DOE/CS/40213-T1

TECHNICAL AND ECONOMIC FEASIBILITY OF MEMBRANE TECHNOLOGY

TECHNICAL PROGRESS REPORT

June 17, 1980 - September 16, 1980

Beet Sugar Development Foundation
P. O. Box 1546
Fort Collins, CO 80522

Andrew Sandre

Report #BSDF 39-4Q-80
The information reported herein has been developed under sponsorship of the U. S. Department of Energy, San Francisco Operations Office. Contract No. DE-AC03-79CS40213.

NOTICE

Available from:

Department of Energy
Technical Information Center
P. O. Box 62
Oak Ridge, TN 37830

BSDF-39-4Q-80

TECHNICAL AND ECONOMIC FEASIBILITY OF MEMBRANE TECHNOLOGY

FOURTH TECHNICAL QUARTERLY PROGRESS REPORT

FOR THE PERIOD

JUNE 17, 1980 - SEPTEMBER 16, 1980

ANDREW M. SANDRE, MANAGER SPECIAL PROJECTS

AMERICAN CRYSTAL SUGAR COMPANY, RESEARCH CENTER, MOORHEAD, MINNESOTA

OCTOBER 1980

WORK PERFORMED UNDER CONTRACT

DE-AC03--79CS40213

BEET SUGAR DEVELOPMENT FOUNDATION

P. O. BOX 1546

FORT COLLINS, CO 80522

TABLE OF CONTENTS

ABSTRACT

OBJECTIVE:

Investigate the potential application of reverse osmosis, ultrafiltration, and electrodialysis to the system of solids concentration in beet sugar process streams.

APPROACH:

All available membranes potentially applicable to the process are being tested for application and durability under typical pH, heat, pressure and recycle conditions.

EXPECTED RESULTS:

Possible reduction of energy requirements for evaporation by 15% - 20%.

PROJECT PHASES:

First Phase (3 months: 9-17-79/12-17-79)

Literature search.

Second Phase (3 months: 12-17-79/3-17-80)

Membrane testing.

Third Phase (20 months: 3-17-80/11-17-80)

Long term on-line testing of membranes and systems.

WORK ACHIEVED DURING THE SECOND PART OF PHASE THREE OF THE PROJECT

The second part of Phase Three has been completed according to schedule, on September 16, 1980.

During this period, more emphasis was put on running reverse osmosis tests with the new prototype machine to select most suitable membranes for our purposes.

The following membranes proved to be adequate, taking also into account their availability on an industrial scale:

A - For the concentration of sugar solutions close to neutrality and at temperatures up to 50° C (especially ion exchange purified juices):

CA-SEPA 97 (Osmonics)
PA-300 (UOP)
FT-30 (Filmtec)

B - For the concentration of sugar solutions at pH 2-9 and temperatures up to 95° C:

PA-300 (UOP)
FT-30 (Filmtec)

However, a 5-10 μ prefiltration of the juice is necessary to prevent mechanical impurities from plugging up the membrane pores. Extensive on-line testing will be carried out during the present beet processing campaign to follow up membrane performances as functions of time.

After trying out the membranes in high pressure cells, they were tested in industrial pressure vessels. All three membrane types proved to be very promising (see tables and graphs).

Some efficient membrane cleaning methods were developed to keep flux and rejection of the used membranes as high as possible. They will also be tried out during the campaign tests on-line, and at regular intervals.

Two semi-industrial reverse osmosis machines have been leased from Osmonics and UOP to test the promising membranes under actual plant conditions. They will be running during the present campaign to collect enough data for an accurate evaluation of membrane performances.

The results obtained with thin juice and ion exchange product under simulated plant conditions with our prototype machine (see tables and graphs) prove that the performances of the high temperature membranes, UOP's PA-300 (Sea & brackish water) and Filmtec's FT-30, are very similar to Osmonics's cellulose acetate SEPA 97 membrane, if not better, with the difference that they can be used at higher temperatures (up to 95° C) and can be cleaned with acidic (to dissolve $CaCO_3$) and alkaline (to hydrolize foulants) solutions.

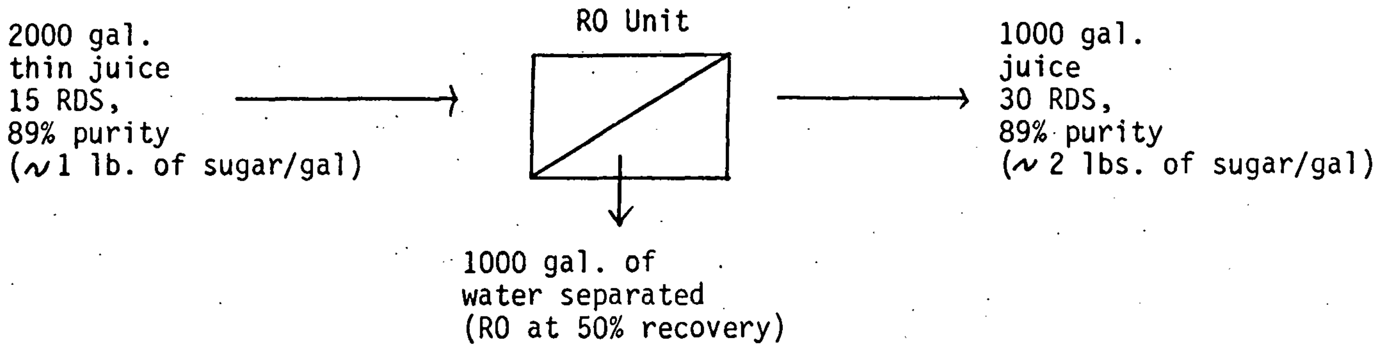
On the other hand, the economics of reverse osmosis applied to the concentration of sea water at room temperature being well established, and assuming that the high temperature membranes will perform well for a time comparable to the lifetime of cellulose acetate membranes (yet to be proved), an attempt was made to predict some reverse osmosis economics in the concentration of thin juice.

The calculations obviously show the advantage of the method and the very high ROI (return on investment), in a time of skyrocketing fuel prices (see following economic analysis). In addition, the water separated from thin juice can be reused in the factory, thus reducing the overall costs of the process.

Conductivity being a measure for the salt rejection performance of a membrane, a development work has been completed to accurately determine the true conductivity of a solution in the presence of sugar, at different RDS. The work proved that sugar molecules have a masking effect on free ions.

PRELIMINARY ECONOMICS OF THIN JUICE CONCENTRATION (up to 30 RDS) BY REVERSE OSMOSIS IN A FACTORY PRODUCING APPROXIMATELY 10^6 GAL/DAY OF THIN JUICE

Considering a sugar and salt rejection of approximately 100% and a thin juice of 15 RDS (usually ~14 RDS):



$$2000 \times 15 = V_1 \times 30$$

$$V_1 = \frac{2000 \times 15}{30} = 1000 \text{ gal. of product \& 1000 gal. of } H_2O \text{ separated}$$

now, to take the RO product up to 60 RDS by evaporation:

$$1000 \times 30 = V_2 \times 60$$

$$V_1 = \frac{1000 \times 30}{60} = 500 \text{ gal. of product \& 500 gal. of } H_2O \text{ separated}$$

or:

2000 gal.	⟶	1000 gal.	⟶	500 gal.
15 RDS	-1000 gal. H_2O	30 RDS	-500 gal. H_2O	60 RDS

Obviously, the first concentration process up to 30 RDS requires the separation of twice the amount of water, as compared to the second step (30 RDS to 60 RDS). Therefore, the impact of a <u>cheaper</u> energy source on the first step is much higher than on the second one. On the other hand, if <u>E energy units</u> are used instead of gallons of water separated, we would get the following equation:

Total energy required (15 RDS ⟶ 60 RDS)

$$\underbrace{2E}_{\text{1st step}} + \underbrace{1E}_{\text{second step}} = 3E$$

And assuming that RO requires ¼ of the evaporation energy for the first concentration step:

$$\therefore \quad \overbrace{\frac{2E}{4}}^{\text{RO instead of evaporation}} + 1E = 1.5\ E$$

thus, half of the energy would be saved in this case (1.5 total energy as compared to 3E).

This, in fact, is the purpose of our investigation. However, RO costs only about 1/10 of evaporation, making the first concentration step even less expensive.

Some variation could be due to changing fuel prices. The following table illustrates the difference of those prices as of September, 1980, in Minnesota:

Lignite:	$\$1/10^6$ BTU
Fuel Oil:	$\$3.40/10^6$ BTU
Gas:	$\$4.84/10^6$ BTU (now $\$5.50/10^6$ BTU)

In case of lignite, the cost of steam produced in the Moorhead plant of American Crystal Sugar Company is 0.14 cts/lb. of steam (quadruple-effect evaporator). To evaporate 1000 Kg of water (1 ton), 500 lbs. of steam are required, at a cost of:

500 x 0.14 cts. = 70 cts.

Now, with a large RO machine (permeation rate of 150 gpm or more), a cost of 70 cts. per 1000 gal. of water separated can safely be considered (Osmonics & UOP communications);

or $\frac{70}{3.8}$ = 18.42 cts. per 1000 kg (1 ton) of water separated (1 gal. of $H_2O \approx 3.8$ Kg),

thus making RO approximately ¼ less expensive than evaporation using lignite as a fuel.

or in other words:

Lignite evaporation	is 3.8 x 1 = 3.8 times more expensive than RO	
Oil "	is 3.8 x 3.40 = 12.92 times more expensive than RO	
Gas "	is 3.8 x 4.84 = 18.39 " " " " "	

As mentioned above, 2000 gal. of the thin juice contain approximately 2000 lbs. of sugar, and the savings achieved by RO could be calculated per ton of sugar processed (concentration of thin juice from 15 RDS to 30 RDS). However, for a better comparison of savings, where different fuels can be used in the evaporation process, daily savings have been calculated for RO vs. evaporation as a first step in water separation and the results have been multiplied by 300 days (continuous work). Of course, actual working days in factories are usually different, as well as the amount of thin juice processed each day. Those variations can then be separately calculated by choosing the correct factor.

To be added is, that RO suppliers consider total operating costs at 70 cts. per 1000 gal. of water separated in a large desalination plant (capacity of approximately 150 gal/min permeation rate and 50% recovery) with a CA membrane and pump life of three years. In our case, and to be on the safe side, we consider a cost of $1.50 per 1000 gal. of permeated water with the FT-30 or PA-300 membranes and under the severe operating conditions in beet sugar plants, in agreement with RO equipment manufacturers. This also includes a possible pretreatment of the sugar solutions. However, for a shorter membrane life, instead of $1.50 roughly $2.00 for two year and $2.50 for one year membrane and pump lifes per 1000 gal. of water separated have been selected.

As an example, the average daily production of thin juice at the Moorhead factory is approximately 10^6 gal. A concentration by RO to 30 RDS would separate ½ x 10^6 gal or 5 x 10^5 gal. of water (two machines with a capacity of approx. 150 - 200 gpm permeate production each and total capital investment of half a million dollars), at a cost of:

for 3-y. membrane & pump life: $\frac{5 \times 10^5}{10^3} \times 1.50$ = \$750 (daily cost)

2-y. " " " " $\frac{5 \times 10^5}{10^3} \times 2$ = \$1000 (daily cost)

1-y. " " " " $\frac{5 \times 10^5}{10^3} \times 2.50$ = \$1250 (daily cost

A quadruple effect evaporation system using lignite as a fuel requires 500 lbs. of steam to evaporate 1000 Kg of water or for 1000 gal. of water:

500 x 3.8 = 1900 lbs. of steam

& for 500,000 gal/day:

1900 x 500 = 95 x 10^4 lbs. of steam at a cost of 0.14 cts/lb of steam.

or total:

95 x 10^4 x 0.14 cts = \$1333

Thus, savings per day with RO vs. different fuels (see also graphs):

a) <u>3-yr. membrane & pump life</u>:

1. RO vs. lignite: 1333 - 750 = \$583
2. RO vs. fuel oil: (1333 x 3.4) - 750 ≅ \$3782
3. RO vs. gas: (1333 x 4.84) - 750 ≅ \$5702

b) <u>2-yr. membrane & pump life</u>:

1. RO vs. lignite: 1333 - 1000 = \$333
2. RO vs. fuel oil: 4532 - 1000 = \$3532
3. RO vs. gas: 6452 - 1000 = \$5452

c) <u>1-yr. membrane & pump life</u>:

1. RO vs. lignite: 1333 - 1250 = \$83
2. RO vs. fuel oil: 4532 - 1250 - \$3282
3. RO vs. gas: 6452 - 1250 = \$5202

And, savings (income) in 300 days would be:

a) 1. \$0.1749 x 10^6	b) 1. \$0.0999 x 10^6	c) 1. \$0.0264 x 10^6
2. \$1.1346 x 10^6	2. \$1.0596 x 10^6	2. \$0.9846 x 10^6
3. \$1.7106 x 10^6	3. \$1.6356 x 10^6	3. \$1.5606 x 10^6

Simple (and basic)* ROI for a capital investment of \$0.5 x 10^6:

a) 1. 35% (22.49%)	b) 1. 20% (14.99%)	c) 1. 5% (5.28%)
2. 227% (118.46%)	2. 212% (110.96%)	2. 197% (103.46%)
3. 342% (176.06%)	3. 327% (168.56%)	3. 312% (161.06%)

