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Environmental Protection Technology Series

DEMINERALIZATION OF SAND-FILTERED SECONDARY EFFLUENT BY SPIRAL-WOUND REVERSE OSMOSIS PROCESS

Municipal Environmental Research Laboratory
Office of Research and Development
U.S. Environmental Protection Agency
Cincinnati, Ohio 45268

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DEMINERALIZATION OF SAND-FILTERED SECONDARY EFFLUENT

BY

SPIRAL-WOUND REVERSE OSMOSIS PROCESS

by

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Contract No. 14-12-150

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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 permiability 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.

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ABSTRACT

A 22.7 cu m/day (6,000 gallons/day) spiral-wound reverse osmosis pilot plant, was operated at the Pomona Advanced Wastewater Treatment Research Facility on the sand-filtered secondary effluent. The pilot plant study was conducted under optimum operating conditions based on previous studies. During the first year of operation, all the system performance parameters, such as salt rejection, water recovery, and product water flux rate, were only slightly decreased from their initial values. However, the salt rejection and product water flux rate were substantially reduced to almost half of their initial values after a two year operation period. During this same two year period, the water recovery was found to decline about 15 per cent of its initial value.

A cost estimate for a 37,850 cu m/day (10 MGD) plant for August, 1973 cost figures indicated that for membranes with only one-year life the process cost was about 16.5¢/1,000 liters (63.6¢/1,000 gallons). However, the cost could be substantially reduced to 12.4¢/1,000 liters (47.5¢/1,000 gallons) for membranes with two-year life. Both cost estimates did not include the costs for sand filtration pretreatment and brine disposal.

This report was submitted in fulfillment of Contract No. 14-12-150 by the County Sanitation Districts of Los Angeles County under the sponsorship of the U.S. Environmental Protection Agency. This report covers a period from July 1971 to June 1973.

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ACKNOWLEDGMENTS

This study was jointly sponsored by the U.S. Environmental Protection Agency and the County Sanitation Districts of Los Angeles County.

Mr. Harold H. Takenaka, former U.S. EPA Project Engineer at Pomona Advanced Wastewater Treatment Research Facility, was instrumental in initiating the pilot plant study.

The advice and suggestions given by Dr. James E. Cruver of Gulf Environmental Systems Company during the course of the study were important contributions to the success of the study.

The untiring efforts of both the operating and laboratory staff of the Pomona Advanced Wastewater Treatment Research Facility are gratefully acknowledged.

SECTION 1

INTRODUCTION

As part of the continuing development program for the application of the reverse osmosis process to wastewater demineralization, the Gulf Environmental Systems Company conducted a special research study, on a short-term basis, on the effects of various pretreatment systems on the membrane performance under the U.S. EPA Contract No. 14-12-831. The study was mostly performed at the Pomona Advanced Wastewater Treatment Research Facility concurrently with other reverse osmosis pilot plant studies.

The Gulf Environmental Systems Company concluded their pretreatment study on the contract expiration date of July 31, 1971. They indicated in their contract final report that an activated carbon adsorption pretreatment was clearly not necessary for a successful reverse osmosis system operation.(1) In the same report, they further demonstrated that a sand filtration process could provide an equally satisfactory pretreatment for a reverse osmosis system operation.

This study was initiated to confirm the findings of the Gulf Environmental Systems Company on an extended long-term basis, and also to achieve the following specific objectives:

A. To establish the effective life of the membrane of a spiral-wound reverse osmosis system in demineralizing a sand-filtered secondary effluent;

B. To determine the reliability of the process performance; and

C. To obtain the operating and design data for making a realistic process cost estimate.

The study was conducted with the same reverse osmosis pilot plant previously used by the Gulf Environmental Systems Company in their pretreatment study. The pilot plant was a spiral-wound membrane system and had a nominal production capacity of 22.7 cu m/day (6,000 gallons/day). The pilot plant had accumulated a total of 2,503 hours of on-stream operation before the Gulf Environmental Systems Company terminated their study on July 31, 1971. This extension study had added another 15,025 hours of operation to the pilot plant to accomplish a total of 17,528 hours (equivalent to a two-year period) of on-stream operation. The study was formally completed as of June 11, 1973.

SECTION 2

CONCLUSIONS

The following conclusions can be drawn from the pilot plant study:

A. The spiral-wound reverse osmosis system operated successfully on the secondary effluent of Pomona Water Reclamation Plant with only sand filtration pretreatment. The effective membrane life for the operation was approximately one year.

B. The pilot plant maintained 90 percent or more of salt rejection, 326 l/sq m/day (8 gal/sq ft/day) or more of product water flux, and 75 percent or more of water recovery under an average operating pressure of 34.5 Kg/sq cm (500 psi) during the first year of operation.

C. The product water flux rate decline was controlled by a daily air-tap water flushing and a three-times-a-week chemical cleaning. Three types of cleaning solutions, namely, Biz enzyme detergent, sodium perborate and sodium ethylenediaminetetraacetate (EDTA), were found to perform equally well. However, as recommended by the membrane manufacturer at the middle of the study, only EDTA cleaning solution was used for the membrane cleaning during the second year of operation.

D. The water flux rate decline could be minimized by maintaining a minimum brine flow of 15 l/min (4 gpm).

E. The water recovery could be enhanced by a partial recycling of the brine to the feed stream.

F. The overall reductions in the salt rejection, water recovery, and product water flux rate during the two years of onstream operation were approximately 51 percent, 15 percent, and 50 percent, respectively.

G. The product water quality prior to the start of serious membrane deterioration was excellent.

H. The total process cost estimate for a 37,850 cu m/day (10 MGD) plant is about 16.5¢/1,000 liters (63.6¢/1,000 gallons). However, if the membrane effective life can be improved from one year to two years, then the process cost can be substantially reduced to 12.4¢/1,000 liters (47.5¢/1,000 gallons). Both cost estimates do not include the costs for sand filtration pretreatment and brine disposal.

I. A comparison of total process costs, including pretreatment costs, between two different pretreatment schemes for the spiral-wound reverse osmosis process indicates that the sand filtration pretreatment scheme is somewhat less expensive than the carbon adsorption pretreatment scheme.

SECTION 3

RECOMMENDATIONS

The relatively short membrane life as concluded from this pilot plant study on the wastewater demineralization is rather discouraging. An optimum membrane life would be three years, if the process is to be practical and economical for the application to the wastewater demineralization.[2] Therefore, it is highly recommended that further studies be pursued primarily in the areas of membrane improvement. Other parameters such as pretreatment methods, membrane cleaning solution and frequency, feed pressure, brine recirculation, membrane module configuration, and brine flow rate should also be thoroughly evaluated and investigated.

SECTION 4

PILOT PLANT DESCRIPTION

The 22.7 cu m/day (6,000 gpd) spiral-wound reverse osmosis pilot plant consisted of four steel pressure vessels 3.0 m (10 ft) long by 10 cm (4 in) in diameter, each of which contained three ROGA spiral-wound membrane modules. Each module had a 5.6 sq m (60 sq ft) of high flux membrane cast on D-601 sail-cloth backing, 1.1 mm (0.045 in) polypropylene Vexar brine spacers, and melamine-treated tricot product water channels. The average water permeability coefficient value of the membrane modules was 2.5×10^{-5} g/sq cm/sec/atm. Figure 1 shows the configuration of the spiral-wound membrane module. The total membrane area in the pilot plant system was 67 sq m (720 sq ft).

The schematic flow diagram of the 22.7 cu m/day (6,000 gpd) reverse osmosis pilot plant is shown in Figure 2. The sand-filtered secondary effluent was chlorinated to provide 1 to 2 mg/l of residual chlorine and acidified to pH close to 5 using sulfuric acid before being fed to the reverse osmosis system. The pilot plant system was in a 2-1-1 array, as shown in Figure 2, to maintain sufficient brine flow rates in the downstream modules.

The brine was partially recycled to maintain an apparent water recovery at the level of 75 to 80 percent. This brine recycling slightly increased the inorganic and organic matters in the feed and thus might cause adverse effect on membrane fouling. A minimum brine flow of 15 l/min (4 gpm) was maintained during the system operation to avoid concentration polarization of the membranes.

An Apco back pressure unit was used to regulate the system operating pressure. Sufficient sample valves were installed on the pilot plant system, so that samples from the raw feed (sand-filtered secondary effluent), blended feed (mixture of sand-filtered secondary effluent, chlorine solution, sulfuric acid, and recycled brine), 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

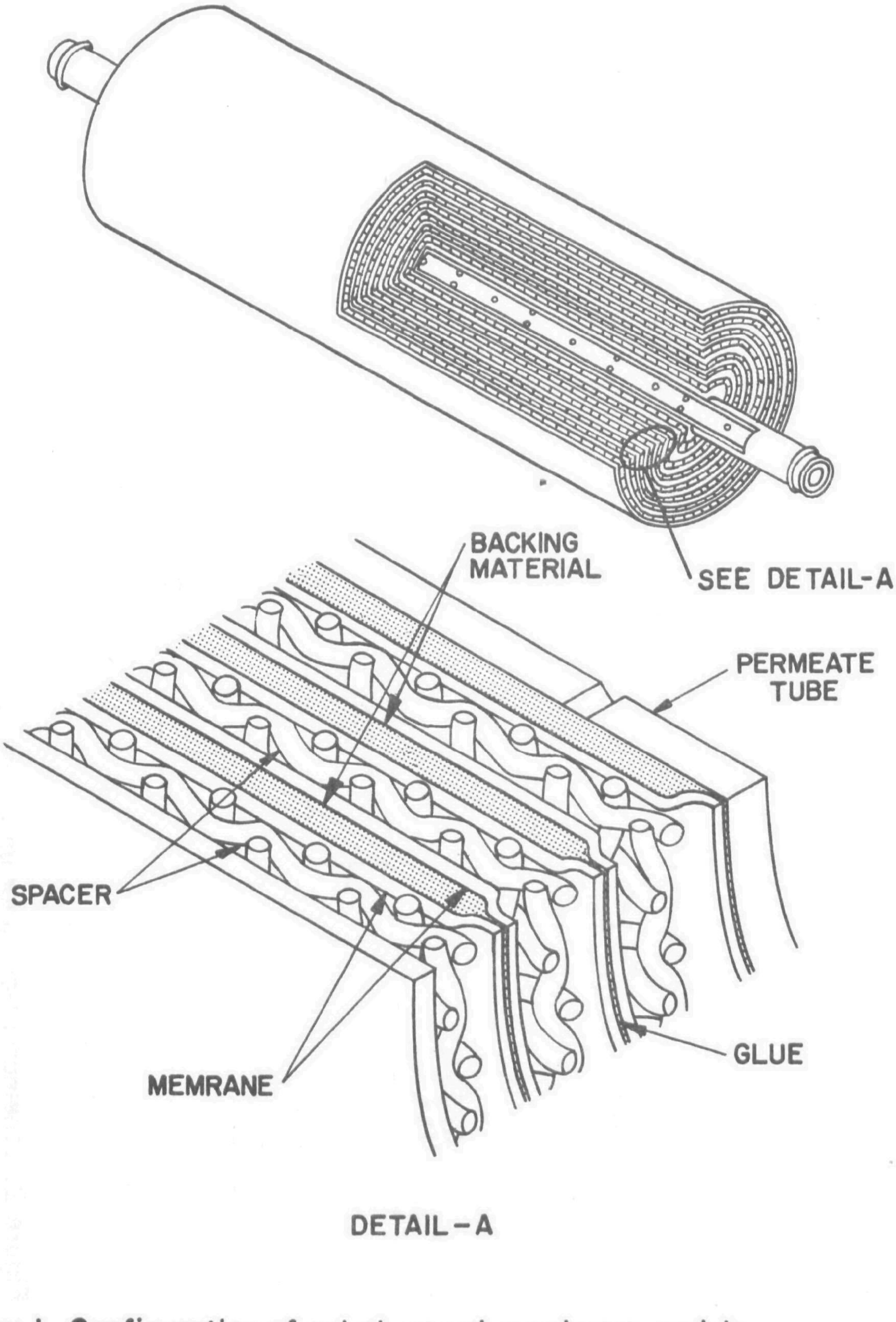


Figure 1. Configuration of spiral-wound membrane module.

