

OIL REMOVAL FROM OILFIELD PRODUCED WATER BY CROSS FLOW  
ULTRAFILTRATION

Brian A. Farnand and Thomas A. Krug

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ABSTRACT

The technical and economic viability has been determined for cross flow ultrafiltration of oilfield produced water with the goal of producing oil-free water for steam generation. Experiments with produced water supplied from various oil producing sites of Western Canada have been performed with successful permeation rates and permeate quality. The resistance to membrane fouling and other factors that decrease permeation rate have been determined in long term experiments. These results have been used to prepare a cost survey of the cross flow ultrafiltration process, and this has been compared to conventional treatment. The results indicate the level of performance that must be maintained in the ultrafiltration process to be economically advantageous.

\* Hydrocarbon Separation Section, Synthetic Fuels Research Laboratory, Energy Research Laboratories, CANMET, Energy, Mines and Resources Canada, Ottawa, K1A 0G1

† Zenon Environmental Inc., 845 Harrington Court, Burlington, Ontario, L7N 3P3

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INTRODUCTION

Enhanced oil recovery (EOR) operations in Western Canada often employ steam to stimulate reservoirs. These steam stimulation operations require large volumes of water for steam generation and injection into the oil bearing reservoirs. They generate large volumes of produced water which is contaminated with oil, dissolved organic compounds, and inorganic material (1). Since many recovery operations in Western Canada are located in water deficient or environmentally sensitive areas, several EOR recovery operations are recycling a portion of their produced water. Oilfield steam boilers that generate 80% quality steam (80% vapour, 20% liquid) can tolerate minor amounts of total dissolved solids (TDS), hardness, and silica, but require total removal of the oil (2). Conventional treatment for the removal of the oil is limited by the presence of very stable oil-in-water (o/w) emulsions, stabilized by the presence of natural or added surfactants (3,4). As well, subsequent operations of silica removal and softening perform best when the oil has been removed.

Some EOR operations use large amounts of water, sometimes as much as 2 to 10 m<sup>3</sup> for every m<sup>3</sup> of recovered oil. Some of this water may be present in the recovered oil as a water-in-oil emulsion while the balance may be present as a distinct water phase with an o/w emulsion. The produced fluid is generally treated in a free water knock out (FWKO) tank and a high temperature separator

(HTS) to separate as much of the oil and water as possible. Water from the FWKO and HTS is usually combined and sent to a skim tank where some residual oil is removed and recycled to the oil recovery part of the process. The remaining water, with some oil, is often disposed of by deep well injection (below the oil bearing formation) or it is sent to further treatment to prepare it for steam generation. Further treatment may consist of final oil removal by induced gas flotation (IGF) and filtration followed by softening operations such as hot lime softening and ion exchange. All of the oil removal operations with the exception of FWKO require large amounts of demulsifying chemicals to destabilize the emulsions. Occasional process upsets and changes in the nature of the emulsions can cause the failure of these units to remove oil from the produced water, and results in the passage of slugs of oil with catastrophic effects on the subsequent deoiling and water softening operations. Thus, the effectiveness and the expense of using chemical demulsifiers for produced water treatment is very dependent upon the smooth operation of the oil recovery operations, which generate a product of greater value to the oil producer and receive priority in operation.

Cross flow ultrafiltration is an efficient and economical means of treating many industrial waste streams containing emulsions such as those formed by metal cutting oil emulsions (5). Since ultrafiltration (UF) is a cross flow system and not a dead end filter, it can handle higher oil contents than a dead end or depth filter, and for the treatment of produced water it can also replace the IGF unit and reduce the loading on the skim tank. This is shown in Fig. 1 where the process schematic of conventional produced water treatment is compared with that of membrane ultrafiltration. The nature of the separation in UF provides an absolute barrier and eliminates oil breakthrough caused by process upset conditions and oil slugs. This work includes bench

scale testing of the UF process for the removal of oil from oilfield produced water (2 000 ppm of oil) with fouling resistant membranes. These results are used to determine the comparative cost of UF and the corresponding conventional treatment processes.

## EXPERIMENTAL

### Analytical Procedures

Samples of produced water were characterized by the methods described below. The results are tabulated in Table 1.

### Total Suspended Oil (TSO) and Total Suspended Solids (TSS)

The sample of known volume (V1) is filtered through a washed, ashed, and preweighed (W1) Whatman glass fibre filter, which is then dried overnight and weighed (W2). The glass fibre filter is then heated with a slow increase in temperature from 150°C to 450°C to volatilize the oil without burning. The temperature is held at 450°C for 1 h and the sample is weighed (W3). The following computations were made.

$$\text{TSO} = (W2 - W3) / V \quad (1)$$

$$\text{TSS} = (W3 - W1) / V \quad (2)$$

It should be noted that the presence of very heavy oil is interpreted as a lower TSO value (high W3 value).

