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RENOVATION OF WASTE SHOWER WATER BY MEMBRANE FILTRATION.(U)
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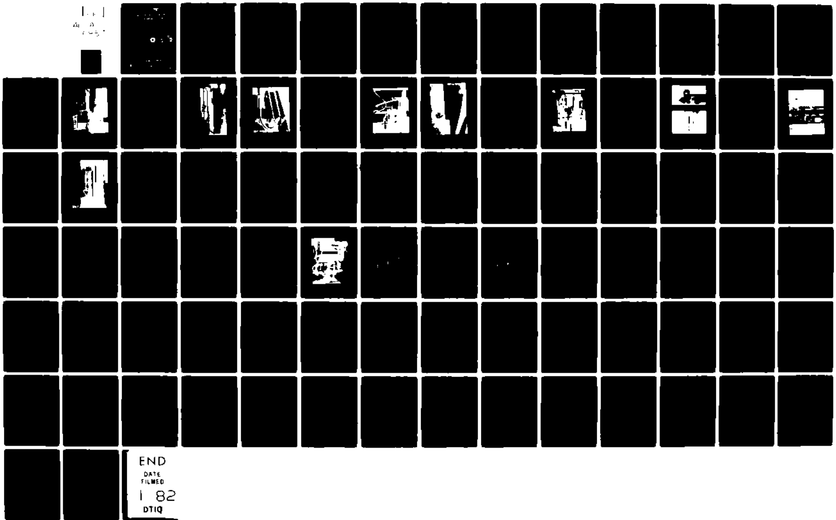

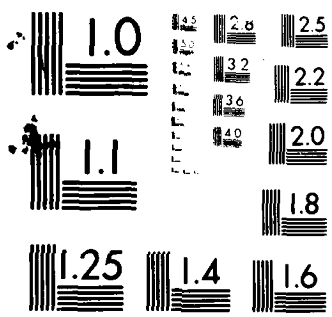

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RENOVATION OF WASTE SHOWER WATER BY MEMBRANE FILTRATION

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U.S. ARMY MOBILITY EQUIPMENT RESEARCH & DEVELOPMENT COMMAND

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Sanitary Sciences Division

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16. Abstract

In accordance with current efforts for environmental protection and energy conservation, the U.S. Coast Guard is considering shower wastewater treatment for reuse as laundry water aboard water craft. A process being investigated for this purpose is ultrafiltration. Five off-the-shelf ultrafiltration systems were considered containing membrane fiber configurations of tubular, spiral-wound, hollow, and plate-and-frame. Ultrafiltration rates (fluxes) along with power requirements were observed to vary significantly depending on the system and the membrane configuration used. The treated water was of suitable quality for reuse as laundry water. Although membrane cleaning could recover flux, the rate of flux decline was faster for cleaned membranes than new membranes. To protect the system, pretreatment is required to remove hair and other fibers from the feedwater prior to application to the membranes. The hollow-fiber membrane configuration is the only membrane configuration not requiring chemicals for membrane cleaning.

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PREFACE

The information in this report was developed by the Sanitary Sciences Division, U.S. Army Mobility Equipment Research and Development Command, Fort Belvoir, Virginia, during the period 1974 to 1976. Authority for the work was Military Interdepartmental Purchase Request (MIPR) No. Z-70099-4-43107, dated 14 May 1974, from Commandant (6-FCP-2/71); U.S. Coast Guard, Washington, D.C. 20590.

The preparation of this report was accomplished under the supervision of Maurice Pressman, Group Leader, Sanitary Sciences Division; Richard P. Schmitt, Chief, Sanitary Sciences Division; and LTC Robert Carnahan, R&D Coordinator, Laboratory 2000.

METRIC CONVERSION FACTORS

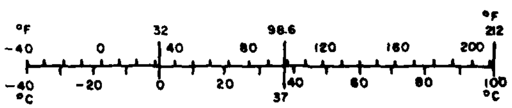
Approximate Conversions to Metric Measures

Symbol	When You Know	Multiply by	To Find	Symbol
		LENGTH		
in	inches	*2.5	centimeters	cm
ft	feet	30	centimeters	cm
yd	yards	0.9	meters	m
mi	miles	1.6	kilometers	km
		AREA		
in^2	square inches	6.5	square centimeters	cm^2
ft^2	square feet	0.09	square meters	m^2
yd^2	square yards	0.8	square meters	m^2
mi^2	square miles	2.6	square kilometers	km^2
	acres	0.4	hectares	ha
		MASS (weight)		
oz	ounces	28	grams	g
lb	pounds	0.45	kilograms	kg
	short tons (2000 lb)	0.9	tonnes	t
		VOLUME		
tsp	teaspoons	5	milliliters	ml
Tbsp	tablespoons	15	milliliters	ml
fl oz	fluid ounces	30	milliliters	ml
c	cups	0.24	liters	l
pt	pints	0.47	liters	l
qt	quarts	0.95	liters	l
gal	gallons	3.8	liters	l
ft^3	cubic feet	0.03	cubic meters	m^3
yd^3	cubic yards	0.76	cubic meters	m^3
		TEMPERATURE (exact)		
°F	Fahrenheit temperature	5/9 (after subtracting 32)	Celsius temperature	°C

*1 in ≡ 2.54 (exactly). For other exact conversions and more detailed tables, see NBS Misc. Publ. 286, Units of Weights and Measures, Price $2.25, SD Catalog No. C13.10:286.

Approximate Conversions from Metric Measures

Symbol	When You Know	Multiply by	To Find	Symbol
		LENGTH		
mm	millimeters	0.04	inches	in
cm	centimeters	0.4	inches	in
m	meters	3.3	feet	ft
m	meters	1.1	yards	yd
km	kilometers	0.6	miles	mi
		AREA		
cm^2	square centimeters	0.16	square inches	in^2
m^2	square meters	1.2	square yards	yd^2
km^2	square kilometers	0.4	square miles	mi^2
ha	hectares (10,000 m^2)	2.5	acres	
		MASS (weight)		
g	grams	0.035	ounces	oz
kg	kilograms	2.2	pounds	lb
t	tonnes (1000 kg)	1.1	short tons	
		VOLUME		
ml	milliliters	0.03	fluid ounces	fl oz
l	liters	2.1	pints	pt
l	liters	1.06	quarts	qt
l	liters	0.26	gallons	gal
m^3	cubic meters	35	cubic feet	ft^3
m^3	cubic meters	1.3	cubic yards	yd^3
		TEMPERATURE (exact)		
°C	Celsius temperature	9/5 (then add 32)	Fahrenheit temperature	°F



FORMULAS USED

Statistical

Given a set of data points $(x_1, x_2, \ldots\ldots\ldots, x_n)$

mean $\bar{x} = \frac{1}{n} \sum_{i=1}^{n} xi$

standard deviation $s_x = \sqrt{\frac{\Sigma x_i^2 - n\bar{x}^2}{n-1}}$

standard error of the mean $s_{\bar{x}} = \frac{s_x}{\sqrt{n}}$

coefficient of correlation $r = \sqrt{\frac{(y-\bar{y})^2 - \Sigma (y - y')^2}{\Sigma (y-\bar{y})^2}}$

where predictions are made with a multiple linear regression equation of the form $y' = a + bx$

General

$\text{Flux} = \frac{\text{Flow}}{\text{Membrane area}}$ $\qquad$ $\text{GFD} = \frac{(\text{GPM})\ (1440\ \text{min/day})}{\text{ft}^2}$

$\text{Power} = \frac{(1.73)\ (\text{I})\ (\text{E})\ (\text{P. F.})}{1000}$ $\qquad$ $\text{kw} = \frac{(1.73)\ (\text{Amps})\ (\text{Volts})}{1000}$

P. F. = 1

CONTENTS

ILLUSTRATIONS

TABLES

LIST OF ABBREVIATIONS

COD	– Chemical Oxygen Demand
E	– Electromotive force (volts)
° F	– Degrees fahrenheit
ft^2	– Square feet
ft/s	– Feet per second
g	– Grams
$gal/ft^2/d$	– Gallons per square foot per day
gal/h	– Gallons per hour
gal/min	– Gallons per minute
h	– Hours
I	– Current (amps)
" or in	– inches
i.d.	– Inner diameter
JTU	– Jackson Turbidity Unit
kW	– Kilowatts
LAS	– Linear Alkylate Sulfonate
mg/l	– Milligrams per liter
MUST	– Medical Unit, Self-Contained, Transportable
o.d.	– Outer diameter
%	– Percent
R	– Coefficient of Correlation
TDS	– Total Dissolved Salts
Temp	– Temperature
TSS	– Total Suspended Solids
TOC	– Total Organic Carbon
UF	– Ultrafiltration

RENOVATION OF WASTE SHOWER WATER BY MEMBRANE FILTRATION

INTRODUCTION

Membrane separation processes are playing an ever-increasing role in water and wastewater treatment. One particular type of membrane process is ultrafiltration. Ultrafiltration is a process for separating ultrafine particles, colloids, emulsions, and even macromolecules by pressure permeation through a special filtration layer of membrane.

The purpose of this study was to evaluate the effectiveness and feasibility of using off-the-shelf ultrafiltration hardware for the shipboard renovation of shower wastewater with subsequent reuse of the effluent as laundry water.

BACKGROUND

Rational thought concerning reuse of wastewater dictates that the degree of treatment necessary is dependent on how the water will be reused. Public health considerations require drinking and kitchen waters to meet the highest standards, followed, in turn, by shower and wash waters, laundry water, and finally, flush water.

Laundry water ranks next to last in reuse standards since the contaminants are not consumed by and do not come in direct contact with personnel. However, contaminants can come in dermal contact as a residue on the dried laundry, with the possibility of dermal irritation or carcinogenic effects. The water must also meet certain minimum physical and chemical standards required by the laundry process.

In considering renovation of a used water for reuse, it is natural to look for a relatively large volume of lightly contaminated wastewater. The renovation of used shower water for use as laundry water follows from this consideration. Membrane filtration was chosen over other physicochemical treatment techniques in an effort to simplify the logistical support problems associated with the treatment process. Power is usually available on board a ship, whereas space for stores of chemicals or spare components is not. Maintenance and repair time must be minimal due to limited availability of manpower.

At present, the U.S. Coast Guard is interested in renovating waste shower water only – not waste shower and washbasin water – for use as laundry water. The decontamination problem could be more complicated by the treatment of both waters, since almost anything could be poured down the drain of a washbasin. If the treatment is particularly successful, the possibility of reusing renovated waste shower water for shower water will also be considered.

The present study was undertaken to determine the effectiveness and feasibility of various systems based on ultrafiltration to process from 600 to 6400 gallons of shower water per day on board a vessel. All systems should be made of off-the-shelf components and have the capability of being scaled up or down, if necessary, for different size shipboard applications.

INVESTIGATION

Nineteen companies selling ultrafiltration equipment representative of the state-of-the-art were contacted for information about the availability of test units for the treatment of shower wastewater. Eleven companies responded with proposals to test units. After a detailed analysis of the proposals by MERADCOM and Coast Guard personnel, four systems were selected for procurement. Two other systems were studied: one which was made available on a rental basis and the other, a test stand already on hand with only the need to procure a module.

The five systems and test stand are summarized in Table 1. A total of eight membrane modules were evaluated on the five systems and one module on the test stand, making a total of nine modules evaluated.

Table 1. Summary of Ultrafiltration Systems

System Number	Operational Mode	Membrane Configuration	Membrane Type	Manufacturer
I*	Feed & bleed	Tubular (porous carbon tubes)	Inorganic	Union Carbide
II	Batch	Plate & frame	Vinyl	Aqua-Chem
III	Batch	Hollow fiber	Noncellulosic	Romicon
IV	Feed & bleed	Spiral wound	Noncellulosic	Abcor
V	Batch	Tubular	Modified Cellulosic	Westinghouse
VI**	Batch	Sprial wound	Noncellulosic	Osmonics

* Rental System

** Test Stand

DESCRIPTION OF SYSTEMS

System I is of the tubular configuration, consisting of two membrane modules each module 6 inches o.d. and 48 inches long operating in series. Each module contains 150 porous carbon tubes ¼ inch i.d. and 48 inches long with an inorganic membrane on the inside of the tube. The total membrane area for the system is approximately 78.5 ft^2. The flow rate through the modules is 300 gal/min, or 2 gal/min per tube, giving an average flow velocity of 15.7 ft/sec. The pressure drop across both modules is approximately 20 lb/in^2g. A simplified flow chart for System I is shown in Figure 1; the actual system is shown in Figure 2. Feed is drawn into the system by a positive displacement feed pump, which raises the pressure of the feed stream. The flow passes through a motorized prefilter to the suction side of the circulating pump. The circulating pump maintains the desired flow velocity through the two modules in the main circulating loop. Pressure is monitored at the circulating pump's suction and discharge and could be regulated by an adjustable pressure relief valve. Permeate passing through the carbon tubes into the shell of the module is piped from the top of the module to the permeate outlet. Concentrate from within the main loop is purged at the top of the loop via a solenoid valve, operated periodically to maintain the desired concentration in the loop. A concentrate by-pass valve is also incorporated for system blow-down. A gas vent and vacuum breaker is located at the top of the loop to provide for discharge and inlet of air when filling and emptying the system. A thermometer in the main circulating loop allows for system temperature readings.

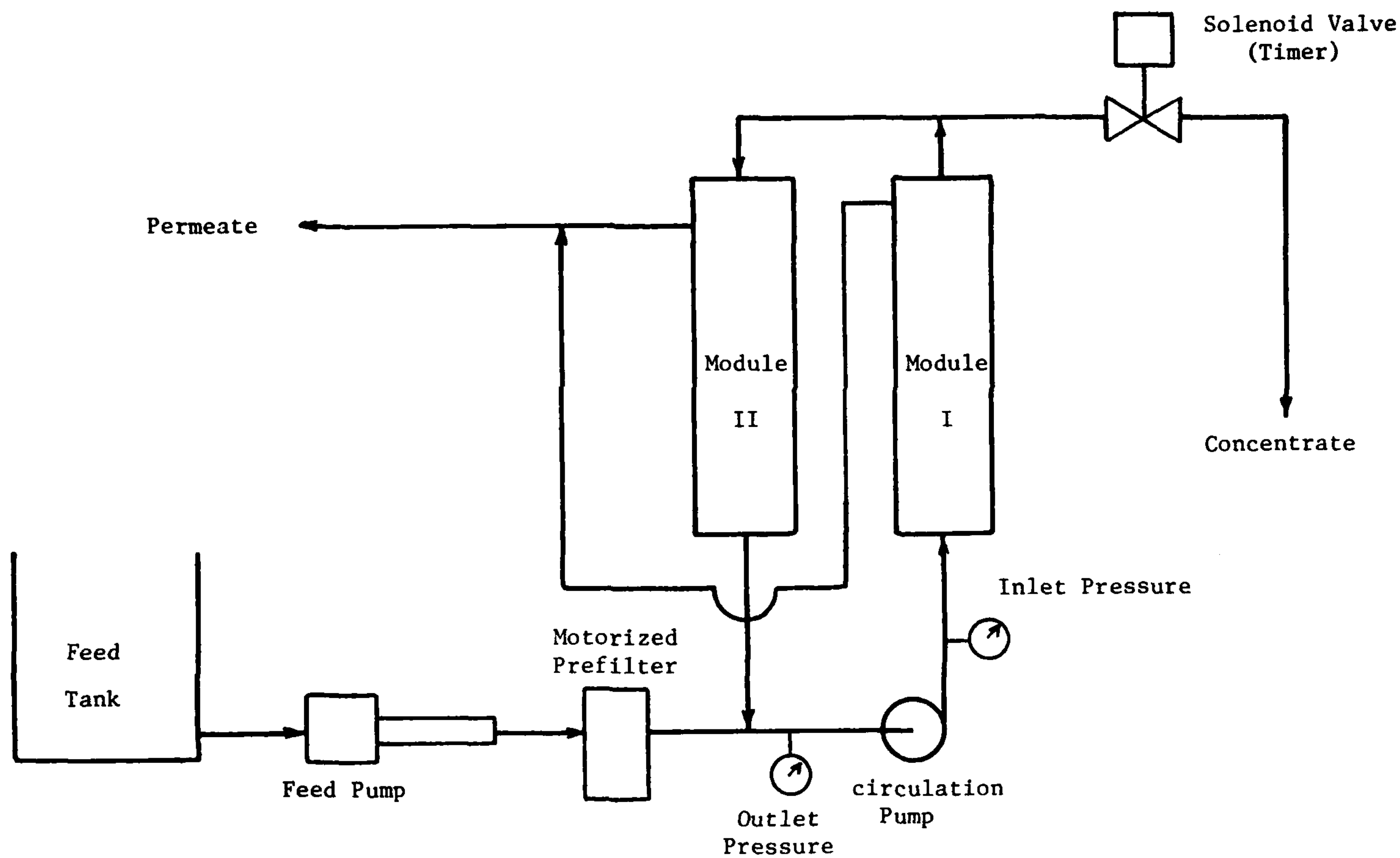


Figure 1. Flow chart for System I.

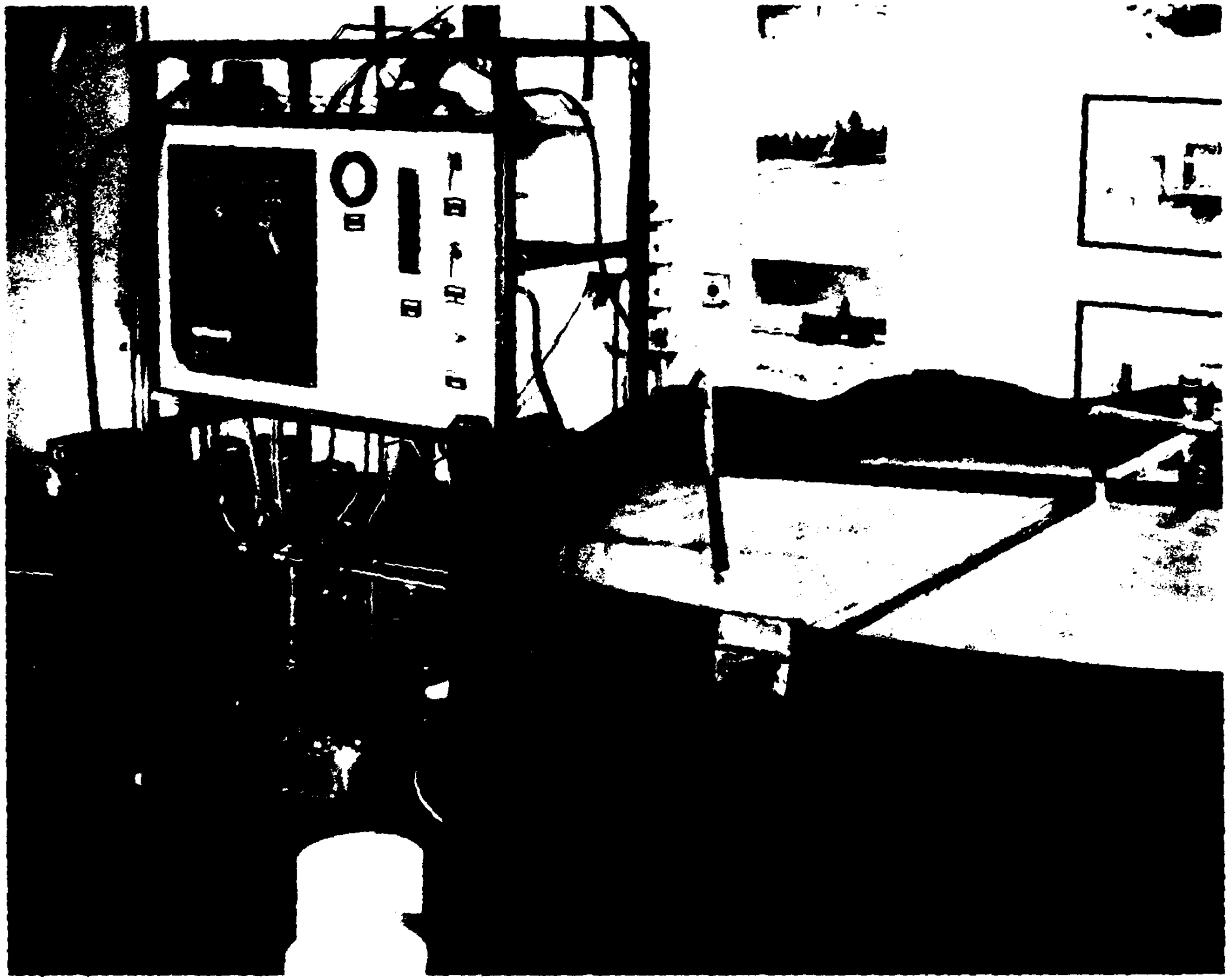


Figure 2. Ultrafiltration System I with samples of permeate, feed, and brine, respectively, from left to right.

System II is of the plate-and-frame configuration, consisting of a module stack measuring 26 by 10 by 10.5 inches. The stack consists of 25 vinyl membranes in flat sheet form arranged in parallel, each with an approximate area of 0.9 ft^2, giving a total module area of 22.5 ft^2. The membrane material is vinyl and has a porosity of approximately 0.02 to 0.05 micron. The basic flow pattern uses a thin channel serpentine path in which feed and permeate streams are contained in discrete flow chambers. Fluid enters the stack end plate and flows through a common channel or feed conduit in all of the cell separators (shown in Figure 3). This feed fluid then is directed at right angles into the channels of the individual cells. In these channels, the fluid passes at a relatively high velocity and is bounded on each side by a controlled porosity ultrafiltration membrane. The fluid in these channels is at a pressure of 20 to 40 lb/in^2 g, while the permeate channels on the opposite side of the membrane are at atmospheric pressure. The permeate channels have the same image as the feed channels. The difference in channels is that the feed cell does not contain any membrane support or turbulence promotion devices, whereas the permeate cell is filled with a plastic mesh material (Figure 3) to give the membrane support at many points. This lack of support in the feed cell prevents the membrane from being backflushed. The permeate is directed to opposite corners of the cell and then removed from the stack (Figure 4). A simplified flow diagram for System II is shown in Figure 5. The system is shown in Figure 6. Feed water is pumped from the feed tank by a single pump through the membrane stack with the brine flow returning to the feed tank. Flow velocity through the membrane, as well as feed pressure, are controlled by two control valves located before and after the stack. A flow meter in the return line measures the flow returned to the feed tank, with the system usually operated at about 40 gal/min. Pressure drop across the stack is usually on the order of 10 to 15 lb/in^2 g. Normal system operation consisted of operating the system until the feed tank volume was decreased by about 85 percent, thereby achieving an 85 percent permeate recovery from the feed system. The temperature was monitored in the feed tank.

System III is based on a hollow fiber cartridge with a noncellulosic membrane coated on the inside of the fiber. Two cartridges were evaluated: one containing 660 fibers with a fiber i.d. of 0.045 inch containing 15 ft^2 of membrane area, and one containing 2940 fibers with a fiber i.d. of 0.020 inch containing 30 ft^2 of membrane area. Both cartridges contained the same membrane material which had a nominal molecular weight cutoff of 50,000. The fibers are encapsulated at both ends in a 3-inch-o.d., 25-inch-long clear, plastic cartridge, the feed flows through the inside of the fibers and the permeate collects in the shell. Fluid can pass in the reverse direction from the outside of the fiber to the inside, making it possible to backflush the membrane. An end view of the 0.020-inch cartridge is shown in Figure 7. Figure 8 is a simplified flow diagram of System III.

The water is pumped from the feed tank through a bag filter. The manufacturer recommends a filter or strainer with at least a 100-micron rating. The flow, at a rate of approximately 16 gal/min at an approximate velocity of 5.9 ft/sec for the 0.045-inch fibers, then enters a three-way valve through which it can be directed to either end of the cartridge. The concentrate coming from the opposite end of the cartridge is returned to the feed tank. The permeate fills a permeate tank of approximately 10-gallon capacity, with overflow becoming permeate flow. A small backflush pump is used for backflushing the membrane with permeate held in the permeate tank. Temperature was monitored in the feed stream. Figure 9 shows the entire system.

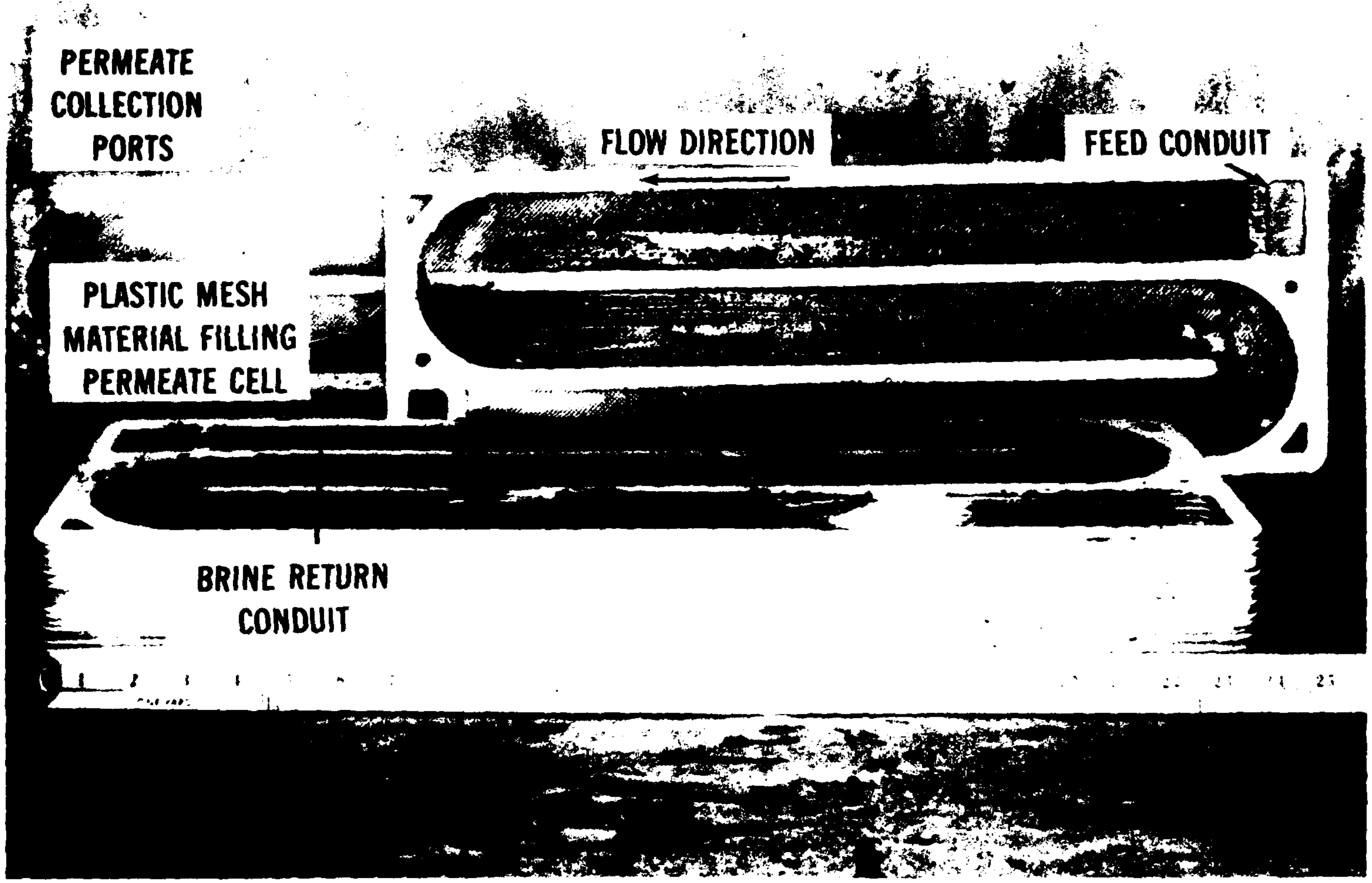


Figure 3. System II disassembled stack showing flow pattern in an individual cell.

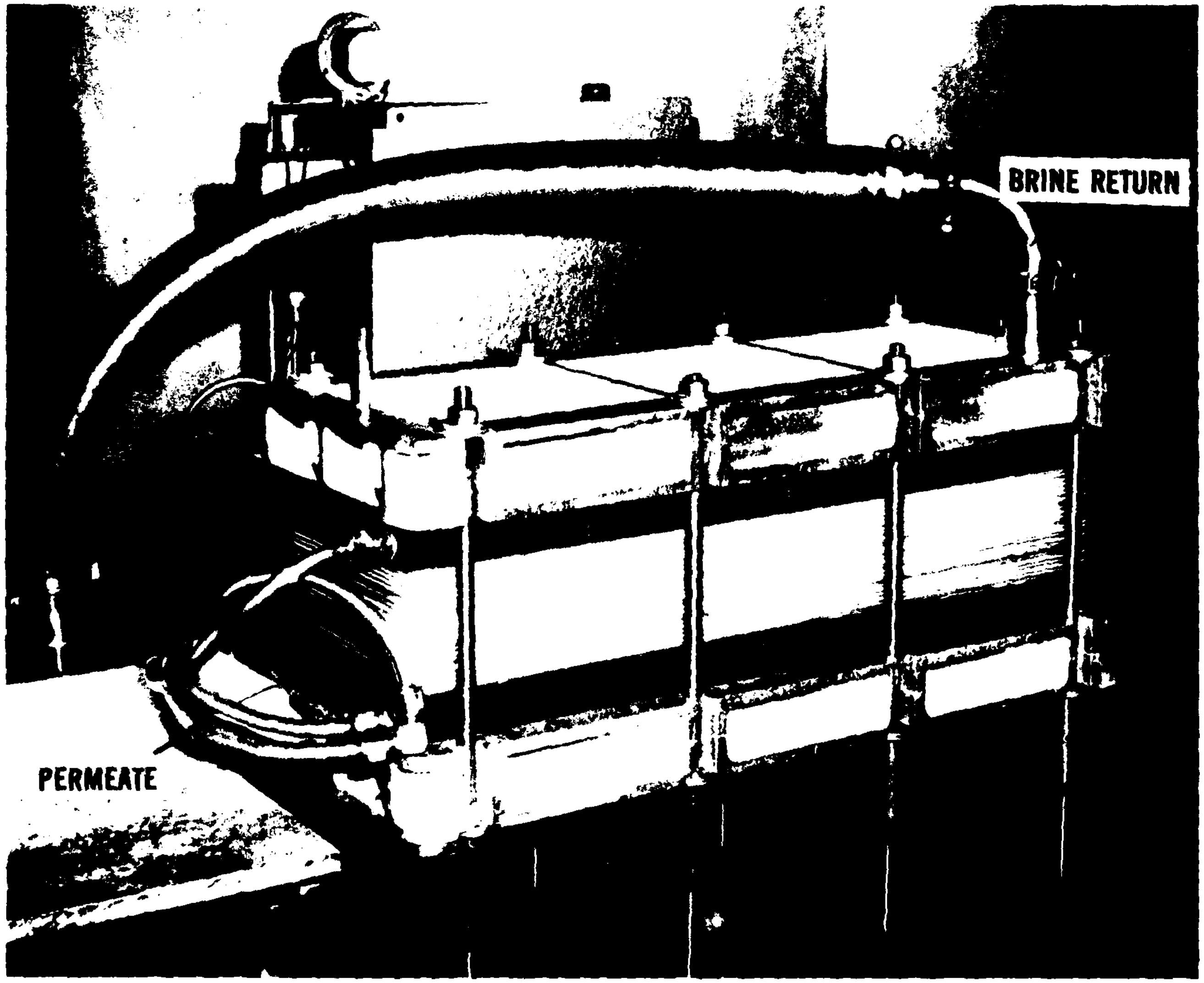


Figure 4. System II membrane stack.