And for a capital investment of $\$0.75 \times 10^6$:

a) 1.	23% (16.66%)	b) 1.	13% (11.66%)	c) 1.	4% (3.52%)
2.	151% (80.64%)	2.	141% (75.64%)	2.	131% (70.64%)
3.	228% (119.04%)	3.	218% (114.04%)	3.	208% (109.04%)

And for a capital investment of $\$10^6$:

a) 1.	17% (13.75%)	b) 1.	9% (9.99%)	c) 1.	3% (2.64%)
2.	113% (61.73%)	2.	105% (57.98%)	2.	98% (54.23%)
3.	171% (90.53%)	3.	163% (86.78%)	3.	156% (83.03%)

Simple (and Basic)* pay-back period in yrs. for 300 working days a year. For a capital investment of $\$0.5 \times 10^6$:

a) 1.	2.86 yrs. (4.5 yrs.)	b) 1.	5 yrs. (6.7 yrs.)	c) 1.	18.94 yrs. (18.9yrs)
2.	0.44 yrs. (0.8 yrs.)	2.	0.47 yrs. (0.9 yrs.)	2.	0.51 yrs. (1.0 yrs)
3.	0.29 yrs. (0.6 yrs.)	3.	0.31 yrs. (0.6 yrs.)	3.	0.32 yrs. (0.6 yrs)

For a capital investment of $\$0.75 \times 10^6$:

a) 1.	4.35 yrs. (6 yrs.)	b) 1.	7.69 yrs. (8.6 yrs.)	c) 1.	28.41 yrs. (28.4yrs)
2.	0.66 yrs. (1.2 yrs.)	b) 2.	0.71 yrs. (1.3 yrs.)	2.	0.76 yrs. (1.4 yrs)
3.	0.44 yrs. (0.8 yrs.)	3.	0.46 yrs. (0.9 yrs.)	3.	0.48 yrs. (0.9 yrs)

For a capital investment of $\$10^6$:

a) 1.	5.88 yrs. (7.3 yrs.)	b) 1.	11.08 yrs. (10 yrs.)	c) 1.	37.88 yrs. (37.9yrs)
2.	0.88 yrs. (1.6 yrs.)	2.	0.95 yrs. (1.7 yrs)	2.	1.02 yrs. (1.8 yrs)
3.	0.58 yrs. (1.1 yrs.)	3.	0.61 yrs. (1.2 yrs)	3.	0.64 yrs. (1.2 yrs)

*Simple ROI has been calculated considering $1.50, $2.00, and $2.50 total cost for the separation of 1000 gal. of water (safe figures for 3-yr, 2-yr, and 1-yr membrane pump-life expectancies plus labor), as compared to the 70¢ indicated in the attached table (from: Wastewater Treatment and Separation Methods by R. P. Ouellette et al, Ann Arbor Science, and private communications).

Basic ROI has been figured out by the standard method, considering savings per year with RO as an income with no built-in depreciation, tax, investment tax credit, energy savings tax credit, or cost of money (tax will vary depending on private company or cooperative tax status).

As an example, for the calculation of the Basic ROI in case of an income (savings) of $\$0.1749 \times 10^6$ per 300 working days and a plant cost of $500,000:

Estimated plant cost	$500,000
Depreciation (10 years, straight line)	$ 50,000
Income (see tables)	$174,900
Income Minus Depreciation	$124,900
Less 50% Tax	$ 62,450
Cash Flow	$112,450
After Tax ROI	22.49%
Payback Period	4.5 yrs.

TABLES AND GRAPHS

(ECONOMIC ANALYSIS)

Some representative R.O. costs are shown in the following table.

R.O. suppliers consider a cost of 70¢ per 1000 gal. of water separated (instead of 55.3¢ or 47.5¢ as indicated in the table) in a desalination process using cellulose acetate membranes. However, we calculated our costs using $1.50 per 1000 gal. of water permeated for membranes different than cellulose acetate, and to be very much on the safe side. (Simple ROI, including labor, energy required, depreciation, tax, etc.)

Basic ROI has been calculated by the standard method as mentioned above.

Energy Consumption of Major Concentration Processes

Energy Required for Withdrawal of 1000 kg Water by Various Concentration Processes

Process for a Concentration of 10 wt % Solution to 35 wt %	kWh
Ultrafiltration	
Efficiency[a] 75%	0.674
Pressure 2.5 atm	
Reverse Osmosis	
Efficiency 75%	18.87
Pressure 75 atm	
Freeze Concentration	
Efficiency 80%	
ΔT (Difference Between Condensor and Evaporator) = 20° C	60.5
ΔT = 40° C	132.0
ΔT = 60° C	260.0
Pervaporation	
Efficiency 90%	746.0
Evaporation	
Efficiency 90%	
Single Effect (Without Aroma Recovery; Quadruple Effect = (205.4 Kwh))	746.0
Double Effect (Without Aroma Recovery; Quadruple Effect = (205.4 Kwh))	373.5
Triple Effect (Without Aroma Recovery; Quadruple Effect = (205.4 Kwh))	249.4
Single Effect (With 90% Aroma Recovery by Distill·α_{aw}=4)	845.0
Double Effect (With 90% Aroma Recovery by Distill·α_{aw}=4)	471.0
Triple Effect (With 90% Aroma Recovery by Distill·α_{aw}=4)	343.5

Competing Costs for Separation Processes

Process	Cost in $/1000 Gallons of Water Removal
Ultrafiltration and Reverse Osmosis	0.20-5.00
Vacuum Evaporation	0.40-15.00[a]
Drum-Drying	25.00
Spray-Drying	50.00
Freeze-Drying	200.00-300.00
Freeze Concentration	0.45-45.00
Selective Precipitation	500.00
Dialysis	640.00-1000.00
Gel Filtration	20.00-100.00
Centrifugation	0.30-10.00
Electrodialysis	0.20-5.00

Energy Consumption and Capital Cost Comparison for Material Separation Processes

Process	Energy Consumption	Capital Costs
Electrodialysis	5-7 kWh/1000 ppm dissolved solids removed per 1000 gal	$500-1500/gpm; $15-25/ft^2 of membrane
Reverse Osmosis	7-12 kWh/1000 gal brackish water	$15-30 ft^2 of membrane
Ultrafiltration	3-6 kWh/1000 gal	$300-600/gpm

Cost Data for a 50-mgd Electrodialysis Plant

plant life	30 years
membrane life	3-5 years
steam factor	90%
fixed charge rate @3.25%	5.27%
insurance	0.25%
brine disposal	$0.02/kgal
power cost	$0.007-$0.01/kWh
Influent TDS	3638 mg/l

Representative Costs of RO Operation

Cost Component	Treatment Cost (¢/1000 gal) 1	10	100 mgd
(Capital cost, 10^6 dollars)	(0.50)	(3.83)	(28.75)
Capital recovery, 8%, 25 yr	14.9	10.9	8.2
Power*	9.5	9.5	9.5
Chemicals	2.5	2.4	2.4
Operation and maintenance	5.0	1.8	0.6
Membrane replacement:			
2-yr life	23.4	14.6	10.5
3-yr life	15.6	9.7	7.0
Total operating cost:			
2-yr life membrane	55.3 } In our	39.2	31.2
3-yr life membrane	47.5 } case:70¢ for a 3-yr M&P life	34.3	27.7

Total Processing Costs for Ultrafiltration

Plant Opening Capacity (gpd)	Application	Cost in $/1000 gal of Ultrafiltrate
> 500,000	Low pressure UF of water and/or effluent waste streams	0.10-0.40
10,000-500,000	Food, drug and fine chemicals processing (corrosion-resistant sterilizable equipment)	1.00-5.00

* Equivalent to 1.5¢ per kWh (industrial rate)

Graph #1 shows the daily savings achieved if RO is used instead of evaporation...10^6 gal/day thin juice from 15 to 30 RDS. Savings on oil or gas are much more significant than on lignite. The effect of membrane/pump life expectancies is not as high as previously thought (see also table showing the different fuel prices).

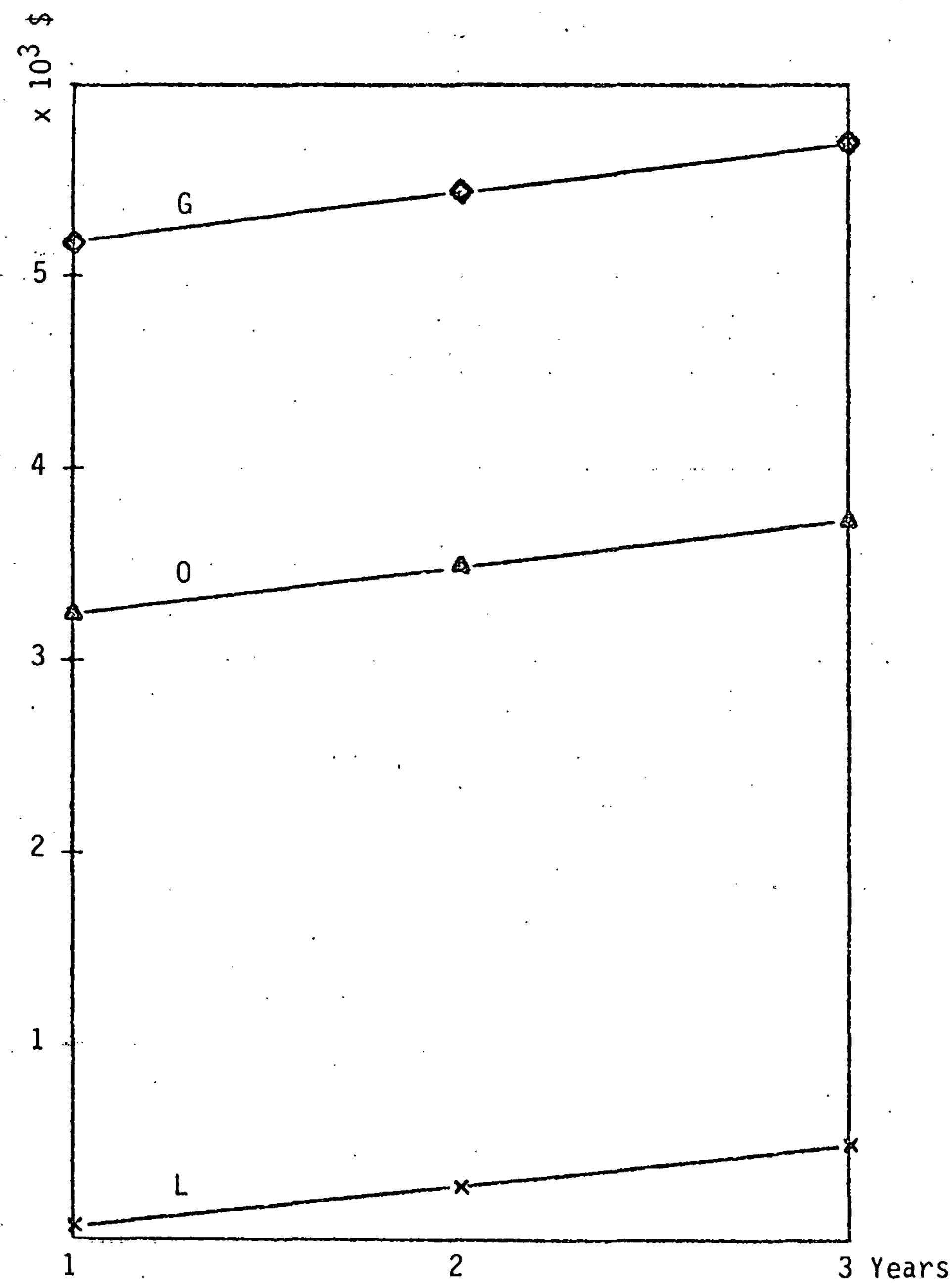
Graphs #2 and 3 show the simple return on investment (ROI) for a reverse osmosis plant treating 10^6 gal/day of thin juice as compared to evaporation by lignite, oil or gas. Returns are quite high in case of oil or gas as a fuel. Membrane/pump life expectancies are insignificant.

Graph #4 illustrates the simple payback period as a function of capital investment for the same purpose as mentioned above. If oil or gas is used in evaporation, payback periods are well below one year (300 working days), and again membrane/pump life expectancies have little effect.

Graph #5 shows the relative simple payback period if the plant is not working three hundred days per year or is not treating less than 10^6 gal/day. Intermediate data can, of course, be extrapolated by using a factor to correct the standard curve (10^6 gal/day).