#### Total Dissolved Solids (TDS) and Total Dissolved Oil (TDO)

The filtrate from the TSS+TSO determination of known volume (V2) is placed in a preweighed (W5) crucible which is dried overnight at 60°C in a desiccator and weighed (W6). The residual in the crucible is heated for 1 h at 450°C and weighed (W7). The following computations were made.

$$\text{TDO} = (W6 - W7) / V \quad (3)$$

$$\text{TDS} = (W7 - W5) / V \quad (4)$$

#### Extractable Oil (Methylene Chloride)

The methylene chloride extractable oil content was measured by vacuum filtering the sample through a 0.45 μm glass fibre filter. The volume of the sample was determined from the filtrate, and the filter was placed in the original sample's container, where it was soaked in methylene chloride for at least 1 h. The solvent was then dried with anhydrous Na<sub>2</sub>SO<sub>4</sub> in a phase separating filter cone and the concentration of bitumen was determined by spectrophotometric absorption at 450 nm wavelength. The zero point was taken as the absorption for pure methylene chloride and the calibration was established by bitumen dissolved in methylene chloride up to 500 mg/L. The sample was diluted with further methylene chloride if it was outside of the calibration solutions' concentration range. The amount of oil in the original

sample was calculated using the original sample volume, the concentration of bitumen in the methylene chloride, and the volume of methylene chloride used for extraction.

The pH was determined with an Extech Model 651 pH meter, the conductivity was determined with a YSI Model 32 conductance meter, the turbidity was determined by an HF Instruments DRT 1,000 Turbidimeter, and the colour was measured by a Hach Colour Test Kit, Model CO-1. Anions were determined by ion chromatography, and cations were determined by direct current plasma. Silica was determined by assuming all silicon was present as silica, and hardness was computed by assuming that all of the calcium and magnesium were present as carbonates.

#### Cross Flow Ultrafiltration

A schematic of the apparatus is shown in Fig. 2. To circulate the produced water through the testing apparatus and to develop the operating pressure, a progressing cavity pump (Moyno Model SSF/AAC) was used. The pump was fed from a 50 L stainless steel reservoir equipped with a jacket for temperature control. The apparatus included control valves to modify the operating pressure and circulation rate, pressure and temperature indicators, an in-line flowmeter, and the capacity to recycle the permeate into the reservoir or to collect the permeate for sampling or volume reduction. The operating pressure for the UF experiments was measured at the inlet of the membrane testing unit. At the circulation rates used with these experiments there was an approximate 100 kPa pressure drop through the membrane testing unit. The membrane testing unit consisted of a bank of 6 membrane support

tubes of nominal 30 cm effective length and a 1.3 cm inner diameter. The flow through the membrane testing unit was serial with recycle to the reservoir. The membrane support tubes were drilled with small holes to allow the membrane permeate to flow out and a small trough was placed under each of the tubes to collect the permeate. Both the membrane support tubes and the troughs were covered to prevent the evaporation of the membrane permeate.

The membranes used in these experiments were tubular with nominal 1.27 cm inner diameters. They are fabricated from modified polysulphone of a proprietary formulation that gives them an oleophobic surface and subsequent resistance to fouling by the oil present in the produced water. They also have a composite structure that permits large permeation rates while taking advantage of the fouling resistance of the membrane surface. Their molecular weight cutoff (MWC0) was determined to be in the range of 10 to 15 kDa as measured with polyethylene glycol. This determination is performed by ultrafiltering aqueous solutions of a single molecular weight polyethylene glycol and reporting the MWC0 as the smallest molecular weight that has a separation in excess of 95%. Polyethylene glycol of staged molecular weights was supplied by both J.T. Baker Chemical Co. and Eastman Kodak.

The membrane experiments were performed by filling the reservoir with 40 L of produced water, circulating the produced water through the membrane testing equipment and recycling the permeate back to the feed tank. When a reservoir temperature of 65°C was reached, permeate was removed from the system and more produced water was added to the tank until a total of 110 L of produced water was added and the total volume was reduced by 90%. This corresponds to 90% volumetric recovery of water from the permeate. At this point, the permeate stream was recycled to the reservoir and determinations of flux decline, permeate quality, and susceptibility to fouling and membrane



cleaning were made. The cleaning formulations consist of blends of detergents and solvents chosen to avoid solvent damage to the membrane as well as to improve the permeation rates of the membranes. Throughout the experiments, the circulation rate through the membranes was held at 20 L/min which corresponds to a Reynolds' number in excess of 30 000. This high level of turbulence in the tube was considered to be sufficient to minimize the formation of a gel layer at the surface of the membrane which reduces flux in less turbulent systems.

## RESULTS

Ultrafiltration experiments were performed in two separate parts. The first was to survey the various sources of the produced water and compare their treatability by UF. The second was to modify the membrane pore sizes and operating conditions to observe any changes in permeation rate or permeate quality. The first set of experiments was performed in a bench scale pilot unit in Western Canada. Samples were collected and the experiments started within 24 to 72 h. The second set of experiments was performed with the same apparatus in Eastern Canada with samples that were not more than 7 days old.

### Effect of Produced Water Source

Several sites were used as sources of produced water to evaluate effluent from conventional treatment operations and to assess the effect of produced water from different regions. Five samples of produced water were collected

and the results of their analysis are shown in Table 1. The operating pressure for these experiments was 690 kPa, and the operating temperature was 65°C. The first case studied was sample 1, site F HTS produced water. The sample had a grey/blue colour with visible suspended solids which did not readily settle and traces of H<sub>2</sub>S were evident as the sample was handled. The change of permeation rates with time are shown in Fig. 3 and the permeate analysis is shown as part of Table 2. In Fig. 3, the decrease in permeation rate with the volume reduction is shown, along with the improvement in permeation rate with cleaning. The change of permeation rate from the 90% volume reduction point at approximately 52 h to the cleaning point at 115 h does not appear to be significant. The average permeation rate during this time is 1.0 m<sup>3</sup>/m<sup>2</sup>/day.