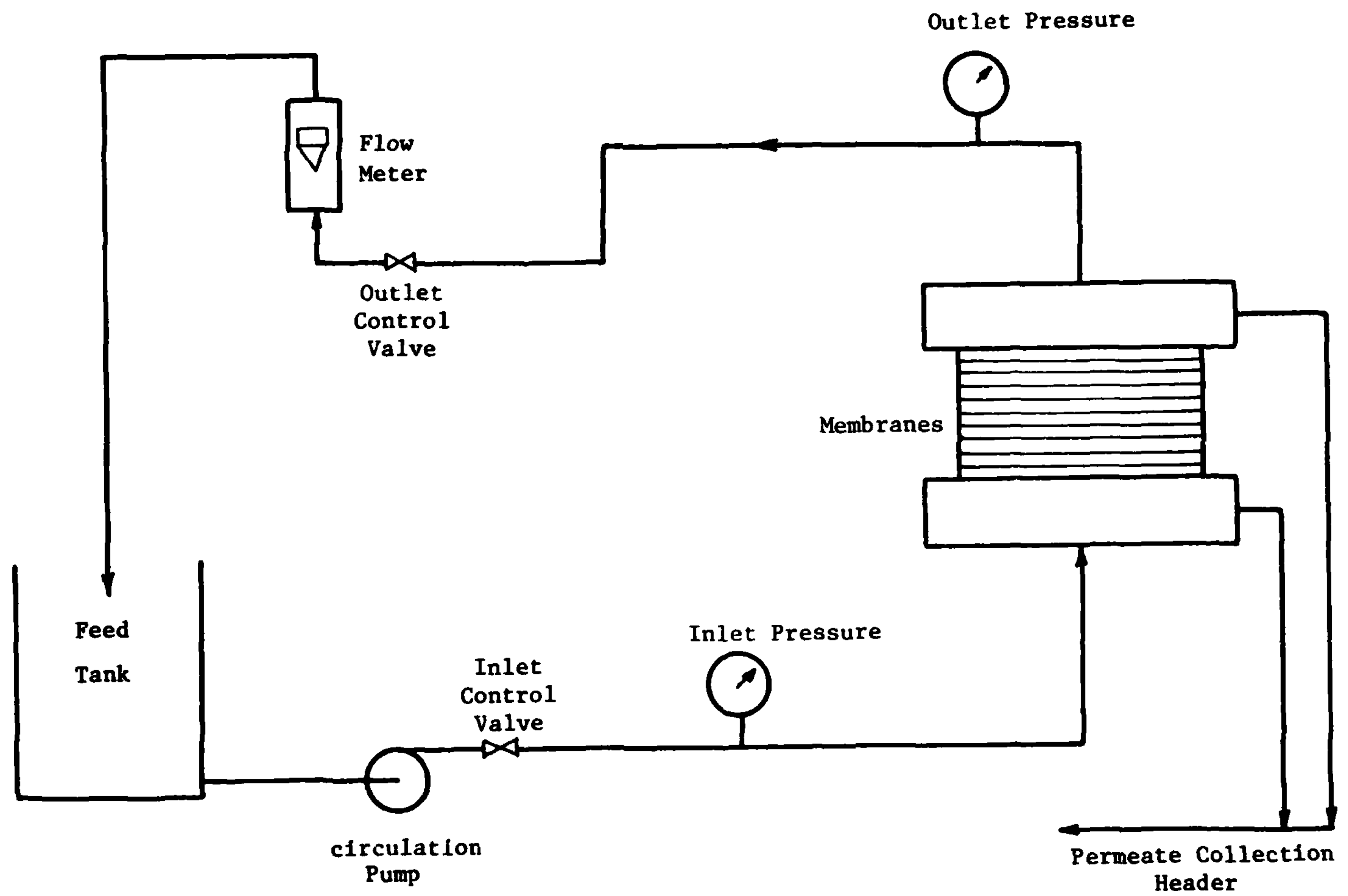


Figure 5. Flow chart for System II.

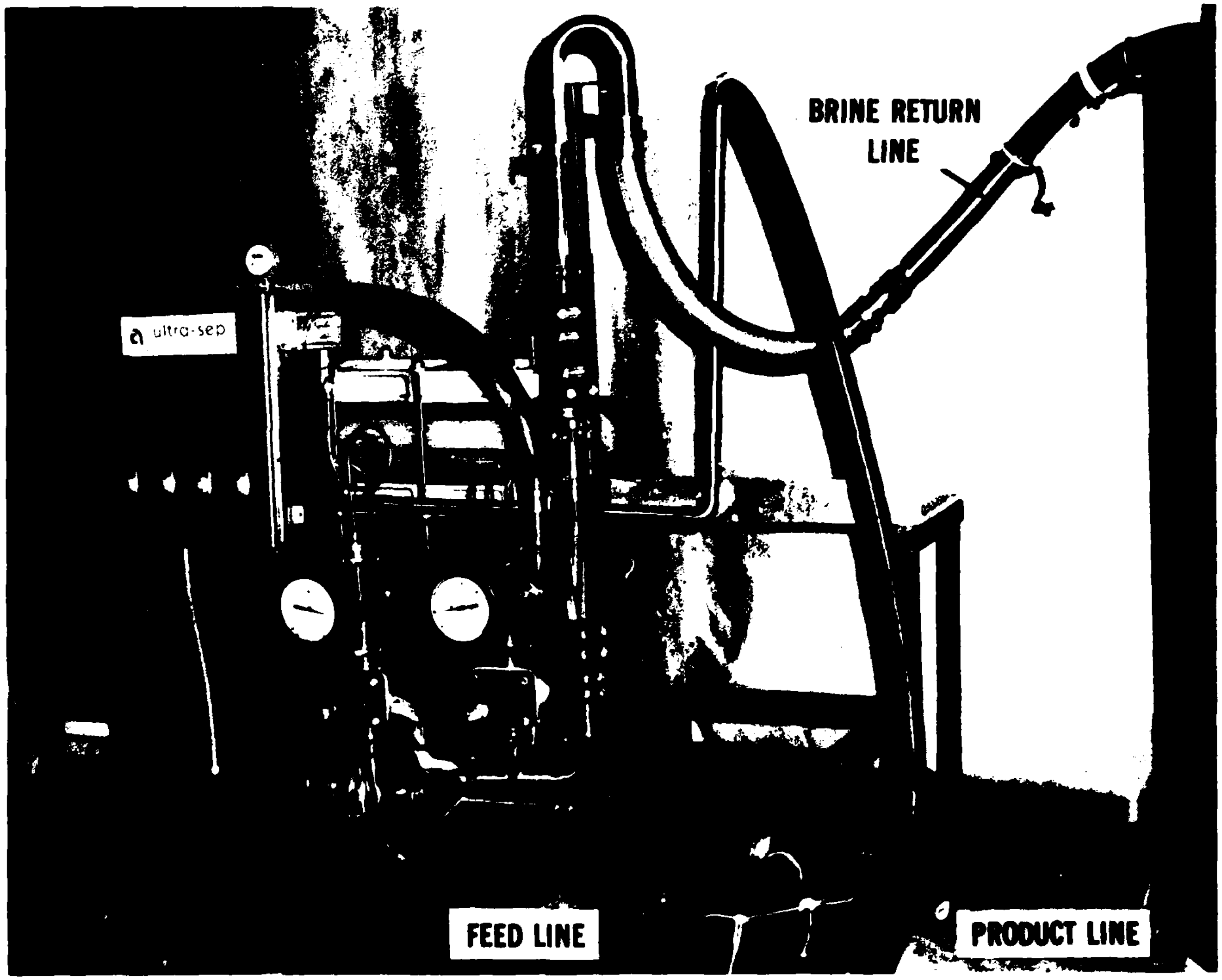


Figure 6. System II – plate-and-frame configuration.

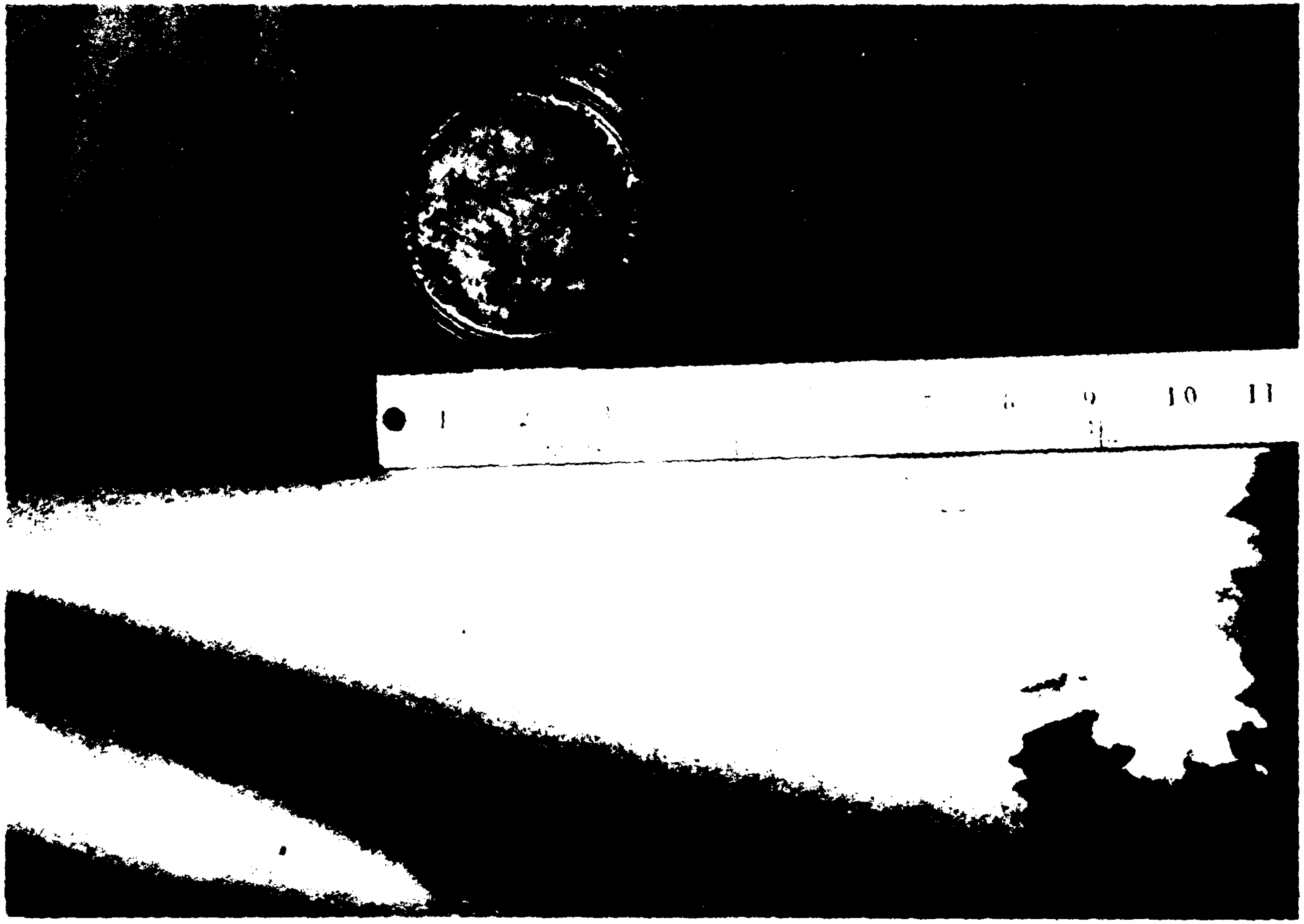


Figure 7. End view of 0.020-inch hollow-fiber cartridge.

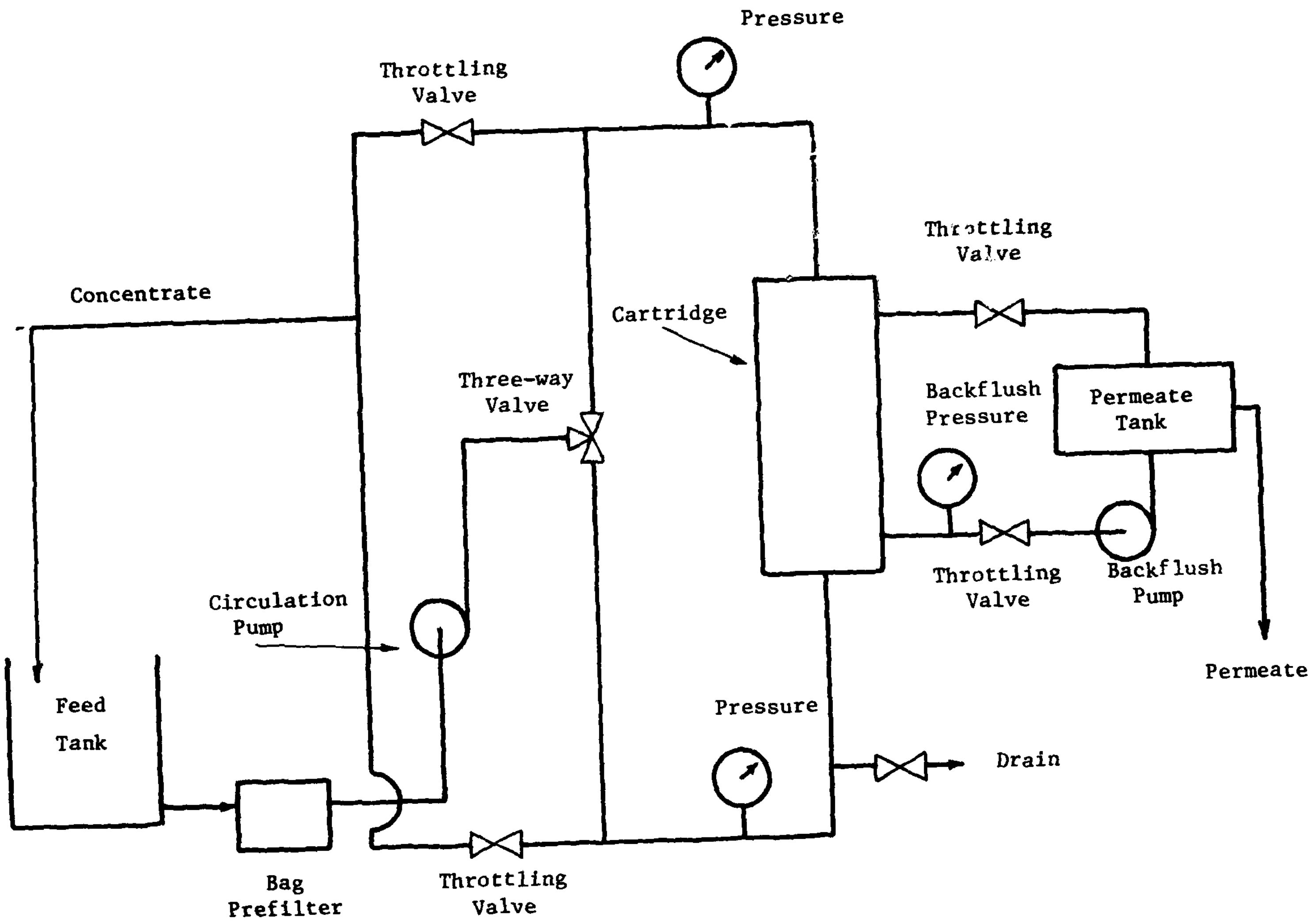


Figure 8. System III flow chart.

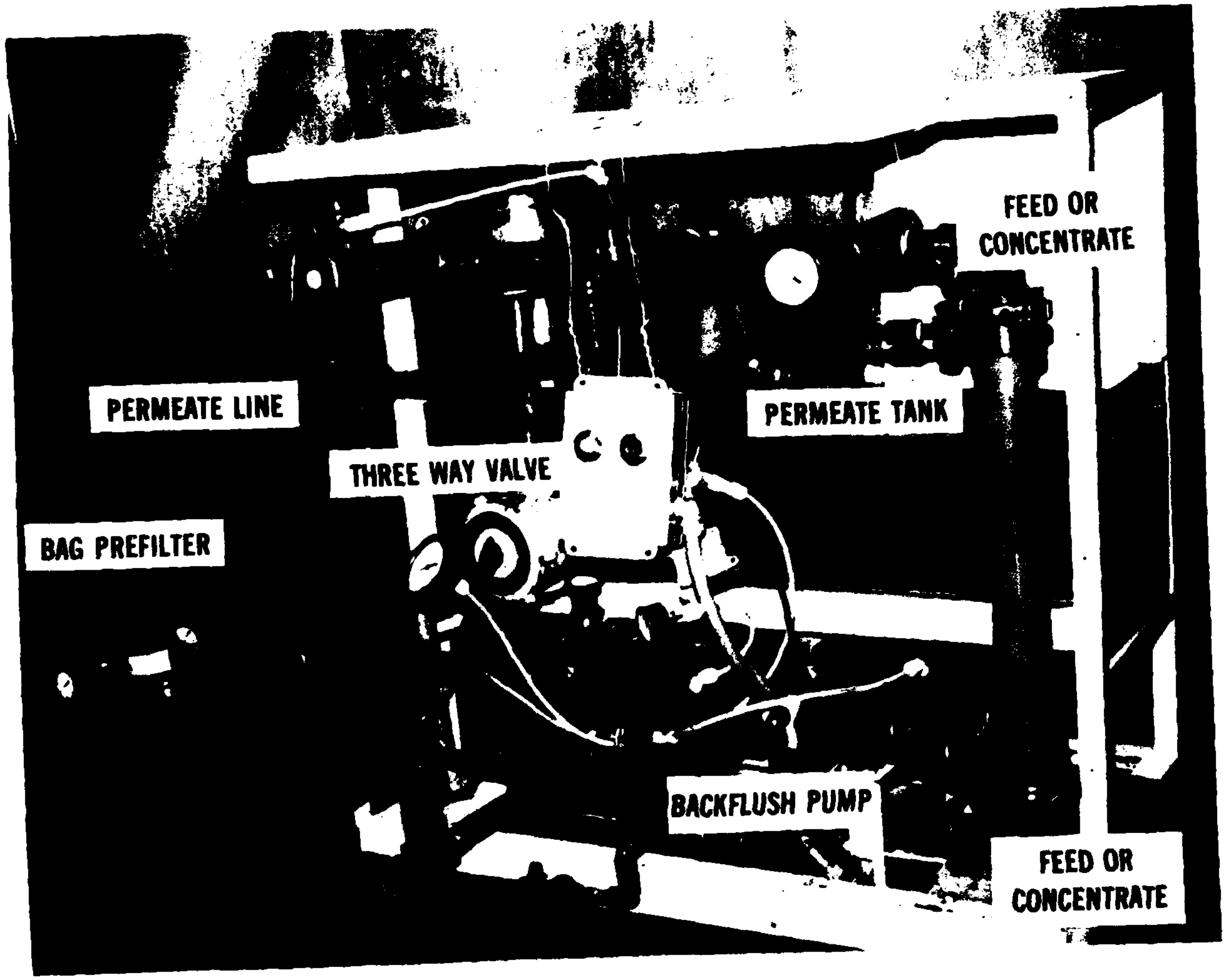


Figure 9. System III – hollow-fiber system.

System IV used modules of the spiral-wound configuration and a noncellulosic membrane. Two types of modules were evaluated: one with a 0.030-inch vexar spacer acting as a turbulence promoter (Figure 10), and one with a 0.080-inch corrugated spacer where the main source of turbulence is the feed water velocity through the module. Both types of modules contained approximately 30 ft^2 of membrane area per module. Figure 11 is a simplified flow chart for System IV. Water is pumped through two prefilters into the circulation pump which maintains the velocity in the circulation loop. The circulation rate through the modules is controlled by the pressure drop across a fixed orifice. Two modules are connected in series, as shown in Figure 12. The circulation rate through the corrugated spacer type module is approximately 60 gal/min and through the vexar modules, 10 to 15 gal/min. Concentrate from within the circulation loop is purged via a solenoid valve operated periodically to maintain the desired degree of concentration in the loop.

System V is of the tubular configuration containing eight modified cellulosic membrane modules. Each module contains 18 resin-bounded sand support tubes ½ inch in diameter, with the membrane on the inside of the log. Two modules are connected in series in each pressure vessel, with the four banks of pressure vessels running parallel. Flow is directed back and forth in sequence through all 18 tubes by flow-directing elements located at the end of the modules. Operation at a system feed flow rate of 25 gal/min yielded an approximate flow velocity of 12.26 ft/sec through the tubes. Figure 13 is a simplified flow diagram of System V. Water is pumped from the feed tank through the tubes and back to the feed tank. Flow rate through the modules is controlled by a by-pass vlave connected between the pump discharge and suction and with the setting of the back pressure valve. Permeate is collected through two permeate draw-off valves located at the top of each module. The system was operated in a batchwise mode to a feed concentration of 85 percent. Figure 14 shows the entire system.

System VI is a test stand designed to evaluate single spiral-wound modules. Figure 15 is a simplified flow diagram at the test stand. Water is pumped from the feed tank through the brine channels of the modules and returned to the feed tank. A by-pass valve, located in the line between the pump discharge and the feed tank, and a needle valve, located in the exit brine stream, were used to control the inlet pressure and the brine flow rate. Temperature measurements were made in the feed tank.

DESCRIPTION OF WASTEWATER

Two different wastewaters were used in this study. The early portion of the study used the MUST synthetic formulation for waste shower water without the hair. Fort Belvoir tap water was the water source. The constituents and their respective concentrations are shown in Table 2. Real shower water used in the latter portion of the tests came from a field shower unit located at Value Engineering Company, Alexandria, Virginia. The chemical characteristics of the two waste streams and Fort Belvoir tap water are shown in Table 3. The mean chemical characteristics for the 12 samples taken from different Coast Guard vessels along with proposed standards necessary for water to be used as laundry water are also shown in Table 3.

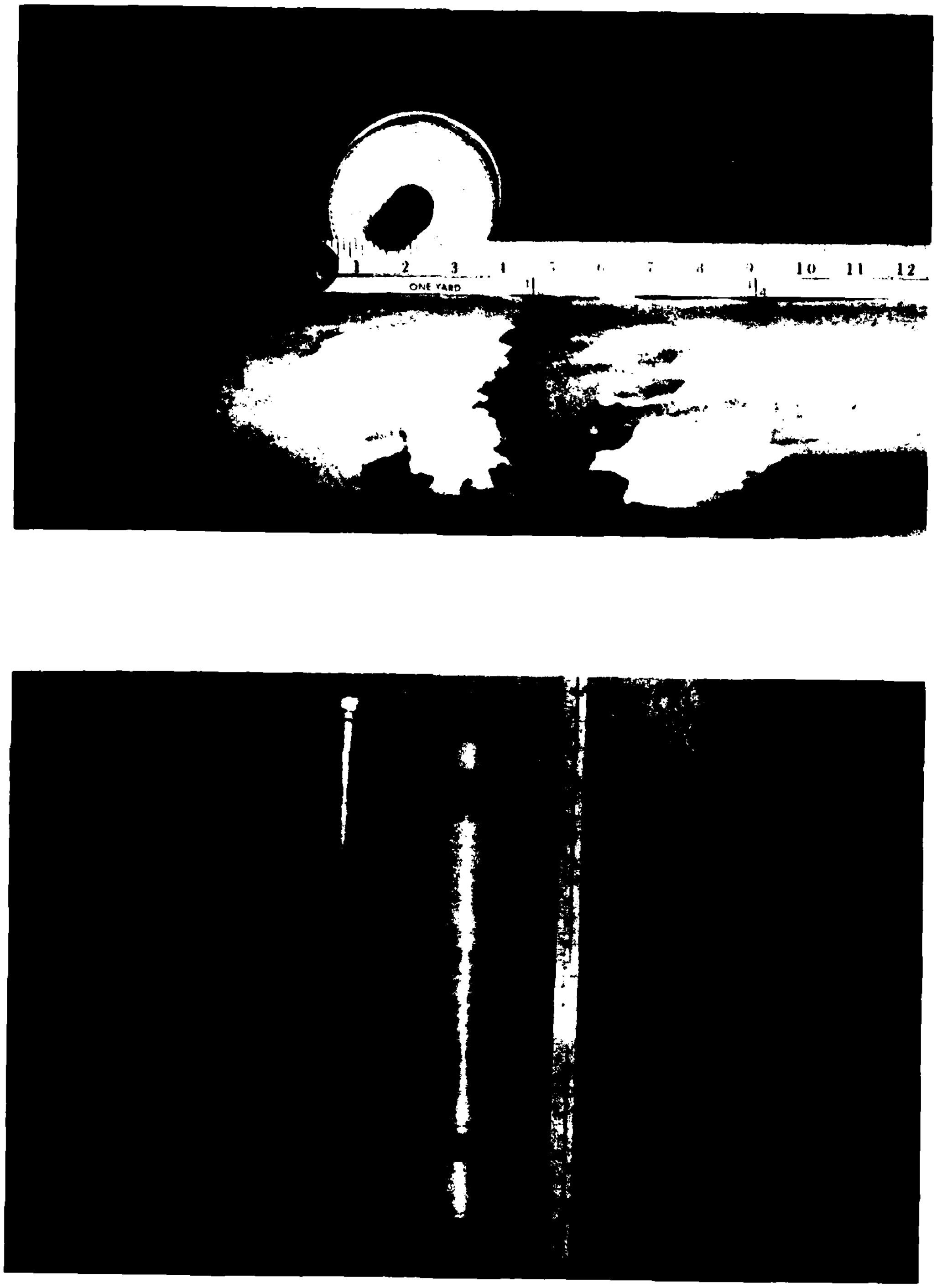


Figure 10. End and side views of spiral-wound module with vexar spacer.

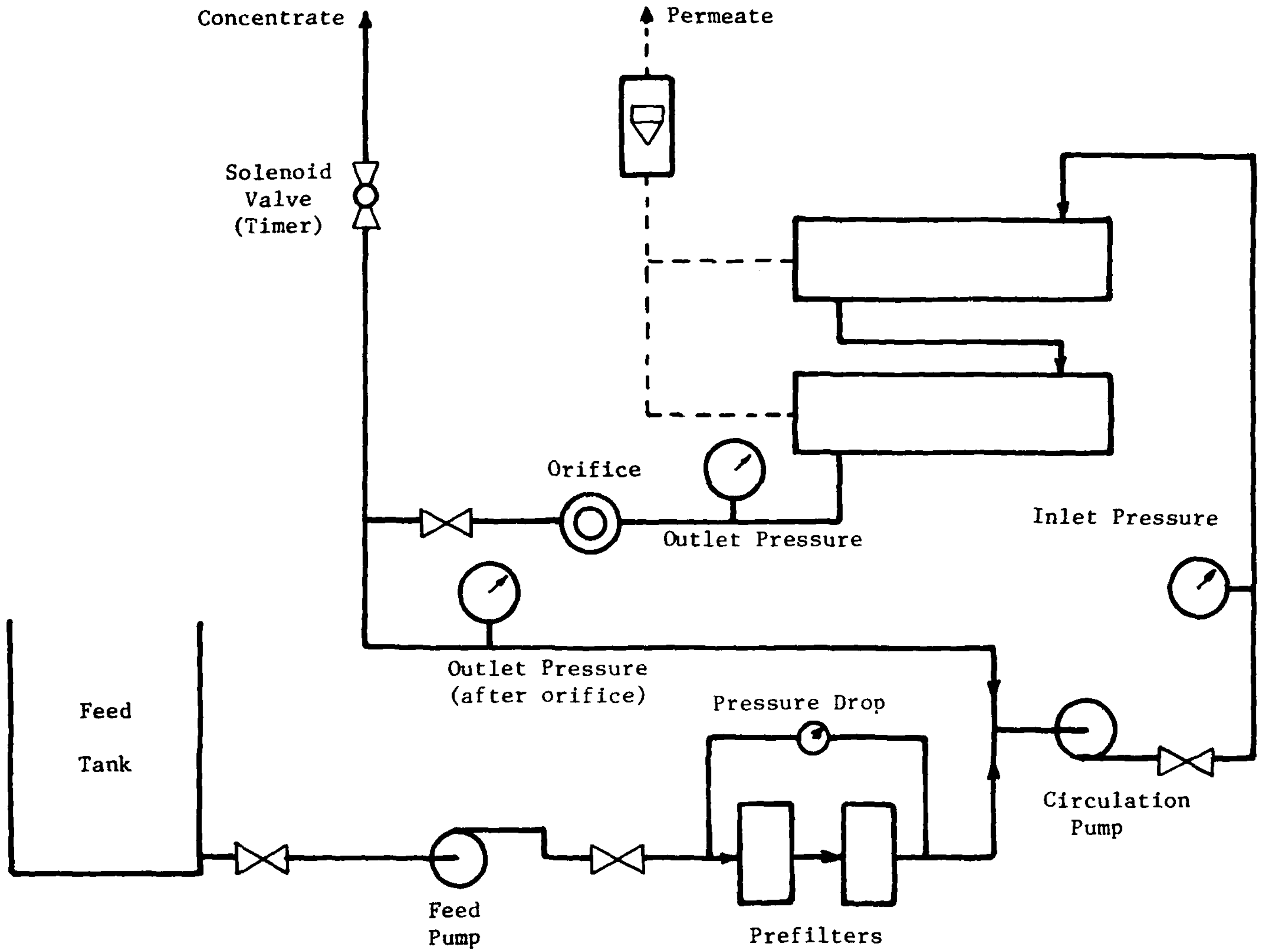


Figure 11. System IV flow chart.

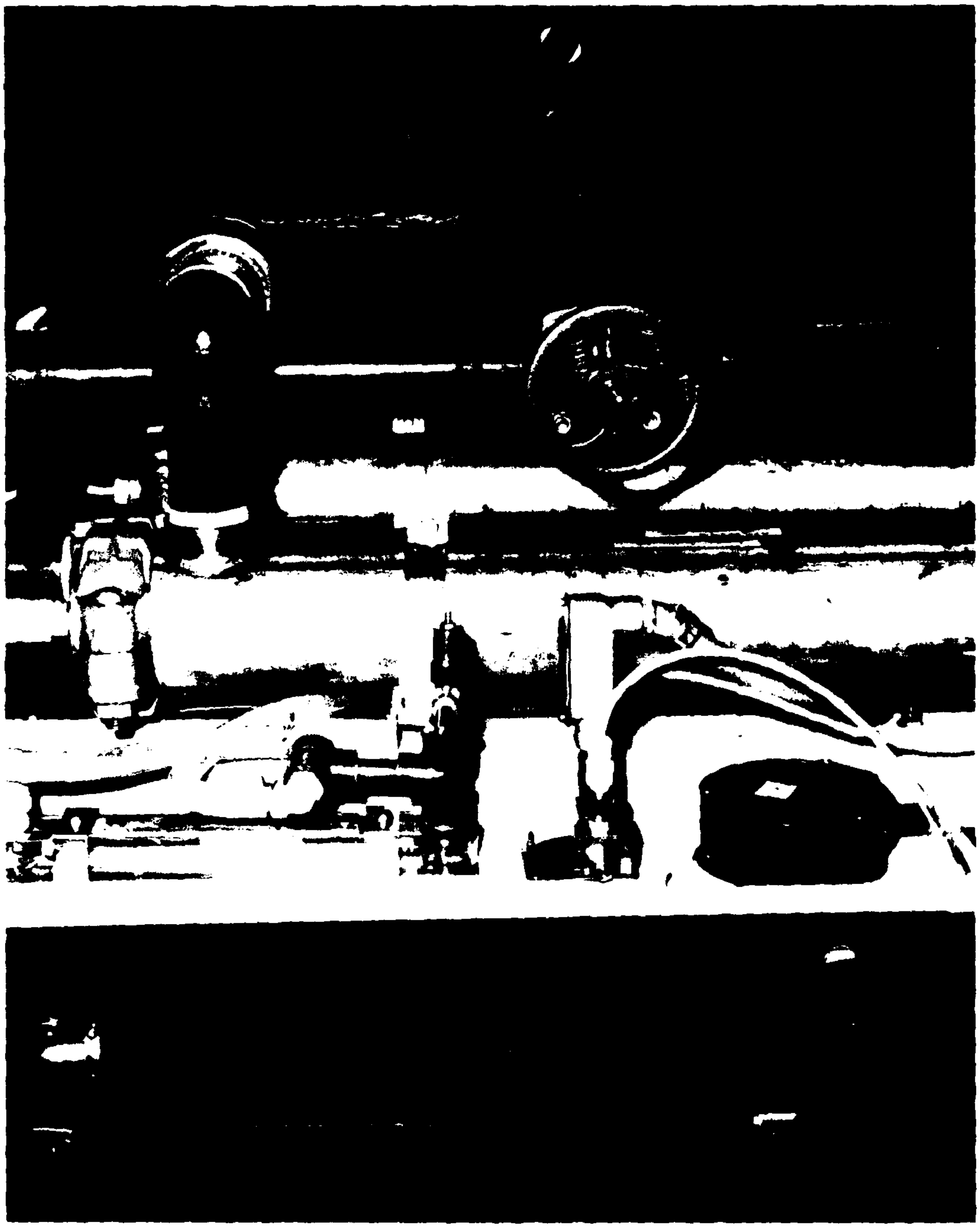


Figure 12. System IV spiral-wound modules connected in series.