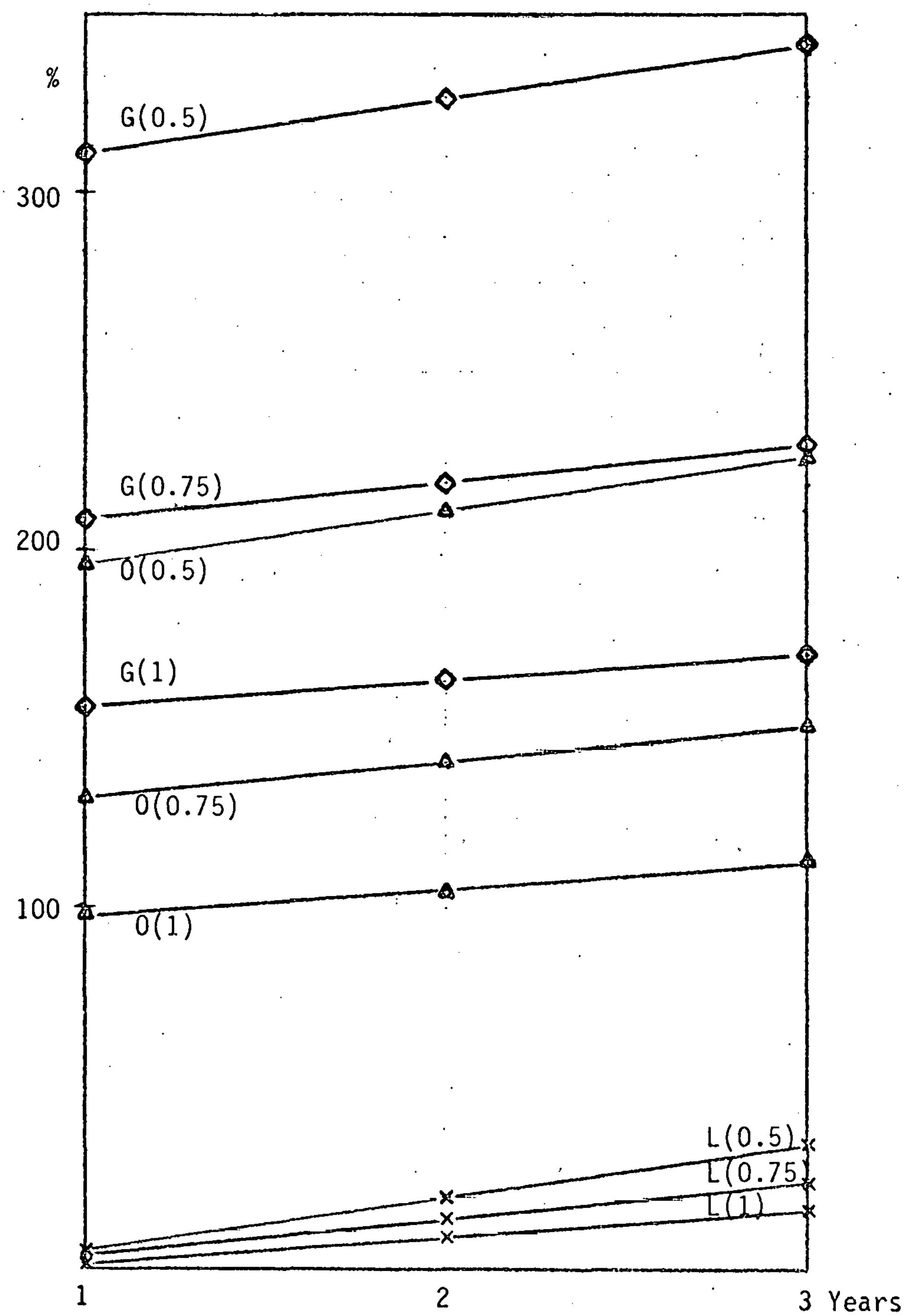
GRAPH #1

Daily Savings As A Function of Membrane/Pump-Life Expectancy. Savings with RO vs. Lignite (L) Oil (O), or Gas (G) as Fuel -- Concentration 15 to 30 RDS & 10^6 gal/day treated thin juice.



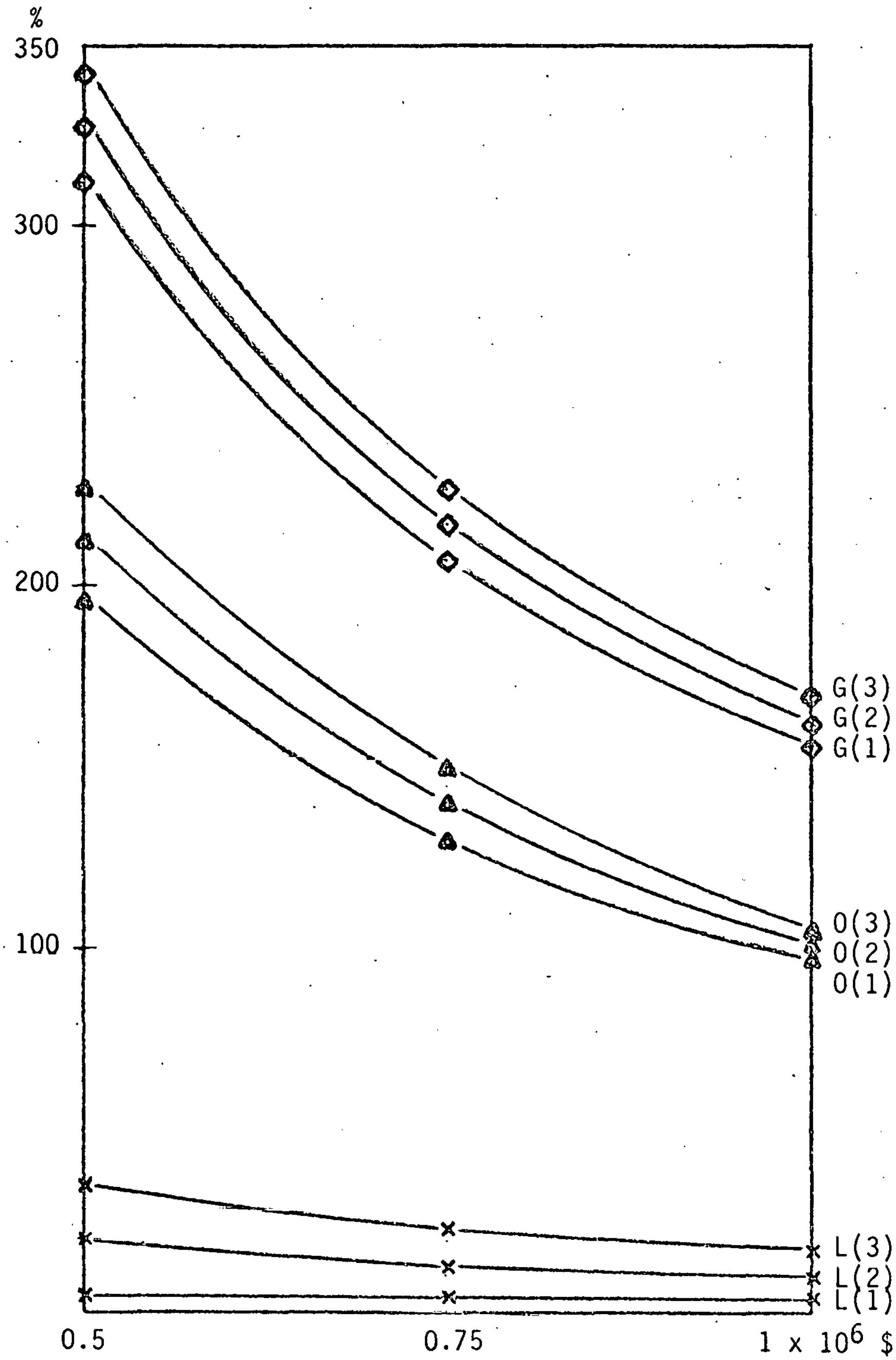
<u>GRAPH #2</u>

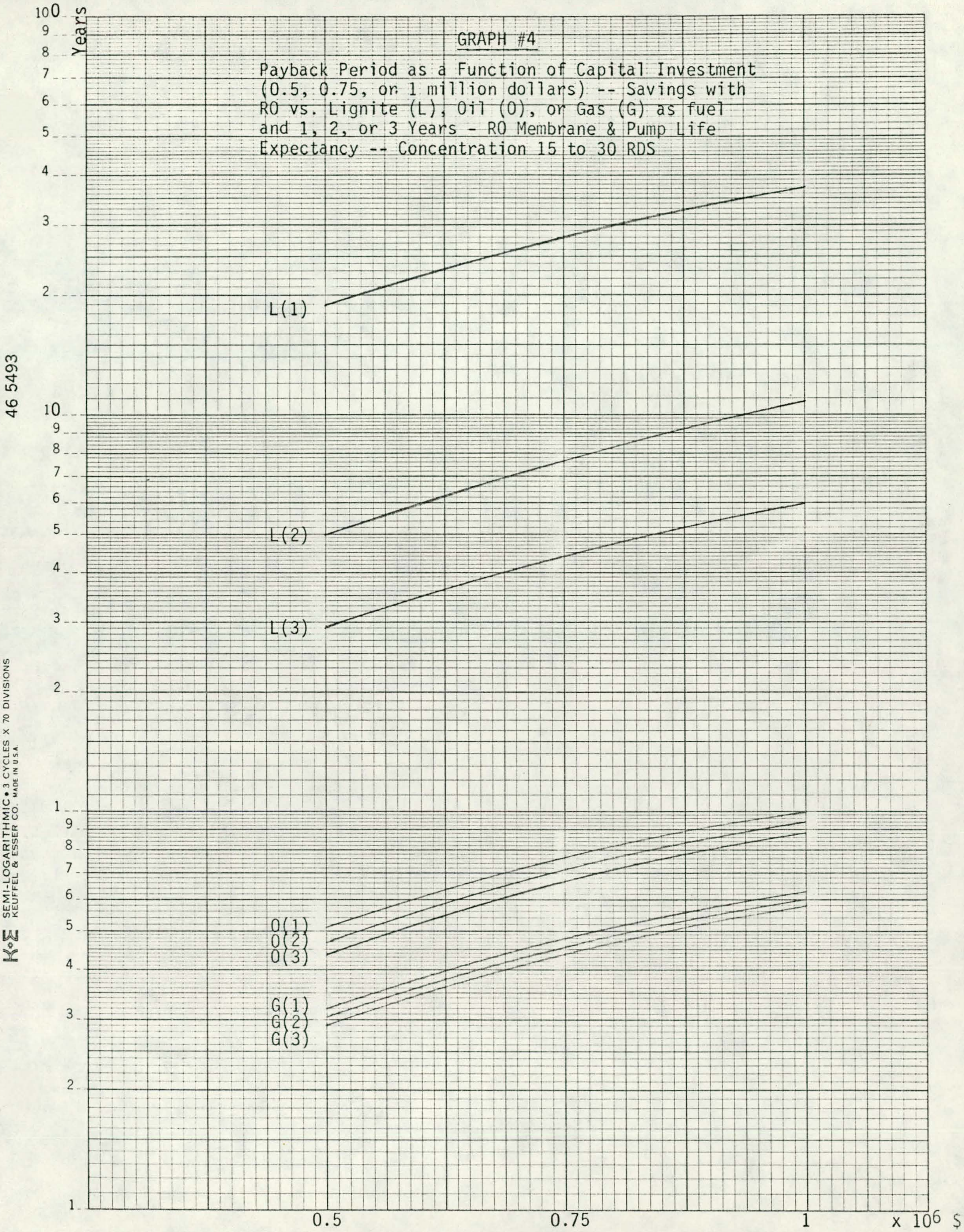
ROI as a Function of RO Membrane/Pump Life Expectancy-Savings with RO vs. Lignite (L), Oil (O), or Gas (G) as fuel and 300 W.D. - 1, 2, 3-year Membrane/Pump Life Expectancy -- 0.5, 0.75, 1 Million Dollars Capital Investment -- Concentration 15 to 30 RDS of 10^6 Gal. Treated Thin Juice