The site E Wellhead sample was obtained just prior to the FWKO unit and some free oil was removed from the sample container. The water part of this sample was very dark brown, and contained a larger amount of suspended material than any of the other samples. The change of permeation rate with time is shown in Fig. 4 and the permeate analysis is reported in Table 2. The average permeation rate at 84.6% volume reduction is 1.3 m<sup>3</sup>/m<sup>2</sup>/day and there is no evidence of decrease of permeation rate at this level of recovery. There was a minimal improvement of permeation rate with cleaning, which suggests that the limits to the permeation rate do not include plugging of the membrane pores.

The site E IGF sample was blue/purple with a small amount of black particles in suspension. The change of permeation rate with time is shown in Fig. 5 and the permeate analysis is reported in Table 2. The average permeation rate is 1.2 m<sup>3</sup>/m<sup>2</sup>/day at 91% recovery, although there appears to be a decrease with time at this level of recovery. The effect of cleaning on performance is temporary, including the physical cleaning obtained by forcing

a ball made of soft sponge material through the membrane tubes. From these last two results, it is apparent that oil content and previous processing of the produced water do not necessarily control UF performance.

The site H FWKO sample was grey and contained some fine black suspended particulate material. The permeation rate with time is shown in Fig. 6, and the permeate analysis is reported in Table 2. The permeation rates of the two sets of tubular membranes are noticeably different, in excess of the 15 to 20% pressure loss observed from the inlet to the outlet of the module, assuming linear permeation with pressure. The average permeation rate for the two sets is used, and at 90% volume reduction, this is  $4.3 \text{ m}^3/\text{m}^2/\text{day}$ . There was a small reduction in permeation rate with time which appeared to stabilize after 30 h of operation. Because of the high permeation rate, no attempt was made to assess the effect of cleaning on membrane performance. These results indicate that large amounts of TDS and hardness do not cause poor permeation rates.

The site D FWKO sample had a muddy brown appearance and contained considerable suspended matter. The permeation rate with time is shown in Fig. 7 and the permeate analysis is reported in Table 2. It should be noted that the operating temperature for this experiment was limited to  $60^\circ\text{C}$  because of a failure in the heating system. In this case, the higher flux tubular membranes were in the second set, as measured by the salt water flux test. The average permeation rate for the entire set of membranes was taken to be  $2.9 \text{ m}^3/\text{m}^2/\text{day}$  at 91% volume reduction. There was a reduction in the permeation rate with time up to 30 h operation, then the permeation rate was considered stable for the duration of the experiment.

UF Optimization

A second study concerning the variation of membrane pore size on the permeation rate of the produced water was made, where the same fouling resistant formulation as that of the previous experiments was used to formulate membranes of different pore sizes. The MWCO's investigated were 2 000 for the FLT-C, 8 000 for the HLT-SC membranes, and 20 000 for the THLT-C membrane. As well, the operating pressure and temperatures were modified to observe any improvements in the permeation performance.

A fresh sample of site C Skim Tank produced water and a fresh sample of site E combined FWKO and HTS produced water were obtained for these experiments. The permeation rates are shown in Fig. 8 and 9, and the analyses of the produced water and the permeates are reported in Table 3. It should be noted that these experiments were performed at a reduced operating pressure of 345 kPa, as part of an evaluation of UF performance at conditions of lower severity. There is a definite variation of the permeation rates with the pore sizes of the membranes, in particular at low volumetric recovery. For the site C produced water, the permeation rates were a direct function of the membrane pore size as determined by the MWCO, at all levels of recovery as shown in Fig. 8. However, for the site E produced water, the smallest pore membrane had the highest permeation rate at 89% recovery as shown in Fig. 9. There was no significant difference in the quality of the permeate for this range of membrane pore sizes.

## DISCUSSION

### Effect of Produced Water Source

Throughout these experiments, the permeation rates obtained were much higher than those obtained earlier (6). The previous work used small scale ultrafiltration cells that were not modified for use with produced water. Their design has poor turbulence and the circulation rate through these membranes was limited by the large pressure drop in the system. The turbulence for the tubular membranes was much larger than the test cells, as shown by calculations presented elsewhere (7). The increase in the permeation rate appears to be caused entirely by the improvement of the system turbulence.

The typical analyses of water quality include absolute information such as oil content and total dissolved solids. This does not offer much information on the processability of the produced water by UF. The history of the produced water including the addition of emulsion destabilizing chemicals has a significant effect on UF performance. This is most notable for the case of FWKO produced water and IGF produced water, where the permeation rate of the FWKO is much greater, despite similar oil content. This is caused by the presence of "reverse emulsion breakers", the chemical additives that encourage the destabilization and coalescence of oil droplets required in HTS, IGF, and skim tank units. Free oil causes a decrease in permeation rate by fouling the membrane pores and reducing their capacity. Membrane cleaners typically contain surfactants that promote the formation of stable oil droplets that re-emulsify the oil from the fouled membrane surface, with the subsequent increases in permeation rate. Thus, by processing produced water streams that

contain very stable emulsions, greater permeation rates are obtained, and less membrane cleaning is required. This is the opposite strategy of conventional produced water deoiling processes that require unstable emulsions.