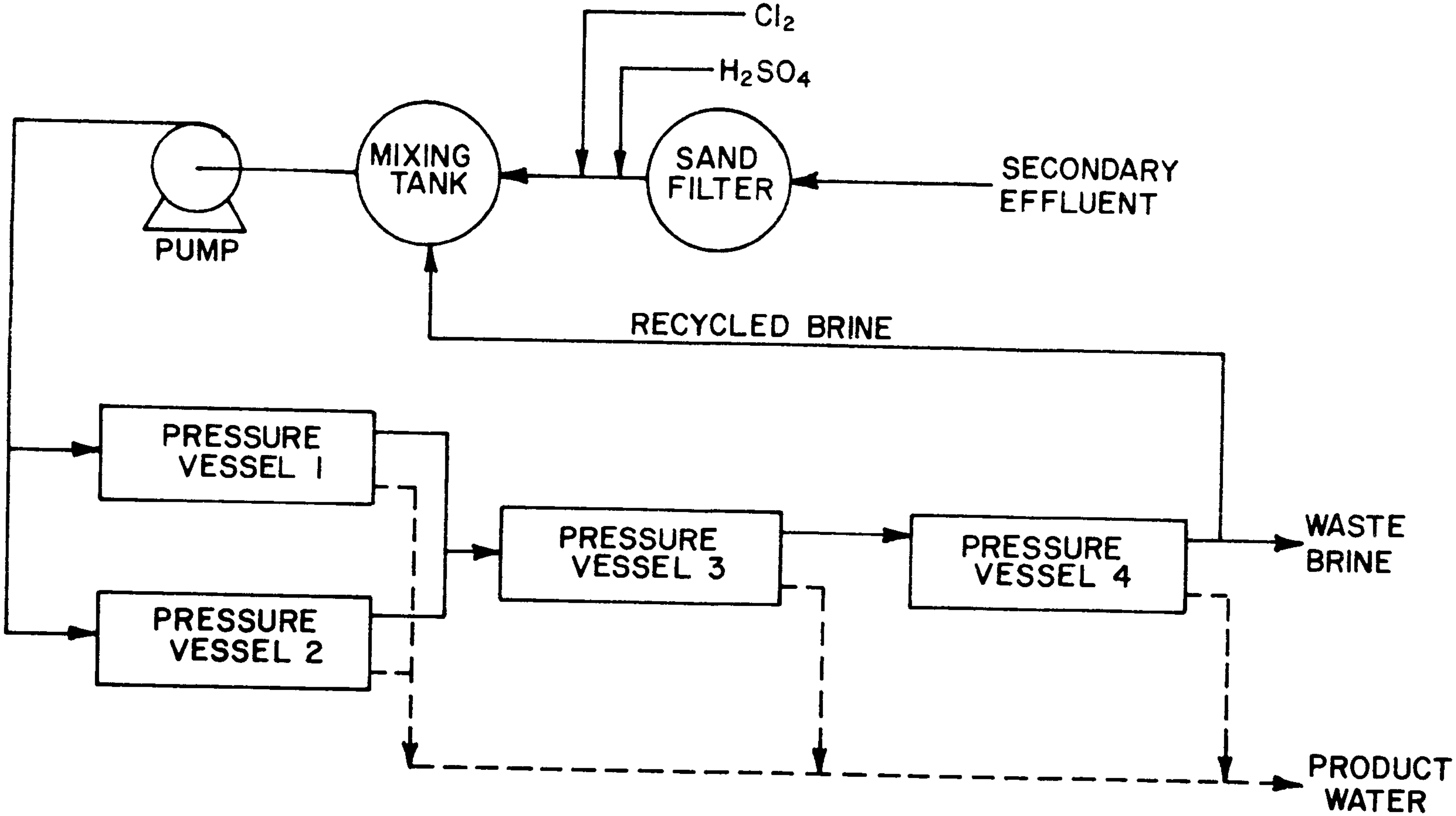


Figure 2. Schematic flow diagram of the reverse osmosis pilot plant.

sulfuric acid to the blended feed stream for pH control. The pump rate was regulated by a pH controller. An Advance gas chlorinator was employed for chlorine addition.

The pilot plant system was designed and constructed to be cleaned regularly without disassembly of manifolds. A chemical cleaning solution was made up in a cleaning tank and was then circulated through the pilot plant by a centrifugal booster pump. Enough valves were provided in the pilot plant system so that each pressure vessel could be cleaned individually or the system could be cleaned as a series-parallel array. Tap water or air-tap water mixture could be introduced for flushing just ahead of the pressure vessel array during cleaning cycle or downtimes.

The pressure sand filter used in the pretreatment system was a standard package designed by L.A. Water Conditioning Company, City of Industry, California. The filter was 76.2 cm (30 in) in diameter and had about 45.7 cm (18 in) depth of sand. The sand bed with an effective size and a uniformity coefficient of approximately 0.5 mm and 1.6, respectively, was supported by a layer of graded gravel. The hydraulic loading rate of the sand filter was maintained at 2 lps/sq m (3 gpm/sq ft). The filter was normally backwashed once a day.

SECTION 5

PILOT PLANT OPERATION

OPERATING CONDITIONS

At the beginning of this study, the pilot plant was operated at a constant product flux rate of 407 l/sq m/day (10 gal/sq ft/day) to simulate the actual plant operation. The operating pressure was frequently adjusted to maintain such constant flux rate operation. The operating pressure was found to vary from 24.2 Kg/sq cm (350 psi) to 48.3 Kg/sq cm (700 psi) to produce a 407 l/sq m/day (10 gal/sq ft/day) flux rate during the first 5,900 hours of operation. Due to the lack of an adequate control mechanism to automatically make the necessary operating pressure adjustment, the constant flux mode of operation was converted to a constant 34.5 Kg/sq cm (500 psi) operating pressure mode of operation starting at 5,900 hours of on-stream operation.

The initial performance parameters under the constant operating pressure operation were as follows:

A. Product water flux rate (adjusted to 25^{o}C) : 488 l/sq m/day (12 gal/sq ft/day).

B. Water recovery (defined as "100 x flow rate of product stream/flow rate of raw feed stream:) : 80 percent.

C. Salt rejection (defined as "100 x conductivity of product stream/conductivity of blended feed stream") : 97 percent.

The sand-filtered secondary effluent was chlorinated to provide 1 to 2 mg/l chlorine residual and acidified to a pH close to 5 using sulfuric acid for biological growth and chemical precipitation controls, respectively. The minimum brine flow was regulated at 15 l/min (4 gpm) and the recycled brine flow was maintained at 10 l/min (2.7 gpm).

MEMBRANE CLEANING PROCEDURES

The reverse osmosis pilot plant operation was started out with a daily tap-water flushing and a twice-a-week chemical solution cleaning cycle. This membrane cleaning schedule was practiced routinely throughout the first 2,500 hours of on-stream operation. However, it was found necessary to increase the frequency of the chemical solution cleaning to three times a week to maintain the desired performance level. This new cleaning frequency was equivalent to a cleaning interval of 815 liters of product water per square meter of membrane surface area (which was about 20 gallons per square foot of membrane surface area).

Three types of chemical cleaning solutions--Biz enzyme-detergent, sodium perborate and sodium ethylenediaminetetraacetate (EDTA)--were tested during the first 7,900 hours of on-stream operation.

However, the Biz enzyme-detergent (contained some small amount of sodium perborate) and the sodium perborate solution were subsequently found corrosive to the cellulose acetate membrane according to Cruver[3] of the Gulf Environmental Systems Company.

Consequently, only the EDTA cleaning solution was used as the cleaning agent for the membranes throughout the rest of the pilot plant operation. The concentrations and constituents of the various cleaning solutions used were as follows:

A. Biz enzyme-detergent solution:

a. 2% of Biz enzyme-detergent

B. Sodium perborate solution:

a. 2% of sodium perborate
b. 0.15% of Triton X-100 (non-ionic detergent)
c. 0.0015% of carboxy methyl cellulose

C. EDTA solution:

a. 1% of EDTA (tetra sodium salt)
b. 0.15% of Triton X-100
c. 0.0015% of carboxyl methyl cellulose

The pH of all the cleaning solutions was adjusted to 7.5 to 8.0 to minimize the membrane hydrolysis reaction. The cleaning solutions were prepared with warm tap water (40°C to 60°C) to improve the cleaning efficiency.

The chemical cleaning solution was first flushed through all the reverse osmosis pressure vessels for five minutes, then the solution was flushed through each individual pressure vessel for 10 minutes, and finally the solution was flushed through the entire system again for another five minutes. Therefore, the total flushing time for the pilot plant was about 50 minutes. The flushing rate for the cleaning solution was regulated at the level of 37.9 l/min (10 gpm). After the chemical solution cleaning cycle, the system was thoroughly rinsed with tap water in the same procedure as the chemical cleaning solution. Therefore, the downtime required for daily membrane cleaning ranged from 50 to 100 minutes depending on whether the chemical cleaning was practiced along with the daily tap-water flushing or not.

SAMPLING AND MONITORING PROCEDURES

During the week days (Monday through Friday), the following operating parameters were monitored routinely:

A. The conductivities of the feed, product, and brine streams of each pressure vessel.

B. The rate of product flow from each pressure vessel.

C. The pressure difference through each pressure vessel.

Daily (including Saturday and Sunday) measurements were performed on the following parameters:

A. The temperature of the blended feed water.

B. The pH values of the blended feed and brine streams.

C. The chlorine residual of the blended feed streams.

D. The total pressure difference through the pilot plant system.

E. The total product flow before and after the membrane cleaning.

F. The conductivities of the blended feed and the final product streams.

G. The total and recycled brine flows.

H. The feed operating pressure.

I. The total on-stream operation time.

Besides the above routine daily monitoring procedures, some grab samples from the raw feed, blended feed, product and brine streams were also taken every Thursday at 8:00 A.M. for water quality analyses.