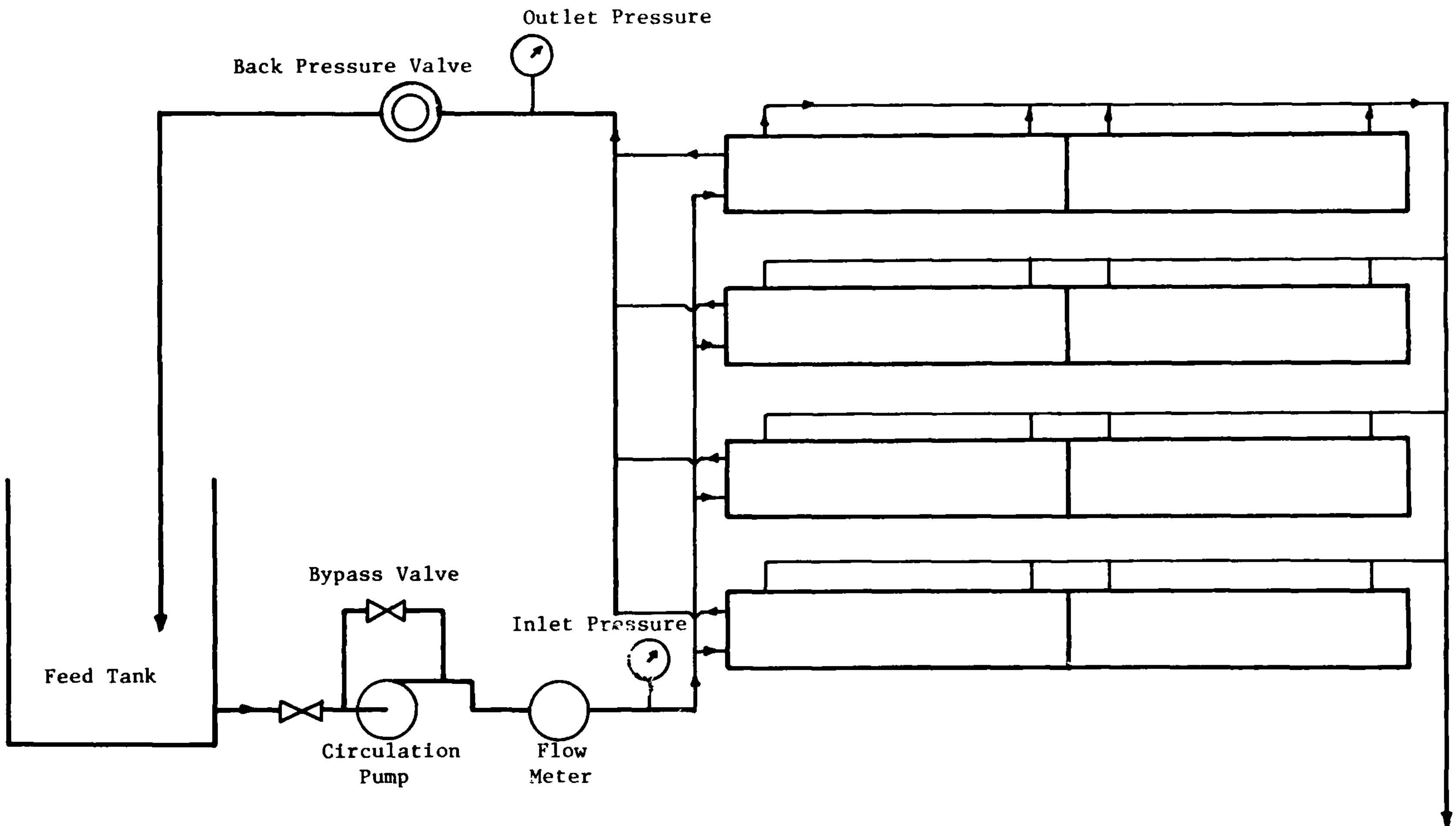


Figure 13. System V flow chart.

Figure 14. System V — tubular sand log configuration.

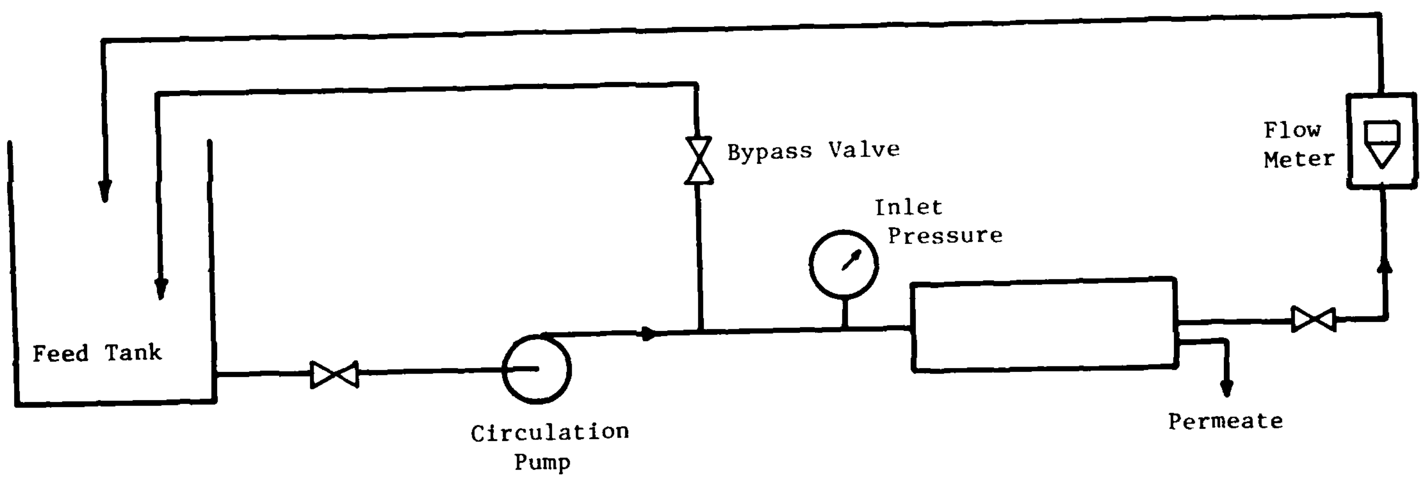


Figure 15. System VI flow chart.

Table 2. Formulation of Synthetic Shower Wastewater

Item	Quantity mg/l	g/180 gal	g/7000 gal
Soap	33	22.5	875
NaCl	40	27.1	1054
Urea	0.5	0.33	12.8
Kaolin	9.1	6.21	242
Talc	9.4	6.42	250
Shower Cleaner	48	32.6	1268
Hair	4.8	3.3	128
Hair Oil	75	51	1983
Hair Gel	18	12.1	471
Shampoo	2.4	1.64	64
Toothpaste	18	12.1	471
Deodorant	0.5	0.33	12.8
DEET	0.5	0.33	12.8
Mouthwash	1	0.64	25
Phisohex	1.5	0.98	38
Hair Dye	0.5	0.33	12.8
Hair Coloring	0.5	0.33	12.8

Table 3. Mean Water Quality of Synthetic and Actual Shower Wastewater

Characteristic*	Actual Shower Waste Water Processed	Ft. Belvoir Tap Water	Synthetic Water MUST Formulation	Proposed Water Reuse Standards	Actual Shower Water Supplied by Coast Guard (12 Samples)
Alkalinity	82	40	39	60	85
Hardness	151	84	92	50	21
TDS	140	95	158	1000	167
Turbidity	64	3.2	39	5	124
pH (units)	5.6	5.8	6.0	6.0–6.8	7.0
Chloride	20.3	–**	35	1000	30
TOC	63	4	50	–	75
COD	–	22	192	–	563
Color (units)	165	35	113	5	408
LAS	1.2	–	0.37	0.5	7.1
TSS	82	–	67	–	115
Barium	ND	–	1.4	1.0	ND
Boron	2.0	–	1.03	30	0.6
Cyanide	–	–	ND	0.2	ND
Fluoride	0.83	–	0.96	1.5	0.53
Manganese	–	–	–	0.2	ND
Cadmium	–	ND	ND	0.05	0.04
Chromium	ND	ND	ND	0.05	0.01
Copper	0.01	ND	ND	0.5	0.32
Iron	1.67	.58	1.8	0.2	1.08
Lead	–	ND	0.01	0.1	0.15
Nitrates	–	–	–	45	14.1
Nitrites	–	–	–	2	0.8
Urea	–	–	–	50	16.7

* Units are mg/l unless otherwise noted
** Not analyzed
ND – not detected

The variability of the tap water which, in turn, causes changes in the synthetic water used as feed must be considered. Two characteristics showing the widest variability were alkalinity with values ranging from 2 to 68 mg/l and hardness with values ranging from 55 to 158 mg/l. This can be related to the alum dosage used at the water treatment plant from which Fort Belvoir gets its tap water. The other characteristics showing variability were iron ND (not detected) to 1.8 mg/l and copper ND to 0.22 mg/l. This is attributable to the age and condition of the Fort Belvoir water delivery system.

Comparison of the mean chemical characteristics for the two feed waters shows little variability between most of the chemical characteristics considered. Alkalinity, hardness, turbidity color, LAS, and barium are the characteristics showing the greatest variability. The effect of tap water variability could explain the difference in alkalinity and hardness. The variability of turbidity might be attributable to different particulate size or makeup rather than quantity. This is demonstrated by the low variability in total suspended solids which is a direct measure of the solids present. Turbidity on the other hand is an optical measurement of refracted light. This turbidity difference could also affect the apparent color, since even a slight turbidity causes the apparent color to be noticeably higher than the true waste solution color. Barium was found in the synthetic water but was not detected in any of the actual waste samples.

A factor known to cause differences in ultrafiltration performances is particulate size and composition. Whether or not the turbidity difference observed between the synthetic waste and the real waste could cause a difference in system performance is not known. Another factor that might affect system performance is the type of organic material rejected by the membrane. This was not considered in this study. Therefore, because of the turbidity variability and the unknown organic makeup of the wastewaters, the question of whether or not system performance on real shower water would be duplicated by the utilization of the MUST formulation for synthetic wastewater cannot be answered.

PROCEDURES

Initial system operational parameters were based on manufacturers' recommendations. However, the limited data base for ultrafiltration operation on shower wastewater made optimization necessary. The area of membrane cleaning seemed to require the most work. Not only was it necessary to determine the cleaning frequency but also the cleaning technique most effective for each membrane configuration.

Daily sampling was done by composing bihourly grab samples of water entering the membrane (brine) and permeate (product). Synthetic feed water was made up in 7000-gallon batches, each of which was sampled at least once. In addition, bihourly grab samples of brine and product were taken and analyzed for turbidity and TOC. Various samples of shower water, generated on board US Coast Guard vessels, were taken by the Coast Guard and analyzed by MERADCOM to compare the waters that the systems actually operated on with shower wastewater from onboard Coast Guard vessels.

RESULTS

Table 4 gives a statistical summary of permeate production data and permeate water quality. A summary of system operation is given in Table 5. The mean permeate water quality for each system and proposed water quality standards for renovated shower water are shown in Table 6. Table 7 summarizes parameters for each UF system. Table 8 shows the statistical values for permeate production of each system. Table 9 gives the complete analysis of shower wastewater samples supplied by the Coast Guard. Graphs of daily flux data for each run and system are given in Appendixes A through F.

DISCUSSION

Generally speaking, the mean permeate chemical parameter data for all of the systems (shown in Table 6) is less than the proposed standards for renovated waste shower water to be used as laundry water. The exceptions are: hardness, barium, and nitrates, with some of the systems exceeding standards for alkalinity, pH, color, and iron. The variability of alkalinity, color, and iron have been discussed previously under wastewater description and are attributable to tap water properties. The pH of the water should be no problem since pH control is achieved easily. Barium was detected only in the synthetic water with none detected in the real shower water or the samples furnished by the Coast Guard. The effects of hardness and nitrate concentrations exceeding proposed standards might be reevaluated, and if found not to be critical, proposed standards could be raised.

A parameter showing variation from system to system is TOC. This parameter appears to be a function of unit operation and related to membrane flux with systems showing higher fluxes also showing higher TOC concentrations in their permeate. This can be explained by membranes that show higher permeation rates pass more carbonaceous material.

UNIT OPERATION

System I, feed-and-bleed system of tubular membrane configuration, was operated at 95 to 98 percent recovery. Inasmuch as the system is of the feed-and-bleed design, the water in contact with the membrane was essentially of constant concentration.

Daily pressure gradients were run for the determination of a theoretical optimum operating pressure. Pressure was plotted against permeate production as shown in Graphs 1 through 4 in Appendix A. A linear increase in permeate production would be observed as the pressure increased to the optimum value, at which point the increase would become nonlinear. Graph 1 shows a linear increase in permeate production for a pressure increase from 40 to 70 lb/in^{2}g, therefore indicating the optimum operating pressure to be greater than 70 lb/in^{2}g. The results of Run 1, as presented in Graphs 5 and 6 of Appendix A, show actual temperature-corrected flux versus time for the three operating pressures of 50, 60, and 70 lb/in^{2}g and corrected flux for 60 lb/in^{2}g, respectively. The vertical lines on the graphs represent the two times when the operating pressure was changed, at approximately 19 hours and 42 hours.

Table 4. Statistical Values Permeate Production

System	Run No.	Duration (Hours)	Duration (Days)	n*	Corrected Flux (GFD) Temp 77° F $\overline{X}$	Corrected Flux (GFD) Temp 77° F Pressure 60 lb/in^{2}g $\overline{X}$	Range
I	1	69	10	56	55.8	60.0	118-29
	2	48	9	41	44.5	44.6	77-20
	3	23	5	19	48.0	48.0	77-25
	4	64	11	46	23.9	36.0	60-24
	5	33	6	36	40.4	60.9	100-30
	6	9	2	9	40.6	60.3	83-48
					Observed Flux (gal/ft^2/d)	Corrected Flux (gal/ft^2/d) Temp 77° F	
II	1	498	51	322	20.5	21.7	60-6
	2	50	8	54	27.8	27.3	58-16
	3	75	11	76	18.0	16.2	63-6
	4	29	5	32	28.8		
III	1	269	36	159	4.7	5.3	30-2
	2	222	27	193	27.4	26.4	75-12
	3	82	12	77	29.0	17.0	56-12
IV	1	265	27	195	10.1	8.0	26-3
	2	87	11	79	9.5	7.2	15-6
	3	69	12	74	8.1	5.0	17-3
	4	118	18	129	21.3	22.9	150-10
V		121	20	131	33.4	40.5	190-15
VI	1	33	6	26	11.9		
	2	54	8	57	8.1		
	3	79	9	16	7.8		
	4	156	9	62	2.1		

* Number of observations.

Table 5. Summary of System Operation

System	Run Number	Duration (hours)	Water Type	Comments	General Comments
I	1	69	Synthetic	System operated at 50, 60, and 70 lb/in², respectively	Membranes cleaned after each run
	2	48	Synthetic	System operated at 60 lb/in²	
	3	23	Synthetic	System operated at 60 lb/in²	
	4	64	Synthetic	System operated at 40 lb/in²	
	5	33	Synthetic	System operated at 40 lb/in²	
	6	9	Synthetic	System operated at 40 lb/in²	
II	1	498	Synthetic	Various cleaning solutions tried during run	Membrane stack disassembled and cleaned between each run
	2	50	Synthetic		
	3	75	Real	Water not prefiltered	
	4	29	Real	Prefilter installed	
III					
0.025-in. channel	1	269	Synthetic	Various cleaning solutions tried during run	Daily backflush
0.045-in. channel	2	222	Synthetic	No cleaning solutions used	Daily backflush and flow reversal
0.045-in. channel	3	82	Real	No cleaning solutions used	Daily backflush, flow reversal, and recycle
IV					
Corrugated spacer	1	265	Synthetic	Various cleaning solutions and techniques tried during run	Extensive high-temperature cleaning with enzyme detergent between each run
	2	87	Synthetic		
	3	69	Real		
Vexar spacer	4	118	Synthetic	No cleaning during run	
V	1	121	Synthetic	Cleaning solution pumped through system after 100 hours operation	
VI	1	33	Synthetic	System operated at 50 lb/in²	Extensive high-temperature cleaning with enzyme detergent between each run
	2	51	Synthetic	System operated at 150 lb/in²	
	3	80	Synthetic	System operated at 150 lb/in²	
	4	160	Sinthetic	System operated at 150 lb/in²	

Table 6. Mean Permeate Water Quality and Proposed Water Quality Standards for *Renovated Shower Water to be Used as Laundry Water**

Characteristic	System I		System II		System III		System IV		System V		System VI		*
	$\bar{X}$	S_x	$\bar{X}$	S_x	$\bar{X}$	S_x	$\bar{X}$	S_x	$\bar{X}$	S_x	$\bar{X}$	S_x	
Alkalinity	34.0	4.8	47	15.6	28	3.06	68	17.24	**	–	17.14	10.64	60
Hardness	91.0	4.8	75	17.4	73	19.4	135	6.46	–	–	76.54	20.42	50
TDS	165	10.4	160	10.0	133	8.02	157	6.23	–	–	137	23.60	1000
Turbidity	1.9	2.4	2.6	1.9	0.5	0.4	2.1	8.1	0.3	0.3	3.0	4.0	5
pH	7.0	0.3	6.8	.07	6.3	0.36	5.9	0.15	–	–	6.7	0.37	6.0–6.8
Chloride	41	1.0	38.5	3.54	39.5	8.1	36.5	2.12	–	–	35.4	2.3	1000
TOC	33	3	10	2	8	2	21	7	27	13	8	6	–
COD	106	14.9	–	–	132	69.8	–	–	–	–	79.6	21.9	–
Color	0.7	2.2	–	–	2.8	4.8	13	9.08	–	–	16.5	32.3	5
LAS	0.11	0.08	0.28	0.14	.07	.02	.06	.02	–	–	.09	.06	0.5
TSS	–	–	10	2.8	1.6	1.6	.4	.89	–	–	7.29	7.95	–
Barium	1.8	0.82	2.7	1.7	2.6	.98	4	2.35	–	–	2.85	2.03	1.0
Boron	1.1	0.28	0.10	.14	.65	.31	5.4	1.14	–	–	.47	.39	30
Cyanide	–	–	–	–	–	–	–	–	–	–	–	–	0.2
Fluoride	1.1	0.32	1.0	0.07	.80	.15	–	–	–	–	.75	.17	1.5
Maganese	–	–	–	–	–	–		–	–	–	–	–	0.2
Cadmium	ND	–	–	–	–	–		–	–	–	–	–	0.05*
Chromium	0.01	0.03	.01	0.02	.01	.00	ND	–	–	–	.01	.00	0.05
Copper	ND	–	ND	–	ND		ND	–	–	–	.03	.05	0.5
Iron	0.5	0.7	.16	0.10	.18	.43	.38	.35	–	–	.37	.50	0.2
Lead	–	–	.02	.02	.02	.03	–	–	–	–	.03	.02	0.1
Nitrates	–	–	7.3	2.95	5.9	0.61	–	–	–	–	6.95	.61	45
Nitrites		–	7.43	4.58	11.55	5.72		–	–	–	9.35	5.29	2
Urea		–	–	–	–	–	–	–	–	–	–	–	50

* Proposed standards

** Not analyzed

Table 7. Summary of Parameters for Ultrafiltration Systems

System Parameter	System Number							
	I	II	III		IV		V	VI
Membrane Configuration	Porous Carbon Tubes	Plate and Frame	Hollow Fiber 0.025-in.	0.045-in.	Spiral Wound Corr.	Vexar	Tubular	Spiral Wound Corr.
Membrane Area (ft²)	78	22.5	30	15	60	60	75	60
System Operating Pressure (lb/in² g)	40-70	25-30	25-30	25-30	70	60	150-170	50 & 150
Circulation Rates (gal/min)	300	40	20	16	70	10-15	19-20	9-13
Maximum Operating Temperature ° F	200	110	123	123	120	140	140	–
Power Consumption Kilowatts PF=1.0*	2.66	2.95	1.44	1.44	3.90	2.88	4.32	–
Mean Permeate Production (gal/h)	169	22	7	17	18	57	127	16
Duration of Test (h)	257	684	587		648		121	346
Actual Production Time (h)	246	652	573		539		121	322
Service and/or Cleaning (h)	11	32	14**		109		0	24
% Uptime	96	95	98**		83		100	93
Mean Power Consumption per 100 gal Permeate (kWh)/100 gal	1.57	13.41	20.57	8.47	21.67	5.05	3.40	–
Operational Mode	Feed & Bleed	Batch	Batch		Feed & Bleed		Batch	Batch
% Recovery	98	85	85		95-98		85	85
Mean TOC Removal (%)	34	80	80	84	72	40	46	84
Mean Turbidity Removal (%)	96	95	98	99	94	98	99	94
Membrane Material	Inorganic	Vinyl	Noncellulosic		Noncellulosic		Modified Cellulosic	Noncellulosic
System Weight	–	525 lb	500 lb		1035 lb		1060 lb	–
System Size	30"x54"x84"	55"x31"x65"	45"x29"x45"		49"x40"x82"		118"x24"x34" or*** 24"x34"x118"	

* A power factor of 1.0 was assumed for comparison.
** Does not include backflushing time.
*** Can be installed horizontally or vertically.

Table 8. Statistical Summary of System Operation

System	Hours of Operation	Mean Temperature Corrected Flux 77° F (GFD)		Mean Permeate Total Organic Carbon (mg/l)*			Mean Permeate Turbidity (JTU)*		
		$\bar{X}$	n	$\bar{X}$	n	S_x	$\bar{X}$	n	S_x
System I	246	51.6**	207	33	4	3	1.9	35	2.4
System II	652	23.5	484	10	47	2	2.6	47	1.9
System III									
0.025-in. channel	269	5.3	159	10	4	2	0.8	6	0.5
0.045-in. channel	304	26.5	270	8	25	2	0.5	25	0.4
System IV									
Corrugated spacer	421	7.2	348	14	31	8	3.1	31	3.0
Vexar spacer	118	22.9	129	30	30	7	0.8	30	13.3
System V	121	40.5	131	27	31	13	0.3	31	0.3
System VI	324	6.4	161	8	25	6	3.0	25	4.0

* Daily composite samples.
** Also pressure-corrected to 60 lb/in^{2}g.
$\bar{X}$ Mean
n Number of observations.
S_x Standard deviation.

Table 9. Laboratory Analysis of Shower Wastewater From Coast Guard Vessels (12 Samples)

Characteristic*	A-1	A-2	A-3	A-4	A-5	A	B	A	B	C	D	1	$\bar{X}$	S_x
Alkalinity	52	48	34	32	104	50	36	62	54	412	108	28	85	106
Hardness	4	6	4	8	16	46	28	20	28	40	28	18	21	14
TDS	100	98	50	68	240	145	100	120	120	600	170	190	167	145
Turbidity	240	240	73	74	470	75	95	48	33	70	30	43	124	131
pH (units)	6.6	6.6	6.4	6.3	7.7	6.3	6.3	6.2	6.3	9.2	9.2	6.9	7.0	1.1
Chloride	18.5	20.0	14.0	23.5	115.0	26.0	26.5	21	17.5	6	19.5	52	30	29
TOC	85	88	23	44	200	38	42	73	85	132	60	28	75	50
COD	640	640	180	320	1500	280	320	550	–**	980	470	310	563	385
Color (units)	990	960	200	225	1400	390	425	35	40	45	30	150	408	459
LAS	3	3	1	1	1	4	3	11.7	10	20	18	9	7.1	6.7
Susp Solids	120	146	40	50	332	84	80	108	–	172	92	40	115	83
Barium	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	1	–	–
Boron	0.4	0.9	0.2	0.6	0.4	1.0	0.7	0.6	0.3	0.8	1.2	0.5	0.6	0.3
Cyanide	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	ND	–	–
Fluoride	ND	–	–	–	–	0.7	0.65	0.5	0.65	0.8	0.95	ND	0.53	0.35
Mangenese	–	–	–	–	–	ND	ND	ND	–	0.03	ND	ND	–	–
Cadmium	0.09	0.08	0.08	0.11	0.09	0.02	ND	ND	ND	ND	ND	ND	0.04	0.05
Chromium	ND	ND	ND	ND	ND	ND	ND	0.04	0.04	0.05	0.03	ND	0.01	0.02
Copper	0.43	0.30	0.13	0.26	0.35	0.57	0.36	0.26	0.43	0.34	0.16	0.22	0.32	0.12
Iron	1.61	1.31	1.40	0.74	0.79	1.31	0.68	0.77	1.75	1.59	0.88	0.61	1.08	0.37
Lead	0.18	0.21	0.13	0.15	0.18	0.12	0.09	0.14	0.17	0.19	0.05	0.15	0.15	0.04
Nitrates	20.5	17.0	7.4	9.6	38	17.0	12.5	8.7	5.5	12.4	6.2	14.5	14.1	8.9
Nitrites	1.0	1.0	ND	ND	3.0	ND	ND	ND	–	–	–	2.4	0.8	1.2
Urea	15	15	12	45	5.4	15	12	–	–	–	–	14.5	16.7	11.9
Conductivity	165	165	84	116	380	170	165	170	–	525	180	300	220	130

* All values in mg/l unless otherwise noted.
** Not analyzed.
ND – Not detected.

The results, as shown on Graph 1, indicate that as the pressure was raised from 50 to 60 lb/in^{2}g and then to 70 lb/in^{2}g the actual flux did, indeed, increase initially; however, within one day operating time period, the actual flux had decreased to the operating range observed at the lower pressure. Considering the greater power consumption, and the little apparent increase in flux, by operating at higher pressures the theoretical optimum pressure as determined by the pressure gradient is not the optimum operating pressure.

Following the conclusion of Run 1, the system was washed. The wash solution consisted of 1 pound nonionic detergent and 1 pound oxalic acid in 30 gallons of water at room temperature. The procedure included rinsing the system with tap water, followed by a 1-hour wash with the cleaning solution, followed by another rinse.

Results of Runs 1 and 2 are shown on Graph 7, Appendix A; the pressure gradients are shown on Graphs 1 and 2, Appendix A. An operating pressure of 60 lb/in^{2}g was chosen for these two runs. The pressure gradients for Run 2 were nearly linear from 40 to 70 lb/in^{2}g; however, for Run 3 a slight decrease was observed above 60 lb/in^{2}g. Graph 7 shows that the rate of flux decline is greater for Run 2 than for Run 1 and much greater for Run 3 than for Run 2. Also apparent is an amount of irreversible fouling, since initial operating fluxes are not obtained after the washes separating the runs.

Toward the end of Run 2, a quick wash was tried where a 120° F oxalic solution was pumped through the system for ½ hour. As seen from Graph 7, this caused an increase in permeate production; however, a rapid decrease followed, indicating insufficient cleaning. At the end of Run 2 the system was washed first with oxalic acid at 90° F, then with nonionic detergent at 150° F for ½ hour each. The modules were allowed to soak in the detergent overnight. The following morning, the cleaning solution was recirculated for an additional ½ hour, followed by a tap water rinse. The rapid flux decline during Run 3 shows that the wash process was probably inadequate or that the operating pressure was too high. The wash following Run 3 consisted of a 150° F rinse with tap water. This was followed by a 2-hour 150° F wash with detergent and a final rinse with tap water.

Run 4, shown in Graphs 2, 8, and 9, was at 40 lb/in^{2}g. Graph 2 shows the pressure gradient, Graph 8 shows the actual flux, and Graph 9 shows the pressure-corrected flux. Pressure gradients for this run remained linear through 60 lb/in^{2}g. The lower operating pressure slowed the rate of flux decline; however, the initial operating flux of the system was not restored. This is the first run that continued long enough for the system to stabilize as far as permeate production. This statilized temperature-corrected flux was between 18 and 19 gal/ft^2/d.

The wash following this run appeared to be most effective; complete membrane regeneration was observed by the high permeate values observed after the wash. The wash consisted of a hot water rinse, followed by ½-hour cleaning with a 145° F nonionic detergent solution. The system was again rinsed with hot water and acid washed for 15 minutes with 150° F solution of oxalic acid and sulfuric acid. After this acid solution was rinsed out, the system was placed back on-line.

Data for Runs 5 and 6 are shown on Graphs 4, 10, and 11. Once again, a linear temperature gradient was observed. The rate of the flux decline for Run 5 was similar to that of Runs 1 and 4 and leveled off at around 18 to 19 gal/ft^2/d. As seen from Graphs 10 and 11, the amount of permeate recovery between Runs 5 and 6 was not as high as between Runs 4 and 5. The only difference in the washes was that the temperature of the latter was about 10° F lower than the earlier one. This, along with earlier data, shows permeate recovery to be a function of the wash water temperature of this system.

The nature of the membrane itself for this system makes it possible to use almost any strength cleaning solution desired. The inorganic membrane is resistant to high-temperature acid and caustic solutions, much more so than an organic-type membrane as used in the other systems. With further experimentation, a cleaning step might be developed which would completely rejuvenate the membrane in a short period of time as compared to the one found which would take about 2 hours.

System II, of plate-and-frame membrane configuration, was operated in a batch mode of operation. Initially, the feed tank was filled with a volume of water. The system was then operated until 85 percent of the water had been treated; therefore, the membranes were washed constantly with water of different contaminant concentration. Typical values of the TOC and turbidity for different degrees of concentration are shown in Table 10.

Table 10. Typical Feed Values of TOC and Turbidity at Different Degrees of Concentration for Batch Mode Operation

Volume Reduction (%)	Turbidity (JTU)	TOC (mg/l)
0	18	13
10	15	13
20	23	18
30	9	17
40	37	18
50	15	17
60	16	17
70	16	20
80	38	29

Table 10 shows that the concentration of contaminants in the feed changes; however, these concentrations would be expected to increase as the waste was reduced in volume. This is not the case, however, which suggests that biological action was taking place in the feed tank as the waste became more concentrated.

Run 1 permeate data are shown on Graphs 1 and 2, Appendix B, where actual flux values are shown in Graph 1 and temperature-corrected flux values are shown in Graph 2. The three vertical lines on the graph represent the times when the system was washed. The four arrowheads before the first wash represent times when the feed tank was dumped (the concentration of the feed solution changed from 85 percent to unconcentrated shower water).

The graphs show that the change in concentration of the feed solution has little effect on the rate of permeate production. The upward variation in permeate production early in the run correlates with times when the unit was shut down. Through a 1-day operation, the flux values are normally high after startup, followed by a slow decline for the rest of the day.

An increase in permeate production was observed after the washes; however, a complete return to initial permeability was not observed. The first wash consisted of an anionic detergent cleaning solution, followed by a tap water rinse. Graphs 1 and 2 show that when the unit was first placed back on line, only a slight recovery in permeate production was observed; however, permeate production continued to increase after the unit was placed back on-line. This indicates that the detergent itself causes some sort of membrane fouling and because of its anionic charge, the detergent is retained by the membrane for an extended period of time.

The second wash consisted of the same detergent wash. This time the detergent was allowed to soak overnight. The following day the system was rinsed with tap water for 3 hours. As seen from the graphs, this rinsing was adequate; however, permeate production recovery was not increased over that observed in the first wash. The same was true for the third wash, in which rinse time was decreased to 1½ hours. After the third wash, extended operation showed the flux leveling off from 12 to 8 gal/ft^2/d.

After the conclusion of Run 1, the membrane stack was disassembled in order to observe how the membranes were fouling. Figure 3 shows the fouling material to be a gelatinous clay material. The material was removed physically from membranes and they were rinsed in tap water. The stack was reassembled and Run 2 was begun.

Graphs 3 and 4, Appendix B, show the flux data for Run 2 where Graph 3 is the observed flux versus time, and Graph 4 is the temperature-corrected flux versus time. Initial flux values were nearly achieved after this cleaning procedure. The rate of flux decline appeared to be about the same as for Run 1, when the different ordinate scales between the graphs for Run 1 and Run 2 are considered. The system was not washed during Run 2. At the conclusion of the run, the system was once again cleaned physically.

Graphs 5 and 6, Appendix B, show the results of Run 3. Once again, initial permeate values were obtained. This run used real shower water as feed. During the early portion of the run, the rate of flux decline appeared to be nearly the same as Runs 1 and 2. However, later flux decline appeared to be more rapid. When the stack was disassembled after the run, the brine channels appeared to be restricted by hair. Because of this, pretreatment in the form of a bag filter was installed before beginning Run 4.

A step-wise regression[1] was performed on these data using flux as the dependent variable and time, temperature, brine flow, pressure drop across the membranes stack, and operating pressure as the independent variables. The program was run in such a way that the independent variables could be entered or removed and the resultant R values would be calculated, where R is the percent variation in flux accounted for by that variable or combination of variables entered or removed.

As would be expected, time was the single most significant independent variable with an R coefficient of correlation value of 74 percent. Adding to this brine flow gave an R value of 78 percent, and adding pressure drop gave an R of 81 percent. When operating pressure was included, R was 90 percent. The interesting point is that the two most significant variables are brine flow and pressure drop, which would be a direct result of the brine channel restriction observed.

[1] Freund, T. E.: "Modern Elementary Statistics." Fourth Edition; Prentice-Hall, Inc.; Englewood Cliffs, NJ (1973).

Run 4 covered about 30 hours of operation. Its purpose was to evaluate screening as pretreatment for hair removal. Constant brine flow for the run indicated that screening was sufficient pretreatment for elimination of the brine flow restriction by hair.