GRAPH #3

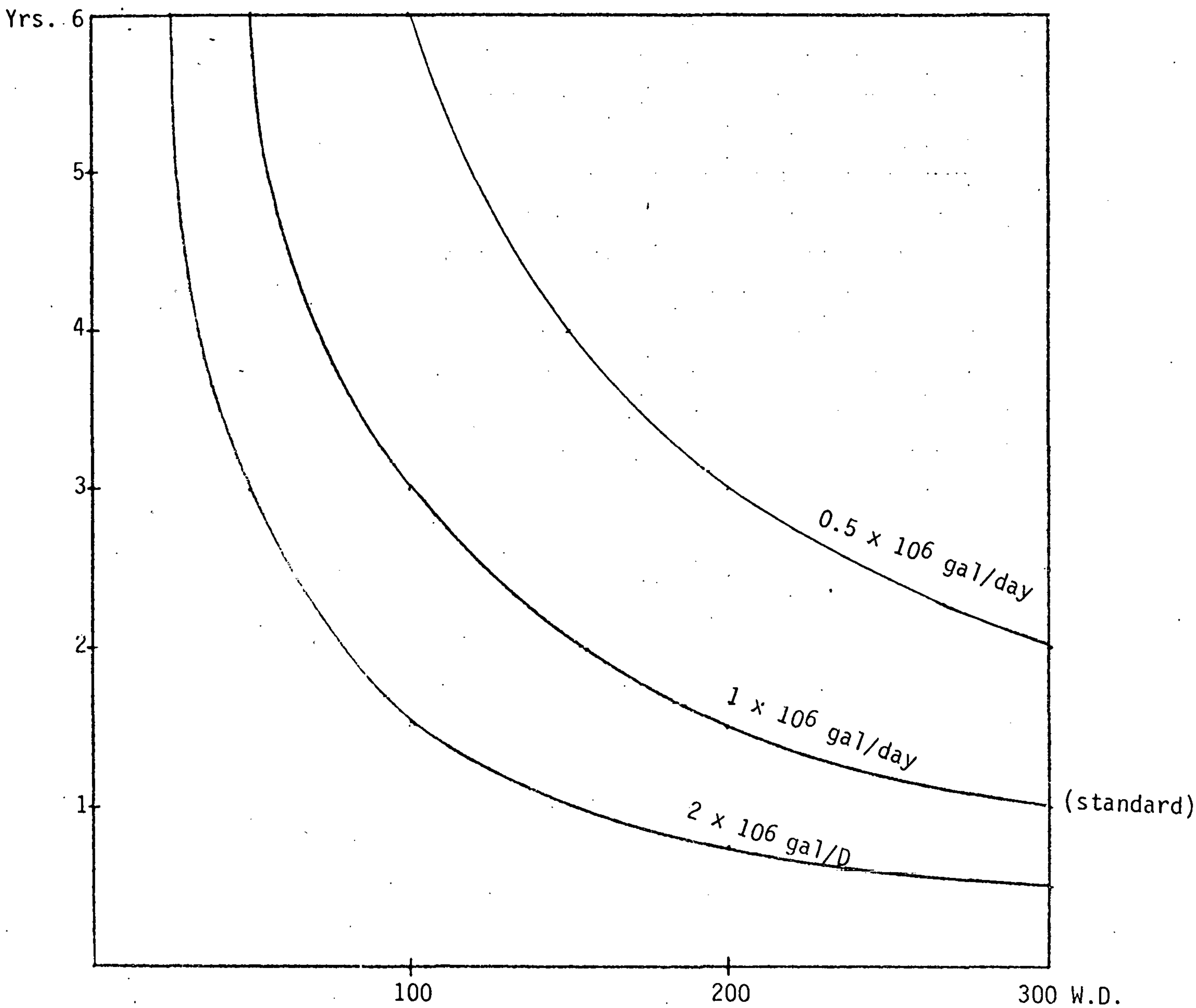
ROI as a Function of Capital Investment-Savings with RO vs. Lignite (L), Oil (O), or Gas (G) as Fuel and 300 W.D. - 1, 2 and 3-year Membrane & Pump Life Expectancy - Concentration 15 to 30 RDS




GRAPH #4
Payback Period as a Function of Capital Investment (0.5, 0.75, or 1 million dollars) -- Savings with RO vs. Lignite (L), Oil (O), or Gas (G) as fuel and 1, 2, or 3 Years - RO Membrane & Pump Life Expectancy -- Concentration 15 to 30 RDS
Years
L(1)
L(2)
L(3)
O(1)
O(2)
O(3)
G(1)
G(2)
G(3)
0.5
0.75
1
x 10^6 $
46 5493
SEMI-LOGARITHMIC • 3 CYCLES X 70 DIVISIONS
KEUFFEL & ESSER CO. MADE IN U.S.A.

GRAPH #5

Relative Payback Period as a Function of Actual Working Days -- In case of a One-Year Maximum Payback Period of 300 W.D. and Different Amounts of Treated Thin Juice



EXPERIMENTAL DATA

(Tables & Graphs)

TABLE I

Flux vs. RDS at different temperatures with simulated thin juice using first Filmtec Ft-30, UOP Seawater, and UOP Brakish water

MEMBRANE TYPE	TEMP. °C	RDS	FLUX L/HR./FT^2	SUCROSE REJECTION %
First Filmtec Ft-30	25	10	2.19	99.80
First Filmtec Ft-30	50	10	4.23	99.6
First Filmtec Ft-30	75	10	5.38	99.75
First Filmtec Ft-30	95	10	5.77	99.85
UOP Seawater	25	10	1.69	99.99
UOP Seawater	50	10	3.0	99.99
UOP Seawater	75	10	3.85	99.98
UOP Seawater	95	10	3.85	99.99
UOP Brakish	25	10	4.2	99.78
UOP Brakish	50	10	5.77	99.73
UOP Brakish	75	10	7.69	99.63
UOP Brakish	95	10	7.31	99.69
First Filmtec Ft-30	25	15	1.12	98.58
First Filmtec Ft-30	50	15	1.85	99.38
First Filmtec Ft-30	75	15	2.5	99.37
First Filmtec Ft-30	95	15	3.15	98.54
UOP Seawater	25	15	1.0	99.89
UOP Seawater	50	15	1.77	99.82
UOP Seawater	75	15	2.38	99.94
UOP Seawater	95	15	2.35	99.93
UOP Brakish	25	15	1.62	99.48
UOP Brakish	50	15	3.0	99.74
UOP Brakish	75	15	4.62	99.6
UOP Brakish	95	15	4.92	99.66
First Filmtec Ft-30	25	20	0.62	99.33
First Filmtec Ft-30	50	20	0.92	99.39
First Filmtec Ft-30	75	20	1.38	99.3
First Filmtec Ft-30	95	20	1.85	99.22
UOP Seawater	25	20	0.69	99.99
UOP Seawater	50	20	1.31	99.99
UOP Seawater	75	20	1.85	99.98
UOP Seawater	95	20	1.85	99.99
UOP Brakish	25	20	1.69	99.91
UOP Brakish	50	20	2.73	98.69
UOP Brakish	75	20	3.38	98.45
UOP Brakish	95	20	3.58	98.54
First Filmtec Ft-30	25	25	0.38	98.58
First Filmtec Ft-30	50	25	0.65	98.24
First Filmtec Ft-30	75	25	0.92	98.25
First Filmtec Ft-30	95	25	1.19	99.18

TABLE I (con't)

Flux vs. RDS at different temperatures with simulated thin juice using first Filmtec Ft-30, UOP Seawater, and UOP Brakish water

MEMBRANE TYPE	TEMP.	RDS	FLUX L/HR./FT^2	SUCROSE REJECTION %
UOP Seawater	25	25	0.85	99.61
UOP Seawater	50	25	1.23	99.91
UOP Seawater	75	25	1.69	99.91
UOP Seawater	95	25	1.62	99.99
UOP Brakish	25	25	1.46	99.62
UOP Brakish	50	25	2.15	99.63
UOP Brakish	75	25	2.62	99.31
UOP Brakish	95	25	2.65	99.74
First Filmtec Ft-30	25	30	0.23	98.40
First Filmtec Ft-30	50	30	0.31	97.89
First Filmtec Ft-30	75	30	0.5	99.38
First Filmtec Ft-30	95	30	0.54	98.83
UOP Seawater	25	30	0.31	99.95
UOP Seawater	50	30	0.46	99.94
UOP Seawater	75	30	0.65	99.99
UOP Seawater	95	30	0.5	99.99
UOP Brakish	25	30	0.73	99.96
UOP Brakish	50	30	0.96	99.98
UOP Brakish	75	30	1.23	99.83
UOP Brakish	95	30	1.08	99.91

TABLE 2

Concentration of IE-Treated thin juice using first SEPA 97 membrane at 800 PSI and 25^{0} and 50^{0}C (113 liters of product concentrated in 3 hours)

MEMBRANE TYPE	TEMP. ^{o}C	RDS	FLUX L/HR./FT^{2}	TIME HOUR	SUCROSE REJECTION %
First SEPA 97	25	9.7	2.21	0	
First SEPA 97	25	12.19	2.12	0.5	99.97
First SEPA 97	25	14.19	1.89	1.0	99.97
First SEPA 97	25	17.76	1.58	1.5	99.97
First SEPA 97	25	21.85	1.26	2.0	99.97
First SEPA 97	25	26.26	0.87	2.5	99.96
First SEPA 97	25	30.34	0.58	3.0	99.95
					SALT REJECTION %
First SEPA 97	50	9.88	2.59	0	95.3
First SEPA 97	50	10.6	2.53	0.5	93.2
First SEPA 97	50	14.0	2.21	1.0	74.6
First SEPA 97	50	21.6	1.42	2.0	66.1
First SEPA 97	50	30.74	0.87	3.0	66.1
First SEPA 97	50	31.18	0.57	3.5	66.1

Sugar Rejections = 99%+

TABLE 3

Apparent and corrected conductivities of solutions in the presence of sugar.

RDS	APPARENT Cond. μmhos	CORRECTED Cond. μmhos	APPARENT ppm NaCl	CORRECTED ppm NaCl
5	3256	3696	1700	1900
10	5632	7392	3000	3940
20	8712	14520	4640	--
30	9592	21824	5480	--
40	8448	29040	4500	--

TABLE 4

Flux of first SEPA 97 membrane at different states

MEMBRANE STATE	FLUX L/HR./FT^2 400 PSI	800 PSI
Original Osmonics Test	1.74	~ 3.2
Our Test of Original	1.26	2.05
After IE Pretreated Test and Rinse	1.1	1.89
After Simulated Thin Juice Concentration Test and Rinse	0.92	1.61
After CS Soap Wash	0.98	1.77
After 10% NaCL Wash	0.89	1.76
After 30 Days	1.01/ Salt Rej: 92.1%	1.74/ Salt Rej: 94.1%

TABLE 5

Flux of first Flimtec Ft-30 membrane at different states (25°C)

STATE OF MEMBRANE	FLUX L/HR./FT^2 400 PSI	800 PSI	SALT REJECTION % 400 PSI	800 PSI
Original	2.72		99.8	
After 24 Hours on-line with 0.2% NaCL	1.77			
After Thin Juice Experiment at 50°C, 75°C, 95°C	1.4	2.5		
After 12D-Wash	1.55	2.6		
After UZ73 Wash	1.55	2.7		
After 1% HCL Wash	1.55	2.75	89.58	92.0
After 1% NaOH Wash	1.65	2.95	98.4	99.23
After Rinsing	1.75	3.1	99.3	99.5
After Thin Juice Experiment at 25°C, 35°C, 50°C, 35°C, 25°C and Concentration at 25°C and 50°C and then rinse	1.25	2.15	98.08	98.25

TABLE 6

Concentration of thin juice, 14 to 17 RDS using first Filmtec Ft-30 at 50°C, 75°C, 95°C,800 PSI.
TEMPERATURE RAISING WHILE CONCENTRATING

MEMBRANE TYPE	TEMP. °C	FLUX L/HR./FT^2	SALT REJ. %	SUGAR REJ. %
First Filmtec Ft-30	50	4.15 at 13.5 RDS	99.1	99.91
First Filmtec Ft-30	75	3.5 at 14.5 RDS	96.41	99.83
First Filmtec Ft-30	95	2.2 at 17 RDS	91.1	99.72

TABLE 7

Comparative salt rejection test using first and second SEPA 97, first Filmtec Ft-30, UOP Seawater, UOP Brakish at 25°C, 400 PSI and 800 PSI

MEMBRANE TYPE	FLUX L/HR./FT^2		SALT REJECTION %	
	400 PSI	800 PSI	400 PSI	800 PSI
First Test				
First SEPA 97	1.01	1.74	92.1	94.1
Second SEPA 97	1.26	2.35	94.4	94.9
First Flimtec Ft-30	1.75	3.1	99.3	99.5
Second Test				
Second SEPA 97	1.12	2.18	92.5	94.3
First Filmtec Ft-30	1.65	2.8	99.0	99.3
Second Filmtec Ft-30	1.85	3.05	99.5	99.5
UOP Seawater	2.69	5.38	98.93	99.17
UOP Brakish	3.18		98.93	

TABLE 8

Comparative membrane test with 0.5% NaCl solution using first and second Filmtec Ft-30, second SEPA 97, and UOP Seawater membranes

MEMBRANE TYPE	TEMP. °C	FLUX L/HR./FT^2 400 PSI	FLUX L/HR./FT^2 800 PSI	SALT REJECTION % 400 PSI	SALT REJECTION % 800 PSI
First Filmtec Ft-30	25	1.5	2.65	98.81	99.1
First Filmtec Ft-30	35	1.8	3.1	98.81	99.05
First Filmtec Ft-30	50	2.3	3.8	98.38	98.91
First Filmtec Ft-30	35	1.85	2.9	98.69	99.29
First Filmtec Ft-30	25	1.5	2.7	98.81	99.21
Second Filmtec Ft-30	25	2.0	3.05	99.28	99.41
Second Filmtec Ft-30	35	2.5	3.9	99.17	99.28
Second Filmtec Ft-30	50	3.0	4.4	99.15	99.28
Second Filmtec Ft-30	35	2.45	3.4	99.17	99.52
Second Filmtec Ft-30	25	1.8	3.1	99.28	99.5
Second SEPA 97	25	1.17	2.19	93.1	95.24
Second SEPA 97	35	1.42	2.53	92.84	94.52
Second SEPA 97	50	1.74	2.75	94.05	95.71
Second SEPA 97	35	1.31	2.21	94.52	96.54
Second SEPA 97	25	1.07	2.05	94.29	95.71
UOP Seawater	25	2.69	5.38	98.81	99.17
UOP Seawater	35	3.39	6.44	98.81	98.81
UOP Seawater	50	4.59	9.35	98.69	98.83
UOP Seawater	35	3.13	6.18	99.15	99.29
UOP Seawater	25	2.52	5.12	98.93	99.29