SECTION 6

RESULTS AND DISCUSSIONS

CONSTANT PRODUCT FLUX OPERATION

Figure 3 shows the variations of the salt rejection and the feed pressure during the first 5,900 hours of on-stream operation. The product flux was maintained at a constant rate of 407 l/sq m/day (10 gal/sq ft/day) during the first 5,900 hours of operation. As indicated in Figure 3, the feed pressure required to maintain the above constant product flux rate was approximately 31.7 Kg/sq cm (460 psi) during the first 1,000 hours of operation. However, the required feed pressure rapidly increased to as high as 47.6 Kg/sq cm (690 psi) during the next 400 hours of operation. Because of this high feed pressure development, the system was depressurized for a 72 hour period after 1,430 hours of operation. During the depressurization period, a constant tap water flushing through the system was maintained. After this depressurization treatment, the feed pressure was reduced to the previous level of 31.7 Kg/sq cm (460 psi) to maintain the 407 l/sq m/day (10 gal/sq ft/day) product flux. The pressure remained at about 27.6 to 34.5 Kg/sq cm (400 to 500 psi) until 2,200 hours of operation.

Between 2,200 to 3,600 hours of operation, the system required 31.1 to 38.0 Kg/sq cm (450 to 550 psi) to maintain the constant flux of 407 l/sq m/day (10 gal/sq ft/day). Then it became necessary to gradually increase the feed pressure to as high as 48.3 Kg/sq cm (700 psi). Two 24 hour and two 72 hour depressurization periods on tap water between 3,600 and 4,450 hours of operation were not successful in reducing the feed pressure to the initial operating level.

At the 3,844 hours of operation, the membrane modules were inspected. Several brine seals were "blown," so they were retaped in place. The "blown" brine seals probably resulted from the high operating feed pressure. At 4,000 hours of operation, the membrane modules were again inspected. The inspection showed that the modules were very clean with the barber-pole type outer wrap tape and the brine seals were in excellent condition. However, the condition of the brine channels could not

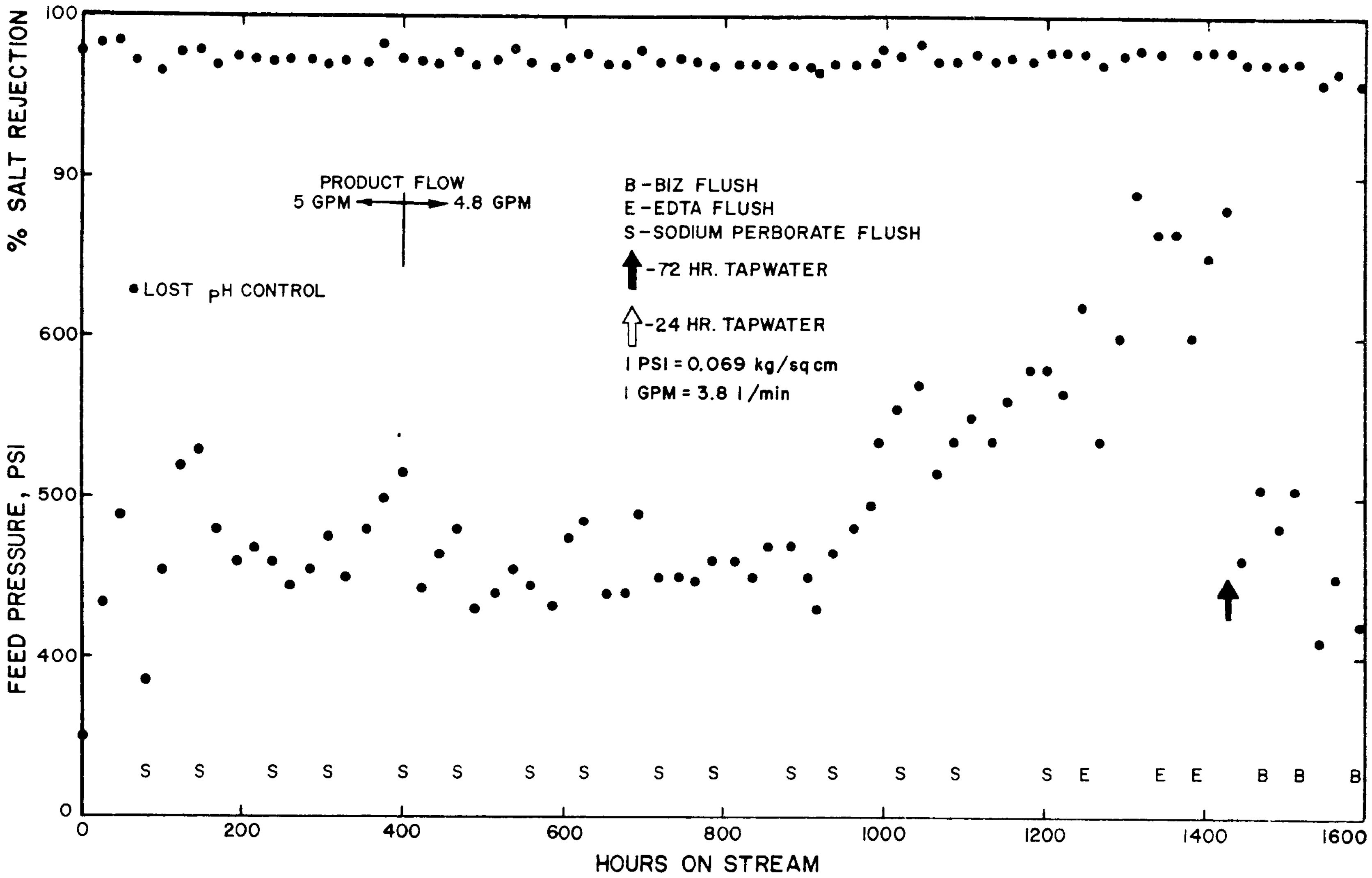
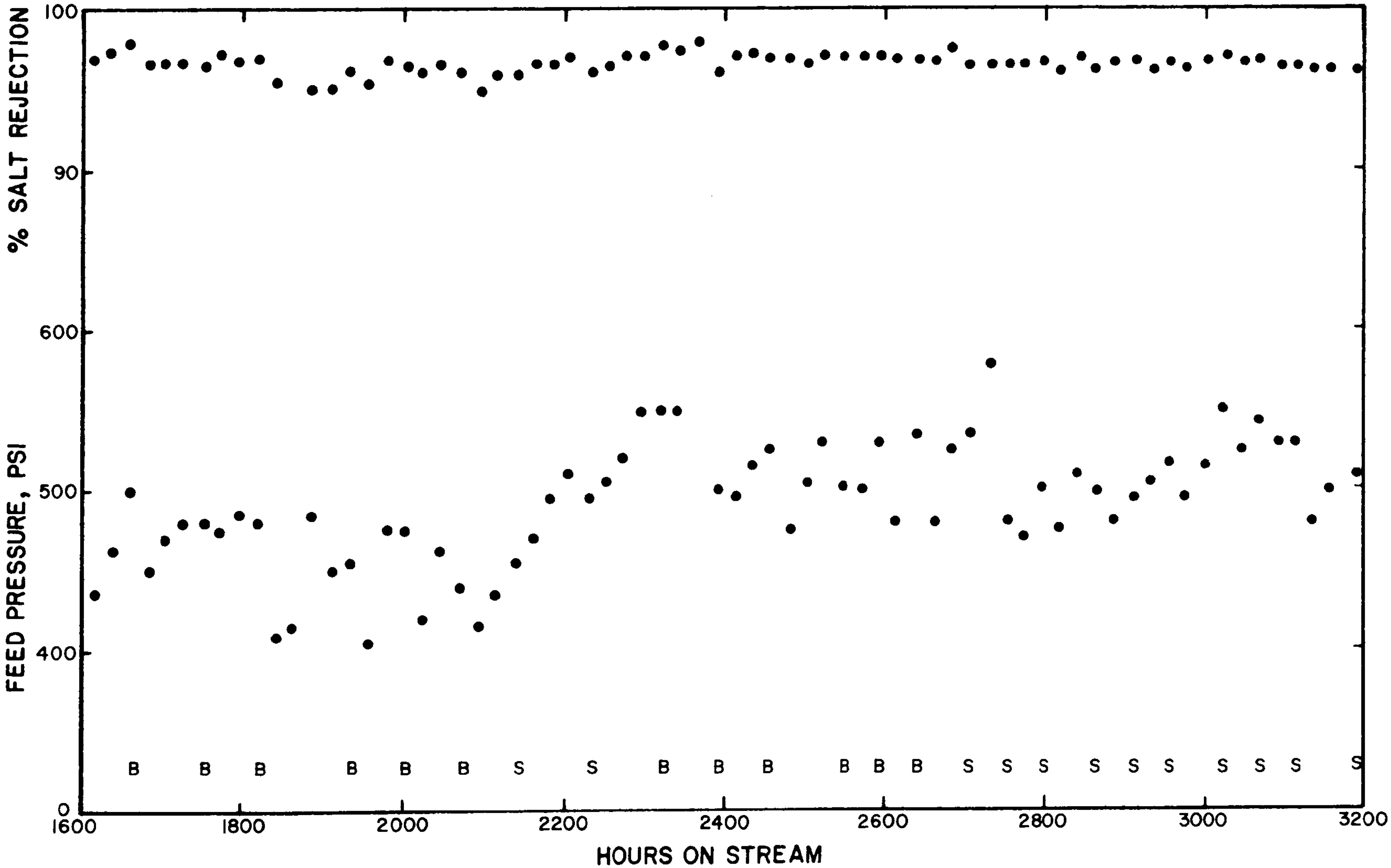


Figure 3. Salt rejection and feed pressure variation vs. operational time under constant flux rate operation.