System III, of hollow-fiber membrane configuration, was operated in a batch mode similar to System II. Therefore, Table 6 would also apply to the feed tank of this system. Initial system operation was operated as the system was supplied by the manufacturer.

Run 1 data, with the 0.020-inch-diameter tubes using synthetic waste as feed, are shown in Graphs 1 and 2, Appendix C, where Graph 1 is the observed flux and Graph 2 is the temperature-corrected flux. As seen from the graphs, initial permeate production was low and the rate of decline was rapid. The initial brine flow was low and also declined rapidly. The membrane cartridge was removed and examined. Fouling of the brine channels at their entrances was observed.

Modifications to the system were then begun. Initially, the membrane cartridge was turned causing brine flow to be reversed through the channels. This helped keep the header clean; however, the brine flow was still low. Cleaning solutions were tried, since the inside of the fibers appeared to be fouled. This helped somewhat, but brine flow was still low.

Finally, cleaning solution was backflushed through the membrane and the unit was repiped so that brine flow could be reversed in the cartridge. An additional bag filter of 150 mesh was installed in the feed line. Since flux values were still unacceptable (3 to 6 gal/ft^2/d) and brine flow was still low, a cartridge containing 0.045-inch-diameter fibers was installed for Run 2.

Graphs 3 and 4, Appendix C, show the results of Run 2 using synthetic water as feed. The unit was backflushed daily or after 24 hours during continuous operation. After backflushing, flow through the cartridge was reversed. At 90, 145, 175, and 220 hours the bag filter was cleaned, which caused an increase in flux due to high module flow rate. The times of backwash and reverse flow are seen on the graphs as the high point in a series of declining flux data points.

No cleaning solution was used during this run. However, since the membrane is chlorine-resistant and chlorine was found during Run 1 to be a somewhat effective cleaning agent, its use in backflush might achieve higher fluxes.

Run 3 used real shower water from a field shower setup. The results are shown in Graphs 5 and 6. Operation for this run consisted of the daily backflushing and flow reversal as in Run 2, plus daily cleaning of the bag filter, plus a 10-minute recycle mode at zero permeate flow. This latter procedure was included to insure removal from the fibers of any fouling material which was loosened by the backflush operation.

The results of this run shown in the graph indicate two areas of membrane permeate production, 33 to 41 gal/ft^2/d and 20 to 15 gal/ft^2/d with rapid change between the areas. One theory for this operation comes from a consideration of the Reynolds number for the flow through the fibers. The theoretical Reynolds number shows the system to be operating in the transition zone between laminar and turbulent flow. A slight change in operational parameters, such as temperature or pressure, could cause a change in the type of flow in the fibers with the end result being an increase or decrease in permeate production. In theory, during turbulent flow there would be a better cleaning action of the membrane surface resulting in less membrane fouling and higher permeate production.

A comparison of the mean permeate production data between the synthetic wastewater and the real wastewater of Runs 2 and 3 show no significant differences in the treatability of the waters. No operational problems were encountered peculiar to the real shower water.

System IV, a spiral-wound membrane configuration, was operated in the feed and bleed mode of operation. Typical concentration ranges for the system were from 95 to 98 percent.

Run 1 used synthetic shower water as feed, and permeate data are shown in Graphs 1 and 2, Appendix D. Various washes were tried throughout the run in search of a suitable cleaning technique. The first wash after about 50 hours of operation resulted in almost complete recovery of permeate production. The wash consisted of a 1-hour wash with hot anionic detergent solution, with the solution allowed to soak overnight, followed by a ½-hour rinse with tap water. Though original permeate production was obtained, the rate of decline in permeate produced was more rapid after this wash than with the new membranes.

The second wash was identical to the first and occurred at approximately 85 hours. The data show that only a slight increase in permeate production, about 5 gal/ft^2/d, was obtained. Furthermore, the rate of flux decline after the wash was very rapid. The modules were disassembled and examined for fouling of the brine channels. They were found to be clear at the headers, so the system was washed. Recovery after this was even less than before. The next cleaning consisted of two 8-hour washes, followed by a thorough rinsing, resulting in little significant membrane rejuvenation. The wash at the run's conclusion consisted of an enzyme detergent wash and overnight soak, followed by a 2-hour rinse. This method gave a significant increase in permeate production.

The data of Run 2, shown as flux versus time, are shown in Appendix D, Graphs 3 and 4. The rate of flux decline as shown in the graphs was still rapid when the difference in ordinate scale is considered. The unit was washed at about 320 hours, with a significant increase in permeate production as shown on the graphs for this run. This wash consisted of an enzyme detergent wash and soak, a citric acid rinse followed by a tap water rinse. Once again, significant permeate recovery was achieved; however, rate of decline was still rapid. At the conclusion of the run, the unit was washed with enzyme detergent, followed by citric acid, followed by a fresh water rinse.

Run 3 data, using real shower water, are shown in Graphs 5 and 6, Appendix D. The data show no significant difference from that obtained on the synthetic water. Washes showed significant improvement in permeate production; however, rate of decline was still rapid. No significant operational problems were encountered with the use of the real water.

At the conclusion of the run, the modules were removed and returned to the manufacturer for evaluation. One of the modules was opened by the manufacturer. The membrane surface was covered with brown slime. An attempt was made to clean the other modules with four separate 1-hour washes. The flux was checked and found to be less than half the original flux. This module was then opened and fou to contain the same brown, fibrous, slimy material covering the membrane.

The fourth run used a different type of spacer material. Instead of the open channel corrugated spacer, the same membrane was used with a turbulence promoting vexar spacer, thereby hopefully eliminating the buildup of fouling material.

The data for Run 4, using synthetic feed water, are shown in Graphs 7 and 8, Appendix D. The graphs show that a much higher flux was obtained with these modules. Furthermore, though the rate of decline was rapid, it appeared to level out at a fixed value at about 12 gal/ft^2/d and held this value for the rest of the run, or about 80 hours. No attempts were made to wash these modules.

System V, of the tubular configuration was operated in the batch mode. Graphs of flux versus time are shown in Graphs 1 and 2, Appendix E. Initial flux values were high, but the rate of decline was rapid with the majority of the run operating in the 30 to 50 gal/ft^2/d permeate production range.

No change in permeate production was observed when the feed tank concentration was changed from its most concentrated feed strength (85 percent) to straight shower water as feed. No attempt was made to wash the system during the run.

This system was operated with no pretreatment of the wastewater. No operational problems were encountered because of this; however, operating time on this system was less than any of the other systems.

System VI, of the spiral-wound configuration, was operated in the batch mode. The flux data for System VI are shown in Appendix F, Graphs 1 through 4, 150 lb/in^2/g. Throughout all runs different cleaning techniques were tried, but to no avail.

The module was removed and examined at the end of the run and found to be plugged with a brown fibrous slime. No pretreatment was used with this module, which could be the reason for this extensive fouling.

GENERAL OBSERVATIONS

Two shower wastewaters were processed during this study: real shower water obtained from a field military shower unit, and synthetic shower water prepared in accordance with the MUST formulation, Table 2. Little difference was observed in permeate water quality and permeate production in processing of the two waters with each test system. It must be remembered, however, that in all cases where real water was processed, synthetic water had been processed extensively first. Therefore, irreversible fouling that had taken place prior to treatment of the real shower water could have been a limiting factor in permeate production as well as the product water quality.

Observation concerning temperature effects should be noted. Not only are higher flux values observed at elevated temperatures, but also the rate of flux decline appears to be lower at higher temperatures. These observations are similar to those made in work done at the David W. Taylor Naval Ship Research & Development Center[2] and in the evaluation of UF systems under the MUST program.[3]

[2] Harris, Lynne R.: "Personnel Communications." David W. Taylor Naval Ship Research and Development Center, Annapolis, MD.

[3] Gollan, A. Z.; "McNulty, K. J.; Goldsmith, R. L.; Kleper, M. H.; Grant, D. C.: "Evaluation of Membrane Separation Processes, Carbon Adsorption, and Ozonation for Treatment of MUST Hospital Wastes." Final Report Contract No. DAMD 17-74-C-4066, AD Number 30057.

Initially, only two systems had pretreatment. During the study, pretreatment was found to be necessary, even with the tubular systems, not only to protect the membrane cartridges from brine channel plugging, but also to keep hair from entangling the pump empellers.

Two modes of operation were used in this study: batch and feed-and-bleed. No significant differences were noted in permeate water quality resulting from either mode of operation. However, power consumption would be less in the feed-and-bleed mode, and the rate of membrane flux decline would be less for batch mode operation. The latter is true because in the batch mode of operation, the membrane only processes the final concentration of waste for a short period of time, whereas in the feed-and-bleed mode, the membrane essentially processes the maximum concentration of waste all of the time. Also elevated temperatures can be obtained by the closed-loop pumping in the feed-and-bleed mode, whereas in the batch mode external heating would be required to achieve elevated temperatures.

Membrane cleaning techniques varied from system to system, depending on membrane formulation. In general, higher temperatures resulted in greater membrane cleaning as observed by higher permeate production values. Anionic detergents were retained by the membranes, causing high TOC values in permeate after cleaning and causing lower permeate output. Greater difficulty in rinsing was also experienced. Chlorine and low pH achieved by acid addition proved effective in membrane cleaning when the membranes could tolerate them. Volume and contaminant levels in cleaning and rinse solutions must be considered in system design.

Tables 6 and 8 show little difference in permeate water quality from system to system. Observed permeate TOC values were higher from systems with higher membrane fluxes, as would be expected due to their larger pore size. Also membranes retaining detergent exhibited higher TOC values after washing, causing higher mean TOC values.

Table 8 shows that the higher mean permeate production was obtained from Systems I and V, both of which were of the tubular configuration. Table 7 shows the better energy effective systems also to be Systems I and V. System V also shows 100 percent uptime and System I, 96 percent uptime. In actual extended operation, however, System I would require a certain percentage of time for cleaning therefore making the uptime percentage lower.

Disadvantages of these systems include their low membrane packing density, as well as their susceptability to breakage by rough handling. Also, their TOC removal was low as compared to the other systems.

The spiral-wound configuration with the corrugated spacer, Systems IV and VI, yielded the lowest permeate production and were the least energy effective. With adequate pretreatment and higher circulation rates, these systems might yield higher permeate production rates as observed with the spiral-wound module with the turbulence promoting vexar spacer.

Advantages of this configuration include high membrane packing density, durability of the modules, and high TOC removal.

Permeate production rates of the remaining systems were similar. The plate-and-frame was the only configuration that could be disassembled and cleaned physically, thereby obtaining complete membrane regeneration. The disadvantage of this system was the need for daily checking of the torque on the membrane stack, meaning that the whole stack must be positioned to be readily accessible.

The hollow-fiber configuration had the unique advantage of membrane backflushing. Generation of large volumes of contaminated waste from membrane cleaning is therefore eliminated. Furthermore, its resistance to chlorine makes chlorine disinfection and cleaning possible. A disadvantage of this system is the high degree of pretreatment necessary for its operation.

The service and/or cleaning time shown in Table 7 represents cleaning time for the membranes and service to the pretreatment portion of the systems. None of the systems experiences any component failures. The amount of service required by the different pretreatment schemes was variable, with the bag filter requiring the most service and the combination of prefilters on System IV requiring the least amount of maintenance.

CONCLUSIONS

Based on the data obtained, this report concludes that:

1. Ultrafiltration is a viable process for the treatment of shower wastewater onboard a watercraft.

2. Because of operational problems likely to be encountered, the tested systems in their present configurations cannot be recommended for installation aboard ship for waste shower water renovation.

3. To protect the system, pretreatment is required to remove hair and other fibers from the feedwater prior to application to the membranes.

4. The hollow-fiber configuration is the only membrane configuration not requiring chemicals for membrane cleaning. This configuration can be cleaned by backflushing with permeate.

5. A system which might prove applicable for shipboard testing can be designed utilizing the best commercially available features of each system studied. This includes hollow-fiber membrane configuration, feed-and-bleed operation, and pretreatment by trapezoidal screen followed by a cartridge filter.

RECOMMENDATIONS

None of the purchased systems would be suitable for direct installation on board a vessel for the treatment of shower wastewater. However, by combining the better features of each system, it is felt that a suitable system could be designed for further testing. This design could meet all criteria for marine use and after extensive land testing, be installed on board a vessel. The system should require minimum operator attention, yet have the capability of monitoring all treatment parameters easily.

The recommended mode of operation is feed-and-bleed, for the reasons discussed earlier. Furthermore, optimization of feed concentration is possible with a system of this design. The size of the circulation loop as well as permeate production rate will determine if a heat exchanger is needed with the system. Another reason is the physical size of the pretreatment system. A smaller pretreatment system is required for treating only makeup water to the circulation loop as compared with treating the total flow through the loop.

Due to waste variability and operational difficulties encountered throughout the study, a pretreatment of the waste is required. The arrangement on System IV (Figure 16) is recommended because it gave trouble-free operation throughout the study. It consisted of a blowdown filter with a spiral-wound element of trapazoidal pore shapes. The filter was cleaned by a sudden release of the driving pressure across the filter. Pressure drop across both filters never exceeded 1 lb/in^2g. These filters were located at the discharge of the feed pump. A strainer, for large items in the feed tank, should also be included to prevent plugging of the piping and feed pump.

The designed system should include a cleaning tank for the membranes. The tank should have the capability of supplying heated wash or backflush water for membrane cleaning. Its location should be such that a separate pump would not be needed.

The grey area for the system design is membrane selection. Two approaches can be taken to this problem, each with its advantages and disadvantages. The first is to pick a membrane configuration and design a system optimized to that configuration. The other approach would be to design a system that could accept modules of any configuration, but give up final optimization of the system.

The problem with the first alternative is which configuration to select. Systems that exhibited lower permeate production values in the study theoretically could be eliminated. However, optimization of each system for shower wastewater was impossible. Therefore, better optimization of a system could mean higher permeate production rates. Eliminating the tubular configuration because of its low packing density or its fragile nature would mean eliminating two systems whose flux was nearly twice that observed for any other system, plus elimination of the two more energy effective systems. Eliminating the plate-and-frame configuration would mean eliminating the only system that could be cleaned physically. Eliminating the spiral-wound module with the vexar spacer would give up the system with the highest packing density. Furthermore, configurations cannot be eliminated because of membranes characteristics, since most membranes are available commercially in a variety of configurations.

If selection of a configuration is necessary, the hollow-fiber configuration would be recommended. The primary reason for this is that the permeate values observed were obtained with no membrane washing. Only daily membrane backflushing was used during the runs, which would eliminate the need for stores of membrane cleaning chemicals. Packing density is high, and the membrane modules are of rugged construction.

The problem with design of a system to accept all of the modules can be seen readily from Table 7. Circulation rates vary from 300 to 8 gal/min, with operating pressures varying from 25 to 150 lb/in^2g. Unless pumps are sized to a particular system, the system cannot be expected to be energy effective. Furthermore, degrees and amounts of backwashing vary from system to system, making optimum sizing impossible.

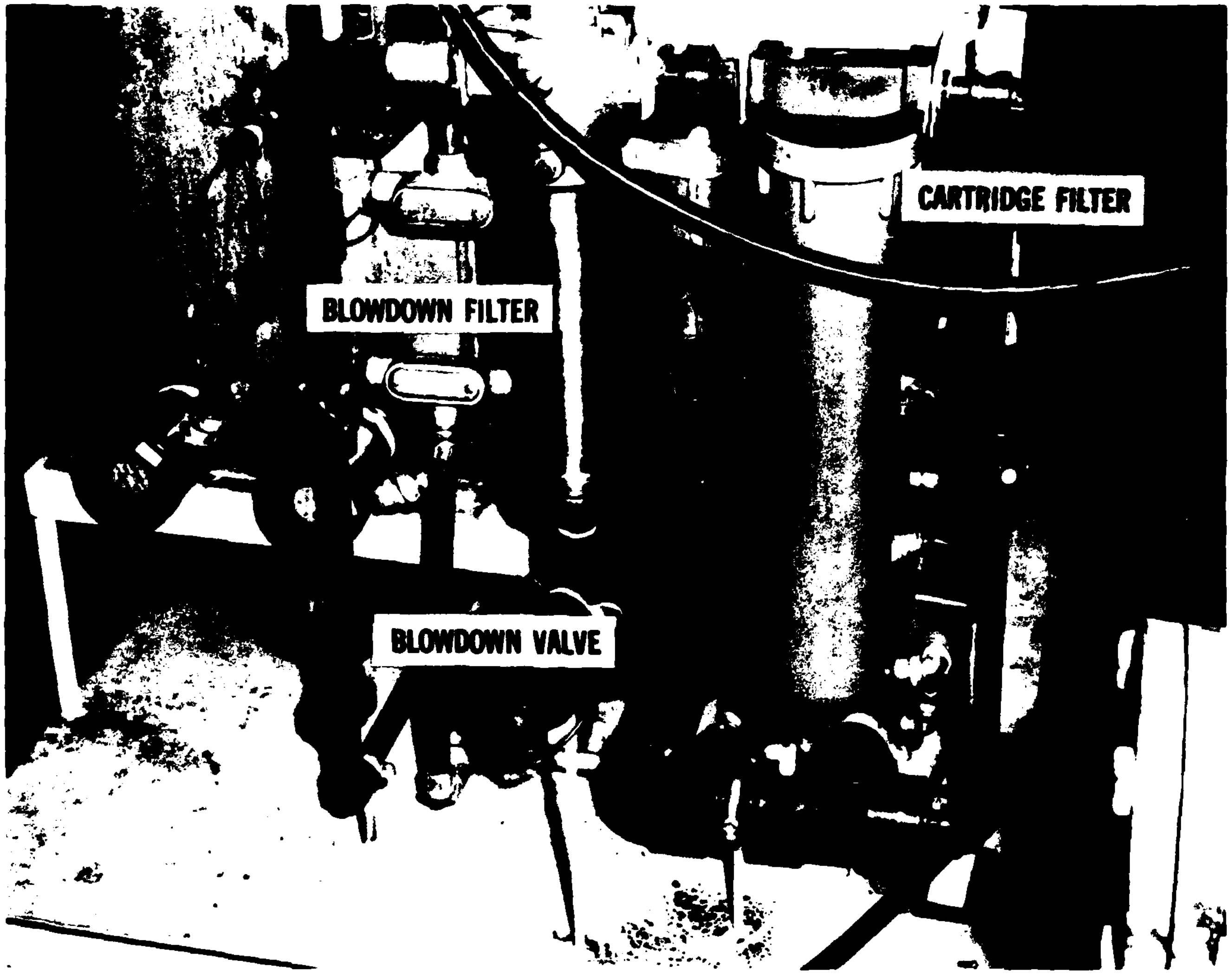
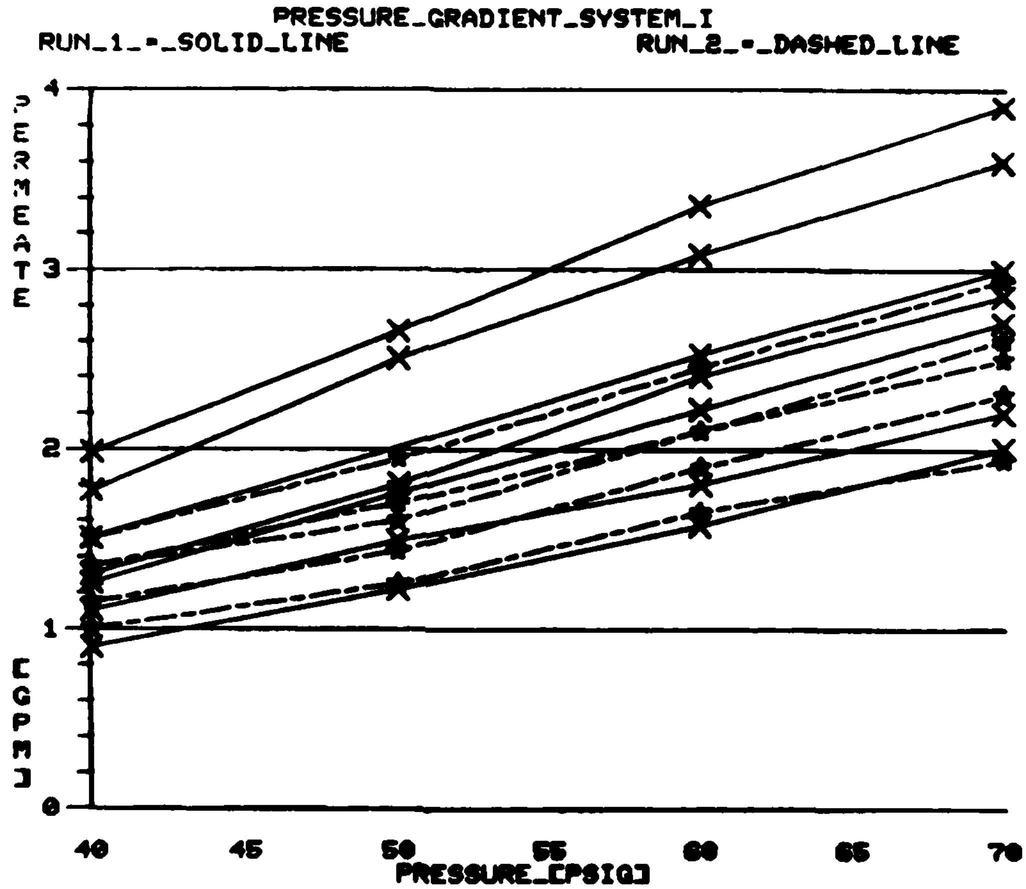
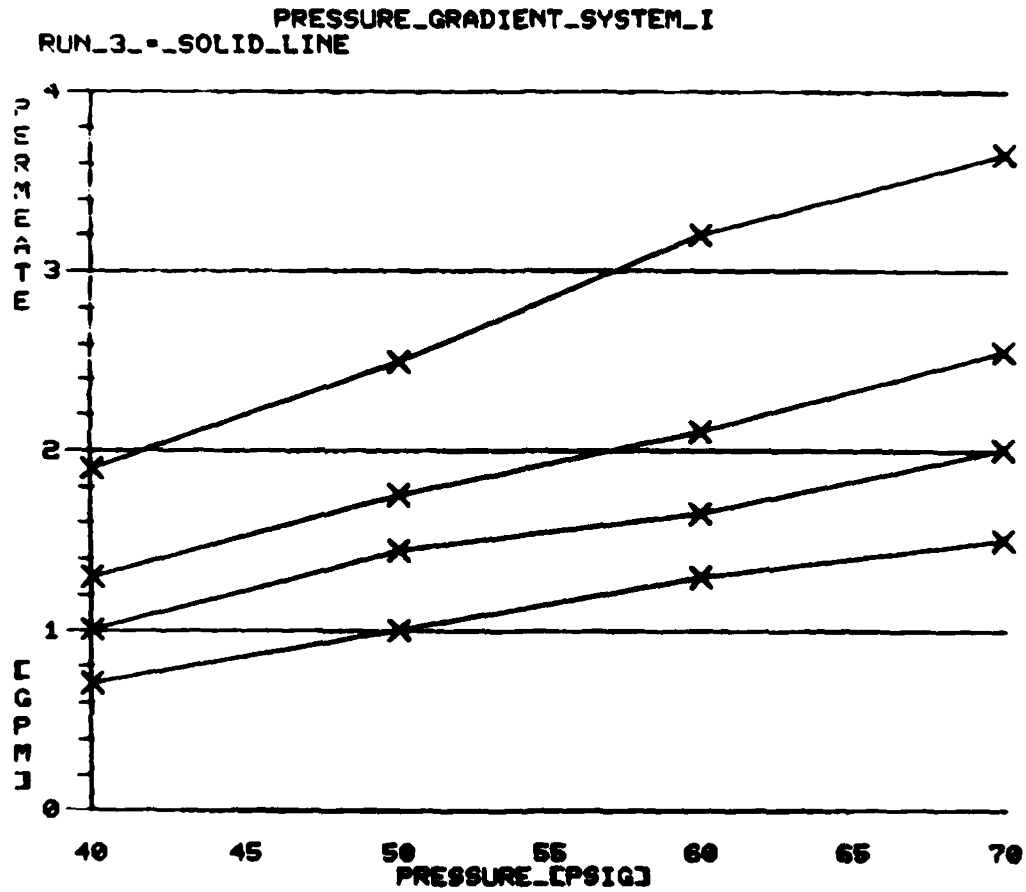


Figure 16. Recommended pretreatment showing cartridge filter and blowdown filter with spiral-wound element.

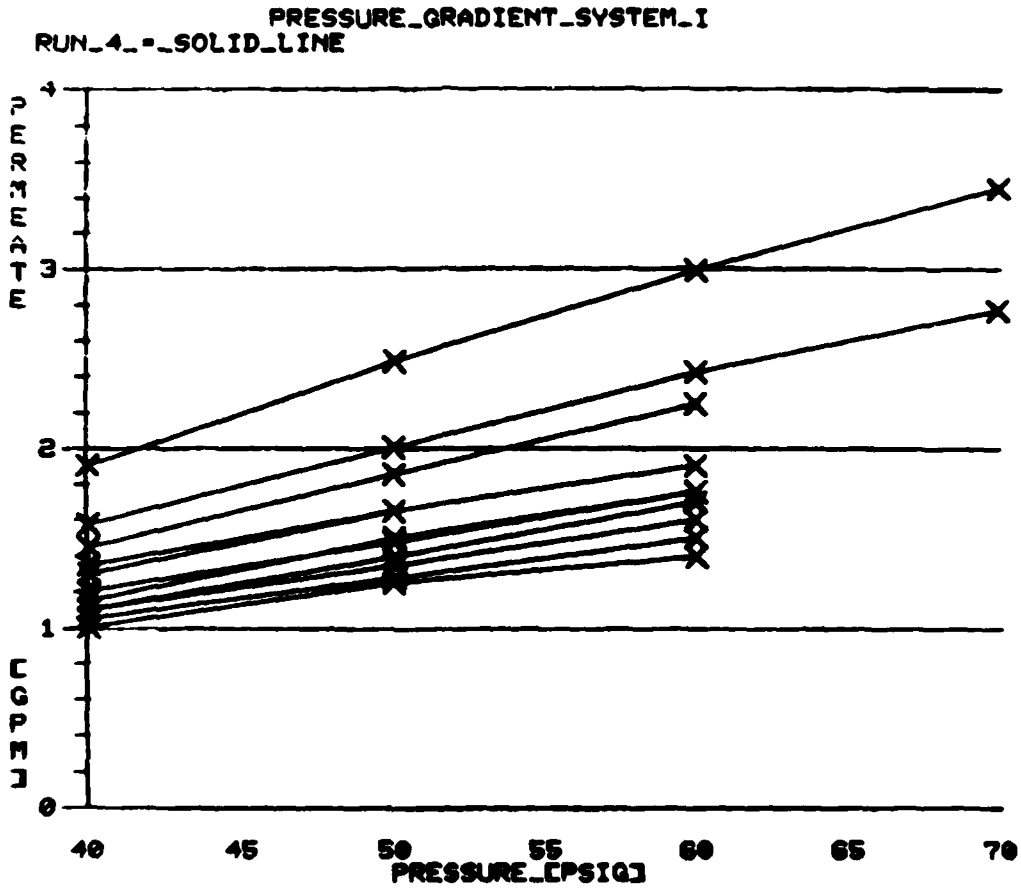
APPENDIX A -- Graphs 1-11



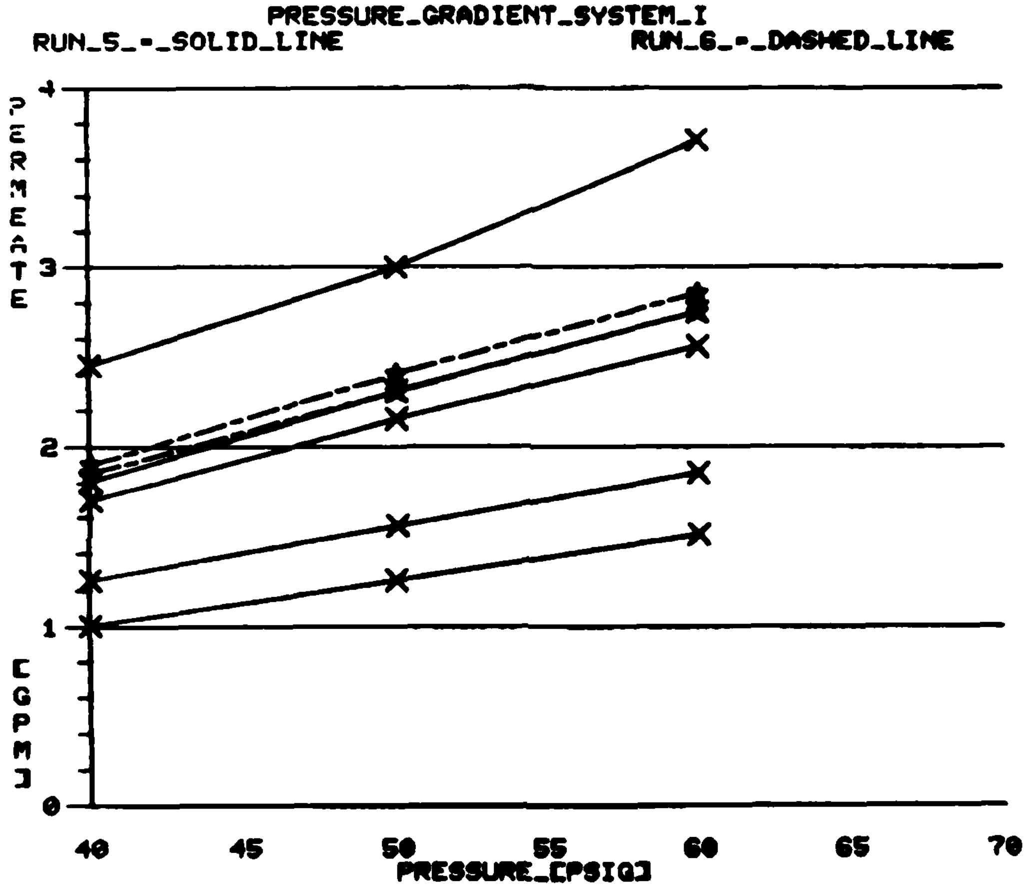
GRAPH 1


PRESSURE_GRADIENT_SYSTEM_I
RUN_3_=_SOLID_LINE
PERMEATE [GPM]
0
1
2
3
4
40
45
50
55
60
65
70
PRESSURE_[PSIG]

GRAPH 2

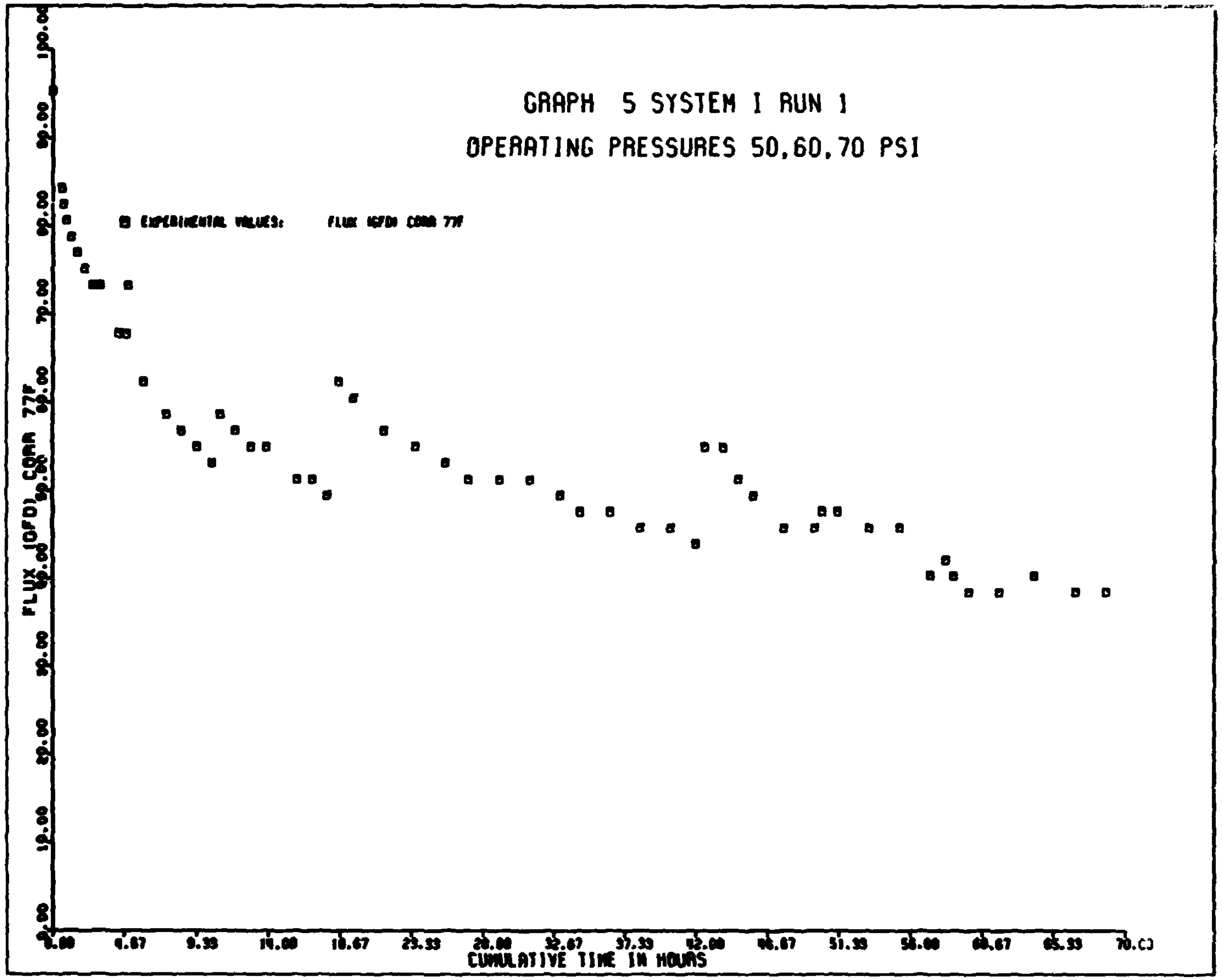

PRESSURE_GRADIENT_SYSTEM_I
RUN_4_=_SOLID_LINE
PERMEATE [GPM]
0
1
2
3
4
40
45
50
55
60
65
70
PRESSURE_[PSIG]