Another consideration is the effect of oil content on permeate quality. The polar membranes used in this work are resistant to oil adhesion as described elsewhere (7). Emulsified oil does not pass through the membrane as shown by the low levels of permeate turbidity reported in Table 2. However, polar hydrocarbons that are dissolved in water can pass through the UF membranes. This effect has been observed and investigated elsewhere (6,8). While these components do not resemble typical petroleum components, they are analyzed as oil content, and presumably have the same deleterious effects on the heat transfer equipment in steam generation as free oil. This requires the use of membranes with smaller pores than those required to simply remove oil droplets from water, as has been observed for other oily water separations (5). The pore sizes of the membranes used in this work have been selected to reduce the permeation of these polar solutes based on our previous work (6).

#### UF Optimization

A significant difference in the two produced waters is the amount of oil they contain. The site E (FWKO+HTS) produced water has almost 10 times the oil content of the site C (skim) produced water, as shown in Table 3. Despite the likely presence of more emulsion destabilizing chemicals in skim tank effluent, the oil content is so low that the effect of gel polarization is very small, even at 90% volume reduction. Thus, for site C, gel polarization and fouling are very small effects and the permeation rates should be less sensitive to volume reduction and approach the permeation rates for pure

water, which are considered to be proportional to MWCO for membranes of similar formulation.

For site E (FWKO+HTS) produced water, the oil content has a greater effect on mass transfer, and the permeation rates change with the amount of volume reduction, with the exception of the smallest pore size membrane. For both produced waters studied in these experiments, the smallest pore membrane's performance was independent of volume reduction. Initially (at low volume reduction) the permeation rate is sufficiently small that gel polarization is not a factor. As the larger permeation rate membranes have gel polarization effects with permeation rates of approximately 5 times greater than the small pore membrane, oil will have a greater chance to coat the surface of the membranes and interfere with the pores. This also creates a site for the adsorption of further oil, and the permeation rate continues to decrease. This suggests that once a membrane surface becomes fouled with oil, it will continue to foul. If fouling can be avoided, then membrane cleaning schedules may be improved.

Variation of the operating temperature was made as shown in Fig. 8, where temperatures of 75° and 85°C were used. The effective viscosity of the produced water is reduced at the higher operating temperatures, and as expected, the permeation rate increases. The reduction of operating pressure in the optimization experiments did not cause as large a decrease in permeation rate as would be expected for pure water. This is the effect of gel polarization at the membrane surface where there is a balance of permeation resistance caused by gel formation and permeation driving force (operating pressure). Less severe operating conditions are an advantage for increased membrane life by avoiding fouling, and operation at 345 kPa is of interest for a commercial process because of reduced membrane supporting hardware

requirements. Similarly, operation at higher temperatures will give better permeation rates, but operation at lower temperatures improve membrane life and also require less extensive membrane supporting hardware.

A further indication of the importance of the stability of the produced water emulsion is the effect of changing the pH. In other work (6) the reduction of pH caused the produced water to destabilize into a clarified aqueous layer and an oily layer. An increase in pH increases the stability of the emulsion by further activating the weak acids that are present in the oil and allowing their participation at the surface of the oil droplets as emulsion stabilizers (8). In short term experiments, a relative increase in permeation rate of approximately 20% was observed for some of the produced waters at elevated pH.

#### Process Economics

The experimental results described above were used to compare the costs of treating oilfield produced water for conventional processing and for ultrafiltration. The design basis is for a typical oilfield 80% quality boiler of 73 MW (250 000 BTU/h) requiring a feed of 1890 m<sup>3</sup>/d (500 000 US gallons/day) of recycled produced water, as described in a review of the production of injection steam for EOR operations (2). The source of the produced water is FWKO or HTS effluent water, and the goal of the separation process is the removal of the oil from the water for the subsequent operations of ion exchange and softening.

The design of the conventional process shown schematically in Fig. 1A was developed on the basis of information available on present practices described in the literature and by representatives of the oil producing industry. The



flow of produced water from the HTS or the FWKO units is first treated in a skim tank with oil skimmed from the top and solids removed from the bottom. Demulsifying chemicals are added before the skim tank to encourage the agglomeration of the oil. The water stream from the skim tank is then treated with additional chemicals and further treated by IGF. An oil froth is skimmed from the IGF and the water stream is chemically treated and filtered in a pressurized sand filter. The filter produces a backwash suitable for sludge disposal and the filtered produced water is then sent for softening. Total direct costs for the process equipment were obtained by applying factors to the purchase cost of the equipment and include shipping, installation, foundations, instrumentation, piping, insulation, electrical, building, services and miscellaneous items (9). The indirect expenses were also calculated by factors and included construction and field expenses, engineering expenses, contractors' fees, and contingencies (9). The chemical requirements and costs for each stage of the process were based upon the information provided by chemical suppliers to the oil industry. The operating costs were determined for chemical demulsifiers, labour, electricity, maintenance and contingencies.

The design of the ultrafiltration separation process shown schematically in Fig. 1B is based on a continuous process operating at 90% recovery and that the retentate would be recycled to either the HTS or FWKO unit to recover the oil. The actual units that would be replaced by the ultrafiltration process are the skim tank, IGF and a sand filter. The costing for the ultrafiltration process is based upon current information and represents the manufacturer's actual selling price, with installation costs calculated with the same correlations as those for the conventional process. The ultrafiltration costs were determined on the basis of three levels of permeation rates, and for

three different types of membrane modules. The module costs for the first case are for currently available tubular membrane modules with inexpensive membrane replacement. The second case is for disposable tubular membrane modules, and the third case is for a small tube membrane module which would also be disposable. The process has an assumed operating pressure of 690 kPa (100 psi), membrane life of 2 years, and a detergent cleaning of the membrane every 2 weeks of operation. The same factors as those of the conventional case were applied to the purchase cost of the membrane equipment to determine the direct cost and the indirect expenses. The operating costs were determined for membrane cleaning and replacement, labour, electricity, maintenance and contingencies.