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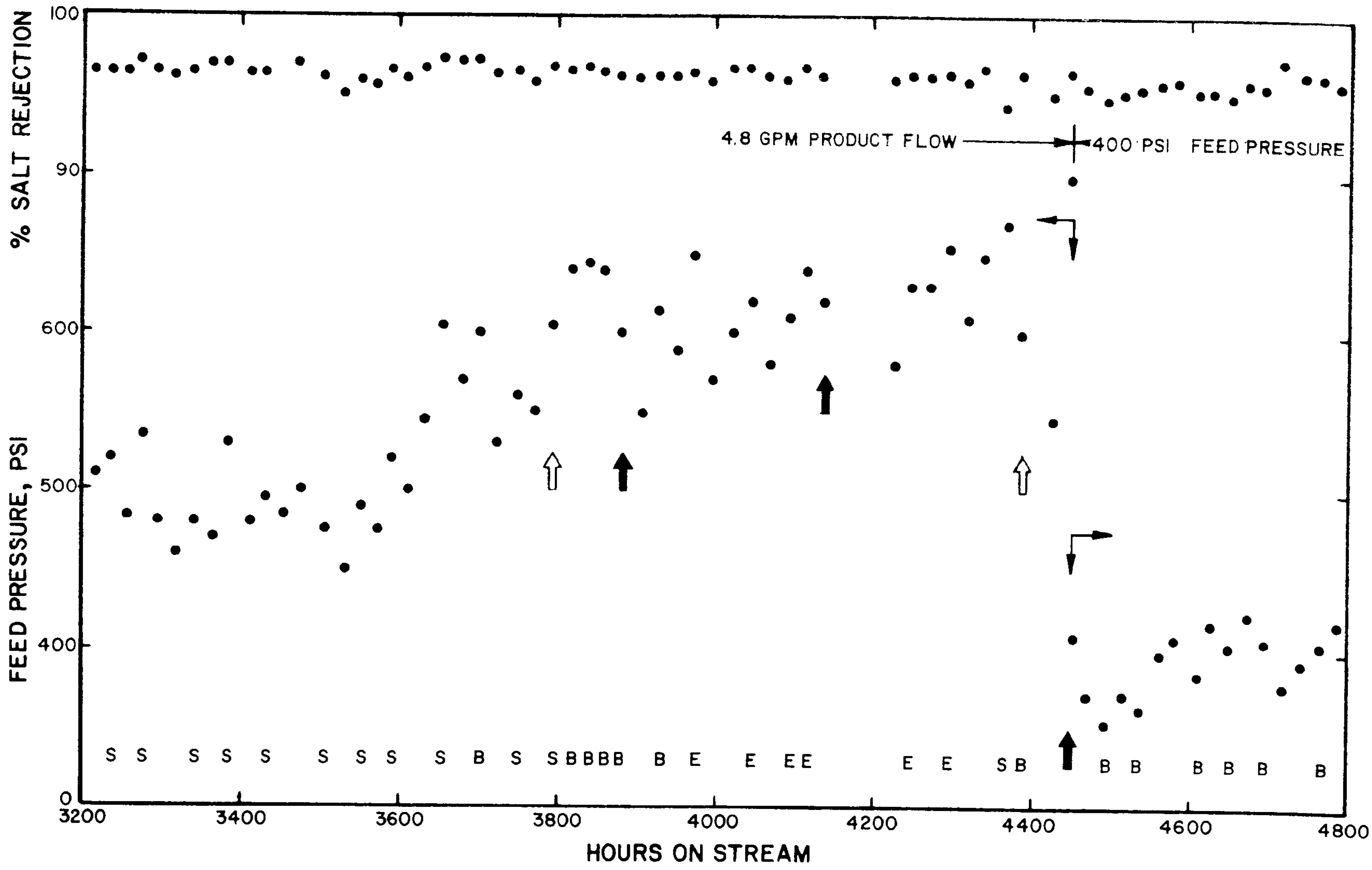

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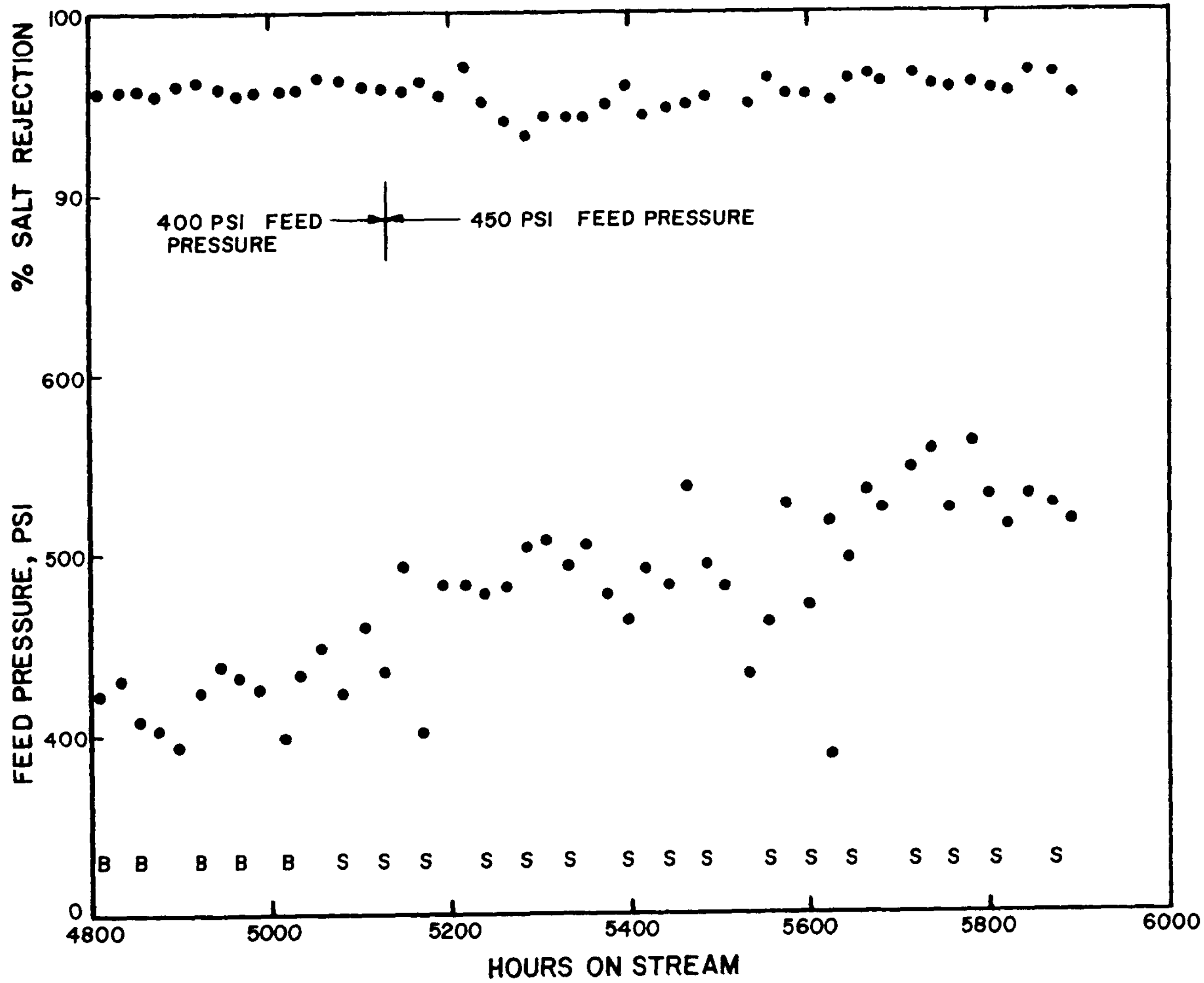


Figure 3. Continued.

be determined without opening the modules. The reason for the high feed pressure required to maintain 407 l/sq m/day (10 gal/sq ft/day) product flux was not readily apparent.

After the 72 hour depressurization at 4,450 hours of operation, it was decided to operate the system at 27.6 Kg/sq cm (400 psi) feed pressure for a month period in an attempt to improve the "A" (water permeability coefficient) value. At this time, the lead module in the pressure vessel No. 1 was replaced with a new module. The old module was sent to the Gulf Environmental System Company for membrane evaluations. The results are shown in Table 1.

During the month long period of constant 27.6 Kg/sq cm (400 psi) feed pressure operation, the product water flux rate (at $25^{0}C$) increased from about 305 to 366 l/sq m/day (7.5 to 9 gal/sq ft/day) which was an encouraging trend. The constant feed pressure was subsequently increased to 31.1 Kg/sq cm (450 psi) for another month of trial operation. The product water flux rate was improved from 366 to 448 l/sq m/day (9 to 11 gal/sq ft/day) as a result of this feed pressure increase. Therefore, the system operation was converted from a constant product flux rate operation to a constant feed pressure operation after 5,900 hours of on-stream operation.

CONSTANT FEED PRESSURE OPERATION

At all times after 5,900 hours of operation, the reverse osmosis pilot plant system was operated at a constant 34.5 Kg/sq cm (500 psi) feed pressure. The variation of the salt rejection and the product water flux rate throughout this series of constant feed pressure operation are shown in Figure 4. As indicated in this figure, the initial product water flux rate was about 448 l/sq m/day (11 gal/sq ft/day). However, the flux rate started to decline at about 7,000 hours of operation.

At 7,266 hours of operation, the No. 2 pressure vessel was inspected because its salt rejection was about 3 percent lower than the other three vessels. The inspection showed that two of the three modules had experienced "blown" brine seals. The three modules looked relatively clean and the outer wrap tapes were in excellent condition. The blown brine seals were retaped in place and the system was put back in operation. Experience at Pomona had shown that the barber-pole type of outer wrap tape could successfully maintain module integrity when used in treating municipal wastewater.

As shown in Figure 4, there was a sudden decrease in product water flux rate at 7,445 hours of operation. This might be the result of excessive membrane fouling due to the poor

TABLE 1

EVALUATION OF THE LEAD MODULE MEMBRANE IN PRESSURE

VESSEL NO. 1 AT 4,450 HOURS OF OPERATION

Feed Solution and Membrane Condition	Feed Pressure (psi)	"A" Value ($g/cm^2/sec/atm$)	Salt Rejection %
2,000 mg/l NaCl Solution Before Membrane Cleaning	415	$1.25X.0^{-5}$	93.4
2,000 mg/l NaCl Solution After Membrane Cleaning	420	$1.81X10^{-5}$	94.3

Notes: 1. The membrane was cleaned with an enzyme-detergent (Biz) at 60°C with pH adjusted to 7.

2. The module was opened after the tests, and there was a brownish film on the membrane with very little, if any, scaling evidence.

3. 1 psi = 0.069 Kg/sq cm.