GRAPH 3



PRESSURE_GRADIENT_SYSTEM_I
RUN_5_-_SOLID_LINE
RUN_6_-_DASHED_LINE
PERMEATE [GPM]
0
1
2
3
4
40
45
50
55
60
65
70
PRESSURE_[PSIG]

GRAPH 4


GRAPH 5 SYSTEM I RUN 1
OPERATING PRESSURES 50,60,70 PSI
EXPERIMENTAL VALUES: FLUX (GFD) CORR 77F
FLUX (GFD) CORR 77F
CUMULATIVE TIME IN HOURS
0.00
4.67
9.33
14.00
18.67
23.33
28.00
32.67
37.33
42.00
46.67
51.33
56.00
60.67
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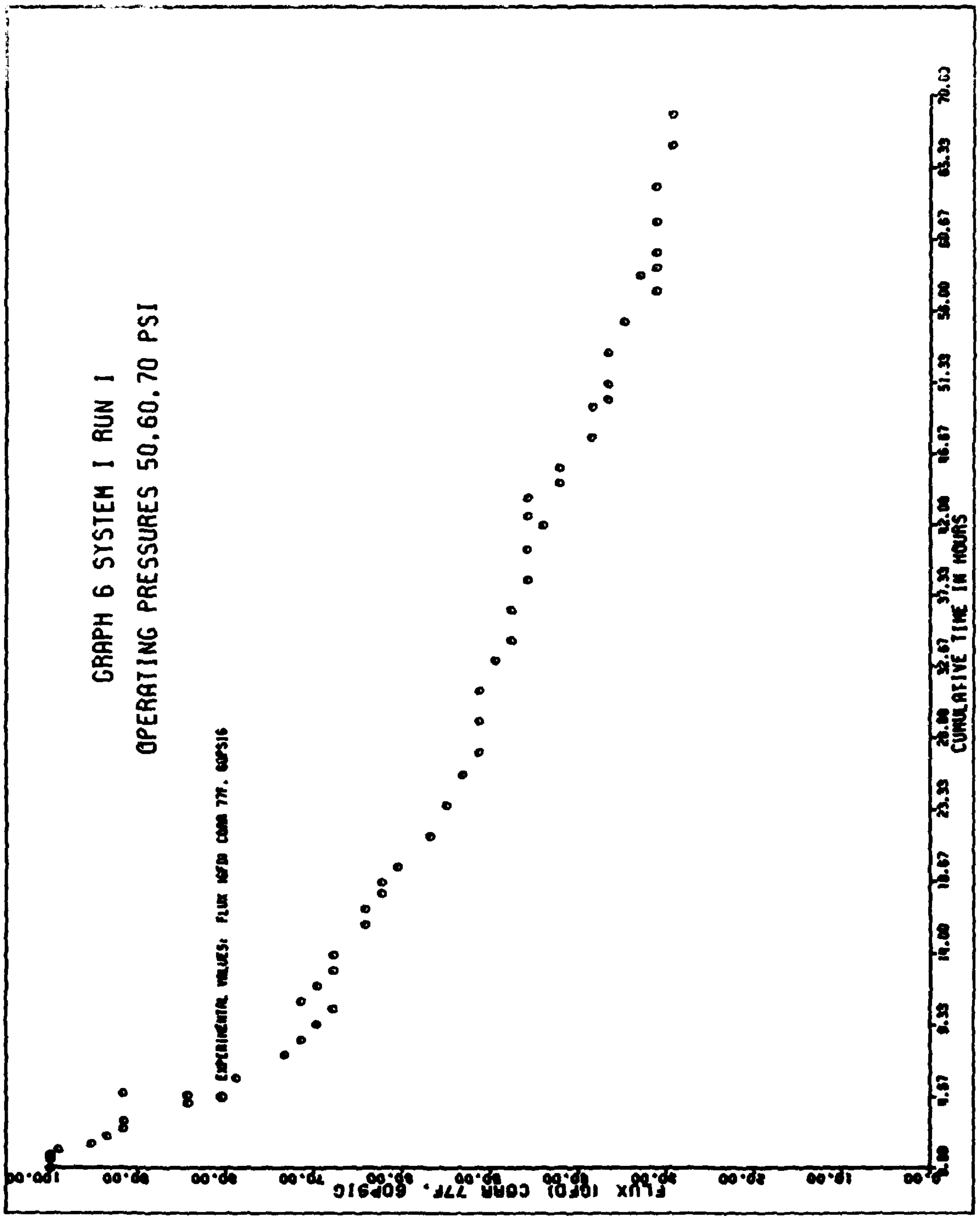
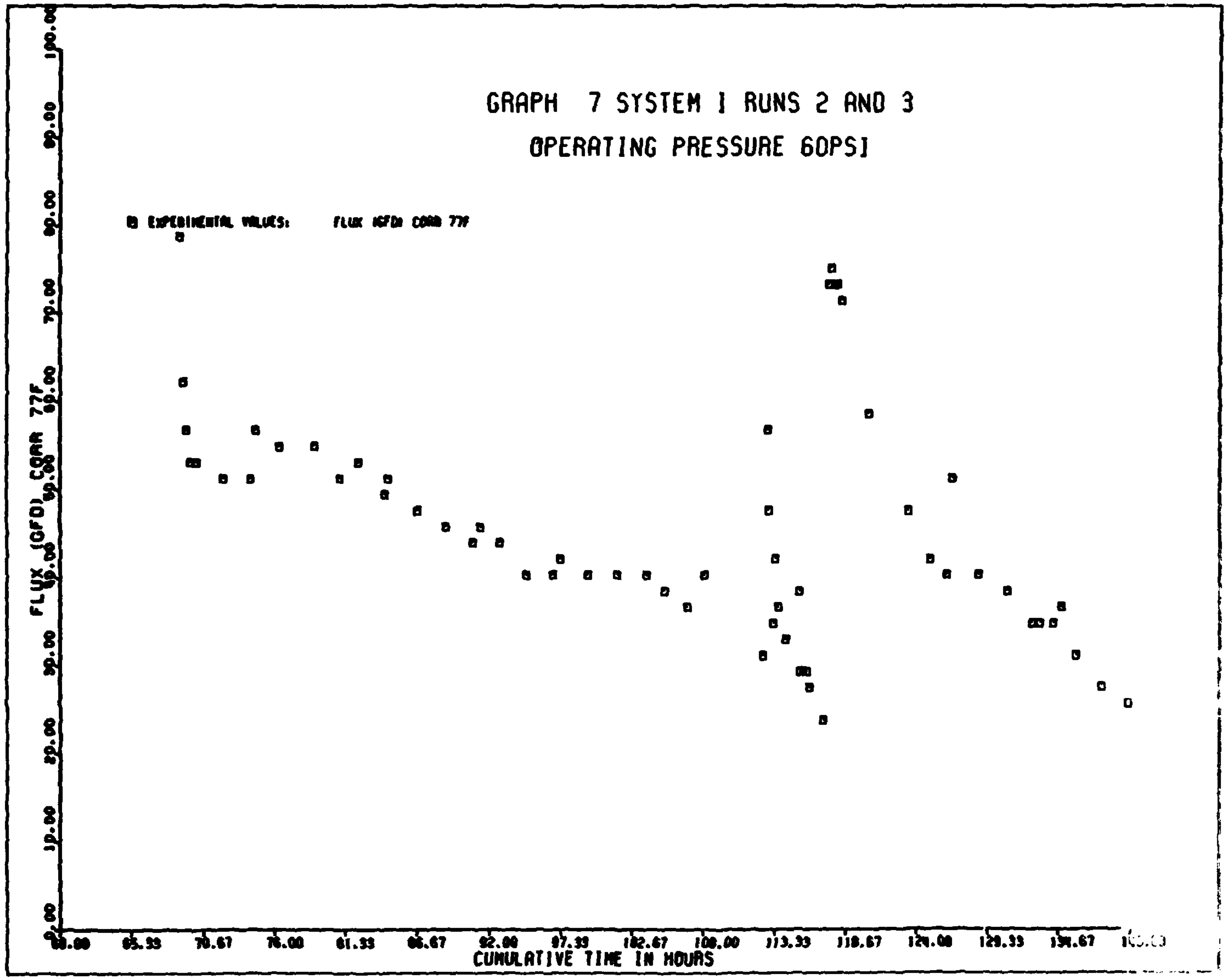
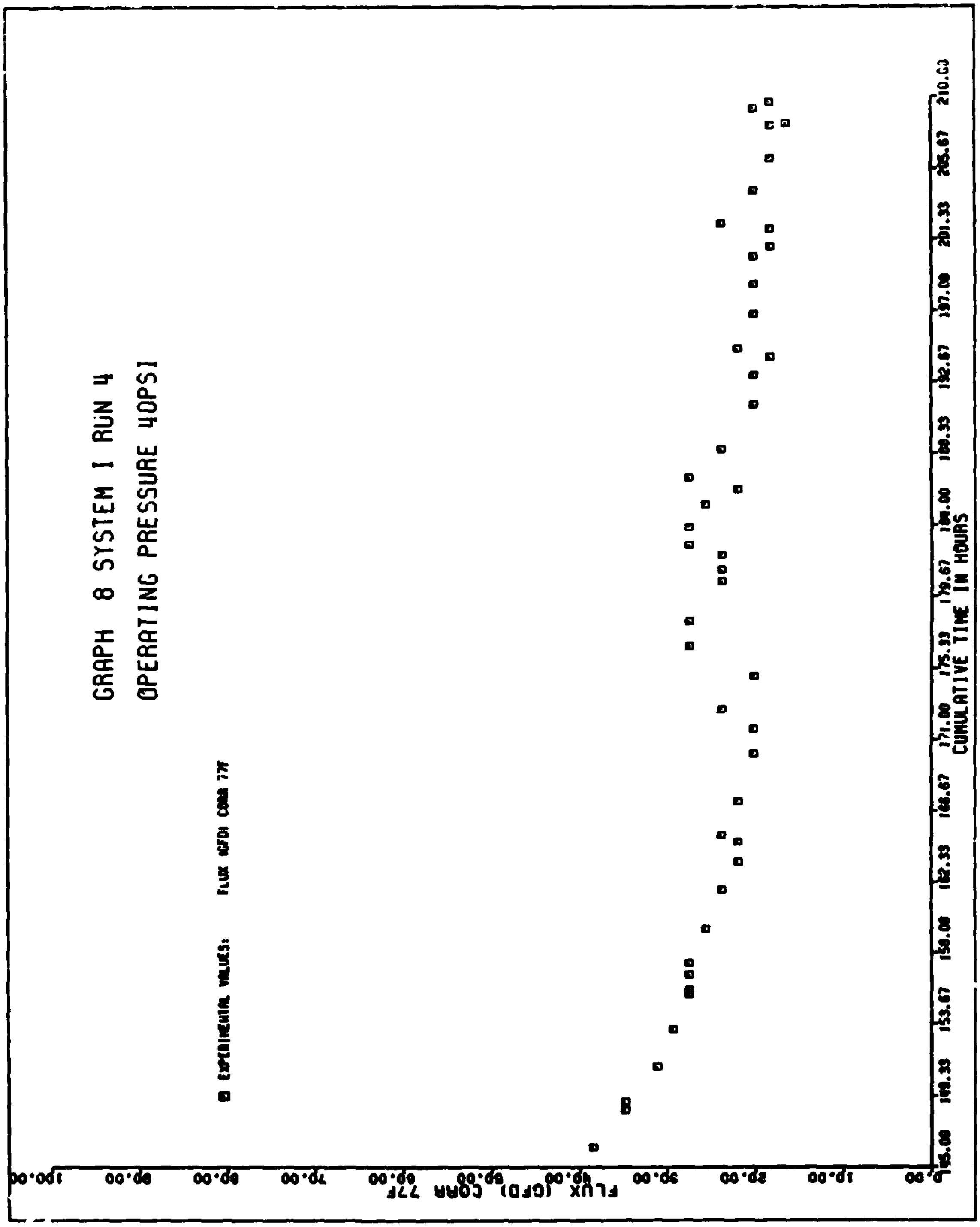
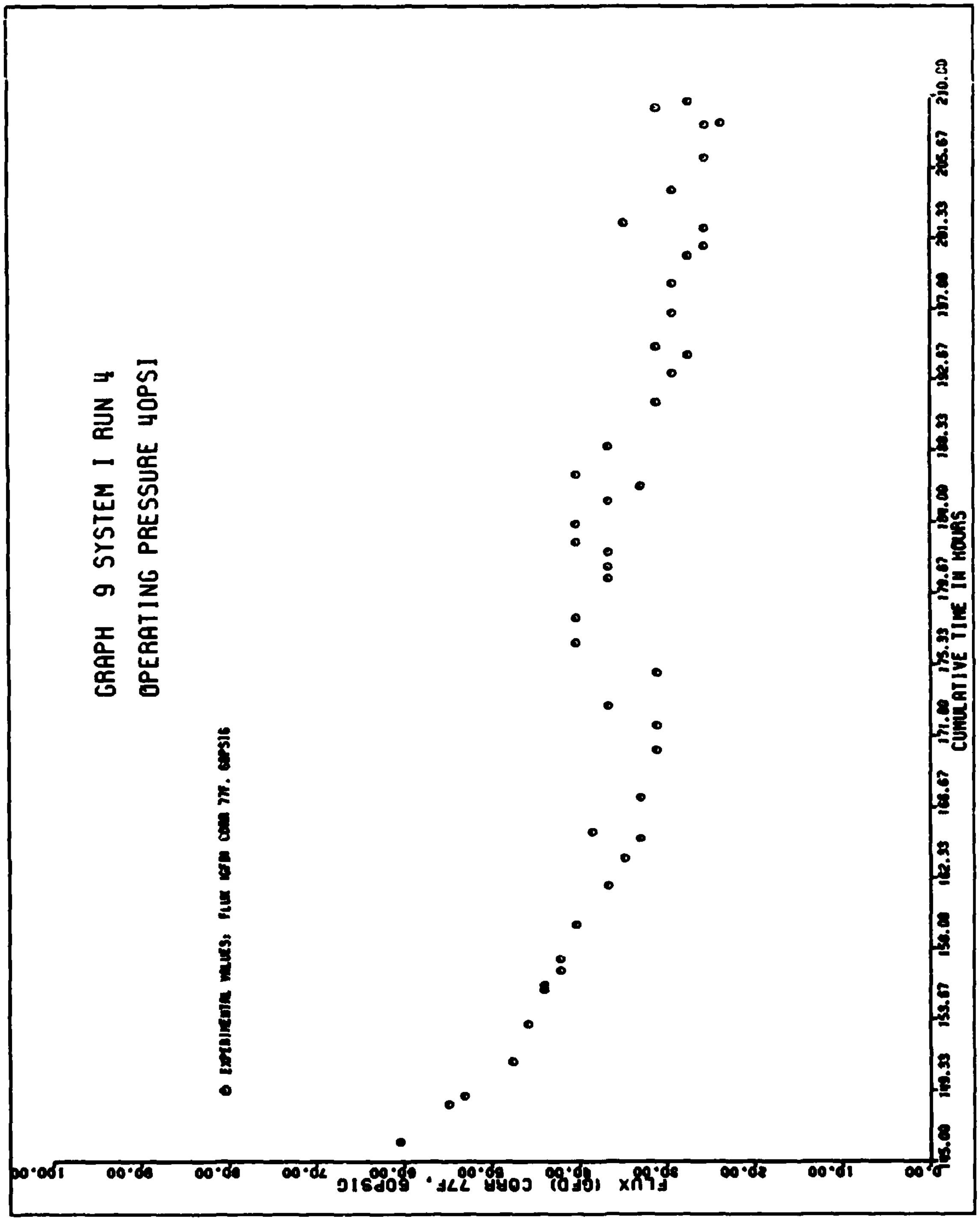
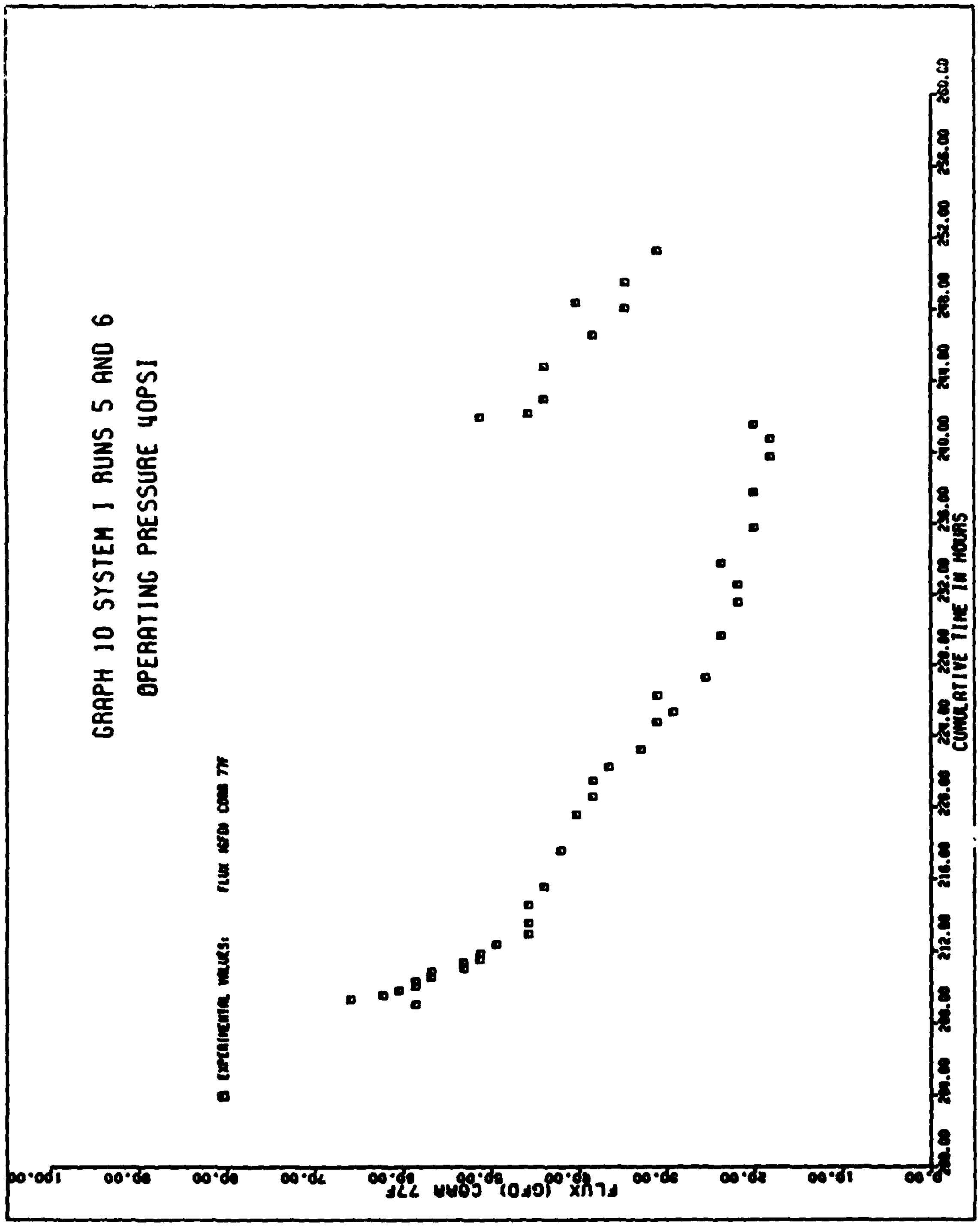
70.00
80.00
90.00
100.00

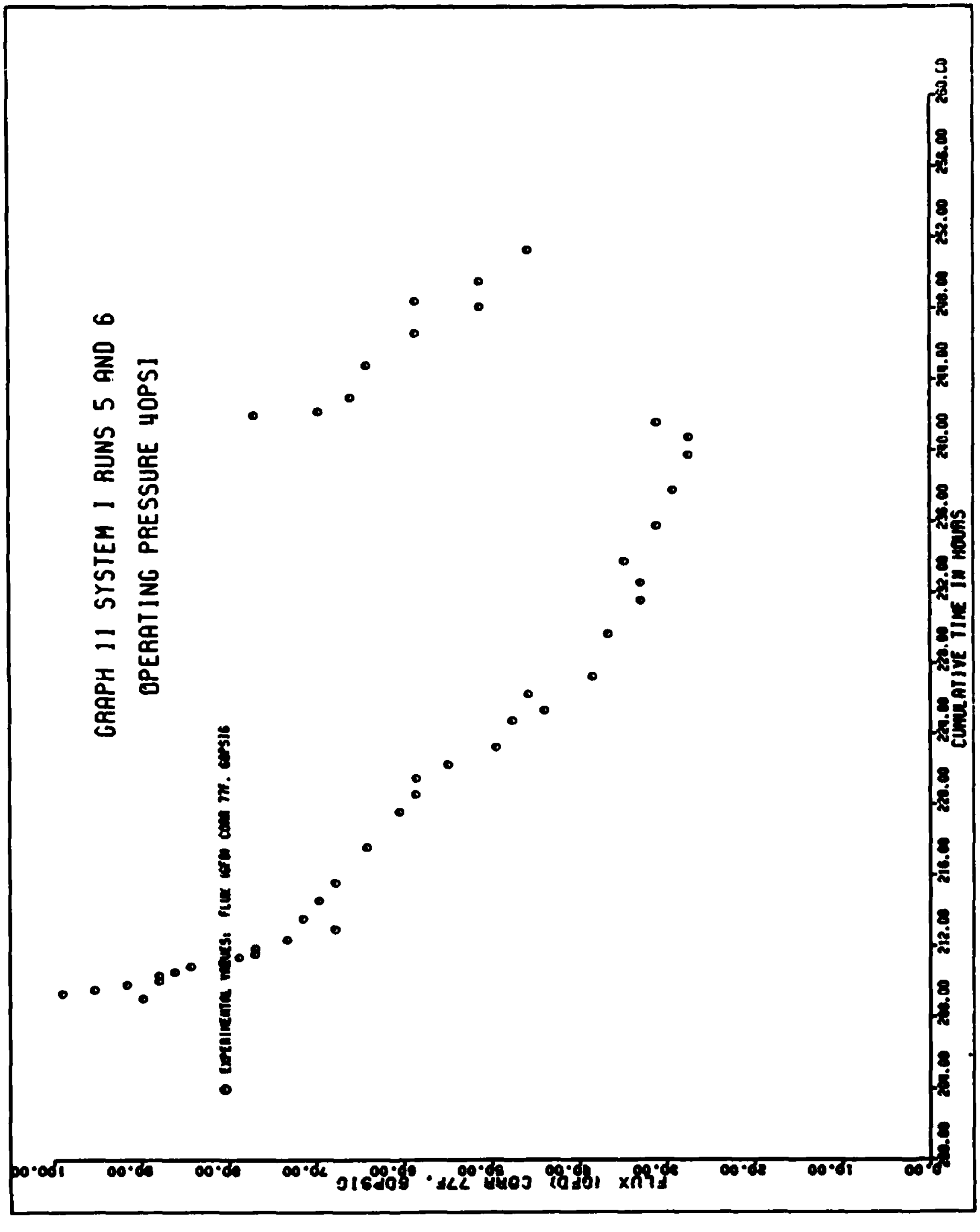

GRAPH 6 SYSTEM I RUN I
OPERATING PRESSURES 50,60,70 PSI
CUMULATIVE TIME IN HOURS
FLUX (GFD) CORR 77F, 60PSIG


GRAPH 7 SYSTEM I RUNS 2 AND 3
OPERATING PRESSURE 60PSI
EXPERIMENTAL VALUES: FLUX (GFD) CORR 77F
FLUX (GFD) CORR 77F
0.00
10.00
20.00
30.00
40.00
50.00
60.00
70.00
80.00
90.00
100.00
60.00
65.33
70.67
76.00
81.33
86.67
92.00
97.33
102.67
108.00
113.33
118.67
124.00
129.33
134.67
CUMULATIVE TIME IN HOURS

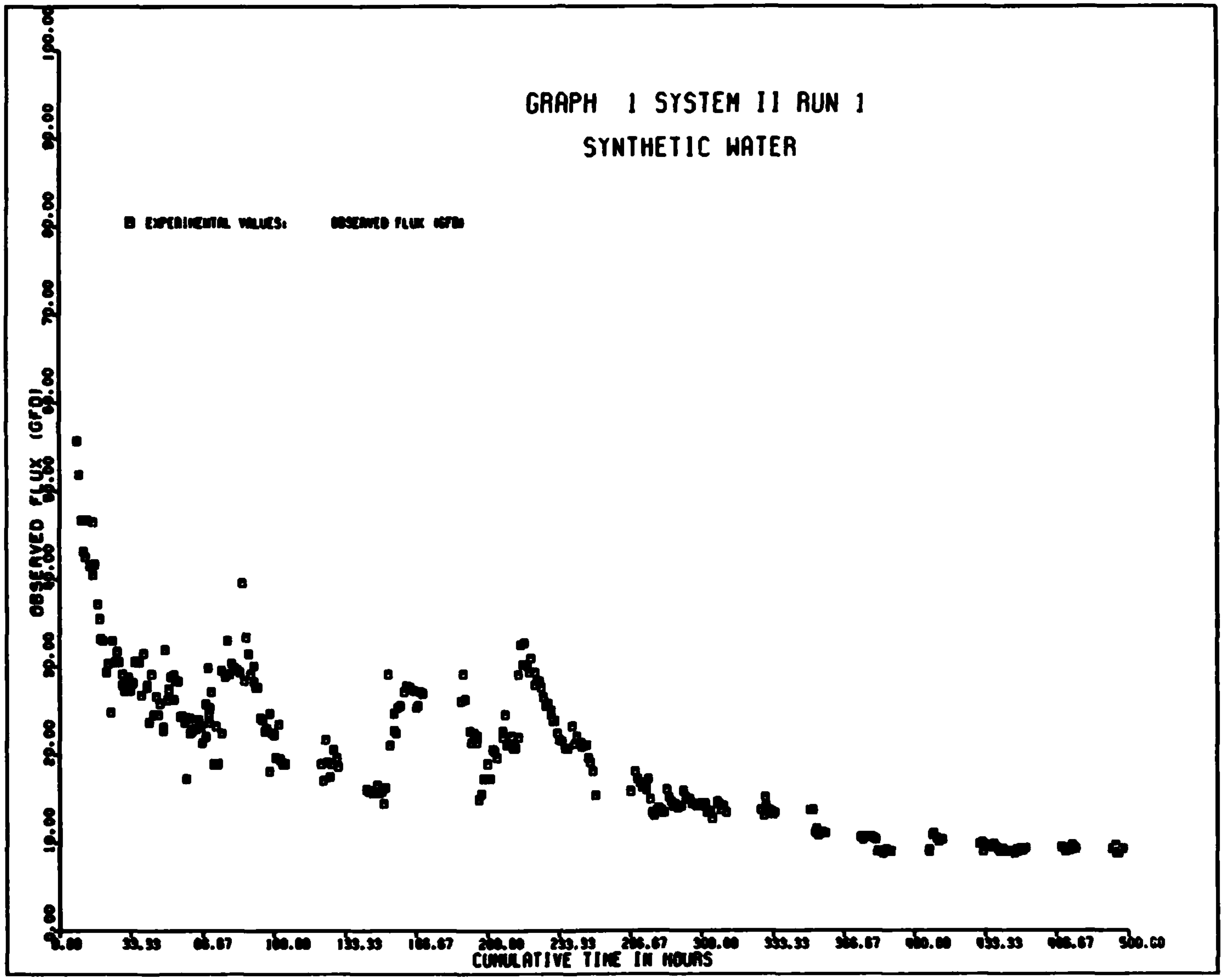

GRAPH 8 SYSTEM I RUN 4
OPERATING PRESSURE 40PSI
EXPERIMENTAL VALUES: FLUX (GFD) CORR 77F
FLUX (GFD) CORR 77F
0.00
10.00
20.00
30.00
40.00
50.00
60.00
70.00
80.00
90.00
100.00
145.00
149.33
153.67
158.00
162.33
166.67
171.00
175.33
179.67
184.00
188.33
192.67
197.00
201.33
205.67
210.00
CUMULATIVE TIME IN HOURS

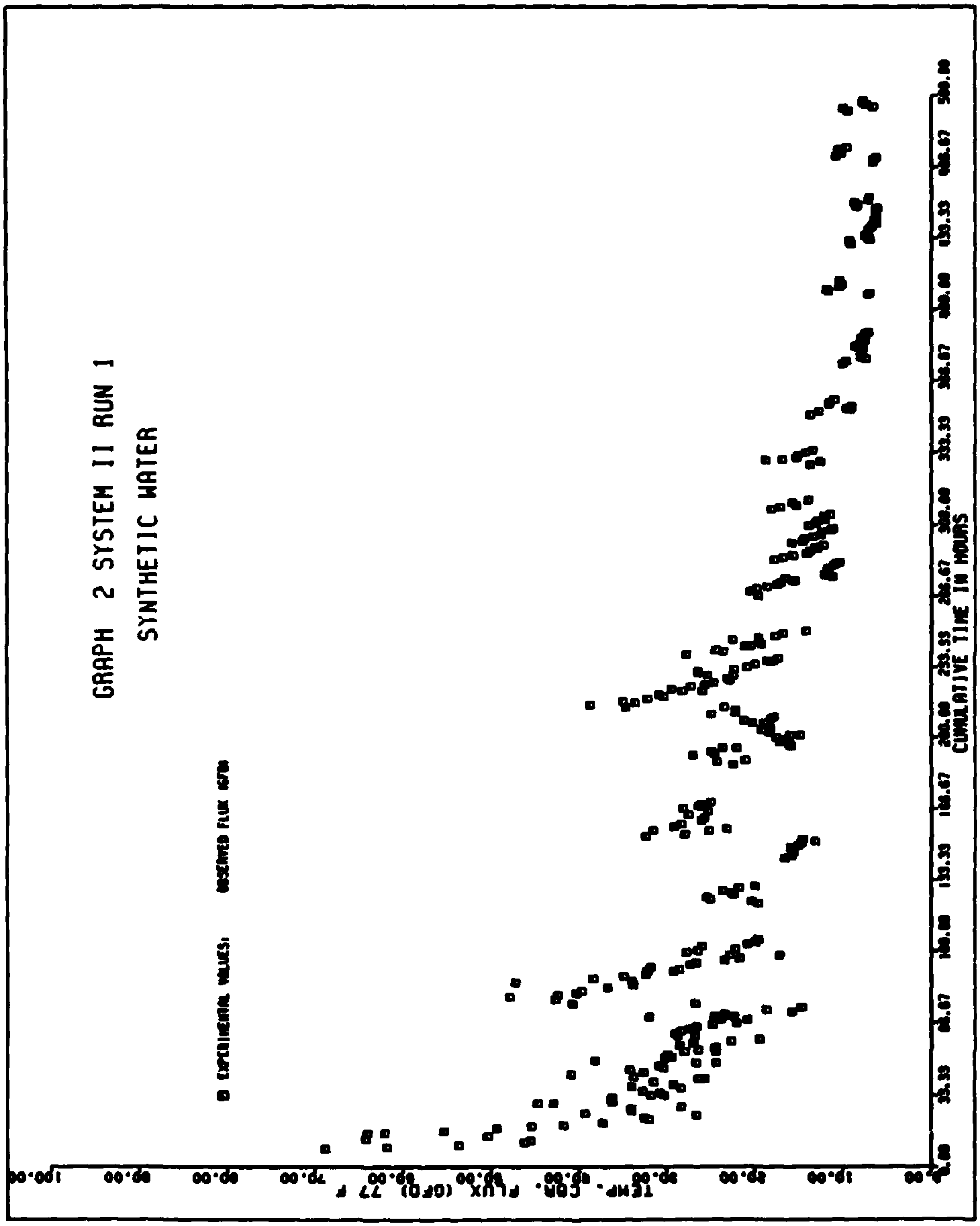

GRAPH 9 SYSTEM I RUN 4
OPERATING PRESSURE 40PSI
EXPERIMENTAL VALUES: FLUX GFD CORR 77F, 60PSIG
FLUX (GFD) CORR 77F, 60PSIG
0.00
10.00
20.00
30.00
40.00
50.00
60.00
70.00
80.00
90.00
100.00
CUMULATIVE TIME IN HOURS
145.00
149.33
153.67
158.00
162.33
166.67
171.00
175.33
179.67
184.00
188.33
192.67
197.00
201.33
205.67
210.00