The capital costs of both the conventional and ultrafiltration cases were considered to be amortized over 5 and 10 years at 11% interest as summarized in Table 4. It is apparent the permeation rate must be in the range of 3.0  $\text{m}^3/\text{m}^2/\text{d}$  for the ultrafiltration process to be less expensive than the conventional process, although lower permeation rates can be tolerated for the lower cost modules. The permeation rate through the membranes is the controlling factor for the economic viability of the process. The major advantage of high permeation rates is the reduction in the capital cost of the process, with a corresponding reduction in the operating costs which include membrane replacement. A graphical presentation of these results are shown in Fig. 10 and 11.

It should be noted that the ultrafiltration process has the advantages of improved operation of downstream units in the absence of oil and solids, and reduced sludge disposal. This was not included in the development of the process costs.

## CONCLUSION

The technical viability of the ultrafiltration process has been demonstrated. The economic viability is directly related to the permeation rate. Future work with the ultrafiltration process shall deal with the testing of actual modules to permit the evaluation of the working life of the membranes and the permeation rates with slightly different flow regimes than those studied in this work. As well, the effect of operation at an oil producer's site will give more process information, including the effect of process upsets and the vagaries of the upstream process.

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Table 1

## Raw feed characterization

PARAMETER	Site F HTS	Site E Wellhead	Site E IGF	Site H FWKO	Site D FWKO
pH	N/A	N/A	7	7	8
TDS (mg/L)	12,000	7,410	7,630	45,400	13,200
TDO (mg/L)	730	1,120	890	2,600	640
TSS (mg/L)	790	2,320	410	440	220
TSO (mg/L)	230	1,230	31	28	100
Turbidity (NTU)	15	7,000	3	24	180
Colour (APHA)	N/A	N/A	500	1,300	2,250
Conductivity (umho)	17	11	9	59	18
Hardness (mg/L)	1,143	149	205	3,713	1,002
Extractable oils (mg/L)	360	>1000	4	18	64
<b>ANIONS (mg/L)</b>					
Bromide	<8	<8	<8	73.0	21.0
Chloride	7,000	5,000	4,100	17,000	6,300
Fluoride	<1	<1	<1	<1	<1
Nitrate (as N)	<2	4	3	<2	<2
Nitrite (as N)	<2	<2	<2	<2	<2
Phosphate (as P)	<8	<8	<8	<8	<8
Sulfate	59	11	87	41	87
<b>CATIONS (mg/L)</b>					
Aluminum	<0.3	<0.3	<0.3	<0.3	<0.3
Barium	7	0	0	3	4
Beryllium	<0.003	<0.003	<0.003	<0.003	<0.003
Boron	24	51	51	13	18
Cadmium	<0.02	<0.02	<0.02	<0.02	<0.02
Calcium	260	58	66	630	220
Chromium	<0.04	<0.04	<0.04	<0.04	1
Cobalt	<0.1	<0.1	<0.1	<0.1	<0.1
Copper	<0.06	<0.06	<0.06	<0.06	4
Iron	6	2	9	0	1
Lead	<0.4	<0.4	<0.4	<0.4	<0.4
Magnesium	120	10	10	520	110
Manganese	1	0	0	1	0
Molybdenum	<0.2	<0.2	<0.2	<0.2	<0.2
Nickel	<0.09	<0.09	<0.09	<0.09	<0.09
Phosphorus	<1	5	<1	1	<1
Potassium	120	100	83	310	96
Silica	46	88	80	5	37
Silver	<0.1	<0.1	<0.1	<0.1	<0.1
Sodium	3,600	2,700	2,700	16,200	4,400
Strontium	20	2	1	37	18
Thallium	<0.2	<0.2	<0.2	<0.2	<0.2
Titanium	<0.1	<0.1	<0.1	<0.1	<0.1
Vanadium	<0.2	<0.2	<0.2	<0.2	<0.2
Zinc	<0.05	<0.05	<0.05	<0.05	<0.05
Zirconium	<0.1	<0.1	<0.1	<0.1	<0.1

TDS - TOTAL DISSOLVED SOLIDS

TDO - TOTAL DISSOLVED OILS

TSS - TOTAL SUSPENDED SOLIDS

TSO - TOTAL SUSPENDED OILS

Table 2 Analysis of produced water and composite permeates\*

	Site F HTS		Site E Wellhead		Site E IGF		Site H FWKO		Site D FWKO	
	feed	perm	feed	perm	feed	perm	feed	perm	feed	perm
Recovery, %	-	90	-	85	-	90	-	90	-	91.3
pH	-	7.7	-	8.4	7.1	8.5	6.9	7.3	7.5	7.3
Cond <sup>b</sup>	16.7	16.0	10.8	10.8	9.1	9.1	59.3	59.1	17.6	18.3
Turb <sup>c</sup>	15.0	1.2	7000	4.0	3.0	1.4	24.0	-	180	-
Colour <sup>d</sup>	-	375	-	550	500	330	1300	95	2250	500
Hardness <sup>e</sup>	1143	1143	185	149	205	179	3713	3514	1000	1068
Silica <sup>f</sup>	99	88	188	77	171	114	10	13	79	66
Ext oil <sup>f</sup>	360	<10	>1000	<10	4	15	18	7	64	5