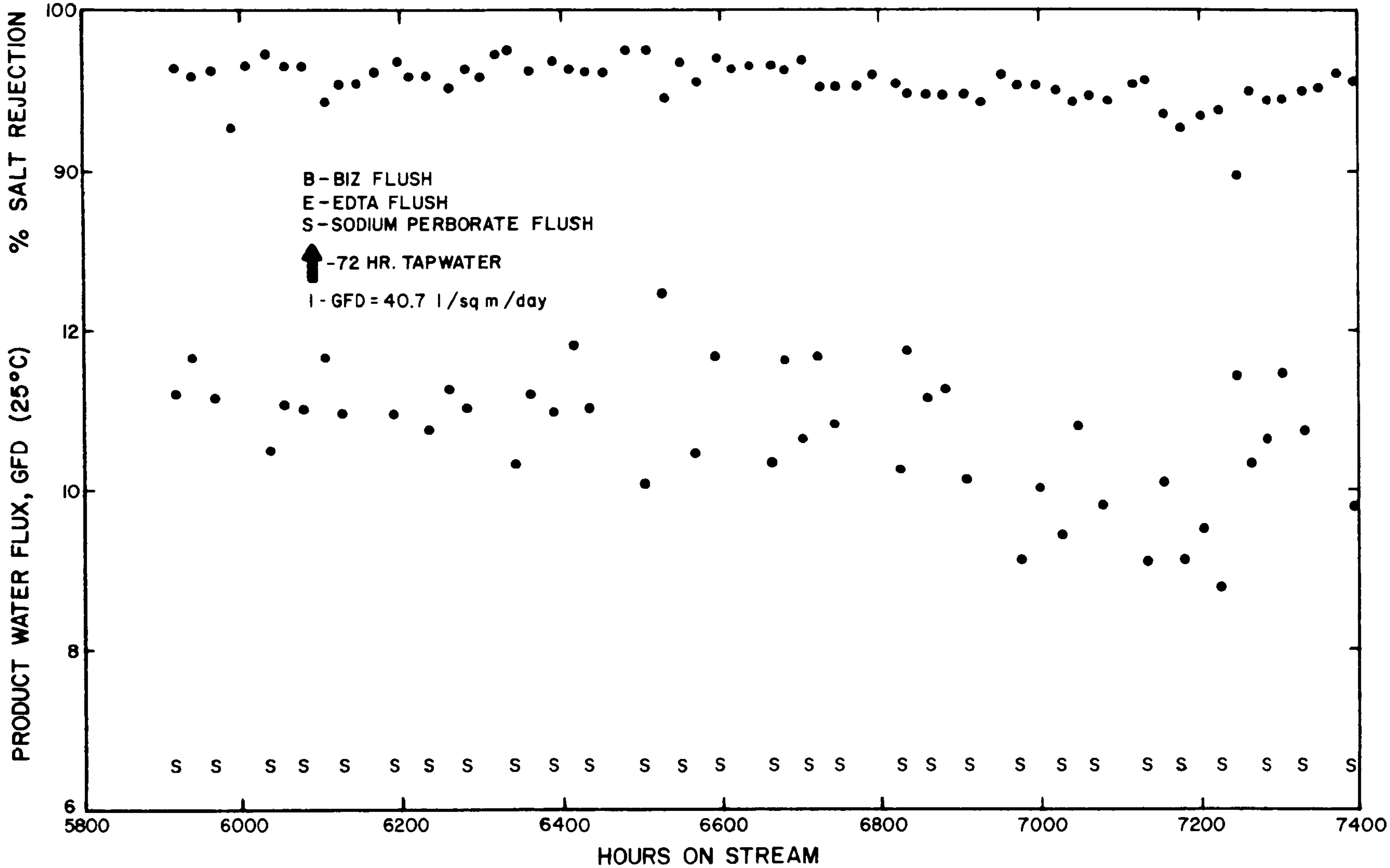
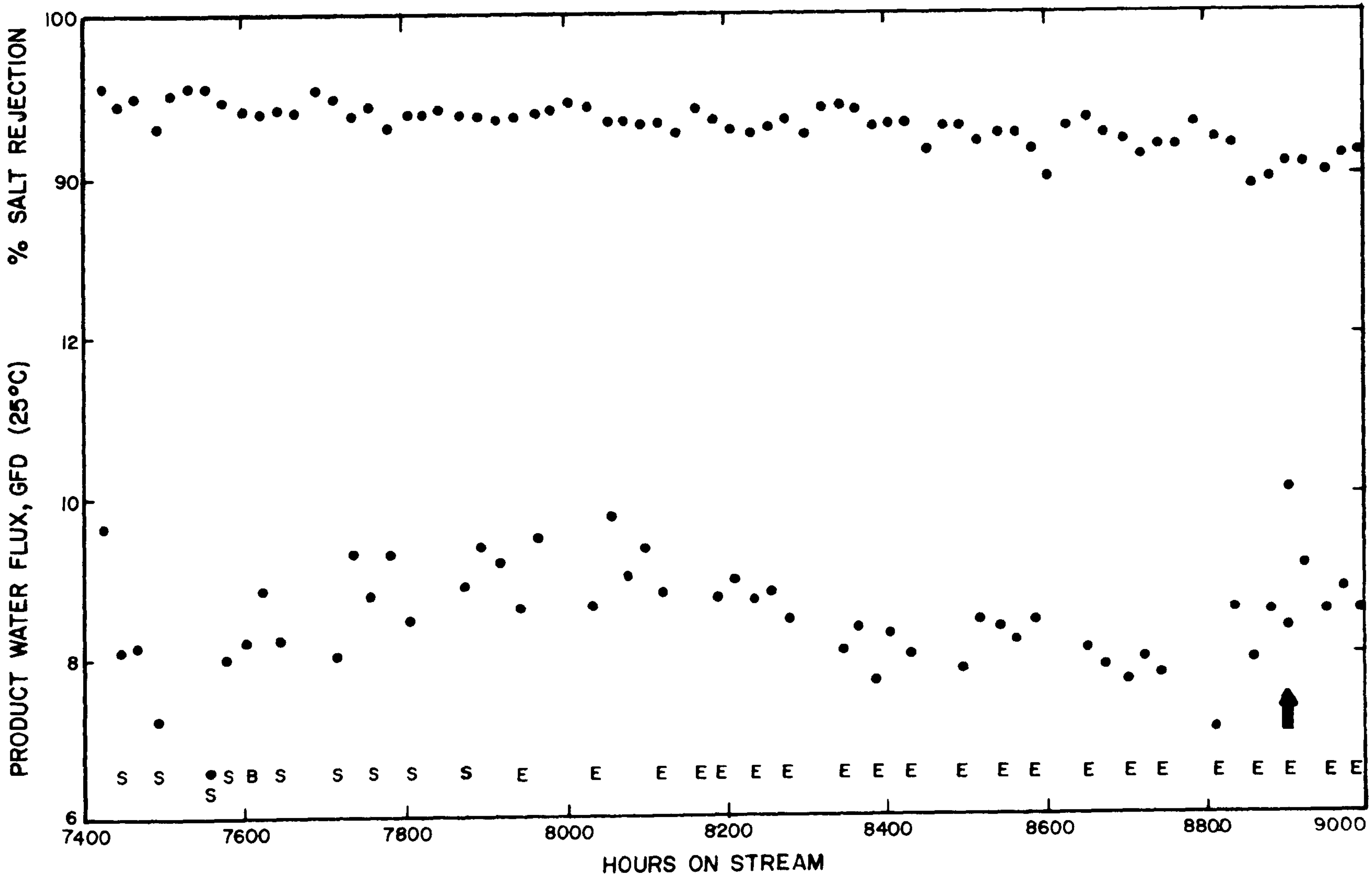


Figure 4. Salt rejection and product water flux variation vs. operation time under constant operating pressure.



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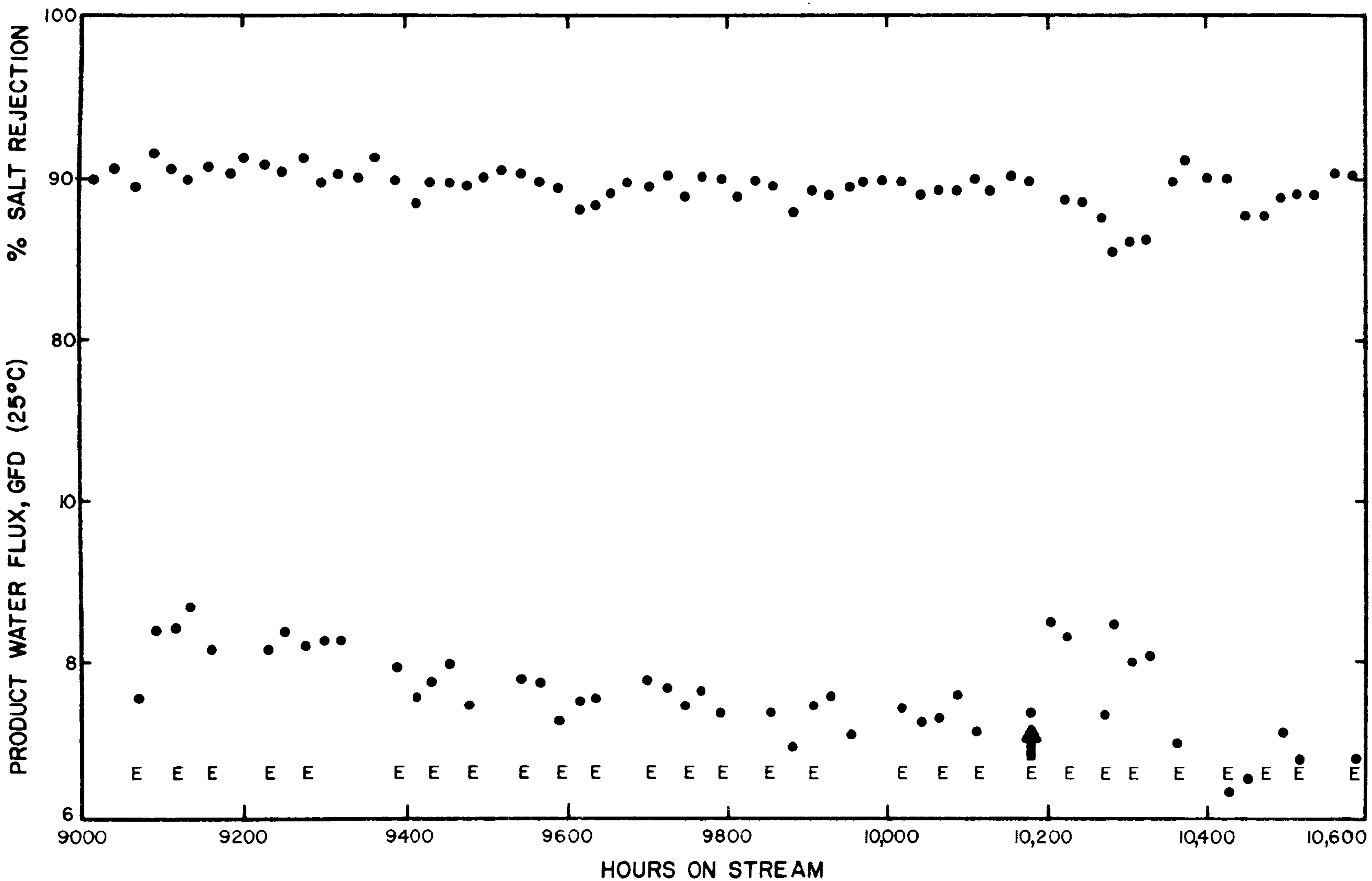