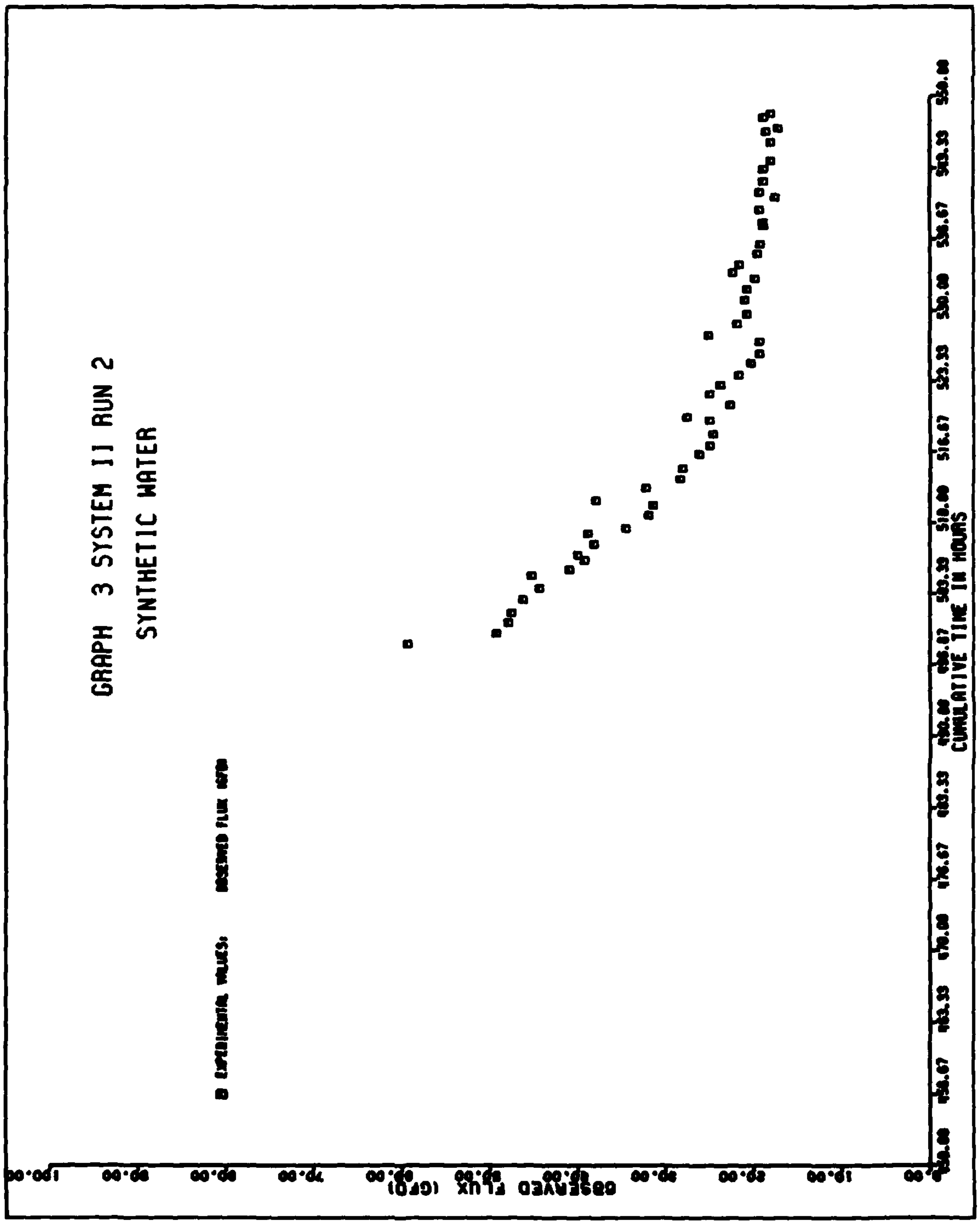

GRAPH 10 SYSTEM I RUNS 5 AND 6
OPERATING PRESSURE 40PSI
EXPERIMENTAL VALUES: FLUX (GFD) CORR 77F
CUMULATIVE TIME IN HOURS
FLUX (GFD) CORR 77F

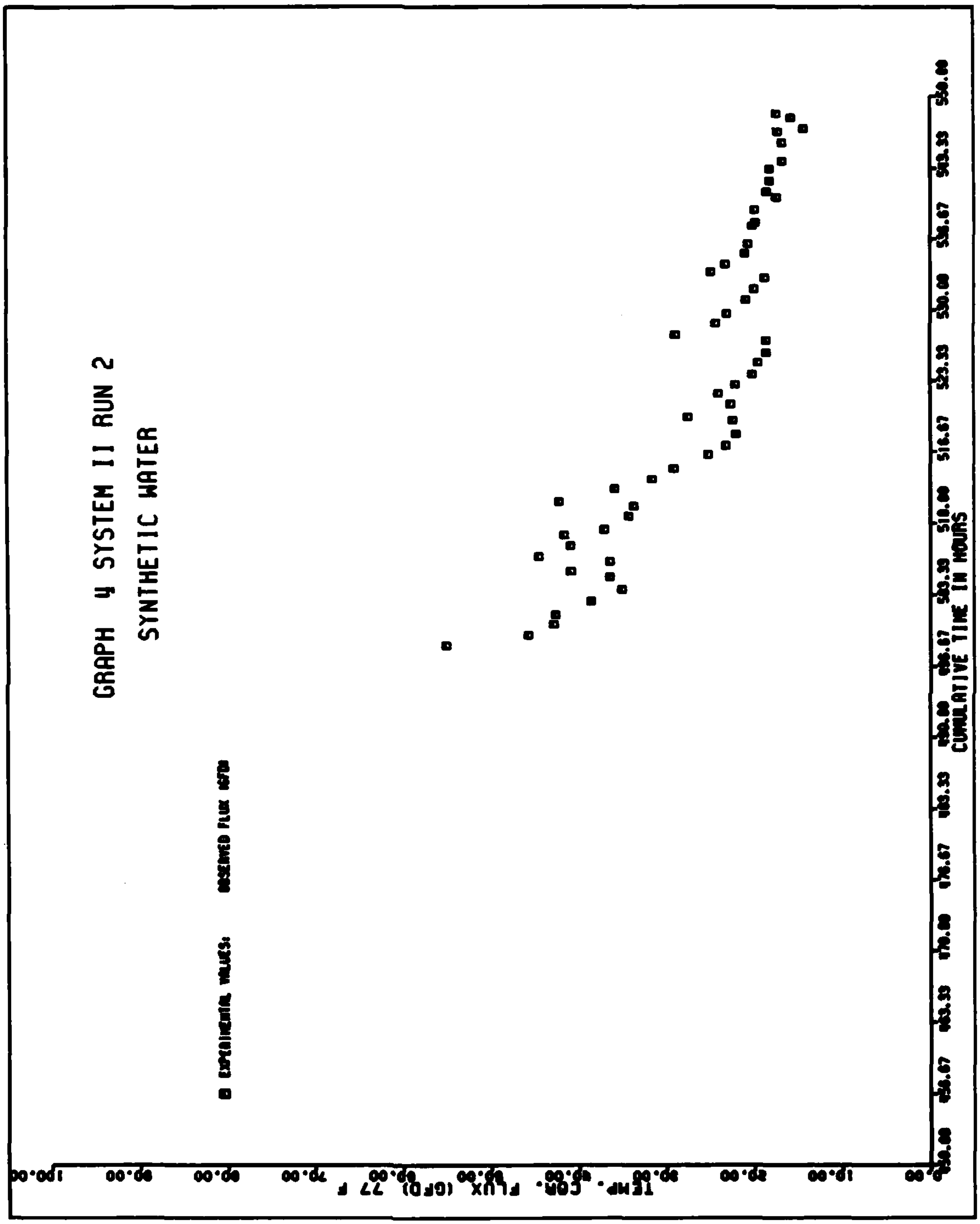

GRAPH 11 SYSTEM I RUNS 5 AND 6
OPERATING PRESSURE 40PSI
CUMULATIVE TIME IN HOURS
FLUX (GFD) CORR 77F, 60PSIG

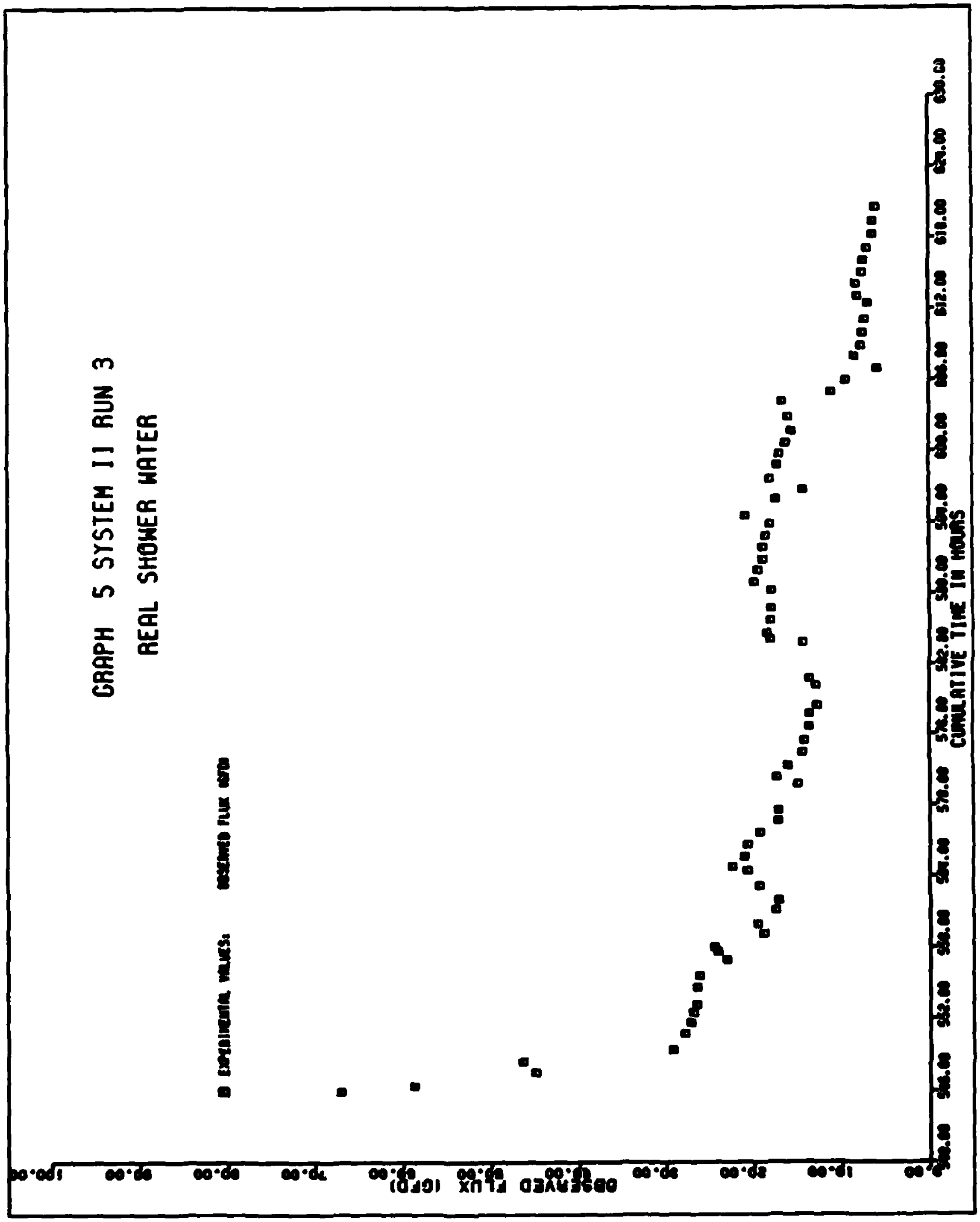
APPENDIX B – Graphs 1-6

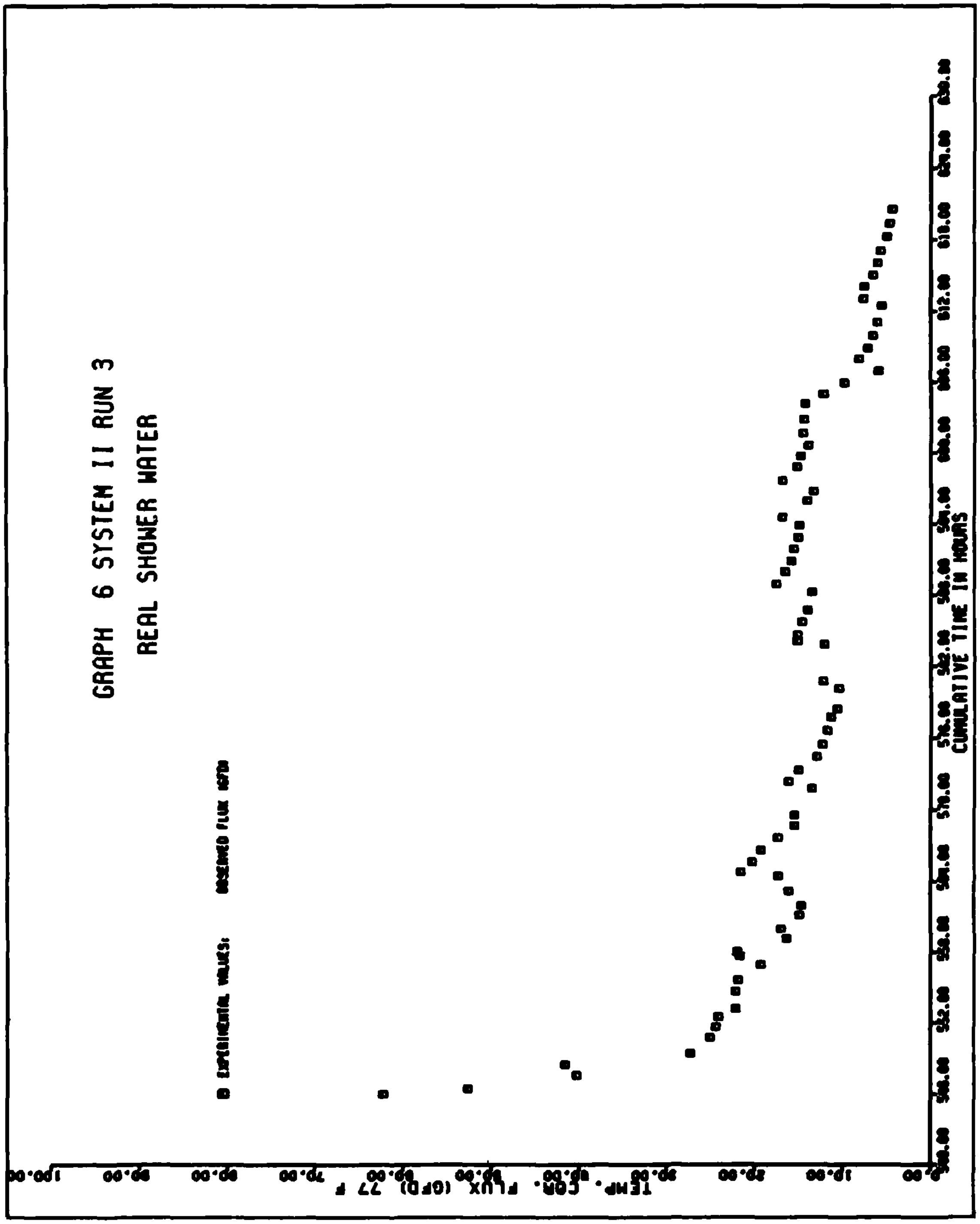



GRAPH 2 SYSTEM II RUN 1
SYNTHETIC WATER
CUMULATIVE TIME IN HOURS
TEMP. COR. FLUX (GFD) 77 F

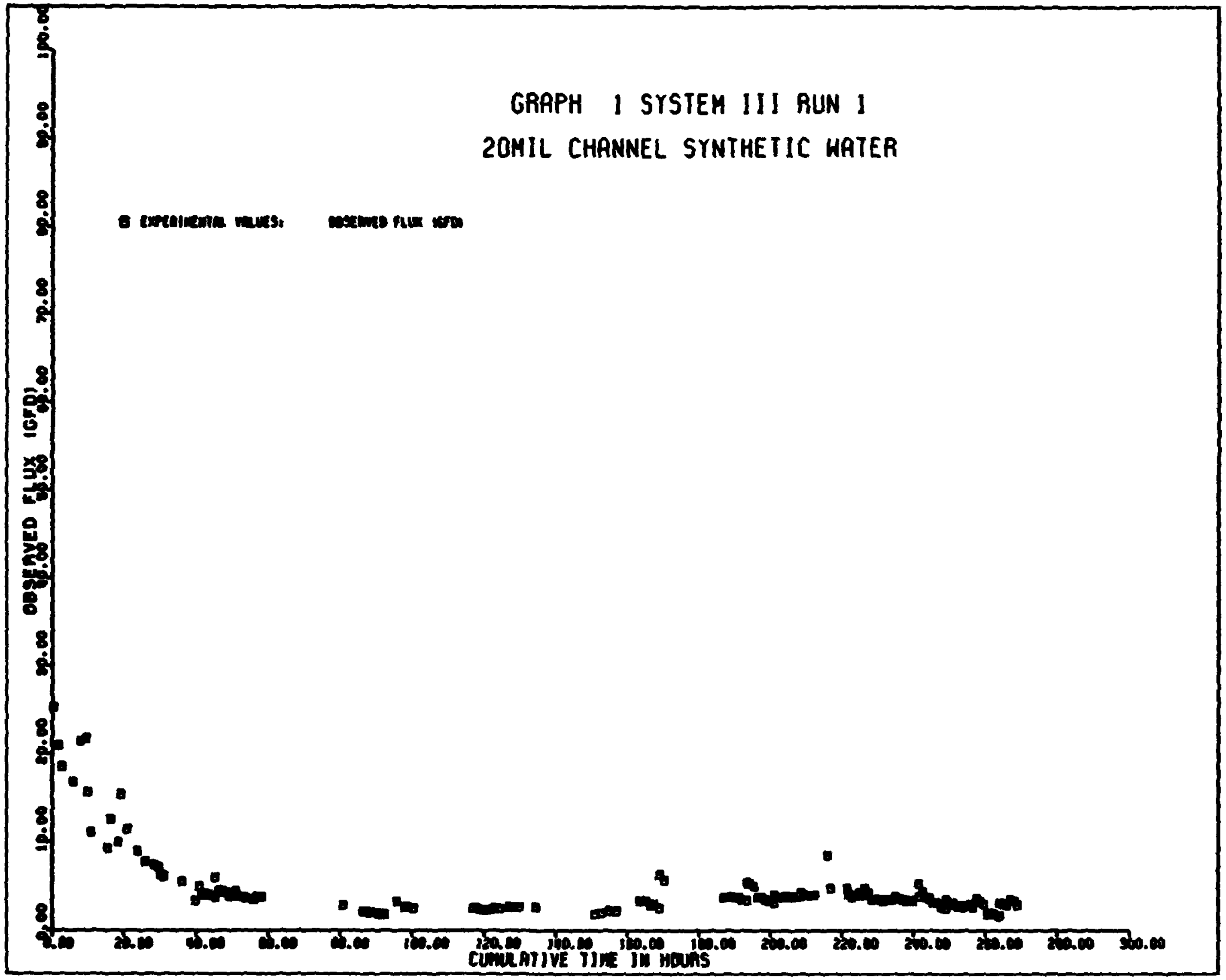

GRAPH 3 SYSTEM II RUN 2
SYNTHETIC WATER
CUMULATIVE TIME IN HOURS
OBSERVED FLUX (GFD)

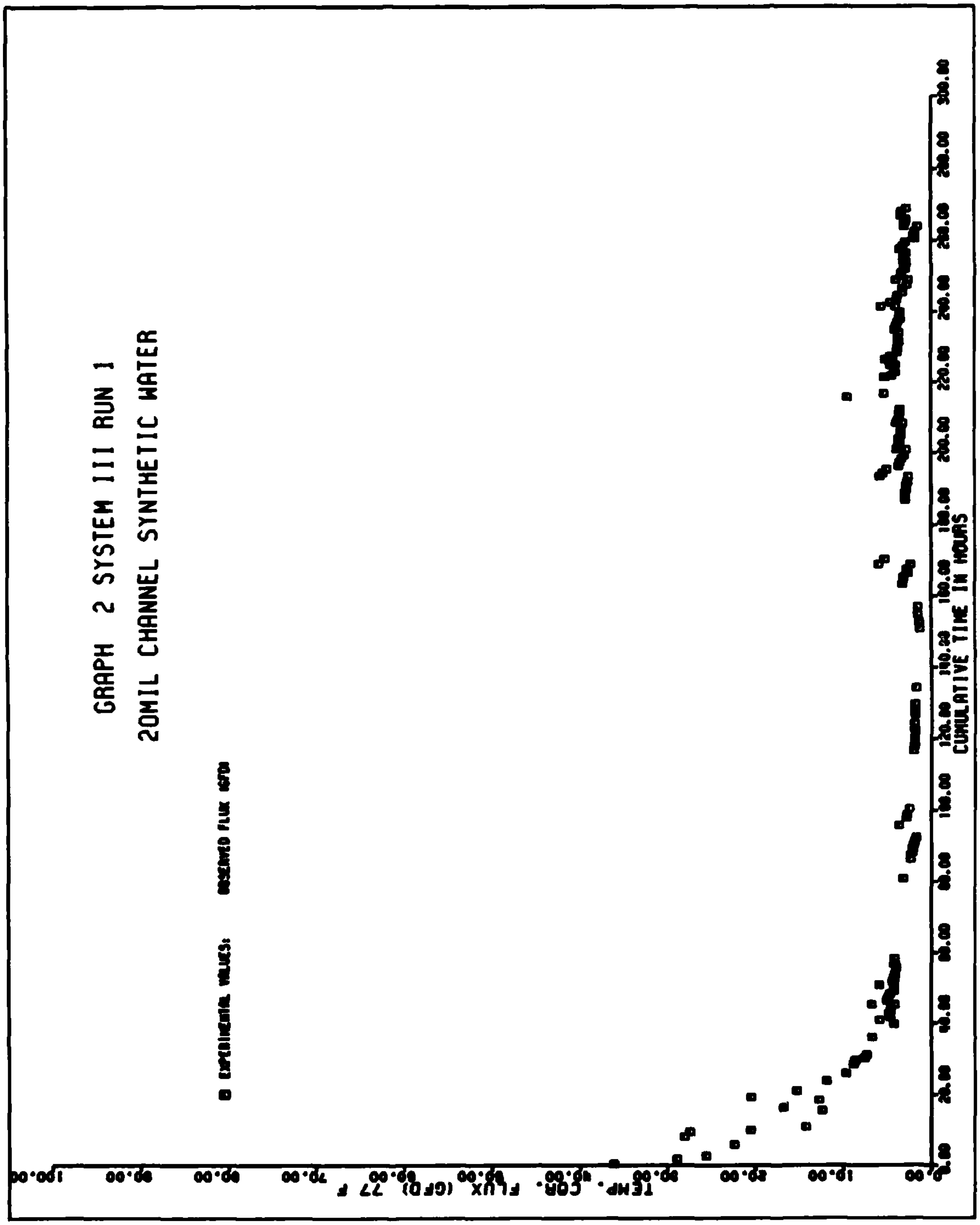

GRAPH 4 SYSTEM II RUN 2
SYNTHETIC WATER
EXPERIMENTAL VALUES: OBSERVED FLUX GFD
CUMULATIVE TIME IN HOURS
TEMP. COR. FLUX (GFD) 77 F

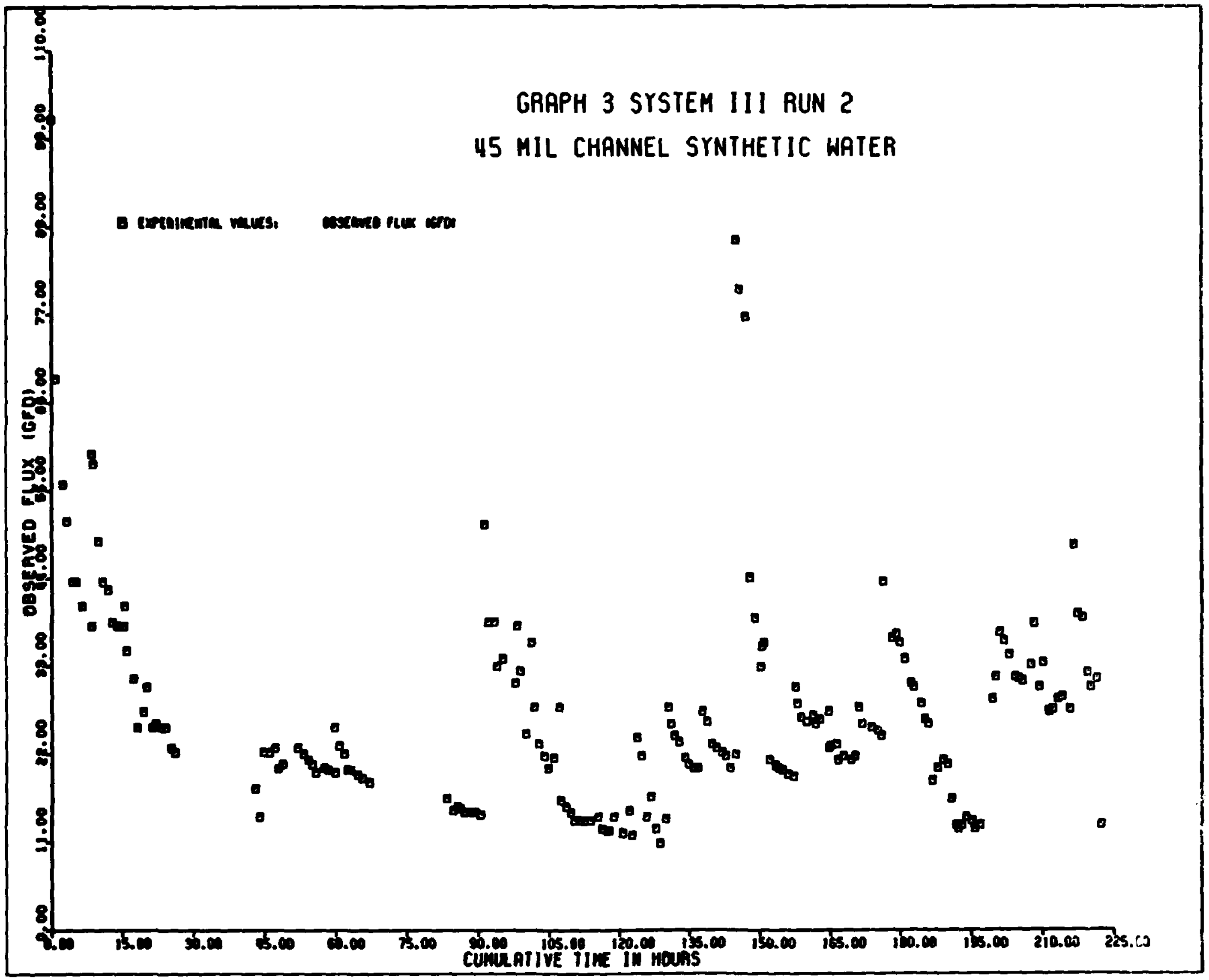

GRAPH 5 SYSTEM II RUN 3
REAL SHOWER WATER
EXPERIMENTAL VALUES: OBSERVED FLUX GFD
OBSERVED FLUX (GFD)
CUMULATIVE TIME IN HOURS


GRAPH 6 SYSTEM II RUN 3
REAL SHOWER WATER
EXPERIMENTAL VALUES:
OBSERVED FLUX (GFD)
TEMP. COR. FLUX (GFD) 77 F
CUMULATIVE TIME IN HOURS
0.00
10.00
20.00
30.00
40.00
50.00
60.00
70.00
80.00
90.00
100.00