TABLE 9

Membrane comparative test with thin juice, raising temperature from 25°C to 35°C to 50°C, then lowering temperature to 35°C and 25°C at 800 PSI and 400 PSI using first and second Filmtec Ft-30, second SEPA 97, and UOP Seawater membranes

MEMBRANE TYPE	TEMP. °C	FLUX L/HR./FT^2		SALT REJ. %		SUGAR REJ. %	
		400 PSI	800 PSI	400 PSI	800 PSI	400 PSI	800 PSI
First Filmtec Ft-30	25	0.5	1.45	97.07	99.10	99.41	99.87
First Filmtec Ft-30	35	0.55	1.85	91.54	99.06	98.91	99.87
First Filmtec Ft-30	50	0.75	2.35	85.68	98.5	99.27	99.83
First Filmtec Ft-30	35	0.6	1.77	91.62	98.48	98.74	99.38
First Filmtec Ft-30	25	0.5	1.45	89.33	99.08	97.72	99.82
Second Filmtec Ft-30	25	0.55	1.7	98.71	99.45	99.74	99.91
Second Filmtec Ft-30	35	0.6	2.05	98.95	99.36	99.79	99.93
Second Filmtec Ft-30	50	0.9	2.55	95.59	98.78	99.27	99.91
Second Filmtec Ft-30	35	0.7	2.0	95.93	98.99	99.16	99.85
Second Filmtec Ft-30	25	0.55	1.65	96.35	99.28	99.11	99.82
Second SEPA 97	25	0.32	1.06	90.26	95.99	99.47	99.82
Second SEPA 97	35	0.39	1.29	87.84	93.24	99.42	99.77
Second SEPA 97	50	0.49	1.52	86.51	93.61	98.89	99.12
Second SEPA 97	35	0.32	1.14	88.39	94.84	99.09	99.07
Second SEPA 97	25	0.27	0.95	88.78	95.65	99.17	99.43
UOP Seawater	25	0.58	2.03	95.89	98.87	99.7	99.98
UOP Seawater	35	0.79	2.57	96.72	98.66	99.83	99.96
UOP Seawater	50	0.97	3.53	95.26	98.37	99.79	99.96
UOP Seawater	35	0.71	2.34	97.06	98.72	99.85	99.94
UOP Seawater	25	0.49	1.72	97.78	99.08	99.91	99.97

TABLE 10

Concentration of thin juice at 25°C, 50°C at 800 PSI using first and second Filmtec Ft-30, second SEPA 97, and UOP Seawater

MEMBRANE TYPE	TEMP. °C	RDS	FLUX L/HR./FT^2	SALT REJ. %	SUGAR REJ. %
First Filmtec Ft-30	25	15	1.3	99.05	99.85
First Filmtec Ft-30	25	20	0.95	98.61	99.81
First Filmtec Ft-30	25	25	0.65	97.75	99.74
First Filmtec Ft-30	25	30	0.25	92.44	99.43
Second Filmtec Ft-30	25	15	1.4	99.05	99.79
Second Filmtec Ft-30	25	20	1.0	98.41	99.53
Second Filmtec Ft-30	25	25	0.65	97.02	99.08
Second Filmtec Ft-30	25	30	0.25	88.91	96.95
Second SEPA 97	25	15	0.87	94.88	99.3
Second SEPA 97	25	20	0.6	93.06	99.25
Second SEPA 97	25	25	0.38	89.34	98.61
Second SEPA 97	25	30	0.17	70.03	94.56
UOP Seawater	25	15	1.24	98.81	99.8
UOP Seawater	25	20	0.69	98.14	99.7
UOP Seawater	25	25	0.42	97.01	99.59
UOP Seawater	25	30	0.17	90.68	99.29
First Filmtec Ft-30	50	15	1.5	95.27	99.64
First Filmtec Ft-30	50	20	1.0	96.04	99.69
First Filmtec Ft-30	50	25	0.65	92.84	99.56
First Filmtec Ft-30	50	30	0.3	85.42	99.53
Second Filmtec Ft-30	50	15	1.7	93.55	98.32
Second Filmtec Ft-30	50	20	0.95	92.07	96.27
Second Filmtec Ft-30	50	25	0.5	86.38	94.74
Second Filmtec Ft-30	50	30	0.25	70.17	93.85
Second SEPA 97	50	15	0.71	79.16	99.48
Second SEPA 97	50	20	0.46	78.81	90.66
Second SEPA 97	50	25	0.32	61.17	78.45
Second SEPA 97	50	30	0.22	33.05	69.15
UOP Seawater	50	15	1.76	97.78	99.91
UOP Seawater	50	20	0.93	96.75	99.99
UOP Seawater	50	25	0.49	92.69	99.72
UOP Seawater	50	30	0.19	85.08	99.43

TABLE 11

Concentration of thin juice at 50°C, 75°C, 95°C, 800 PSI with first and second Filmtec Ft-30 membranes

MEMBRANE TYPE	TEMP. °C	RDS	FLUX L/HR./FT^2	SALT REJ. %	SUGAR REJ. %
First Filmtec Ft-30	50	15	1.5	95.27	99.64
First Filmtec Ft-30	50	20	1.0	96.04	99.69
First Filmtec Ft-30	50	25	0.65	92.84	99.56
First Filmtec Ft-30	50	30	0.3	85.42	99.53
Second Filmtec Ft-30	50	15	1.7	93.55	98.32
Second Filmtec Ft-30	50	20	0.95	92.07	96.27
Second Filmtec Ft-30	50	25	0.5	86.38	94.74
Second Filmtec Ft-30	50	30	0.25	70.17	93.85
First Filmtec Ft-30	75	15	1.95	95.87	99.66
First Filmtec Ft-30	75	20	1.3	93.03	99.53
First Filmtec Ft-30	75	25	0.8	90.85	99.41
First Filmtec Ft-30	75	30	0.4	66.52	98.95
Second Filmtec Ft-30	75	15	0.95	93.44	99.35
Second Filmtec Ft-30	75	20	1.3	88.42	99.18
Second Filmtec Ft-30	75	25	0.65	85.01	98.96
Second Filmtec Ft-30	75	30	0.25	57.31	98.74
First Filmtec Ft-30	95	15	1.95	91.75	98.66
First Filmtec Ft-30	95	20	1.3	92.38	99.4
First Filmtec Ft-30	95	25	0.8	83.99	99.18
First Filmtec Ft-30	95	30	0.25	47.79	93.91
Second Filmtec Ft-30	95	15	0.95	90.07	99.19
Second Filmtec Ft-30	95	20	1.3	78.4	98.87
Second Filmtec Ft-30	95	25	0.4	88.62	99.16
Second Filmtec Ft-30	95	30	0.2	38.05	93.19

TABLE 12

Thin juice test using first and second Filmtec Ft-30, and SEPA 97 at 30 RDS, 800 PSI

MEMBRANE TYPE	TEMP. °C	FLUX L/HR./FT^2	SALT REJECTION %	SUGAR REJECTION %
First Filmtec Ft-30	25	0.3	99.22	99.49
First Filmtec Ft-30	50	0.3	84.59	99.61
Second Filmtec Ft-30	25	0.15	89.01	99.35
Second Filmtec Ft-30	50	0.15	76.88	99.39
Second SEPA 97	25	0.19	74.39	99.67
Second SEPA 97	50	0.16	54.81	99.63

TABLE 13

Salt rejection test with 0.5% NaCL solution

MEMBRANE TYPE	TEMP. °C	FLUX L/HR./FT^2 400 PSI	FLUX L/HR./FT^2 800 PSI	SALT REJECTION % 400 PSI	SALT REJECTION % 800 PSI
First Filmtec Ft-30	25	1.25	2.15	98.08	98.25
First Filmtec Ft-30	50	1.9	3.3	98.38	98.75
Second Filmtec Ft-30	25	1.15	1.0	97.8	98.16
Second Filmtec Ft-30	50	0.9	1.5	91.51	97.69
SEPA 97	25	0.91	1.25	92.01	93.9
SEPA 97	50	1.07	1.74	95.35	95.46
UOP Seawater	25	1.68	3.44	97.61	98.69
UOP Seawater	50	3.09	4.94	98.47	98.61

Graph #1 (A through F)

Show higher fluxes with the UOP and Filmtec membranes as compared to the SEPA 97 (cellulose acetate) fluxes.

Graph #2

Shows a very steady concentration gradient of ion exchange juice with the Sepa 97 membrane at 25 & 50^{o} C (maximum allowable temperature for cellulose acetate membranes). This means that the membrane is quite adequate for that purpose (if the juice is close to neutrality and free of foulants). Long-term on-line testing is necessary to confirm our preliminary results.

Graph #3

Shows apparent and true conductivities of thin juice as a function of RDS. They have been determined for the same amount of salt, with and without sugar, in a simulated thin juice. It is, of course, important to know the true conductivity of a feed, as membrane salt rejection characteristics are determined by the following equation:

$$\text{salt rej. \%} = 100 - \frac{\text{permeate conductivity}}{\text{feed conductivity}} \times 100$$

(in case of a high-capacity module, feed true conductivity will be replaced by:

$$\frac{\text{feed true conductivity} + \text{concentrate true conductivity}}{2})$$

The permeate conductivity is always true because of the absence of sugar.

Graphs #4 & 5

Show the behavior of two FT-30 membrane elements ($\sim$6 ft^2 of membrane each) during a concentration test of thin juice from 15 to 30 RDS--the irregularities are due to some wrapping problems of the prototype elements. However, the membranes showed in general good fluxes and sugar rejections.

Graph #6

Shows a thin juice concentration test at 25 & 50^{o} C using the Sepa 97 membrane. It would be possible to use latter membrane for that purpose if thin juice wasn't at 95^{o} C and pH 8-9.

Graph #7

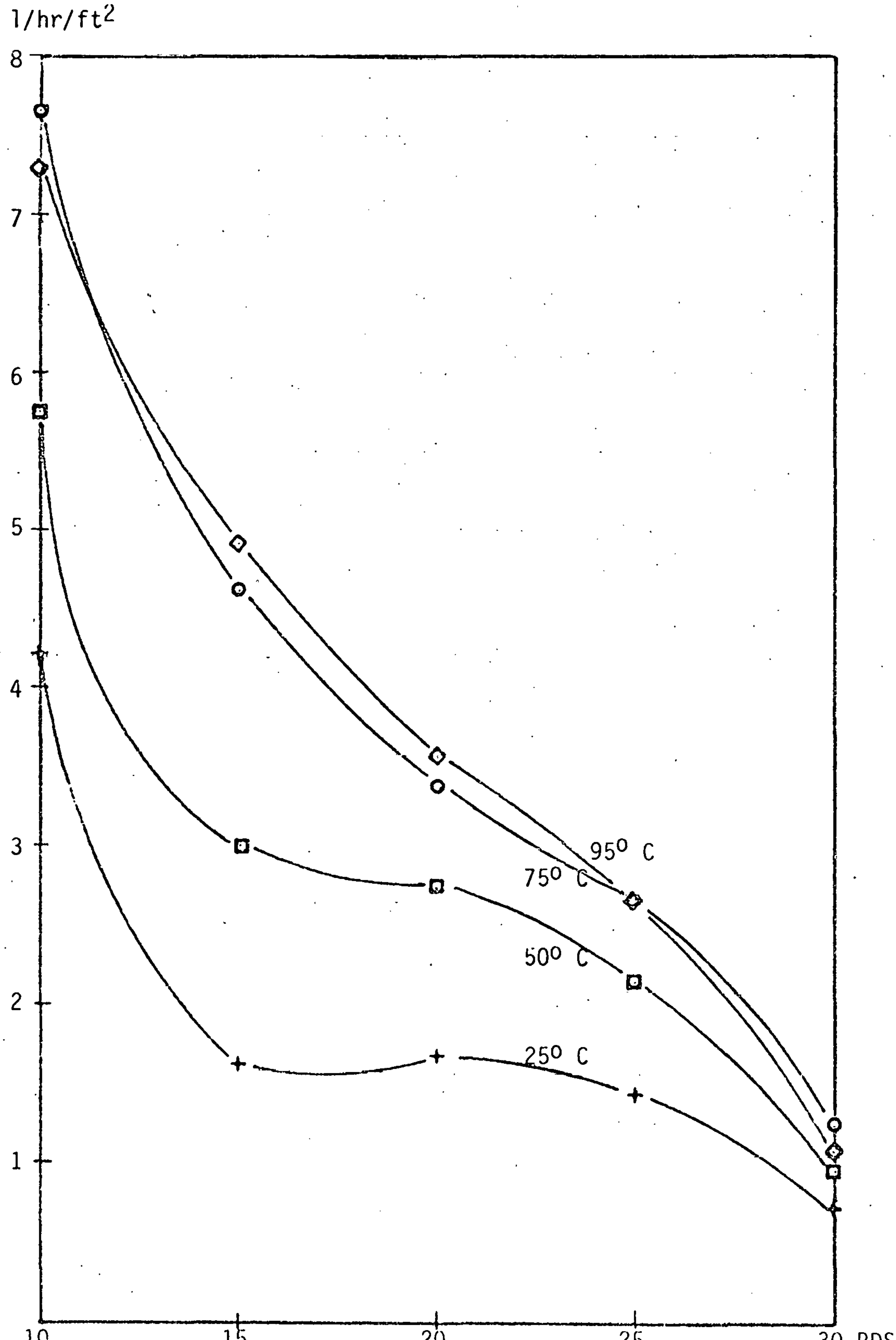
In this experiment, the UOP sea water membrane is being compared to the Sepa 97 under the same conditions, during thin juice concentration at 25 and 50^{o} C.