\* composite permeates from initial to the final recovery

<sup>b</sup> in  $\mu$ mho units

<sup>c</sup> in NTU units

<sup>d</sup> in APHA units

<sup>e</sup> in mg/L units

Table 3 Analysis of optimization experiments

	Site E				Site C			
	feed	permeate <sup>a</sup>			feed	permeate <sup>a</sup>		
Membrane	-	FLT-C	THLT-C	THLT-C	-	THLT-C	FLT-C	HLT-SC
Recovery %		84	84	84		87	87	87
pH	7.4	8.1	8.2	8.2	7.2	8.0	8.0	8.0
Cond <sup>b</sup>	7.1	6.6	6.5	6.4	12.2	11.8	11.6	11.6
Turb <sup>c</sup>	70	4.6	3.4	4.7	82	2.5	2.7	2.4
Colour <sup>d</sup>	7500	375	425	425	1750	500	325	370
Hardness <sup>e</sup>	97	82	82	83	257	216	208	215
Silica <sup>e</sup>	195	176	180	176	137	173	188	171
Ext oil <sup>e</sup>	720	<10	<10	11	26	<10	<10	<10
TSO <sup>e</sup>	490	39	52	49	120	36	51	49
TDO <sup>e</sup>	330	280	290	260	430	370	400	400
TDS <sup>e</sup>	6200	5160	5300	5190	9530	9290	9060	9230
TSS <sup>e</sup>	540	42	47	47	68	30	27	29

<sup>a</sup> composite permeates from initial to the final recovery

<sup>b</sup> in  $\mu$ mho units

<sup>c</sup> in NTU units

<sup>d</sup> in APHA units

<sup>e</sup> in mg/L units

**Table 4 Capital and operating costs for conventional and membrane based processes for oil removal**

Oil Removal Process	Module Cost	Assumed flux (m <sup>3</sup> /m <sup>2</sup> d)	Capital cost (\$1000)	Operating costs (\$1000/yr.)	Yearly costs (over 10 yr.) (\$1000)	Yearly costs (over 5 yr.) (\$1000)
Conventional	---	---	1,573	434	702	859
Membrane	High	4.0	1,563	221	487	643
		3.0	2,084	255	609	818
		2.0	2,932	382	881	1,174
	Med	4.0	1,004	254	425	525
		3.0	1,339	298	526	660
		2.0	1,884	447	768	956
	Low	4.0	814	223	361	443
		3.0	1,085	257	442	550
		2.0	1,527	386	645	798

To develop yearly costs, capital costs are amortized at 11% interest over 10 years and 5 years.