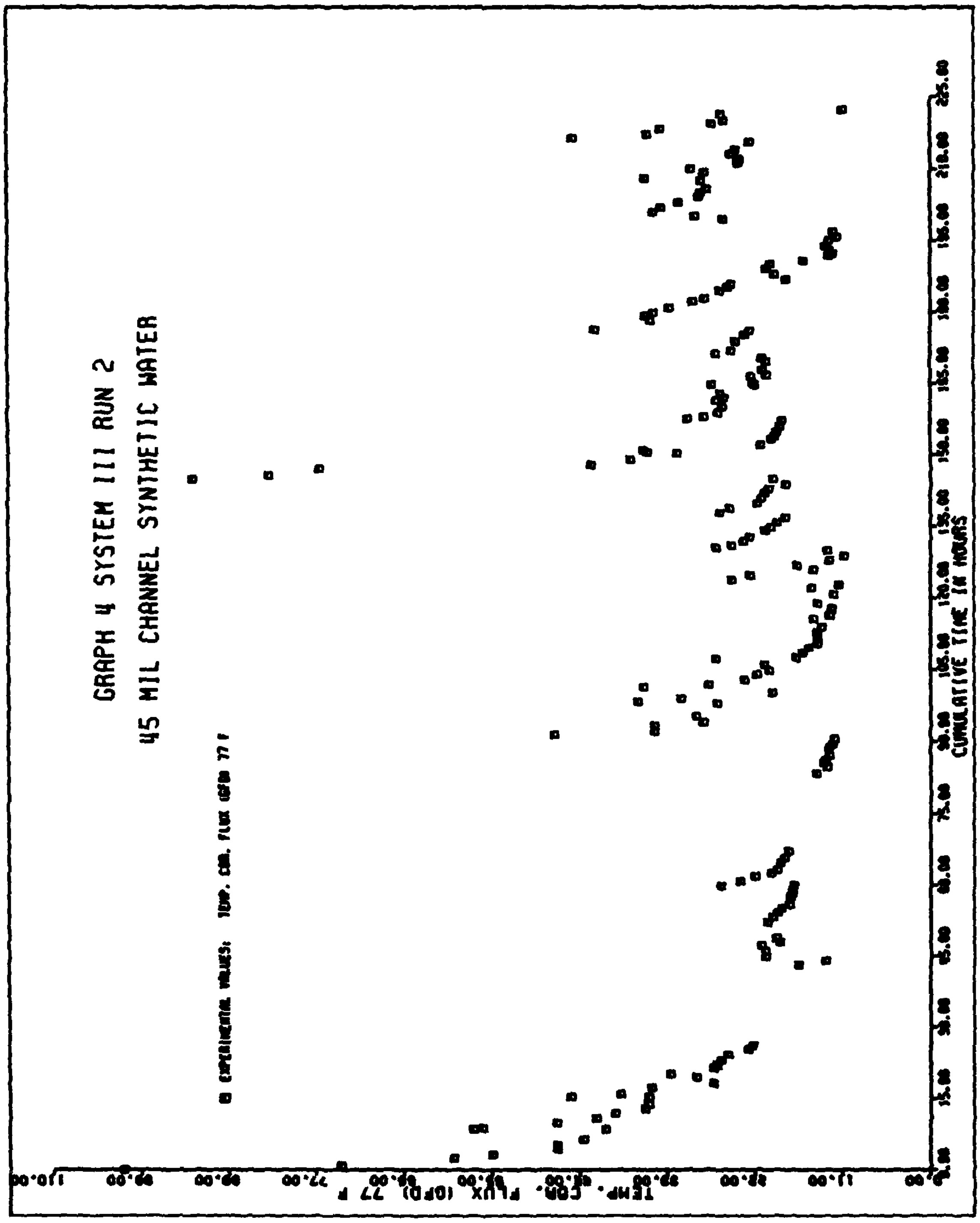
APPENDIX C – Graphs 1-6

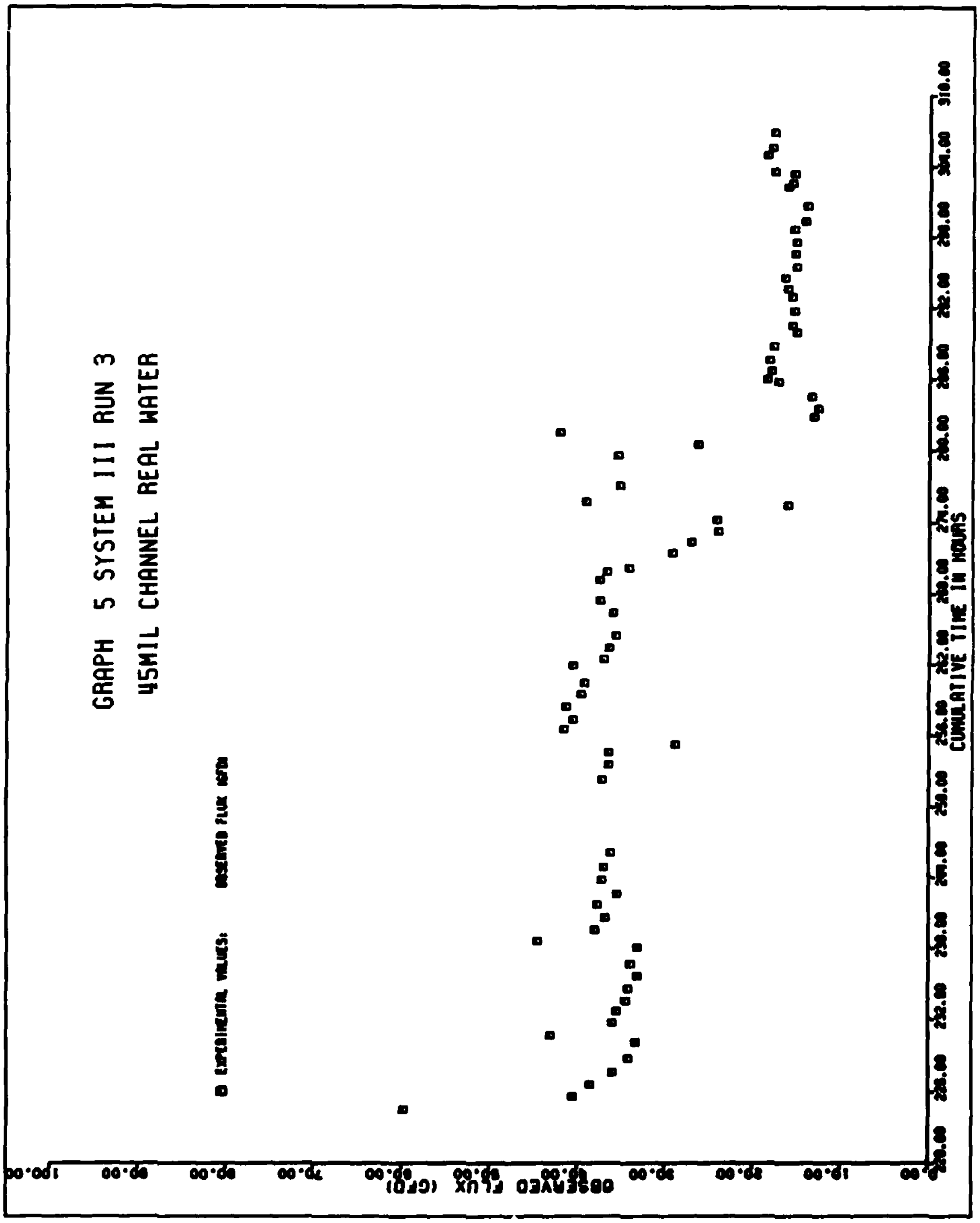


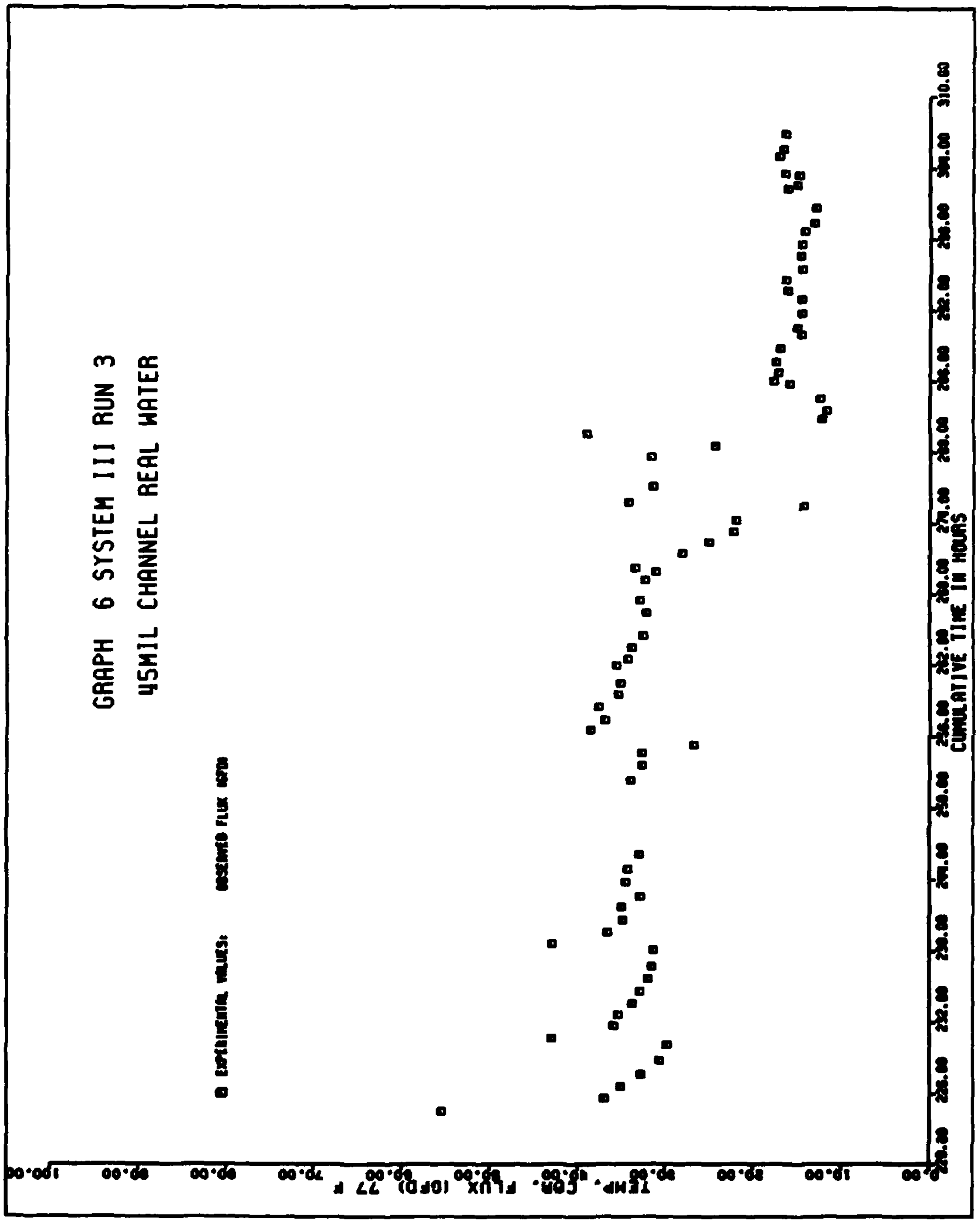

GRAPH 2 SYSTEM III RUN 1
20MIL CHANNEL SYNTHETIC WATER
EXPERIMENTAL VALUES: OBSERVED FLUX (GFD)
CUMULATIVE TIME IN HOURS
TEMP. COR. FLUX (GFD) 77 F



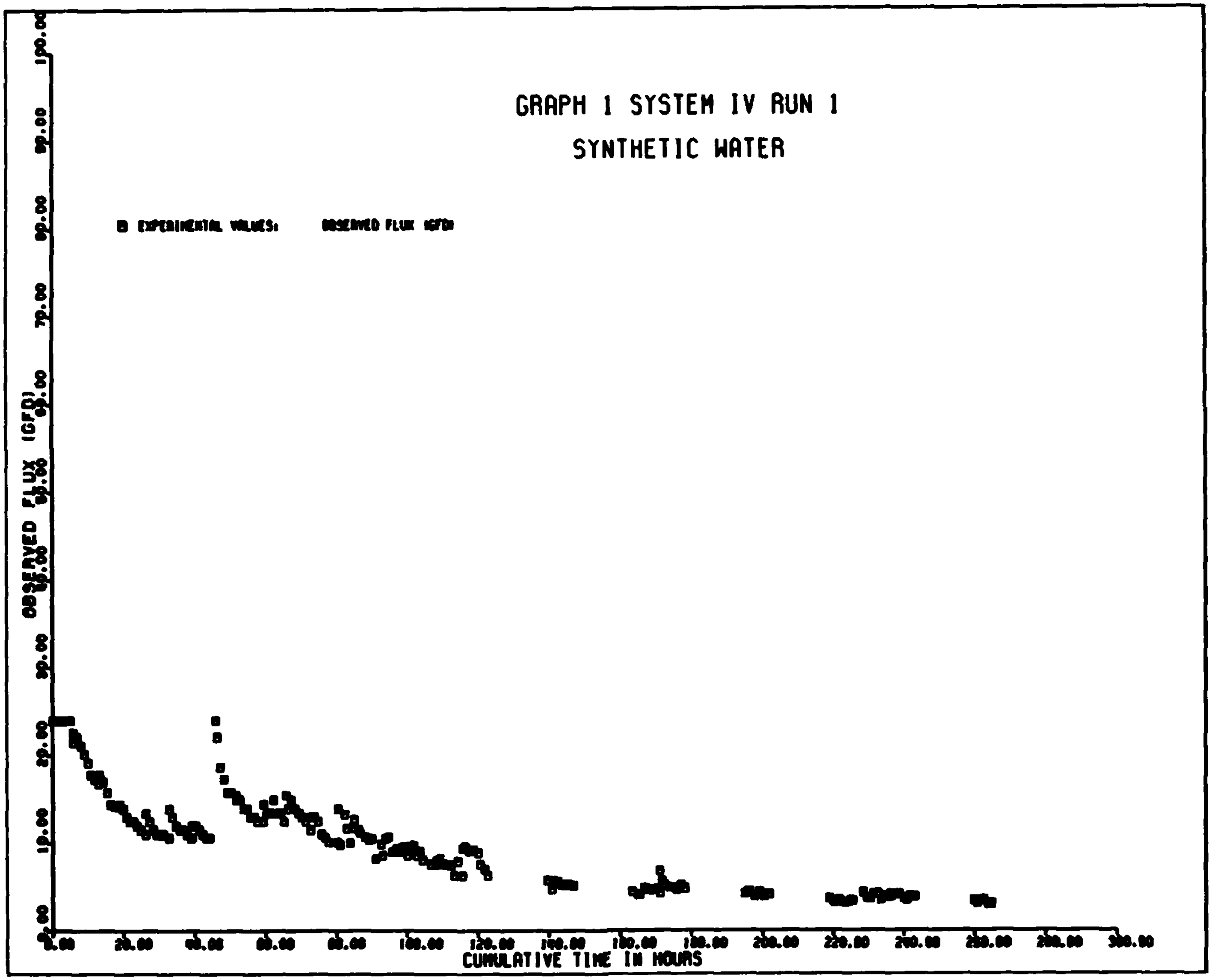
GRAPH 3 SYSTEM III RUN 2
45 MIL CHANNEL SYNTHETIC WATER
EXPERIMENTAL VALUES: OBSERVED FLUX (GFD)
OBSERVED FLUX (GFD)
0.00
11.00
22.00
33.00
44.00
55.00
66.00
77.00
88.00
99.00
110.00
0.00
15.00
30.00
45.00
60.00
75.00
90.00
105.00
120.00
135.00
150.00
165.00
180.00
195.00
210.00
225.00
CUMULATIVE TIME IN HOURS

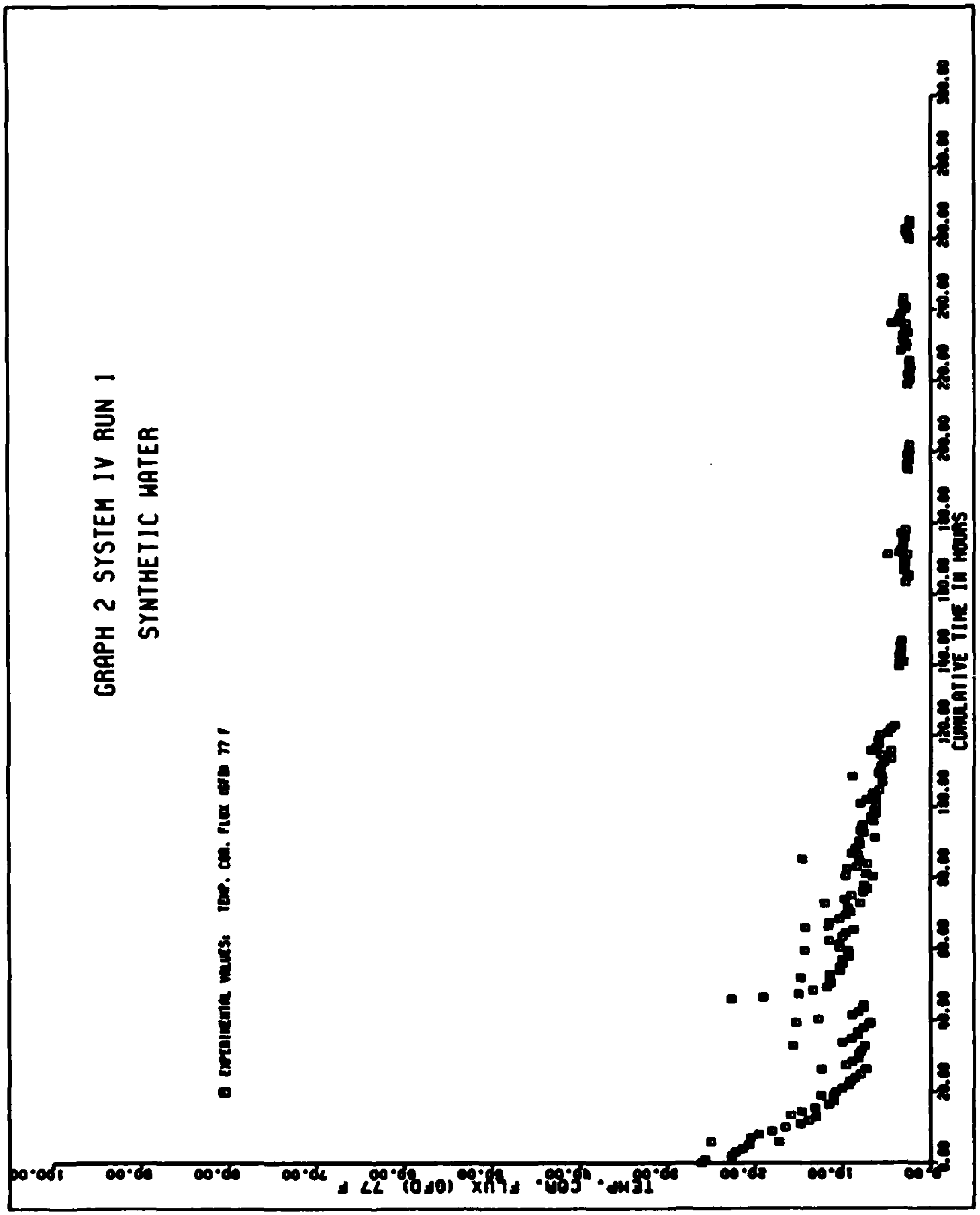

GRAPH 4 SYSTEM III RUN 2
45 MIL CHANNEL SYNTHETIC WATER
CUMULATIVE TIME IN HOURS
TEMP. COR. FLUX (GFD) 77 F


GRAPH 5 SYSTEM III RUN 3
45MIL CHANNEL REAL WATER
EXPERIMENTAL VALUES: OBSERVED FLUX (GFD)
CUMULATIVE TIME IN HOURS
OBSERVED FLUX (GFD)

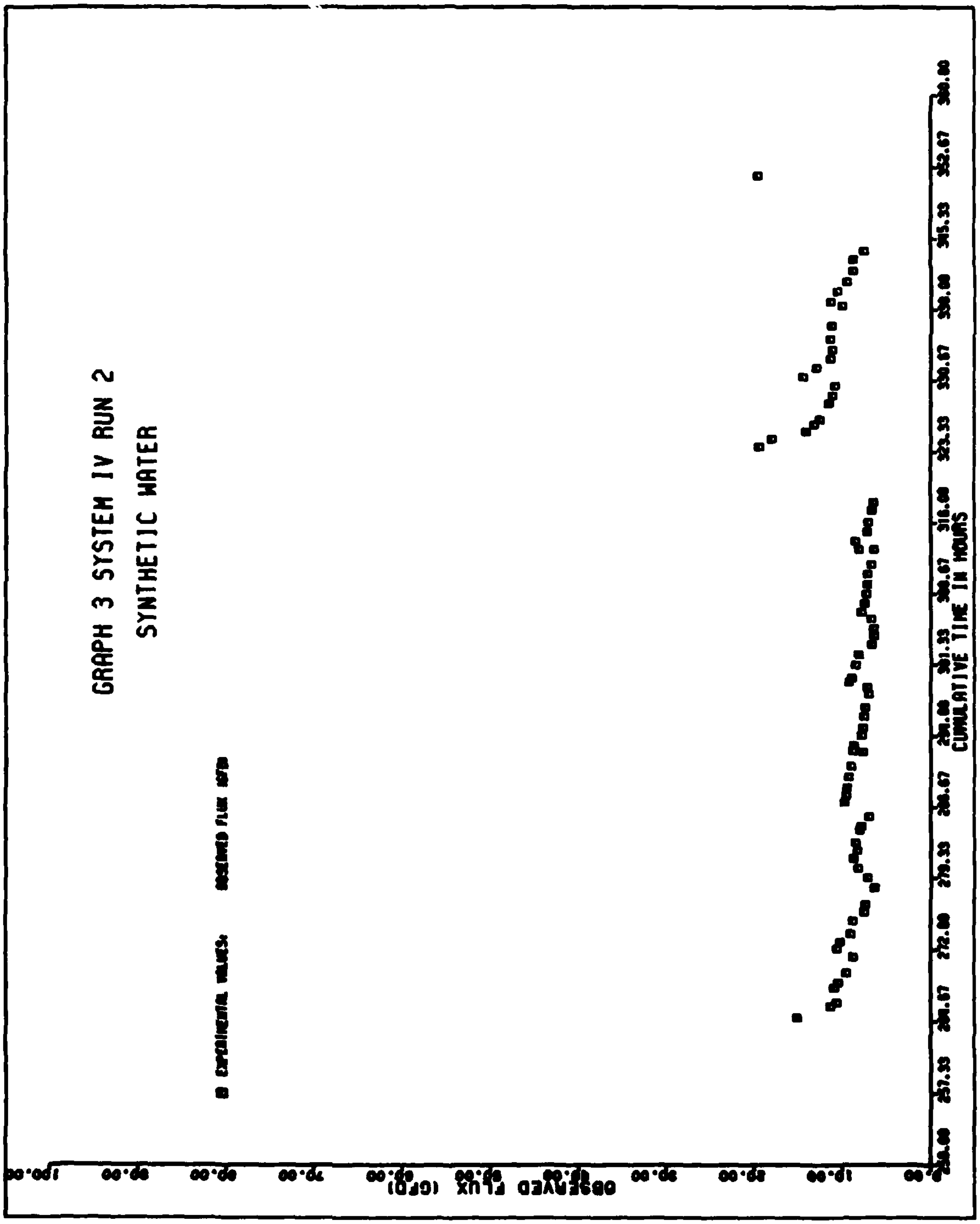

GRAPH 6 SYSTEM III RUN 3
45MIL CHANNEL REAL WATER
EXPERIMENTAL VALUES: OBSERVED FLUX GFD
CUMULATIVE TIME IN HOURS
TEMP. COR. FLUX (GFD) 77 F

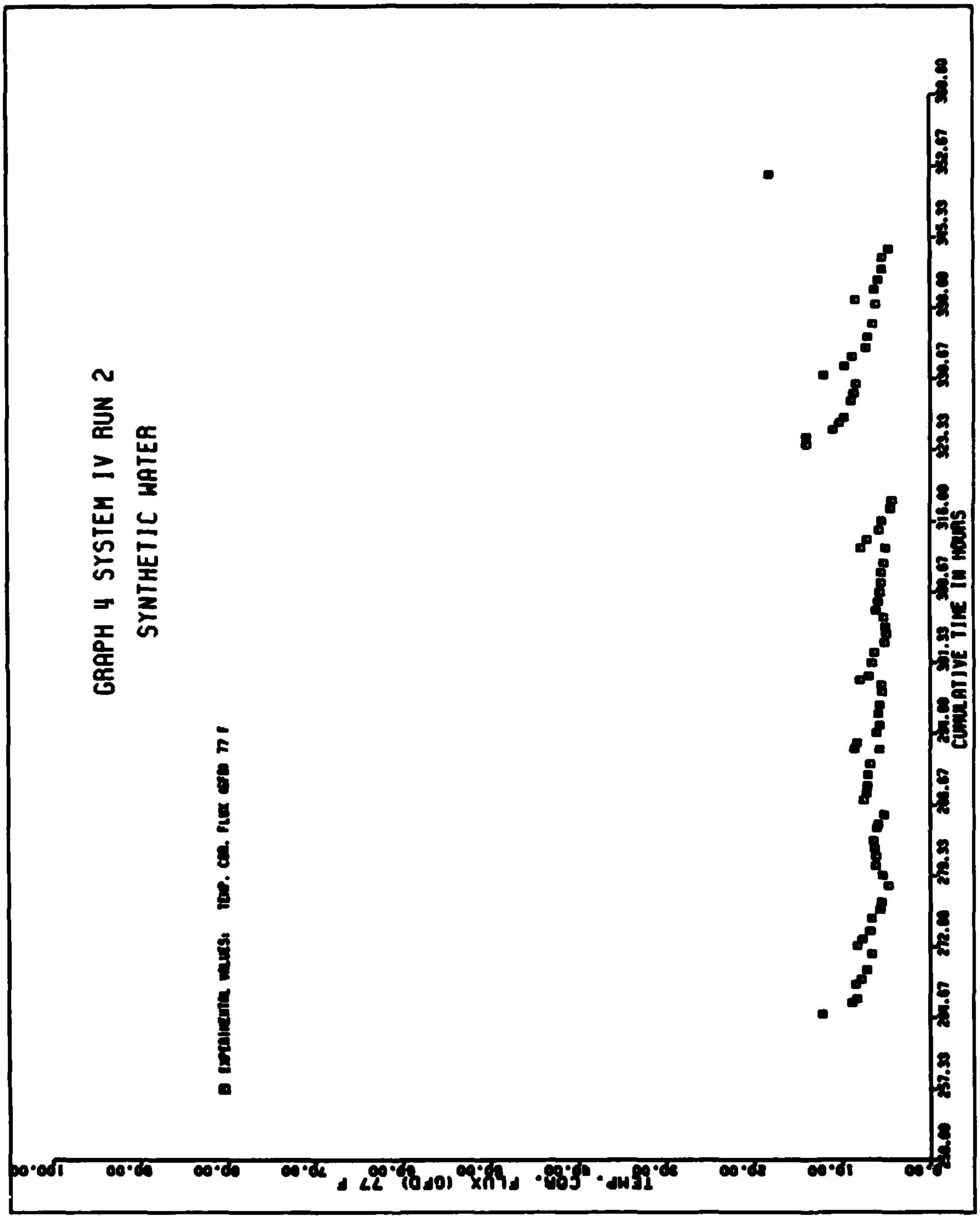
APPENDIX D – Graphs 1-8

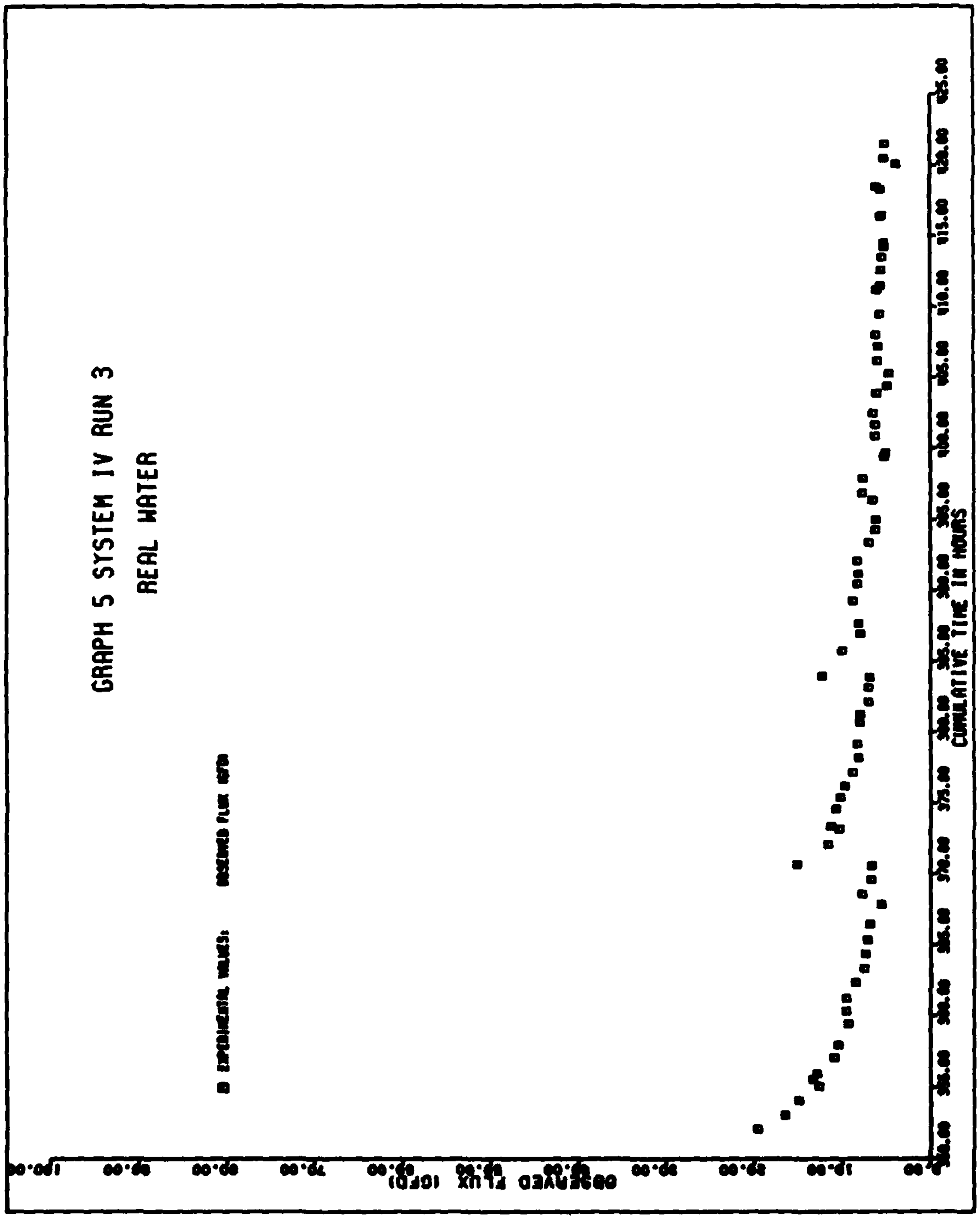


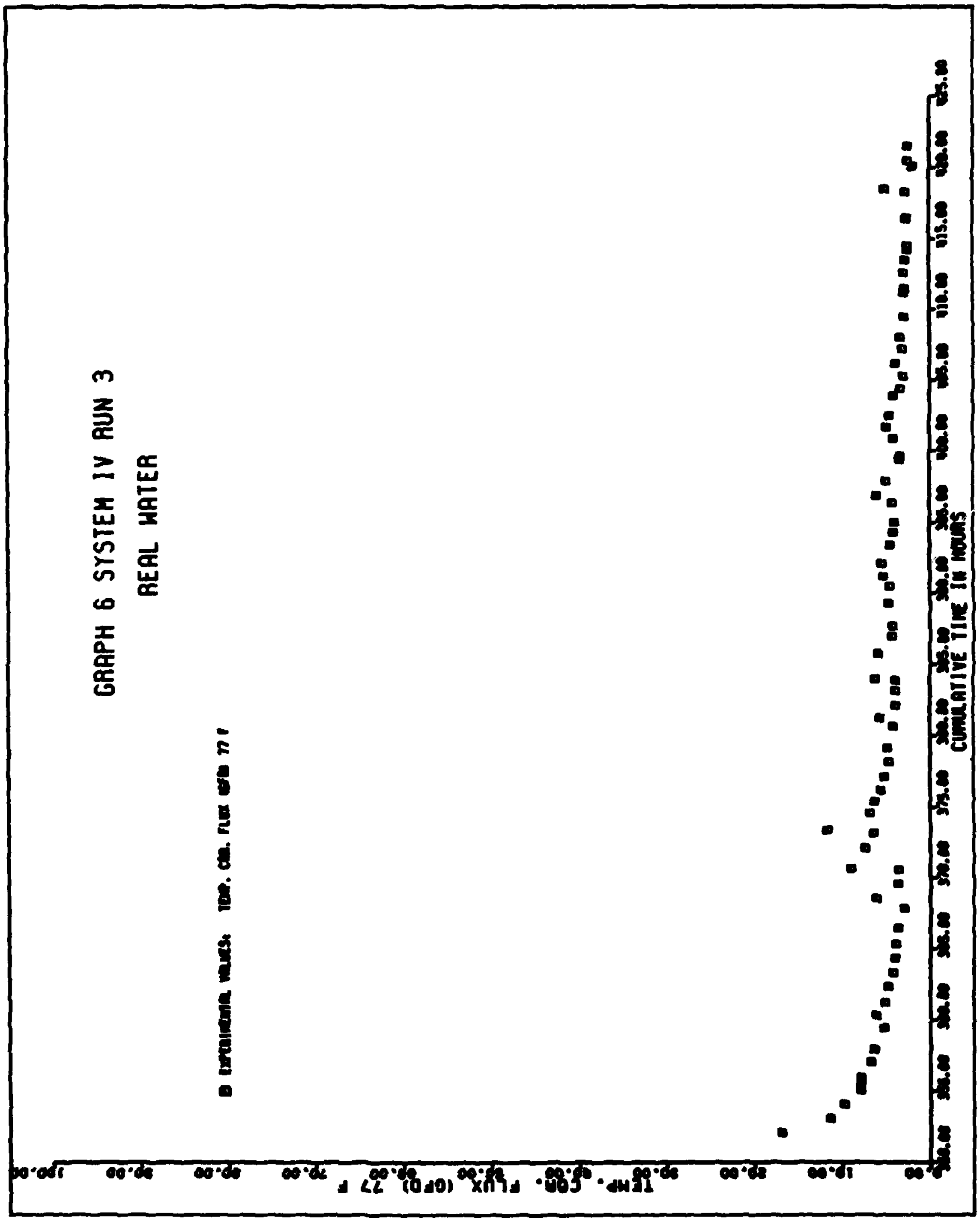


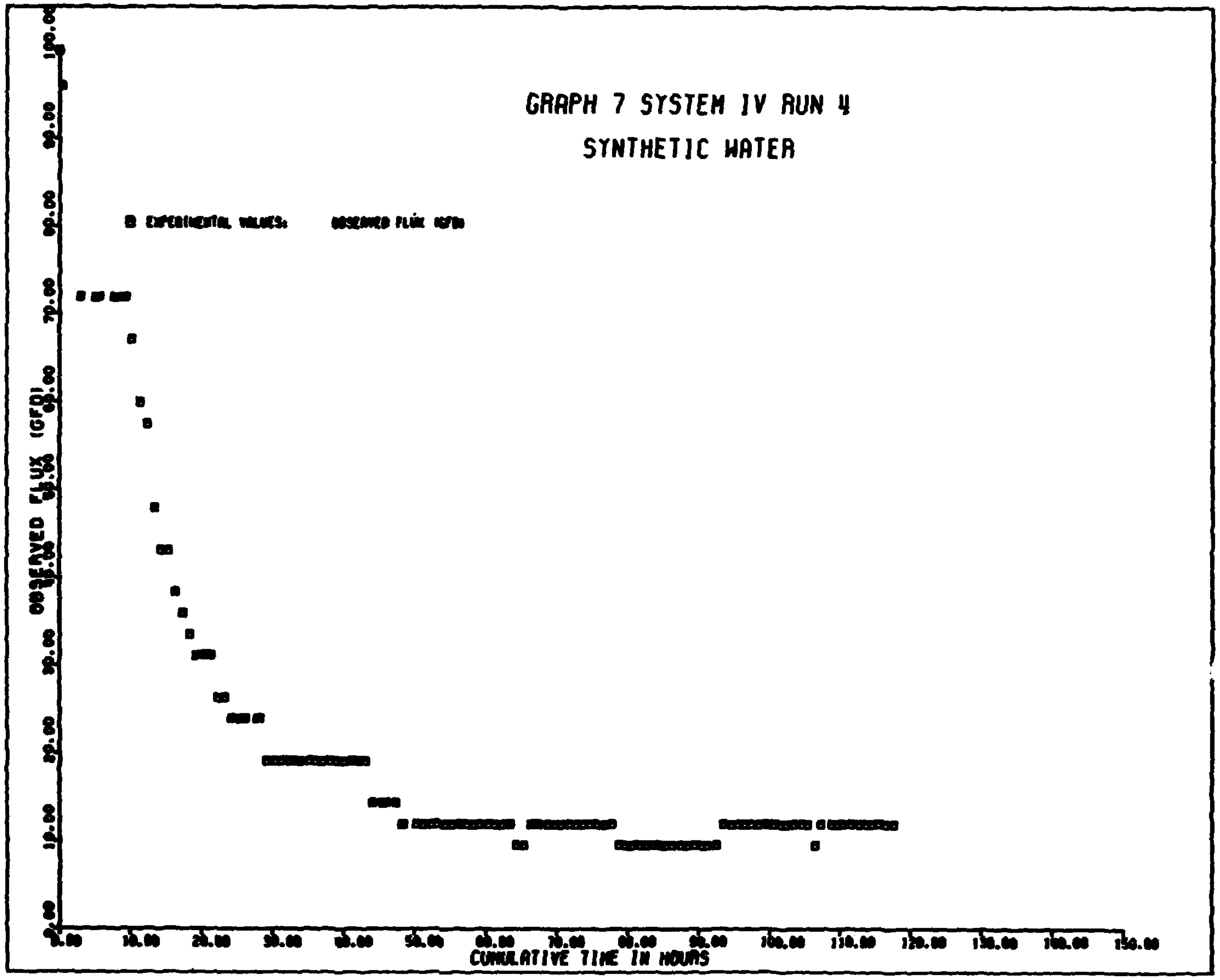
GRAPH 2 SYSTEM IV RUN 1
SYNTHETIC WATER
TEMP. COR. FLUX (GFD) 77 F
CUMULATIVE TIME IN HOURS