It is clear that better fluxes and sugar rejections are obtained with the UOP membrane (see tables).

GRAPH 1A (from Table 1)

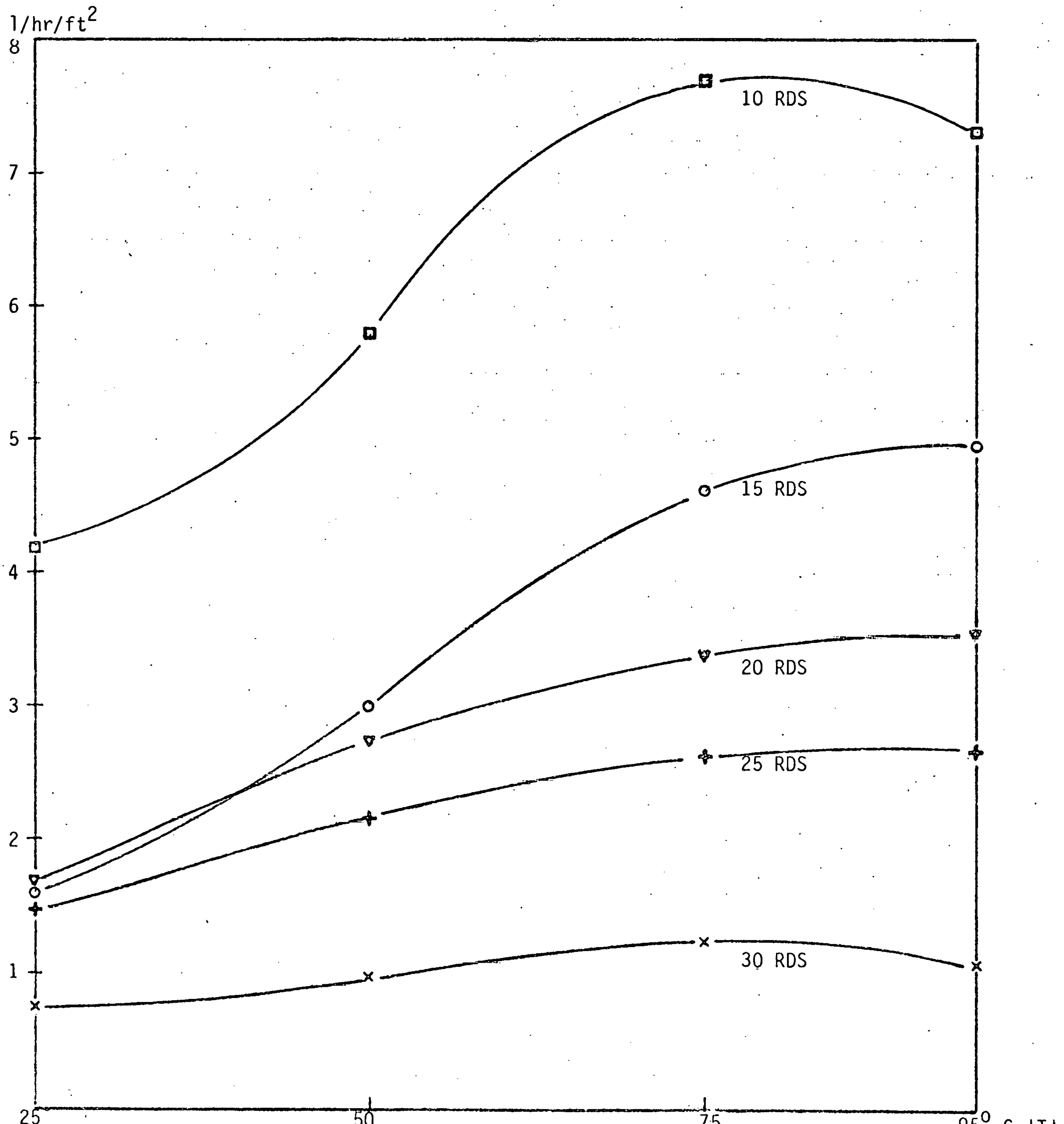
- UOP brackish water M. in test cells with thin juice at 800 psi -

Flux vs. RDS at different temperatures



GRAPH 1 B (from Table 1)

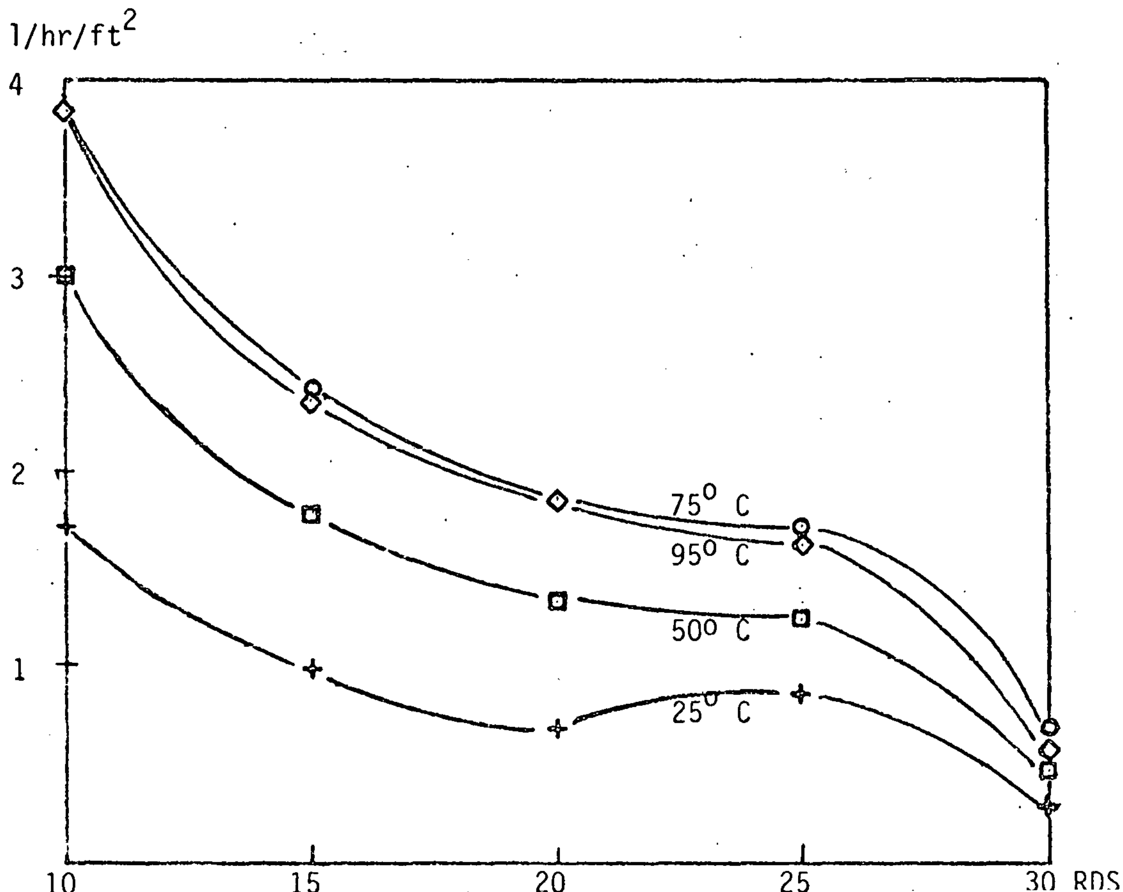
- UOP brackish water M. in test cells with thin juice at 800 psi -- Flux vs. temperature at different RDS



GRAPH 1 C (from Table 1)

- UOP sea water M. in test cells with thin juice at 800 psi -

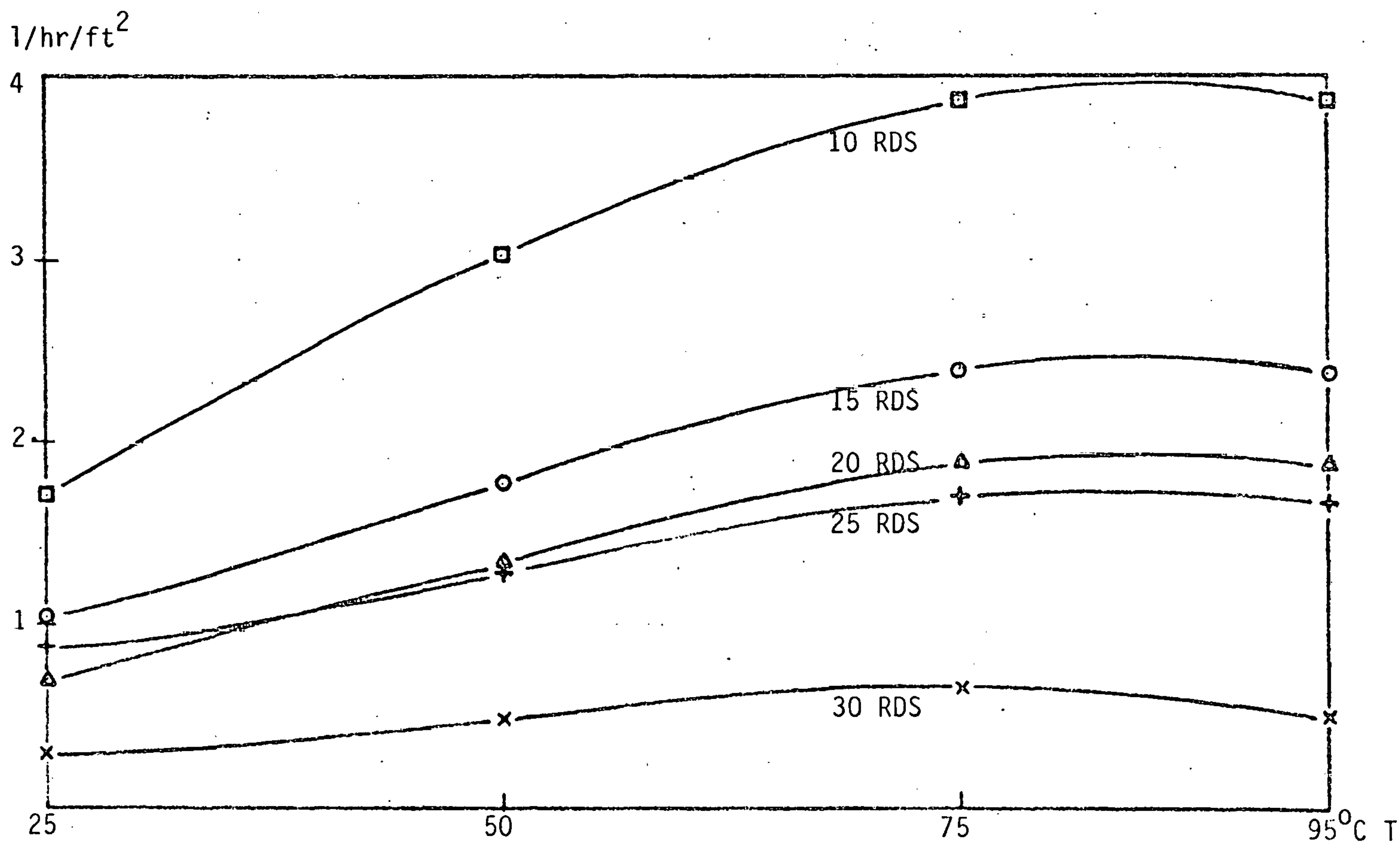
Flux vs. RDS at different temperatures



GRAPH 1 D (From Table 1)