Figure 1A Conventional Produced Water Treatment Process

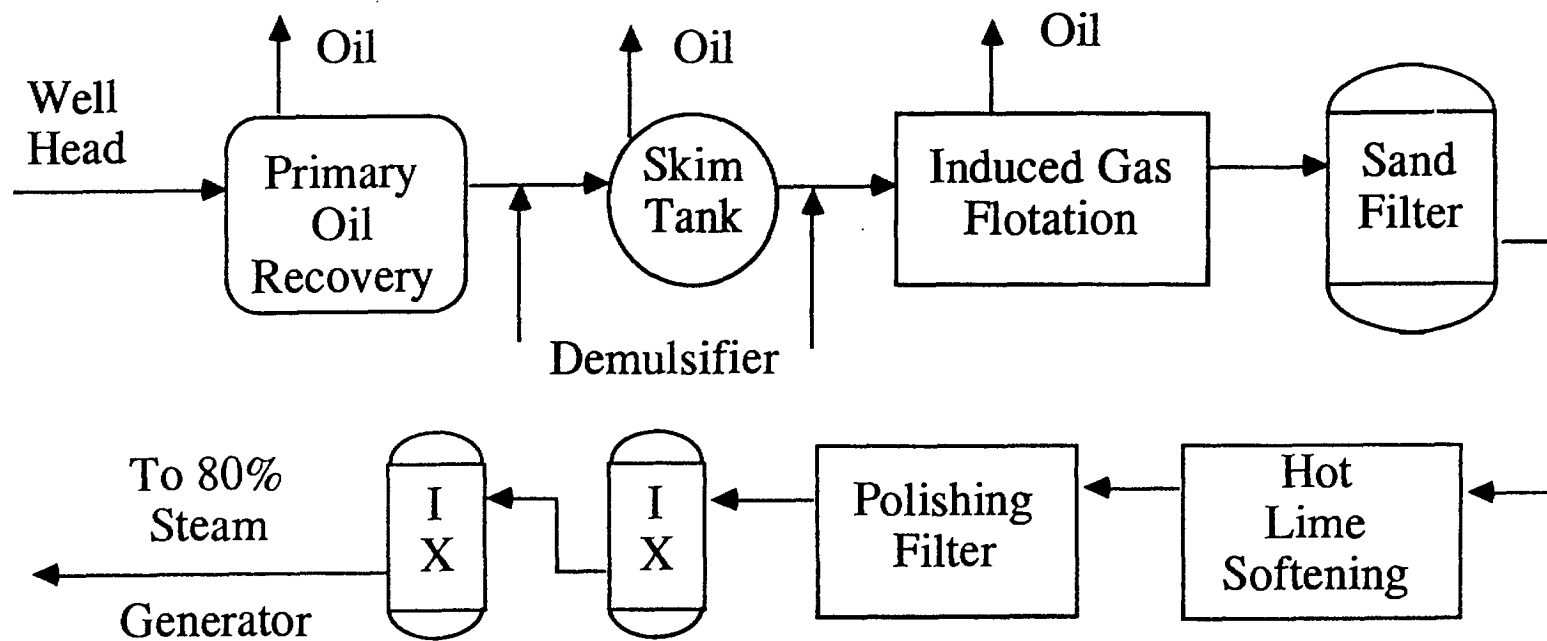


Figure 1B Potential Membrane Treatment Process

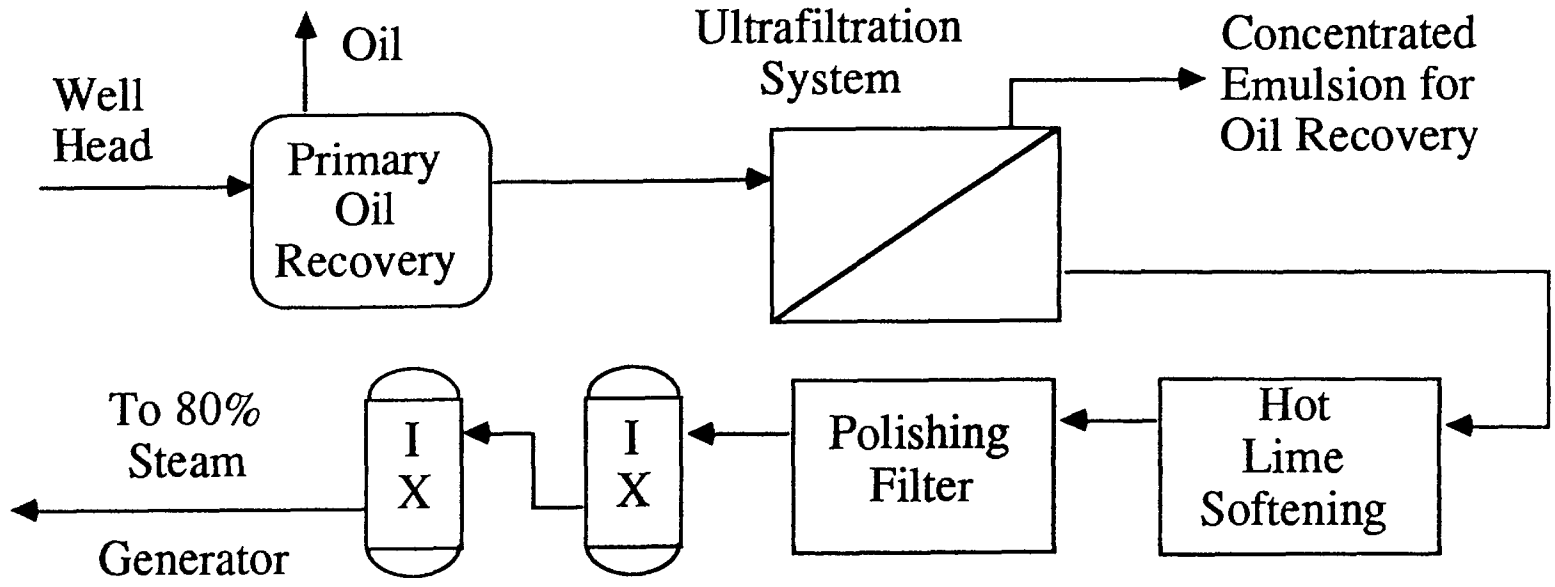


