot plant study, all the spiral-wound modules were removed from the system and sent to the Gulf Environmental Systems Company for testing to determine which modules had lost salt rejection ability and why this had occurred. The results of the GESCO tests are summarized in Table 10. The tests were conducted at 55.2 Kg/sq cm (800 psi) with 10,000 mg/l sodium chloride solution.

As indicated in Table 10, the salt rejection of the lead modules definitely fell off, while some of the modules in the pressure vessels 8 and 9 still had rejections above 80 percent. The number of distribution of salt rejection range is indicated below:

Salt Rejection (%)	Number of Modules in Each Range
0 - 9	0
10 - 19	1
20 - 29	7
30 - 39	1
40 - 49	1
50 - 59	4
60 - 69	4
70 - 79	4
80 - 89	5
90 - 99	0
	27

Table 10 also shows the water permeation coefficient before and after the pilot plant study, which accumulated a total of 9,475 hours of on-stream operation. In all cases except three, the permeability coefficient dropped below the initial value. No significant location dependence of the decline was demonstrated.

Four modules among the twenty-seven modules were selected for further dye checking, visual inspection, and membrane sample testing. The results of these observations are shown in Table 11.

The GESCO membrane tests did not show the exact cause of the membrane deterioration. However, three possible fouling mechanisms were postulated:

A. Hydrolysis of the membrane caused by the high pH of the enzyme-detergent cleaning solution.

B. Some trace substances in the feed water attacked the membrane.

TABLE 10. RESULTS OF MODULE TESTS CONDUCTED AT THE END OF CONSTANT FEED PRESSURE OPERATION STUDY

Pressure Vessel	MODULE #1				MODULE #2				MODULE #3			
	A		R		A		R		A		R	
	Ti	Tf	Ti	Tf	Ti	Tf	Ti	Tf	Ti	Tf	Ti	Tf
1	2.27	1.96	93.1	19.0	2.29	2.25	92.7	20.0	2.04	1.92	94.9	56.2
2	2.15	2.04	94.3	24.6	2.25	1.96	94.7	21.0	2.23	1.91	94.3	28.9
3	2.05	2.06	93.5	26.0	2.29	2.25	93.8	27.3	2.23	2.85	91.0	42.0
4	2.23	1.72	93.3	56.5	2.26	1.85	92.6	27.4	2.04	3.13	94.6	37.2
5	2.30	1.82	94.8	61.7	2.28	2.03	94.3	53.5	1.94	1.66	94.5	66.4
6	2.17	1.60	93.7	61.1	2.33	2.07	93.0	67.5	2.18	1.60	95.3	72.0
7	2.23	1.87	91.8	56.1	1.94	1.69	96.3	81.1	1.92	1.34	94.6	74.9
8	2.07	1.69	95.9	88.2	2.21	1.44	94.2	77.5	2.26	1.77	91.1	82.3
9	2.12	1.57	93.2	83.1	2.14	1.49	90.1	81.8	2.21	1.66	94.1	75.6

Notes:
1. A = Water permeability coefficient, (g/sq cm/sec/atm) X 10^5,
2. R = Salt rejection, %.
3. Ti= Initial value; Tf = Final value after 9,475 hours of operation.

TABLE 11. RESULTS OF DYE CHECKING, VISUAL INSPECTION AND MEMBRANE SAMPLE TESTING

MODULE	TEST RESULTS
Module #2 Pressure Vessel No. 9	Integrity was good except for a small product tube leak and a membrane pinhole caused by a crease in the product water channel material. The module appeared to be quite clean.
Module #1 Pressure Vessel No. 1	No leaks were observed and module integrity was good. There was visible evidence of fouling and membrane rejecting surface attack (indicated by dye pickup). Some local areas did not pick up dye.
Module #3 Pressure Vessel No. 4	No leaks were observed and module integrity was good. Moderate fouling and membrane rejecting surface attack were evident. There seemed to be more membrane surface attack near the product tube.
Module #2 Pressure Vessel No. 5	No leaks were observed and module integrity was good. There was some evidence of fouling and membrane surface attack.