- UOP sea water M. in test cells with thin juice at 800 psi -

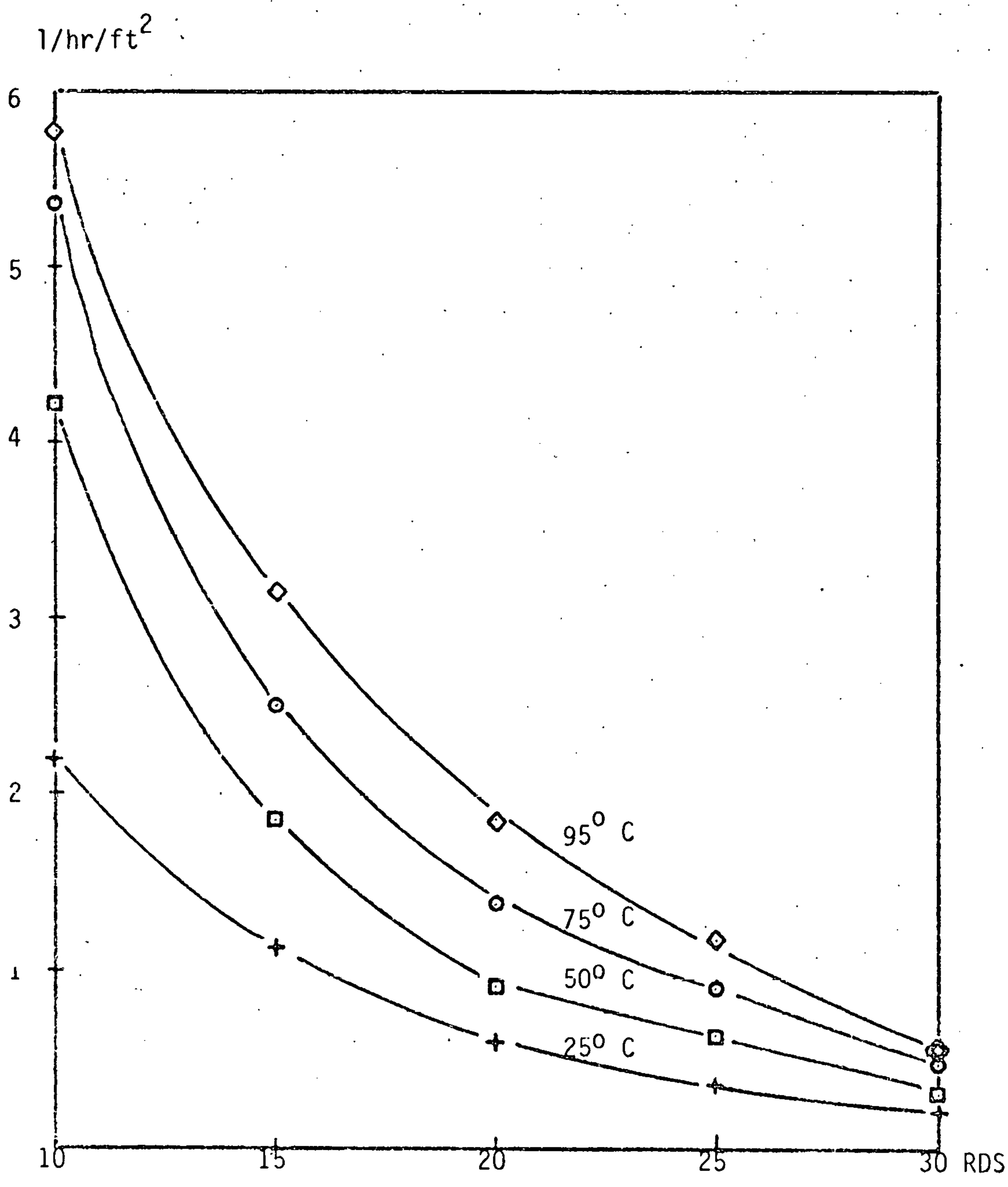
Flux vs. temperature at different RDS



GRAPH 1 E (from Table 1)

- FT-30 M. in test cells with thin juice at 800 psi -

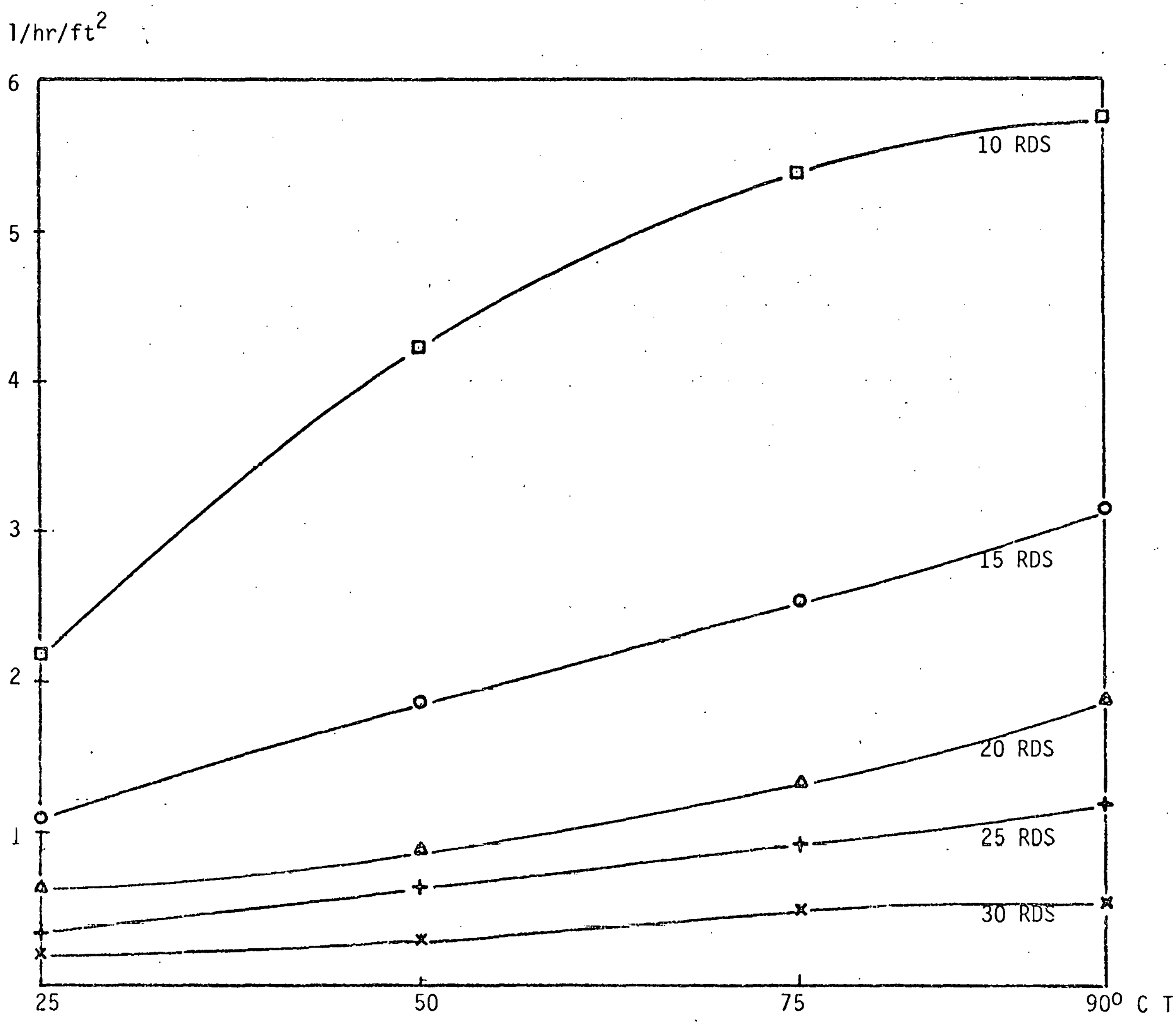
Flux vs. RDS at different temperatures



GRAPH 1 F (from Table 1)

- FT-30 M. in test cells with thin juice at 800 psi -

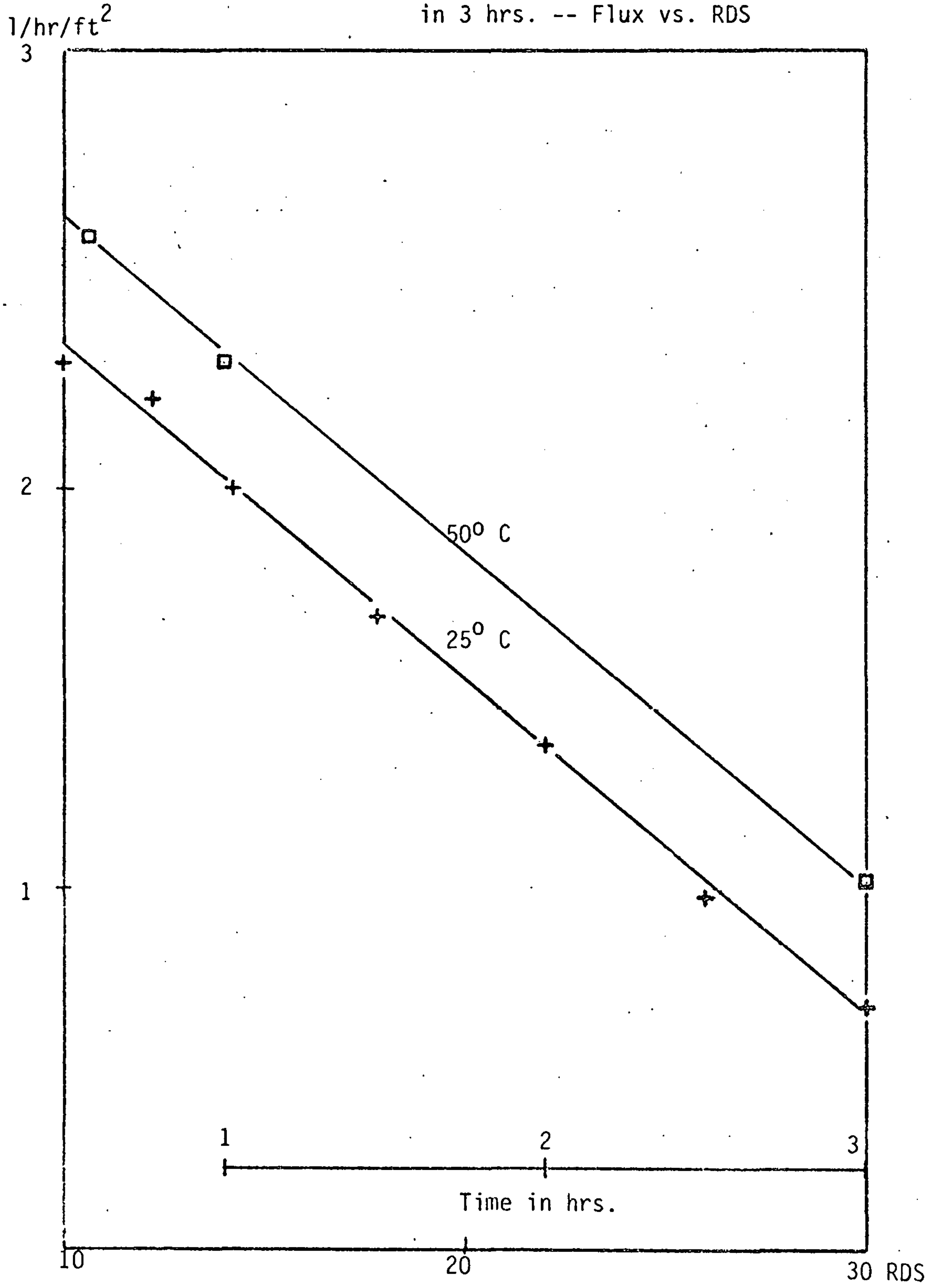
Flux vs. Temperature at different RDS



GRAPH 2 (from Table 2)

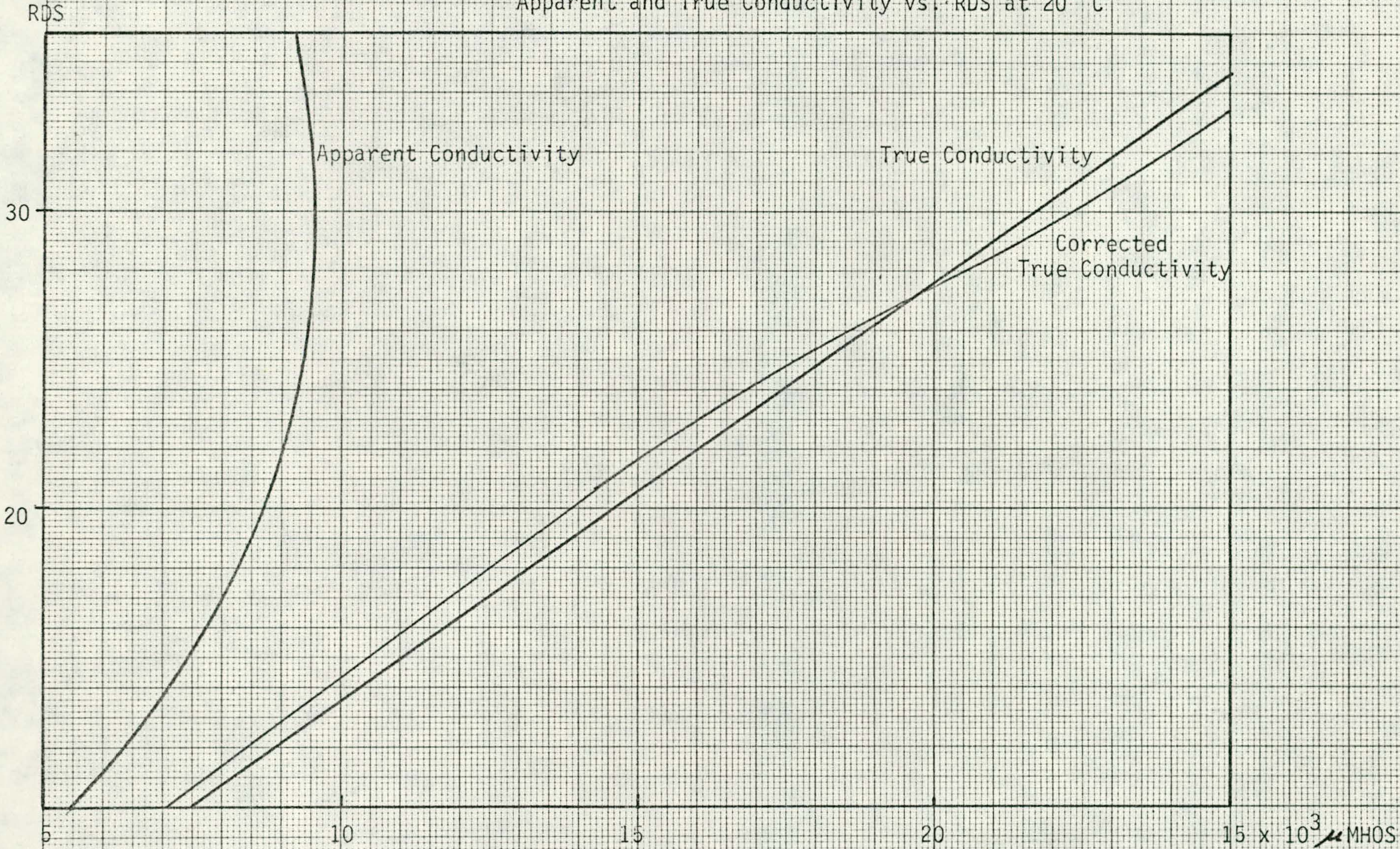
- Sepa 97 M. in element (19 ft^2) with IE juice at 800 psi, 25 & 50^o C -