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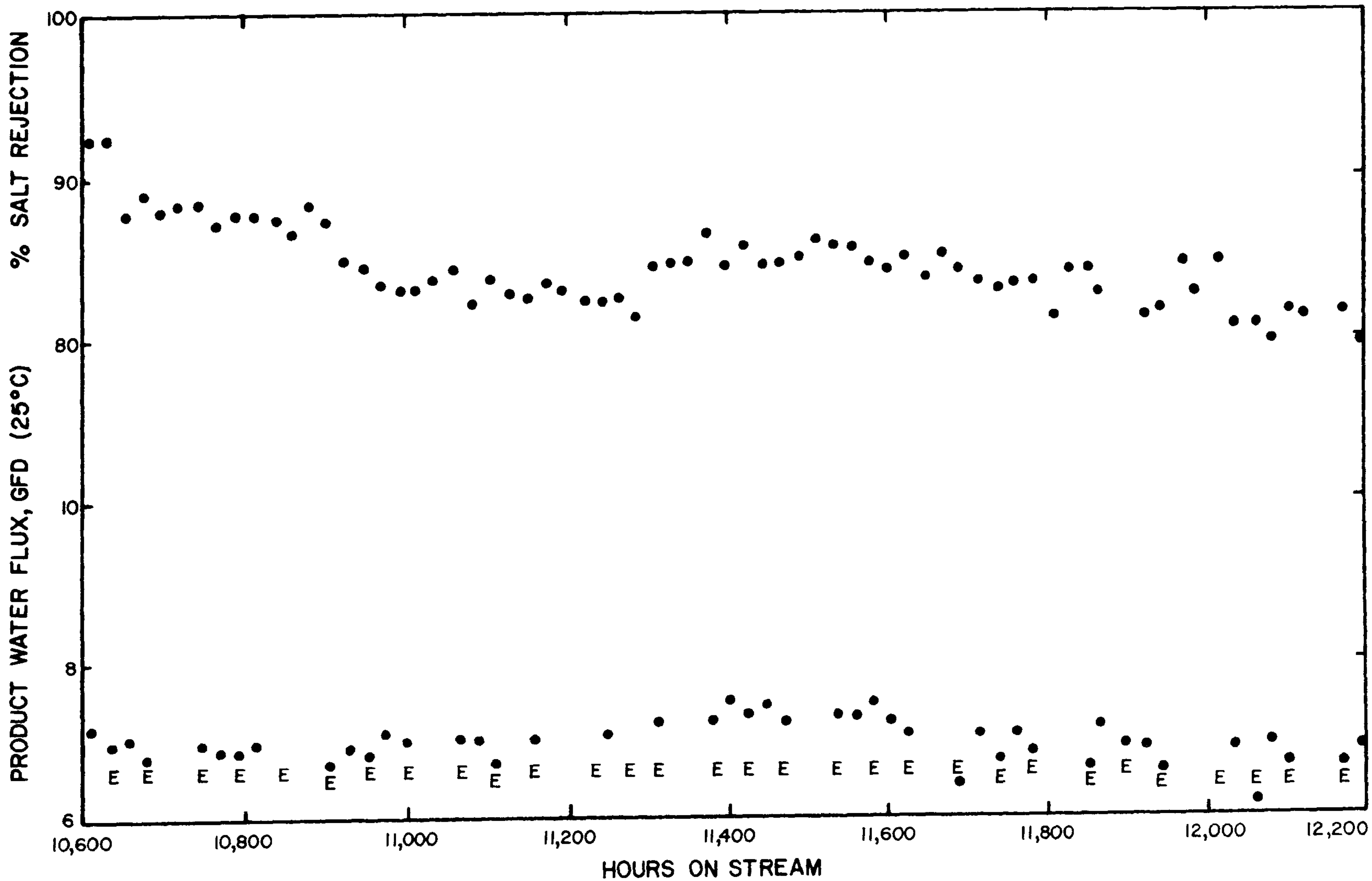

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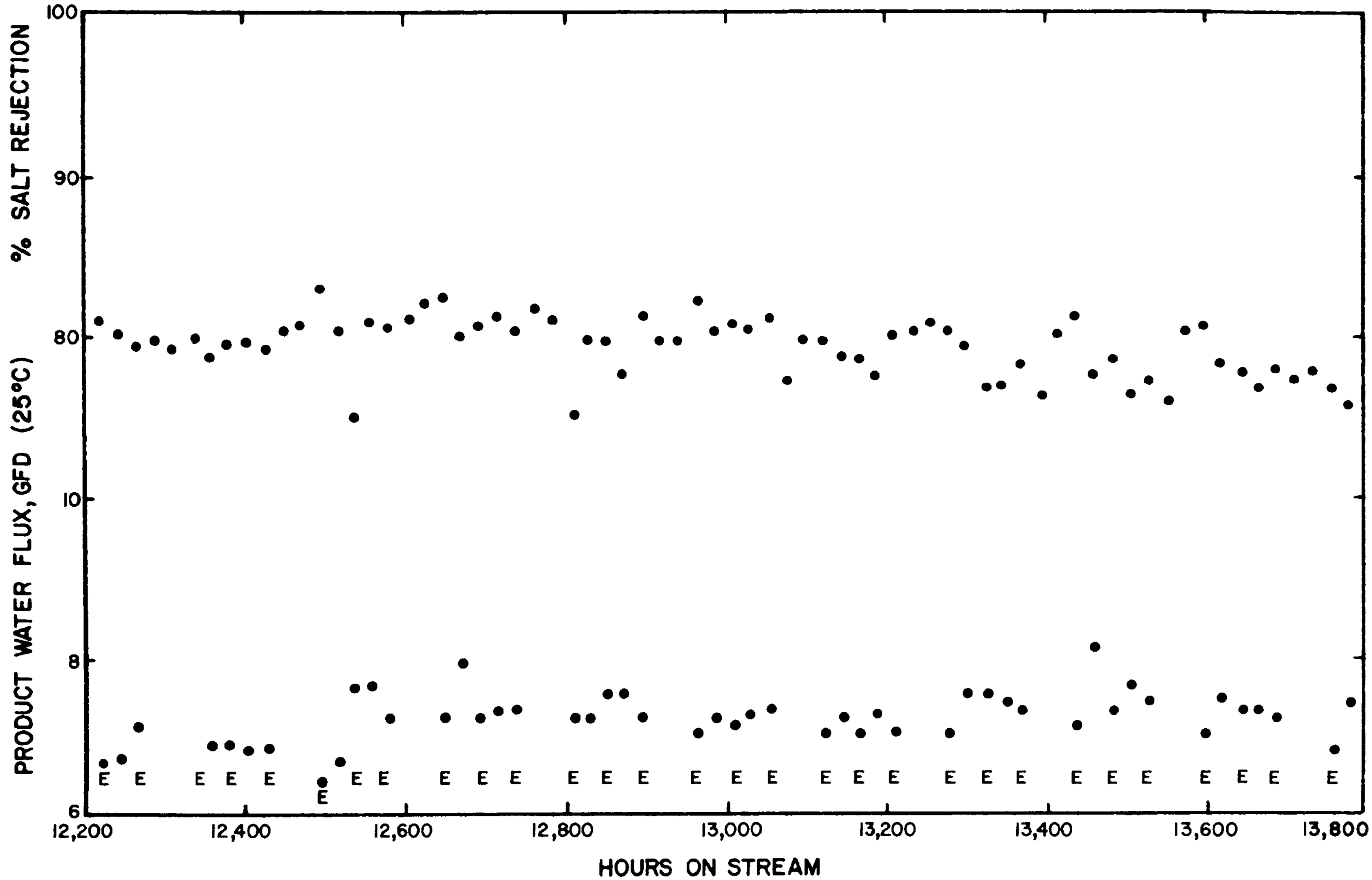

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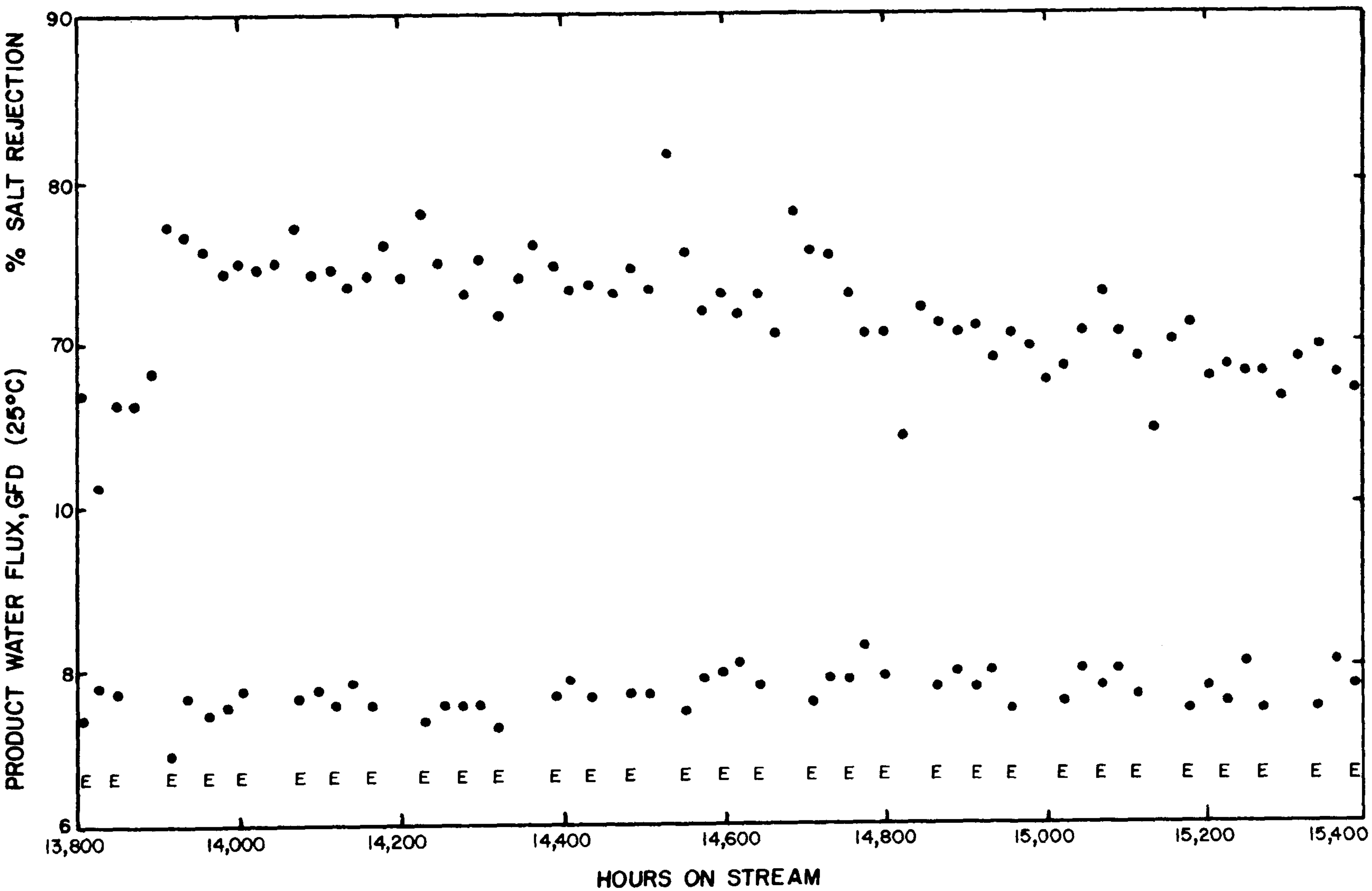