C. Some substances in the fouling layer attacked the membrane.

The hydrolysis was believed to be the most probable cause for the deterioration of the membrane surface.

Under Constant Flux Rate Operation

During the second phase of the pilot plant study, three membrane modules, one from each of the first three pressure vessels, were removed at 5,678 hours of on-stream operation and sent to the GESCO for membrane evaluation. The results of the membrane evaluation are summarized below:

A. Module from the Pressure Vessel No. 1:

Testing Feed	"A", g/sq cm/sec/atm	Rejection, %
Tap Water (No pH adjustment)	2.62×10^{-5}	79.2
2,000 mg/l NaCl	---	57.3

"A" = Water permeability coefficient

The module was probed to check for possible leaks, but the results indicated a uniformly poor rejection over the entire module length. The module was also dye-checked and opened for visual inspection. A substantial amount of dirt was present, both in the brine spacer and on the membrane. A white, flaky deposit (probably calcium sulfate) was noted near the product water tube. Some areas of the membrane seemed to adsorb the dye more than others, indicating poorer rejection in these areas.

Membrane samples were taken from the two types of areas and tested. The results are as follows:

Sample	"A", g/sq cm/sec/atm	Rejection, %
Heavy Dye area	5.41×10^{-5}	40.1
Light Dye area	4.14×10^{-5}	52.4

B. Module from the Pressure Vessel No. 2:

Testing Feed	"A", g/sq cm/sec/atm	Rejection, %
Tap Water (No pH adjustment)	1.57×10^{-5}	95.7
2,000 mg/l NaCl	---	88.4
Tap Water (pH adjusted to 5.5 to 6)	1.51×10^{-5}	94.6

C. Module from the Pressure Vessel No. 3:

Testing Feed	"A", g/sq cm/sec/atm	Rejection, %
Tap Water (No pH adjustment)	1.76×10^{-5}	94.1
2,000 mg/l NaCl	---	82.4
Tap Water (pH adjusted to 5.5 to 6)	1.69×10^{-5}	93.3

The membrane tests indicated that the membranes were hydrolyzed. The degree of hydrolysis seemed quite severe. Visually, it was difficult to determine the extent of membrane degradation. Exposure to high pH feeds of cleaning solutions could cause hydrolysis. However, this did not occur in this instance. Bacterial action was another factor, but continuous chlorine addition with 1.5 to 2.0 mg/l chlorine residual should be an adequate preventative. Therefore, the real cause for the loss in the rejection ability of the membrane could not be clearly determined.

SECTION 7

PROCESS COST ESTIMATE

Based on the pilot plant studies conducted at Pomona Advanced Wastewater Treatment Research Facility, a cost estimate has been prepared for a 37,850 cu m/day (10 MGD) spiral-wound reverse osmosis plant in demineralizing a carbon-treated secondary effluent. The major assumptions made for this cost estimate are listed in the following:

A. The TDS of the blended feed for the reverse osmosis plant is about 1,200 mg/l;

B. The water recovery for the process is about 80 percent;

C. The product water flux rate is approximately 407 l/sq m/day (10 gal/sq ft/day) at 25°C;

D. The process is capable of reducing or rejecting 90 percent of the blended feed TDS;

E. The sodium perborate at 2 percent concentration is used as the membrane cleaning solution, with pH of the solution adjusted to 7.5 with sulfuric acid;

F. The membrane cleaning is performed once a week, or at an interval of 3,054 liters of product water per square meter of membrane area (75 gallons per square foot of membrane area);

G. The effective membrane life is only one year;

H. The capital cost is amortized for 20 years at 5 percent interest rate; and

I. The reference date of the cost estimate is August, 1973.

The initial capital cost including the feed pumps, membranes, pH controllers, chlorinators, chemical feed systems, booster pumps, brine recirculation pumps, and a post treatment system for final pH adjustment is about 3.66 million dollars for a 37,850 cu m/day (10 MGD) spiral-wound reverse osmosis plant. The total membrane cost is about 1.15 million dollars. Since the membrane has to be replaced every year, the cost for membrane replacement is approximately 8.2¢/1,000 liters (31.5¢/1,000 gallons) of product water. If the membrane life can be improved to two years, then the membrane replacement cost will be substantially reduced to 4.0¢/1,000 liters (15.4¢/1,000 gallons).