Concentration of 113 l from 10 to 30 RDS in 3 hrs. -- Flux vs. RDS



GRAPH 3 (from Table 3)

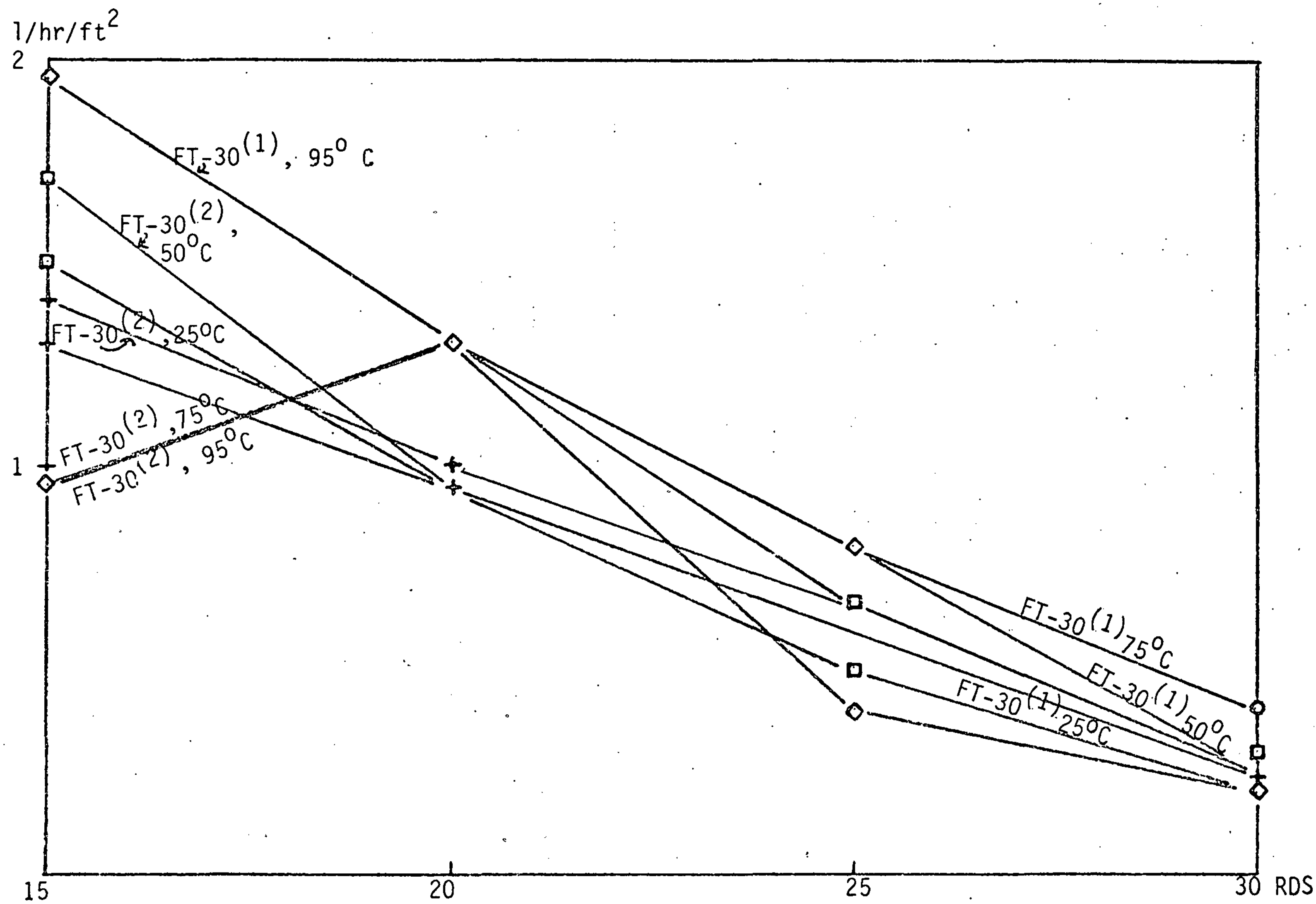
Apparent and True Conductivity vs. RDS at 20° C



GRAPH 4 (from Tables 10 & 11)

- FT-30(1)&(2) M. in elements (6 ft^2) with thin juice at 800 psi - Concentration from 15 to 30 RDS --

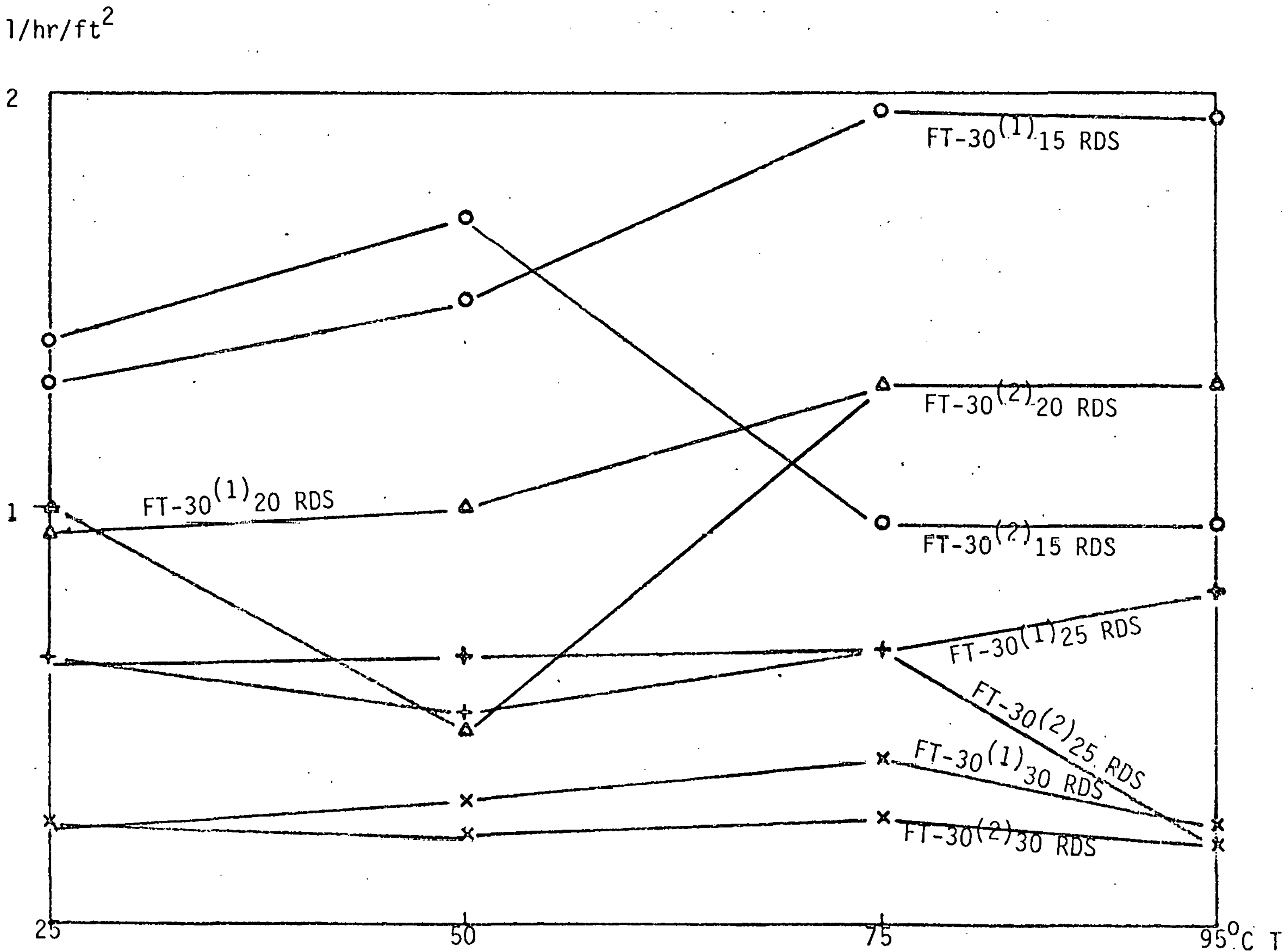
Flux vs. RDS at different temperatures -



GRAPH 5 (from Tables 10 & 11)

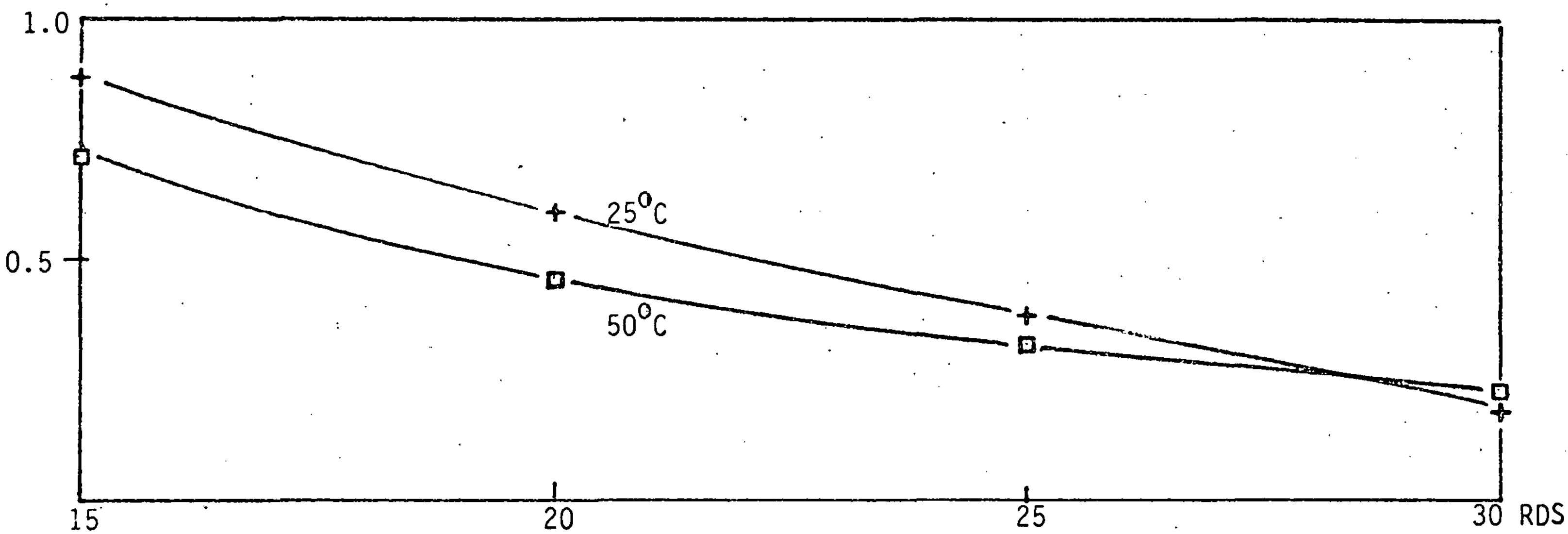
- FT-30(1)&(2) M. in elements (6 ft^2) with thin juice at 800 psi - Concentration from 15 to 30 RDS -

Flux vs. Temperature at different RDS



GRAPH 6 (from Table 10)

- Sepa 97(2) M. in element (19 ft^2) with thin juice at 800 psi -- Concentration from 15 to 30 RDS -- Flux vs. RDS at 25 & 50° C



GRAPH 7 (from Table 10)

- UOP sea water M. in element (68 ft^2) with thin juice at 800 psi - Concentration from 15 to 30 RDS - Flux vs. RDS 25 & 50^o C.