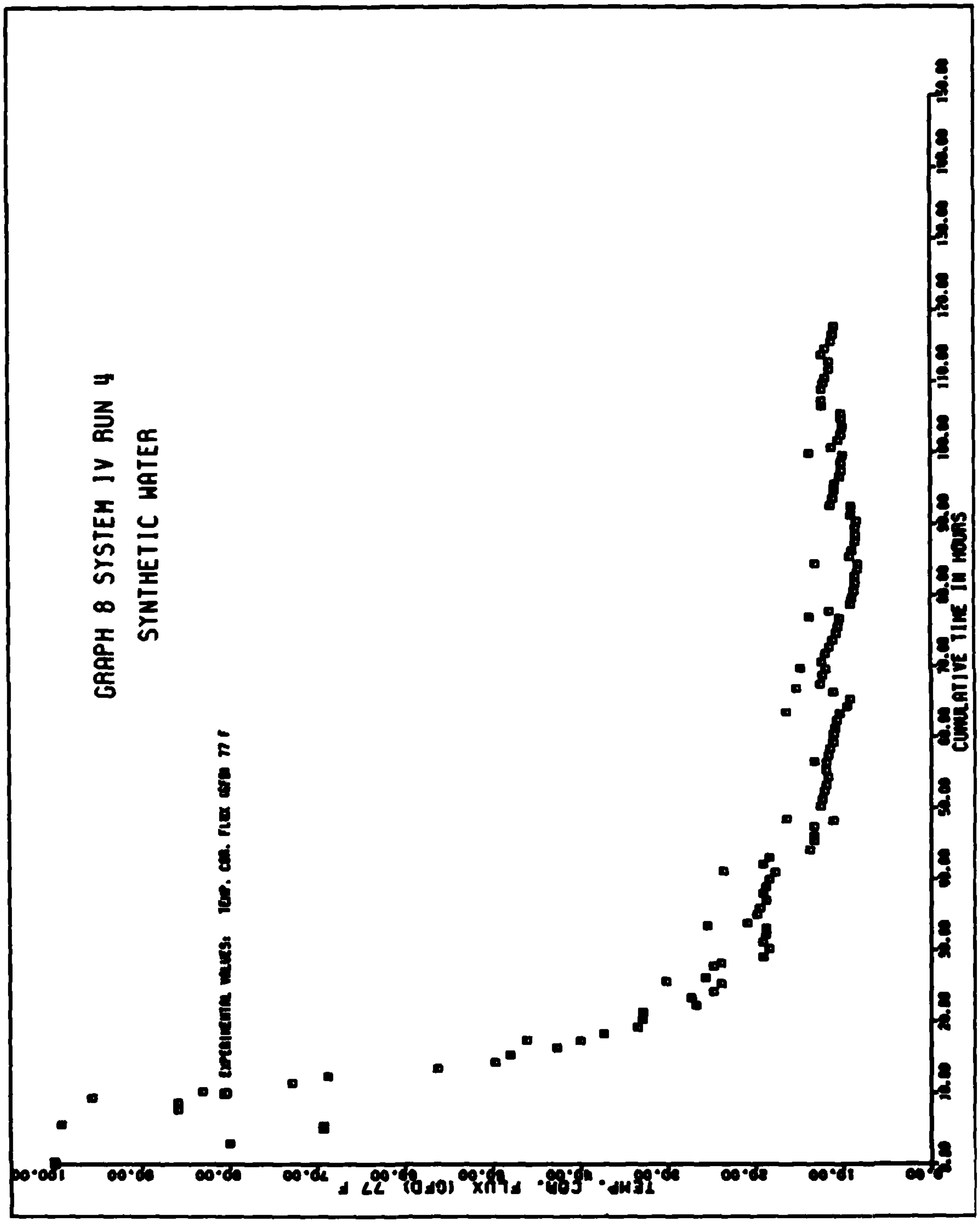

GRAPH 3 SYSTEM IV RUN 2
SYNTHETIC WATER
CUMULATIVE TIME IN HOURS
OBSERVED FLUX (GFD)


GRAPH 4 SYSTEM IV RUN 2
SYNTHETIC WATER
EXPERIMENTAL VALUES: TEMP. COR. FLUX GFD@ 77 F
TEMP. COR. FLUX (GFD) 77 F
0.00
10.00
20.00
30.00
40.00
50.00
60.00
70.00
80.00
90.00
100.00
250.00
257.33
264.67
272.00
279.33
286.67
294.00
301.33
308.67
316.00
323.33
330.67
338.00
345.33
352.67
360.00
CUMULATIVE TIME IN HOURS

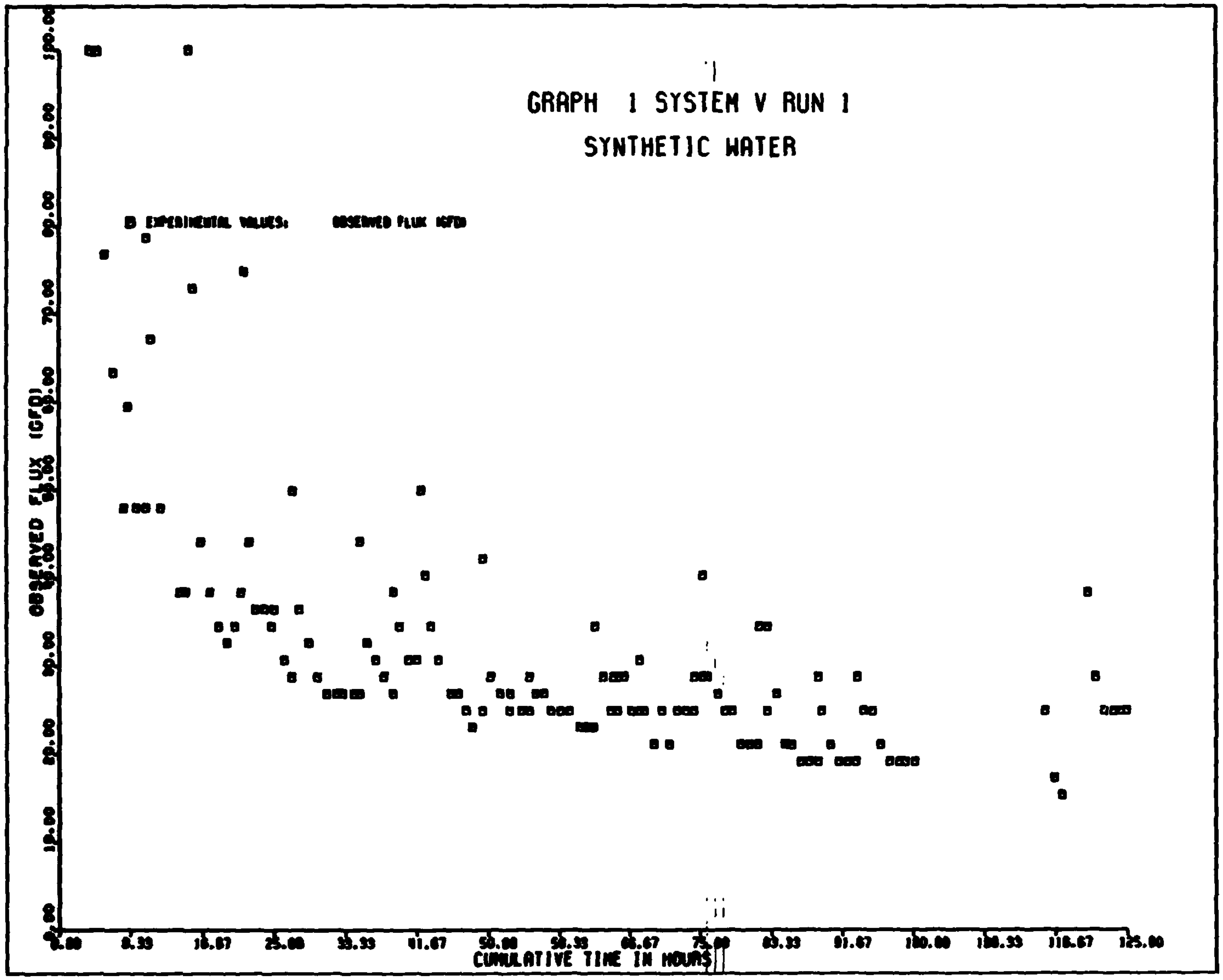

GRAPH 5 SYSTEM IV RUN 3
REAL WATER
EXPERIMENTAL VALUES:
OBSERVED FLUX DATA
CUMULATIVE TIME IN HOURS
OBSERVED FLUX (GFD)

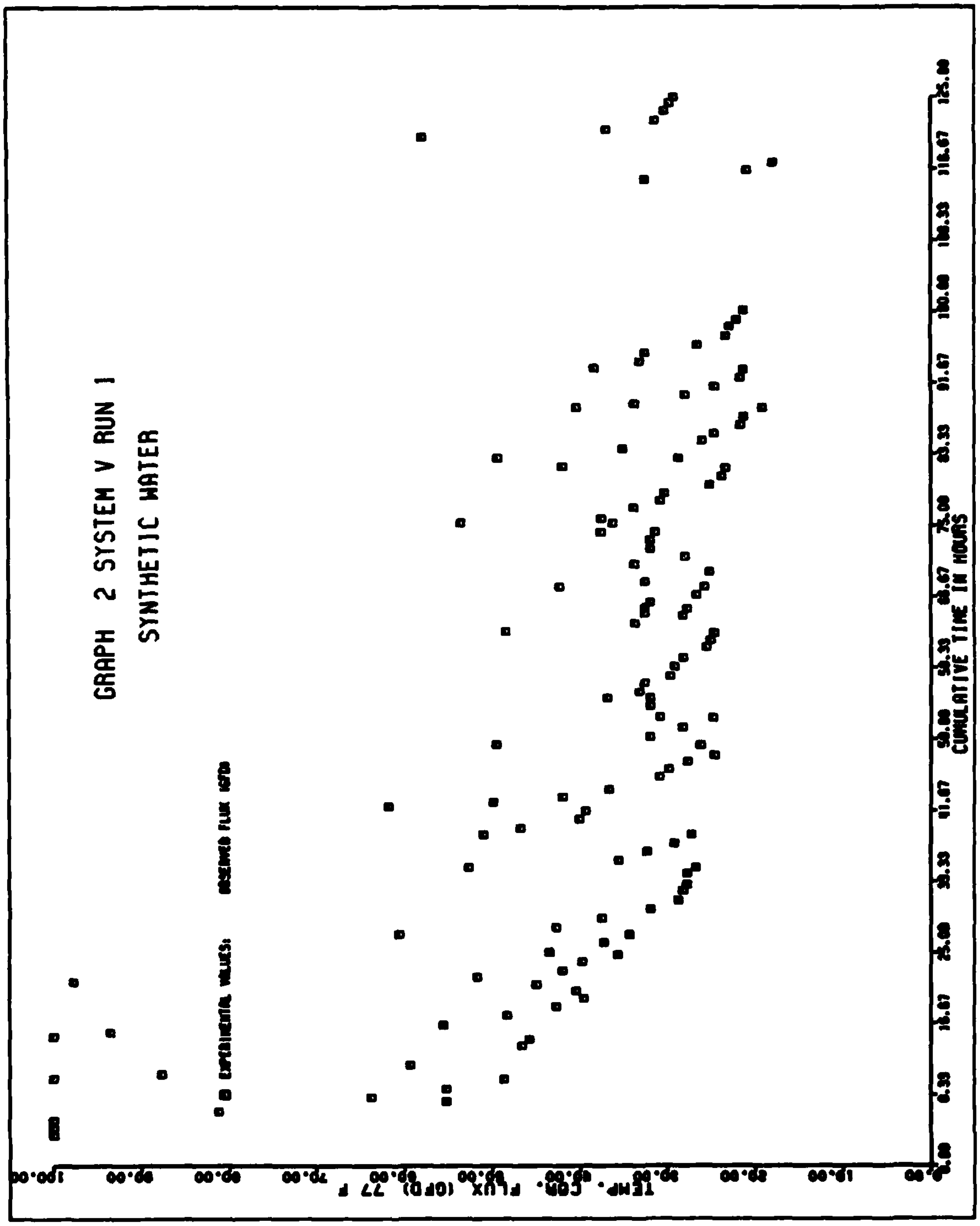

GRAPH 6 SYSTEM IV RUN 3
REAL WATER
EXPERIMENTAL VALUES: TEMP. COR. FLUX GFD 77 F
TEMP. COR. FLUX (GFD) 77 F
CUMULATIVE TIME IN HOURS


GRAPH 7 SYSTEM IV RUN 4
SYNTHETIC WATER
EXPERIMENTAL VALUES: OBSERVED FLUX (GFD)
OBSERVED FLUX (GFD)
0.00
10.00
20.00
30.00
40.00
50.00
60.00
70.00
80.00
90.00
100.00
0.00
10.00
20.00
30.00
40.00
50.00
60.00
70.00
80.00
90.00
100.00
110.00
120.00
130.00
140.00
150.00
CUMULATIVE TIME IN HOURS

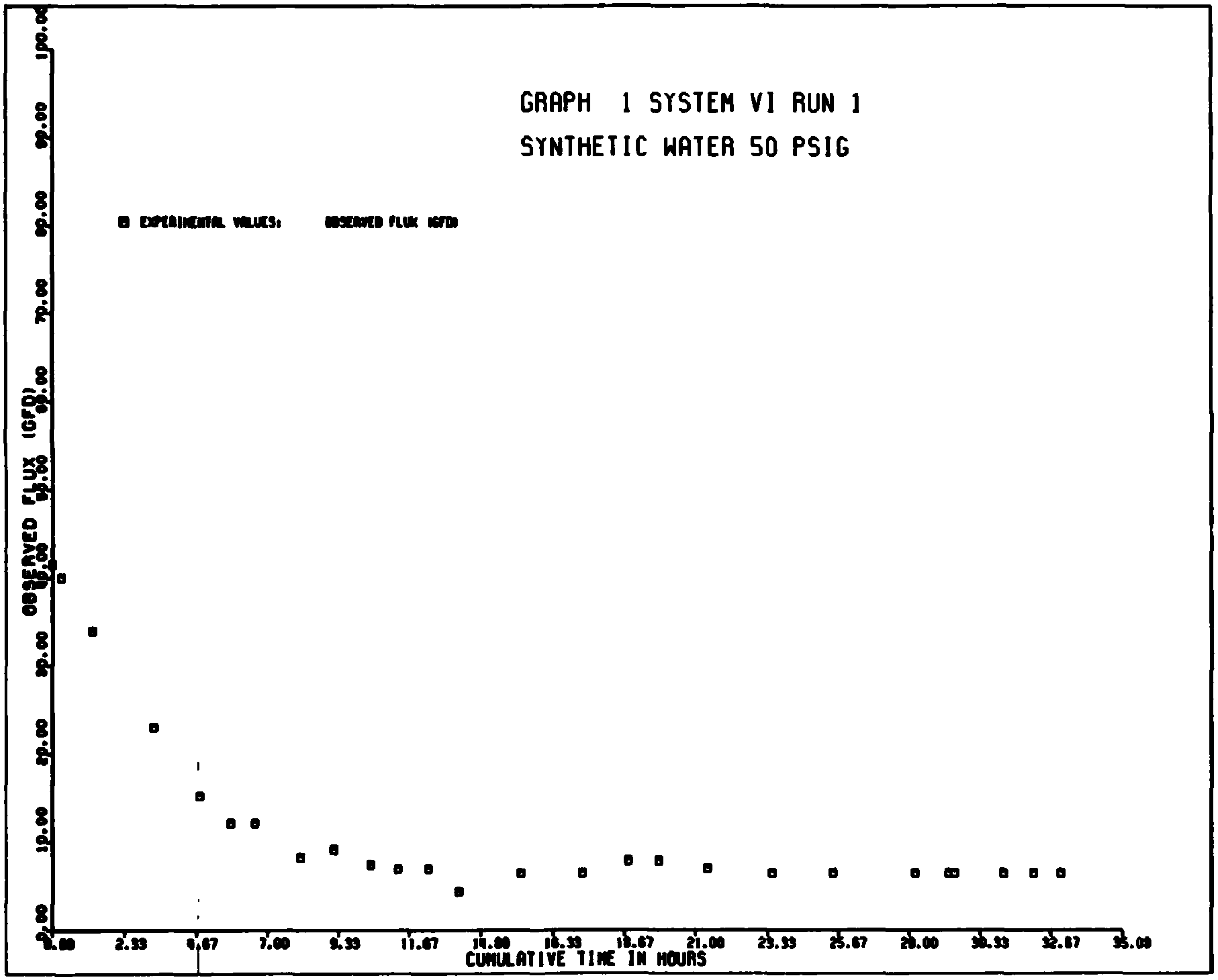

GRAPH 8 SYSTEM IV RUN 4
SYNTHETIC WATER
EXPERIMENTAL VALUES:
TEMP. COR. FLUX (GFD) 77 F
CUMULATIVE TIME IN HOURS
0.00
10.00
20.00
30.00
40.00
50.00
60.00
70.00
80.00
90.00
100.00
110.00
120.00
130.00
140.00
150.00

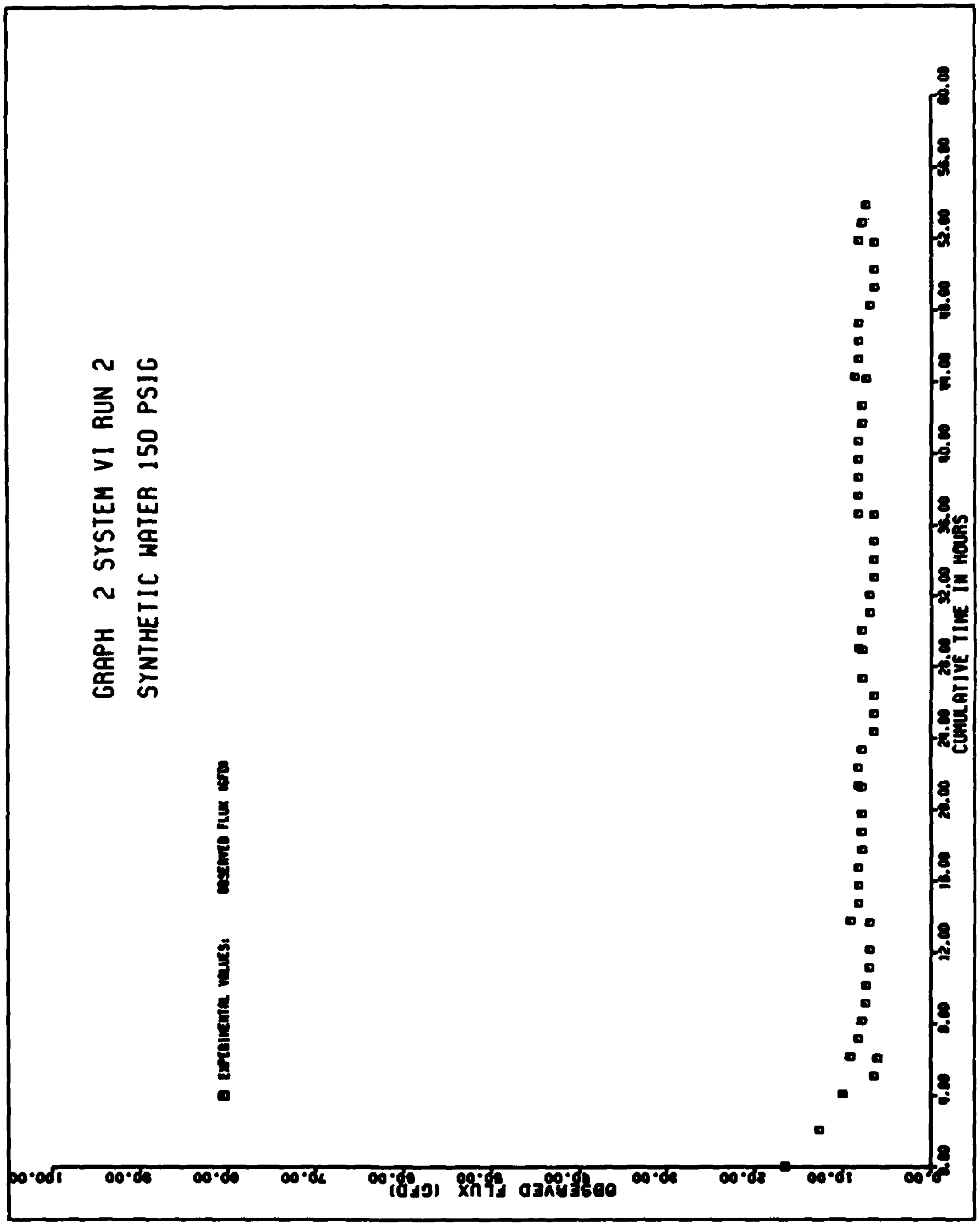
APPENDIX E – Graphs 1-2

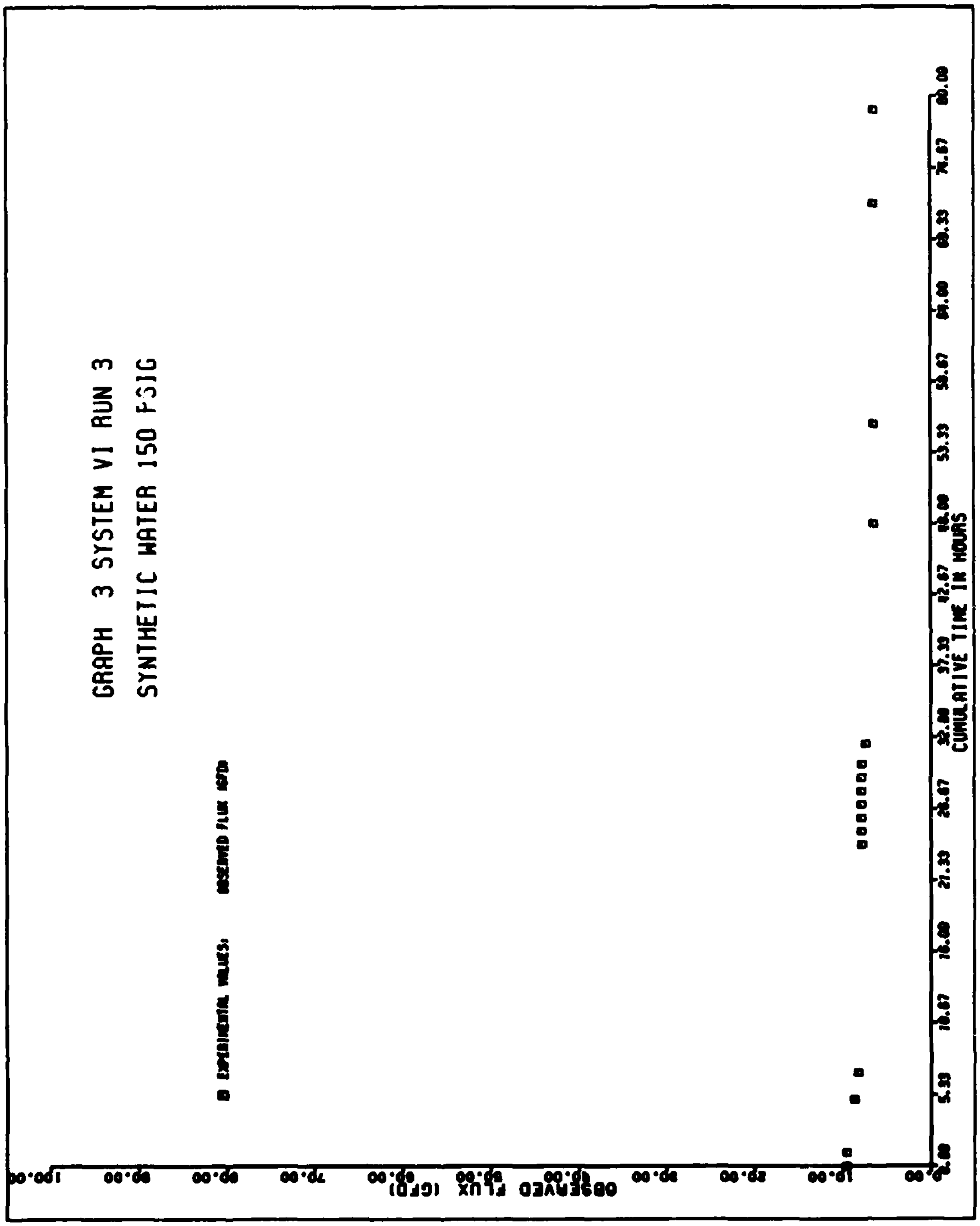



GRAPH 2 SYSTEM V RUN 1
SYNTHETIC WATER
CUMULATIVE TIME IN HOURS
TEMP. COR. FLUX (GFD) 77 F

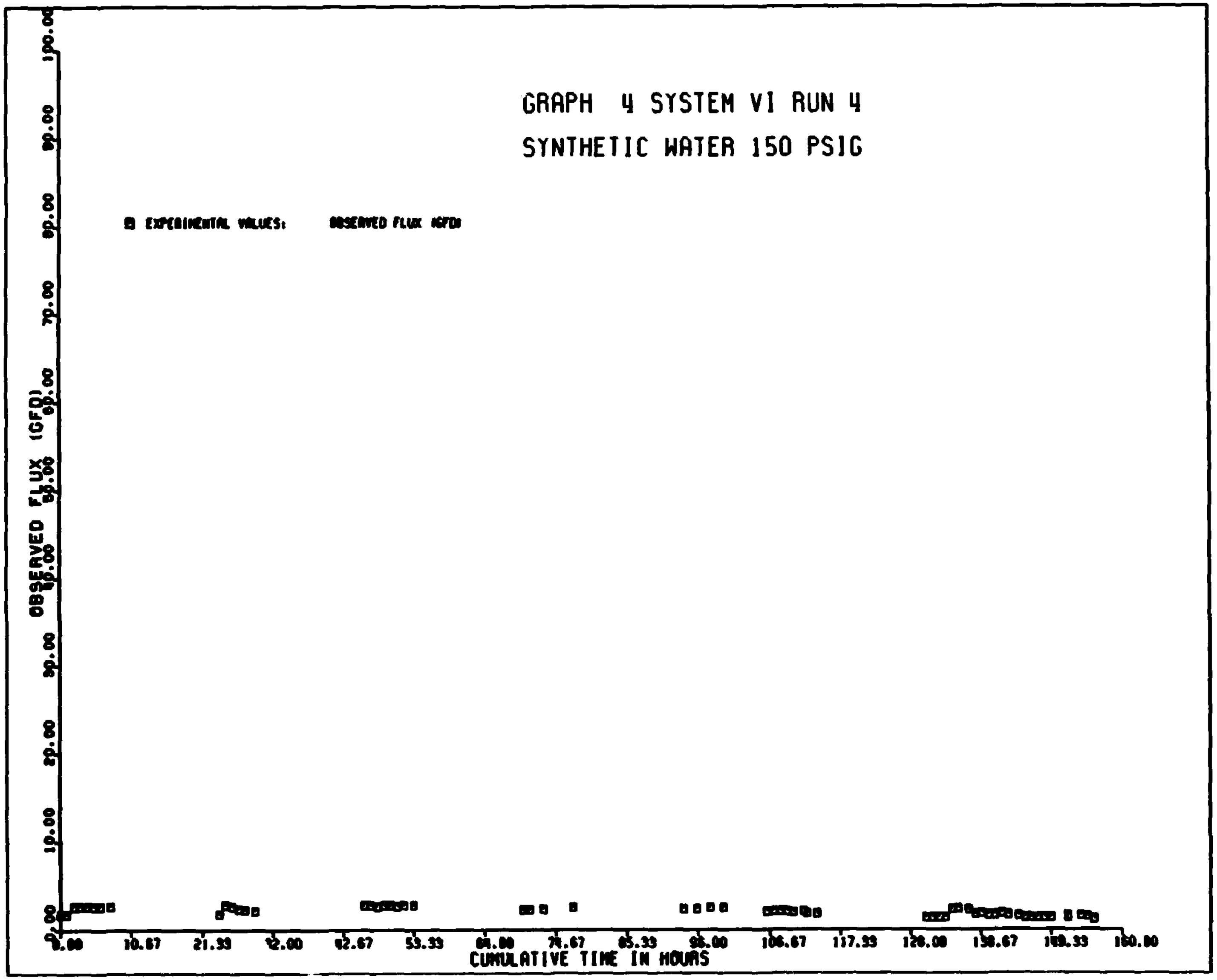
APPENDIX F – Graphs 1-4




GRAPH 2 SYSTEM VI RUN 2
SYNTHETIC WATER 150 PSIG
EXPERIMENTAL VALUES: OBSERVED FLUX (GFD)
OBSERVED FLUX (GFD)
CUMULATIVE TIME IN HOURS



GRAPH 3 SYSTEM VI RUN 3
SYNTHETIC WATER 150 PSIG
EXPERIMENTAL VALUES: OBSERVED FLUX GFD
OBSERVED FLUX (GFD)
0.00
10.00
20.00
30.00
40.00
50.00
60.00
70.00
80.00
90.00
100.00
CUMULATIVE TIME IN HOURS
0.00
5.33
10.67
16.00
21.33
26.67
32.00
37.33
42.67
48.00
53.33
58.67
64.00
69.33
74.67
80.00


GRAPH 4 SYSTEM VI RUN 4
SYNTHETIC WATER 150 PSIG
EXPERIMENTAL VALUES: OBSERVED FLUX (GFD)
OBSERVED FLUX (GFD)
CUMULATIVE TIME IN HOURS
0.00
10.00
20.00
30.00
40.00
50.00
60.00
70.00
80.00
90.00
100.00
10.67
21.33
32.00
42.67
53.33
64.00
74.67
85.33
96.00
106.67
117.33
128.00
138.67
149.33
160.00

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