The annual maintenance material cost is based on 5 percent of the capital cost, excluding the cost of membranes. The labor requirements include:

A. One man-hour per cleaning schedule for a 378.5 cu m/day (0.1 MGD) section of the plant; and

B. Three man-years for operating the 37,850 cu m/day (10 MGD) plant.

The total power cost (1¢/kwh) for the 37,850 cu m/day (10 MGD) plant operation is estimated to be about 2.0¢/1,000 liters (7.8¢/1,000 gallons). The unit costs for the various chemicals used in the reverse osmosis process are estimated as follows:

A. Sodium perborate = \$0.37/Kg (\$0.17/lb);

B. Triton X-100 non-ionic detergent = \$0.84/Kg (\$0.38/lb);

C. Carboxy methyl cellulose = \$0.97/Kg (\$0.44/lb);

D. Sulfuric acid = \$0.04/Kg (\$0.02/lb); and

E. Chlorine = \$0.09/Kg (\$0.04/lb).

According to the above chemical unit costs, the total expenses for process chemicals will amount to approximately 1.1¢/1,000 liters (4.3¢/1,000 gallons).

Table 12 summarizes the various parts of the total process cost estimate. As indicated in the table, the total process cost is approximately 14.9¢/1,000 liters (57.4¢/1,000 gallons) for one year membrane life. The cost can be reduced to about 10.7¢/1,000 liters (41.3¢/1,000 gallons) by improving the membrane life to two years. Both cost estimates do not include the costs for the carbon adsorption pretreatment and the brine disposal.

TABLE 12

PROCESS COST ESTIMATE FOR 37,850 cu m/day (10 MGD) SPIRAL-WOUND REVERSE OSMOSIS PLANT

Amortization of Capital	¢/1,000 gallons	¢/1,000 liters
$3.66 x 10^6; 20 years @ 5%	8.8	2.3
Operation and Maintenance		
Chemicals (H_2SO_4, Cl_2 and cleaning agent)	4.3	1.1
Membrane Replacement		
One-year membrane life	31.5	8.2
Two-year membrane life	15.4	4.0
Maintenance Materials	3.4	0.9
Power	7.8	2.0
Labor	1.6	0.4
Total Process Cost:		
One-year membrane life	57.4	14.9
Two-year membrane life	41.3	10.7

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15. SUPPLEMENTARY NOTES

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16. ABSTRACT A 56.8 cu m/day (15,000 gallons/day) spiral-wound reverse osmosis pilot plant was operated at the Pomona Advanced Wastewater Treatment Research Facility on the carbon-treated secondary effluent. The specific objectives for this study were (a) to establish the effective membrane life for wastewater demineralization with carbon adsorption pretreatment; (b) to determine the reliability of the process performance; and (c) to derive a realistic process cost estimate. The study was first conducted on a constant feed pressure basis, and then it was run on a constant product water flux rate basis. During the first phase of the study, pH adjustment was not practiced for the weekly enzyme-detergent membrane cleaning procedures. However, this was practiced in the second phase of the study. The results from both phases of studies substantiated the fact that the membrane effective life was only about one year in demineralizing the carbon-treated secondary effluent. A cost estimate for a 37,850 cu m/day (10 MGD) reverse osmosis plant indicated that for membranes with only one-year life the process cost was about 14.9¢/1,000 liters (57.4¢/1,000 gallons). However, the cost could be substantially reduced to 10.7¢/1,000 liters (41.3¢/1,000 gallons) for membranes with two-year life. Both cost estimates did not include the costs for carbon adsorption pretreatment and brine disposal.

17. KEY WORDS AND DOCUMENT ANALYSIS

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