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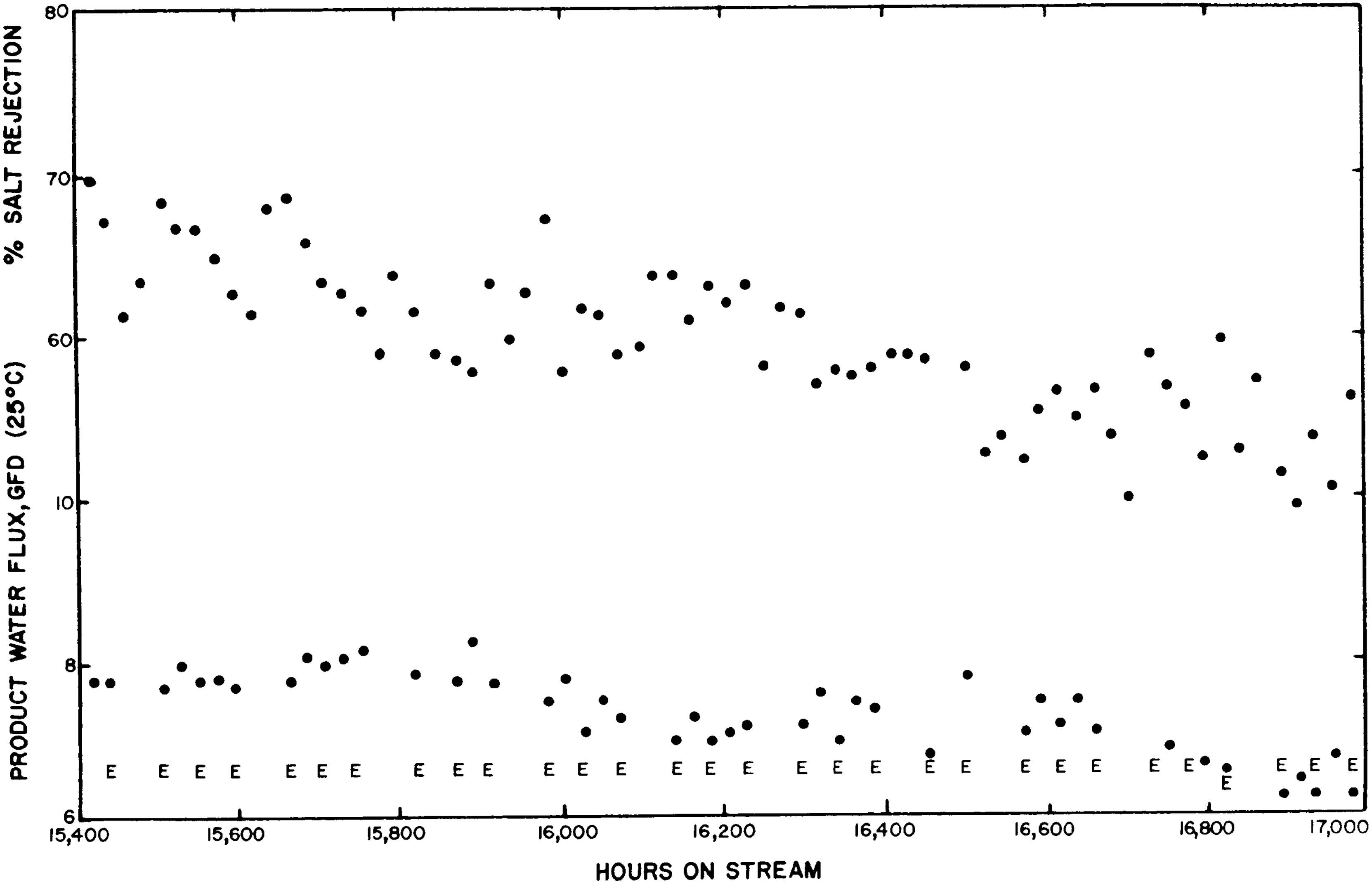

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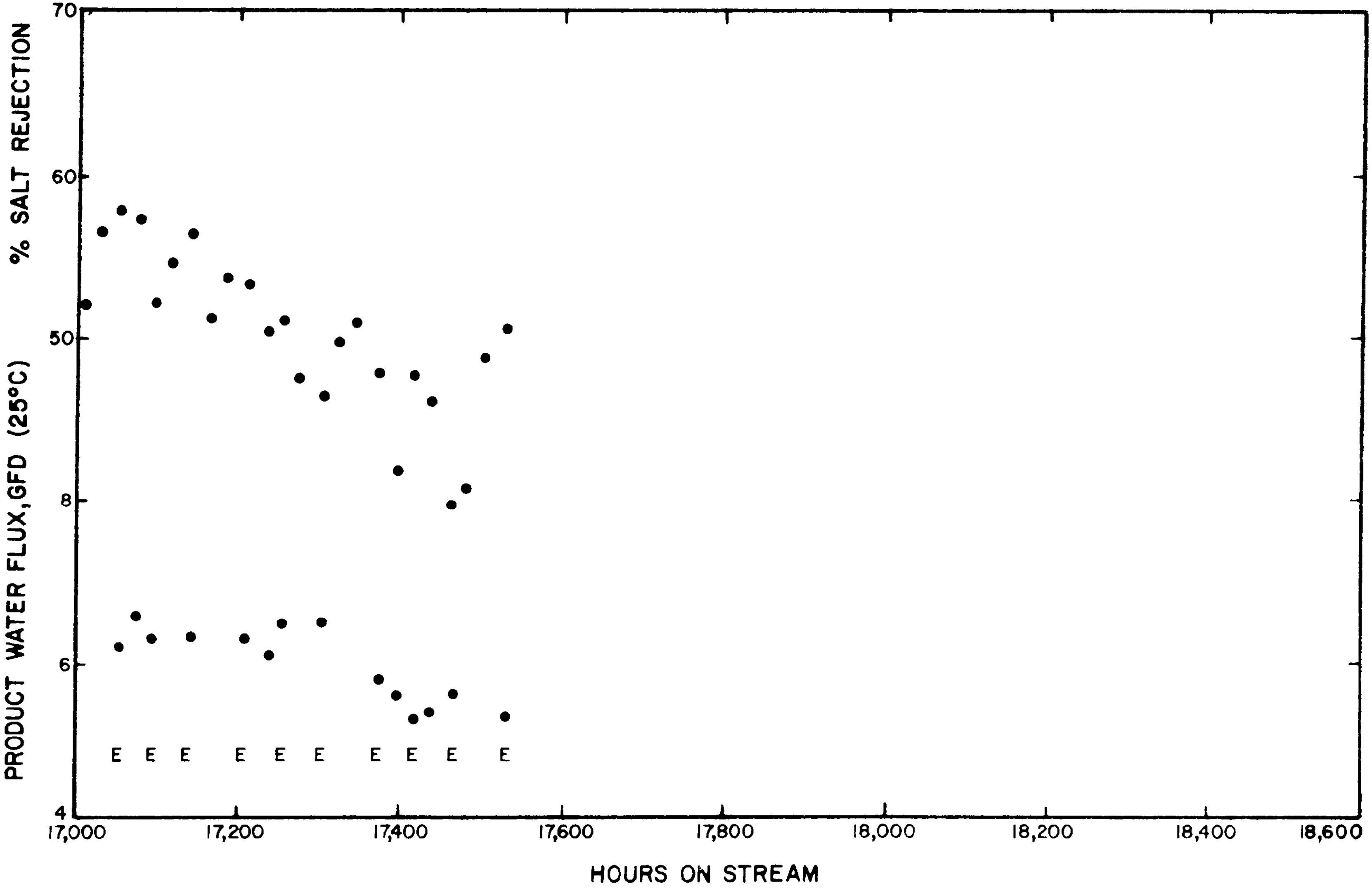

Figure 4. Continued.

secondary effluent quality. The turbidity and the dissolved chemical oxygen demand (DCOD) of the sand-filtered secondary effluent were as high as 22 JTU and 76 mg/l, respectively, during the two week activated sludge plant upset period between 7,421 and 7,737 hours of pilot plant operation. In spite of the poor quality secondary effluent, the module cleaning, using sodium perborate solution, was able to maintain the product water flux rate between 265 to 366 l/sq m/day (6.5 to 9 gal/sq ft/day).

At 8,096 hours of operation, the membrane modules in the system were inspected again. The results of the inspection showed that feed ends of the lead modules in pressure vessels No. 1 and No. 2 were covered with suspended solids and other debris, which were easily removed by applying a high pressure stream of tap water. No serious scaling to the extent of causing a severe restriction of the brine flow was found. The brine seals and the outer wrap tape were in good condition. A 72 hour depressurization on chlorinated tap water allowing the module inspection was able to improve the product water flux rate slightly.

As indicated in Figure 4, the salt rejection maintained at 95 percent or more for the first 7,400 hours of on-stream operation. Since then, the salt rejection gradually decreased to about 90 percent at 9,000 hours of operation. Similarly, the product water flux rate was maintained at 407 l/sq m/day (10 gal/sq ft/day) or higher practically throughout the first 7,400 hours of operation, and then it decreased to about 326 l/sq m/day (8 gal/sq ft/day) at 9,000 hours of operation. This amounted to a 6 percent reduction in salt rejection and a 20 percent reduction in product water flux rate during the first year pilot plant operation.

The salt rejection was maintained at the level of 90 percent between 9,000 and 10,600 hours of operation. However, the product water flux rate was found to decrease from the level of 326 l/sq m/day (8 gal/sq ft/day) to the level of 285 l/sq m/day (7 gal/sq ft/day) during the same operation period.

During the period between 10,600 and 12,600 hours of operation, the product water flux rate was found to hold steady at the level of 285 l/sq m/day (7 gal/sq ft/day) while the salt rejection was found to decrease rapidly from 90 percent to 80 percent.

The product water flux rate started to reverse its down-trend at 12,600 hours of operation (in the month of November, 1972), as shown in Figure 5. This up-trend continued for a period of 3,200 hours, and it was finally decreased again at 15,800 hours of operation.

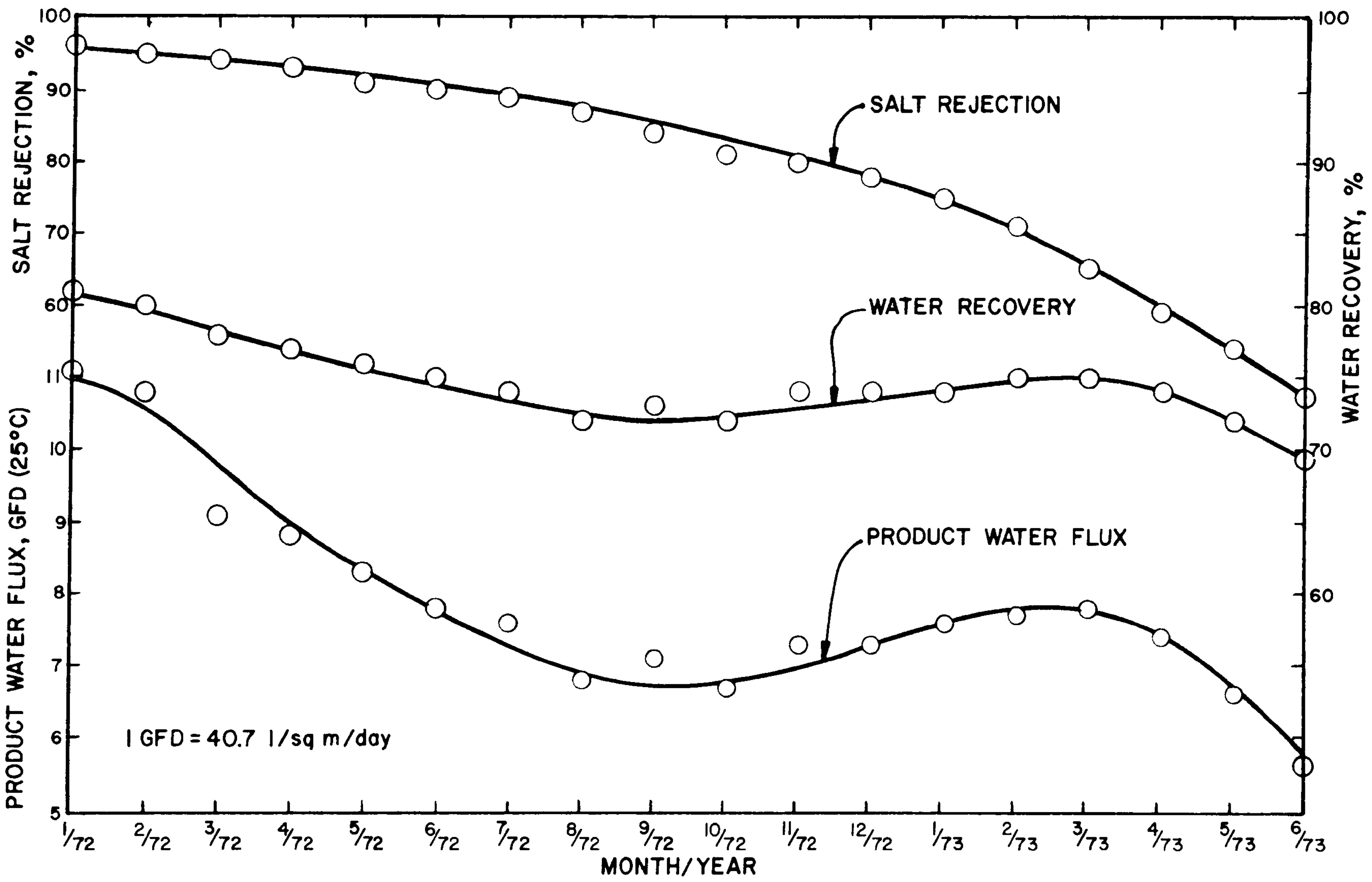

Figure 5. Monthly averages of the R.O. pilot plant performance parameters (corresponding to the period of 5900 to 17528 hours of operation).

The phenomenon of the product water flux rate reversal could very well indicate that some physical destructions of the membrane structure had taken place inside the membrane modules. The damaged membranes could lose the ability to function. Consequently, some brine water would be able to flow through the membrane and contaminate the product water, thus causing a poor salt rejection. After 15,800 hours of operation (corresponding to the month of March, 1973), the salt rejection, product water flux rate, and the water recovery were all found to decrease rapidly, as shown in Figure 5. The pilot plant was finally terminated on June 11, 1973, after a total of 17,528 hours of on-stream operation. The total reduction in salt rejection, water recovery, and product water flux rate during the two years of on-stream operation period were approximately 51 percent, 15 percent, and 50 percent, respectively.

WATER QUALITY CHARACTERISTICS

The typical water quality characteristics of the sand-filtered secondary effluent, which was used as the raw feed to the reverse osmosis pilot plant study, are shown in Table 2. The concentration of the nitrate ion in the raw feed was so low that the 30 percent rejection, as indicated in Table 2, might not demonstrate the true rejection capability of the reverse osmosis system. It is also shown in Table 2 that both organic matter and turbidity were quite effectively removed. However, those high rejection values during the initial period of operation were found to decrease rapidly after approximately 9,000 hours of on-stream operation.

Table 3 shows the typical ion rejection values at three different periods of operation. As indicated in Table 3, [illegible] chloride and nitrate rejections dropped to zero after about two years of operation.

TABLE 2

AVERAGE WATER QUALITY CHARACTERISTICS

(October 15, 1971 to March 2, 1972)

Analysis	Raw Feed (mg/l)	Blended Feed (mg/l)	Product (mg/l)	Brine (mg/l)	Rejection (%)
Na	94.5	175.0	13.8	323.0	92
K	12.0	21.9	1.4	41.2	94
Ca	57.2	103.9	0.94	211.3	99
Mg	11.4	21.8	0.34	41.8	98
Cl	80.6	155.4	18.9	301.0	88
SO_4	65.0	534.0	4.9	1,038	99
PO_4-P	10.3	19.8	0.26	37.5	99
NH_3-N	18.6	33.1	1.8	63.6	95
NO_3-N	0.09	0.10	0.07	0.16	30
TCOD	31.7	57.1	0.95	99.8	98
DCOD	25.1	50.0	0.56	85.8	99
TDS	496	1,127	52	2,099	95
Turbidity (JTU)	2.1	3.3	0.03	8.0	99

Notes: 1. Raw Feed is sand-filtered secondary effluent.

2. Analysis on once a week grab samples taken at 8:00 A.M.

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

TABLE 3

TYPICAL ION REJECTION VALUES (%) AT

DIFFERENT PERIOD OF OPERATION

Ion	Approximate Hours on Stream		
	6,500	12,000	17,500
Sodium	90	72	43
Potassium	94	80	38
Ammonia Nitrogen	95	75	39
Calcium	99	90	40
Magnesium	98	85	48
Chloride	84	33	0
Nitrate Nitrogen	30	13	0
Sulfate	99	88	47
Phosphate	98	85	33
Total COD	99	85	48
Dissolved COD	99	86	37
TDS	95	79	41

SECTION 7

PROCESS COST ESTIMATE

The pilot plant studies at Pomona Research Facility have successfully demonstrated the technical feasibility of applying the spiral-wound reverse osmosis system to the demineralization of the sand-filtered secondary effluent. A process cost estimate for this application is prepared on the basis of the pilot plant operation results. The major assumptions made for this cost estimate are listed as follows:

A. The blended influent TDS for the reverse osmosis system is about 1200 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 rejecting 90 or more percent of the influent TDS;

E. One percent EDTA solution is used as the membrane cleaning solution, with pH of the solution adjusted to 7.5 to 8.0 with sulfuric acid at a temperature of 40 to 60°C;

F. The membrane cleaning is performed three times a week, or at an interval of 815 liters of product water per square meter of membrane area (20 gallons per square foot of membrane area);

G. The membrane life is one year;

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

I. The estimate is based on August, 1973 material and construction costs.

The initial capital cost including the feed pumps, membranes, pH controllers, chlorinators, chemical feed pumps, 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 membrane cost alone is about 1.15 million dollars.

Since the membrane has only one year of useful life, the membrane replacement cost is estimated to be about 8.2¢/1,000 liters (31.5¢/1,000 gallons). However, if the membrane life can be improved to two years, then the membrane replacement cost can 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 with a labor rate of $10,000 per year; and

B. Three man-years for operating the 37,850 cu m/day (10 MGD) plant at the same labor rate of $10,000 per year.

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. EDTA = $1.21/Kg ($0.55/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 ($0.040/lb).

According to the above chemical costs, the total expenses for the process operation will amount to 2.44¢/1,000 liters (9.1¢/1,000 gallons). This total chemical cost can be broken down into the following three different categories:

A. Membrane cleaning - 1.5¢/1,000 liters (5.6¢/1,000 gallons);

B. Acidification for chemical precipitation control = 0.8¢/1,000 liters (3.0¢/1,000 gallons); and

C. Chlorination for biological growth control = 0.1¢/1,000 liters (0.5¢/1,000 gallons).

Table 4 summarizes the various items of the process cost estimates. As indicated in Table 4, the total process cost is approximately 16.6¢/1,000 liters (63.6/1,000 gallons) for one year membrane life. The cost can be reduced to about 12.4¢/1,000 liters (47.5¢/1,000 gallons) by improving the membrane life to two years. Both cost estimates do not include the costs for the sand filtration pretreatment and the brine disposal.

This pilot plant study has shown that the secondary effluent from an activated sludge plant can be successfully demineralized by a spiral-wound reverse osmosis process with only sand filtration pretreatment, instead of the carbon adsorption pretreatment.(4) The difference in the total process costs, including the pretreatment costs, between the sand filtration (1¢/1,000 liters or 4¢/1,000 gallons) and the carbon adsorption (2.8¢/1,000 liters or 10.8¢/1,000 gallons) pretreatment schemes is shown in the following comparison:

A. With carbon adsorption pretreatment

a. One-year membrane life = 17.7¢/1,000 liters (68.2¢/1,000 gallons)

b. Two-year membrane life = 13.5¢/1,000 liters (52.1¢/1,000 gallons)

B. With sand filtration pretreatment:

a. One-year membrane life = 17.6¢/1,000 liters (67.6¢/1,000 gallons)

b. Two-year membrane life = 13.4¢/1,000 liters (51.5¢/1,000 gallons)

TABLE 4

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)	9.1	2.4
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	3.0	0.8
Total Process Cost:		
One-year membrane life	63.6	16.6
Two-year membrane life	47.5	12.4

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16. ABSTRACT

A 22.7 cu m/day (6,000 gallons/day) spiral-wound reverse osmosis pilot plant, was operated at the Pomona Advanced Wastewater Treatment Research Facility on the sand-filtered secondary effluent. The pilot plant study was conducted under optimum operating conditions based on previous studies. During the first year of operation, all the system performance parameters, such as salt rejection, water recovery, and product water flux rate, were only slightly decreased from their initial values. However, the salt rejection and product water flux rate were substantially reduced to almost half of their initial values after a two year operation period. During this same two year period, the water recovery was found to decline about 15 per cent of its initial value.

A cost estimate for a 37,850 cu m/day (10 MGD) plant for August,1973 cost figures indicated that for membranes with only one-year life the process cost was about 16.5¢/1,000 liters (63.5¢/1000 gallons). However, the cost could be substantially reduced to 12.4¢/1,000 liters (47.5¢/1,000 gallons) for membranes with two-year life. Both cost estimates did not include the costs for sand filtration pretreatment and brine disposal.

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