Figure 2 Flow Schematic for Membrane Test Equipment

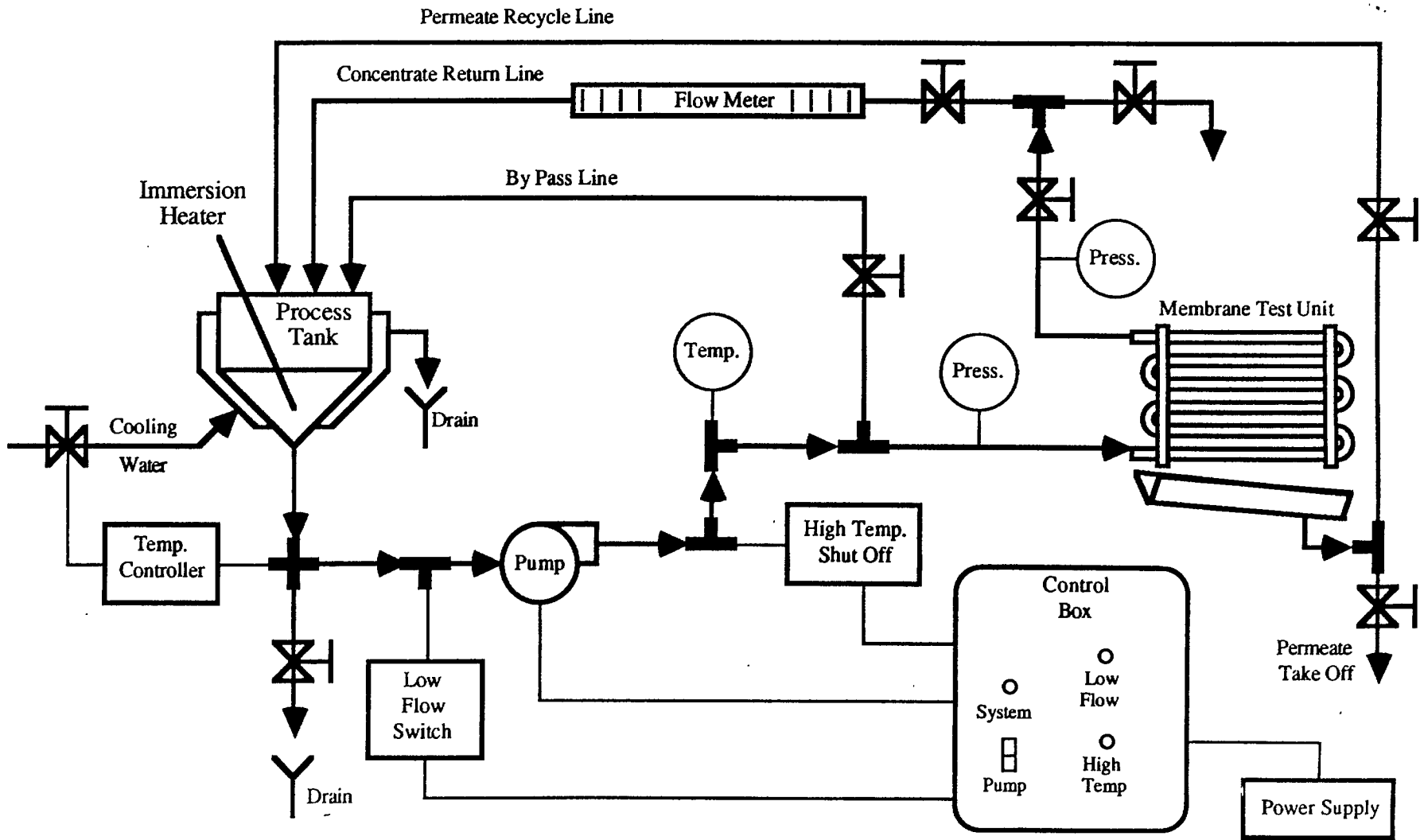


FIGURE 3

FLUX RATE VS TIME  
SITE F, HTS

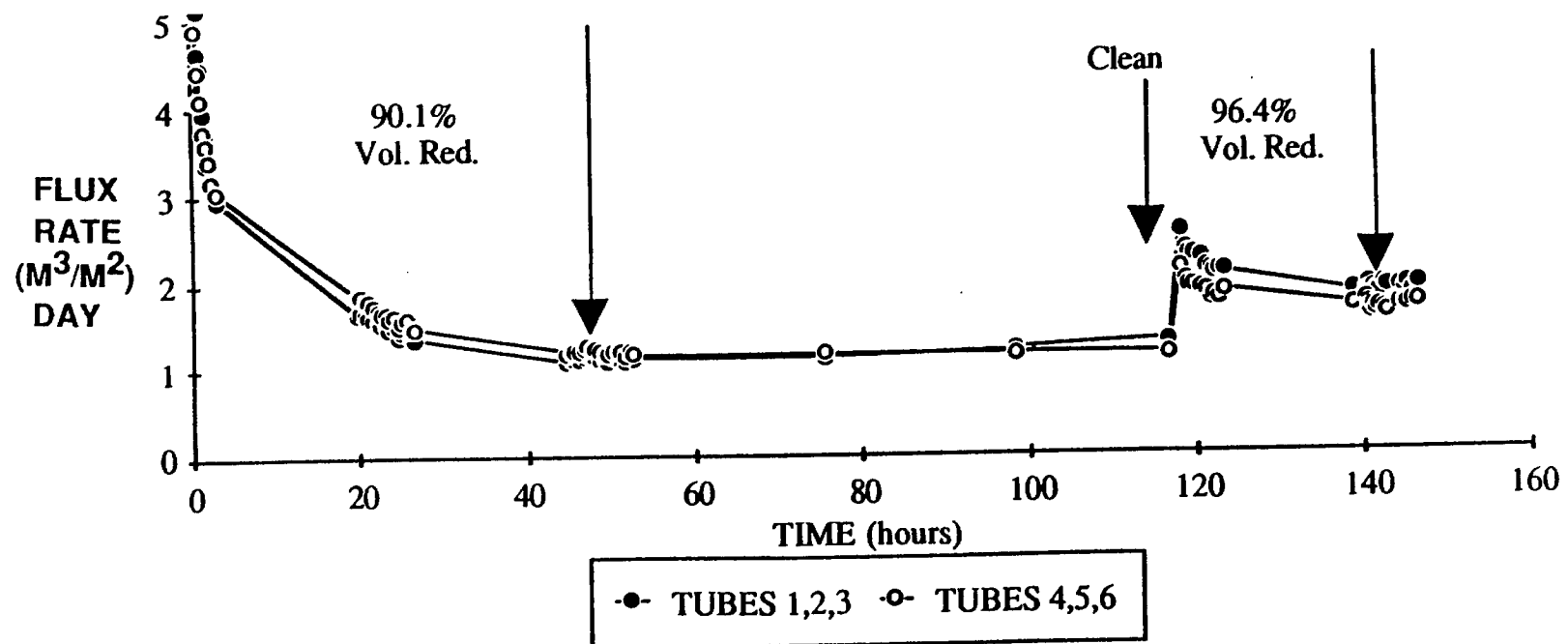


FIGURE 4  
SITE E, WELLHEAD

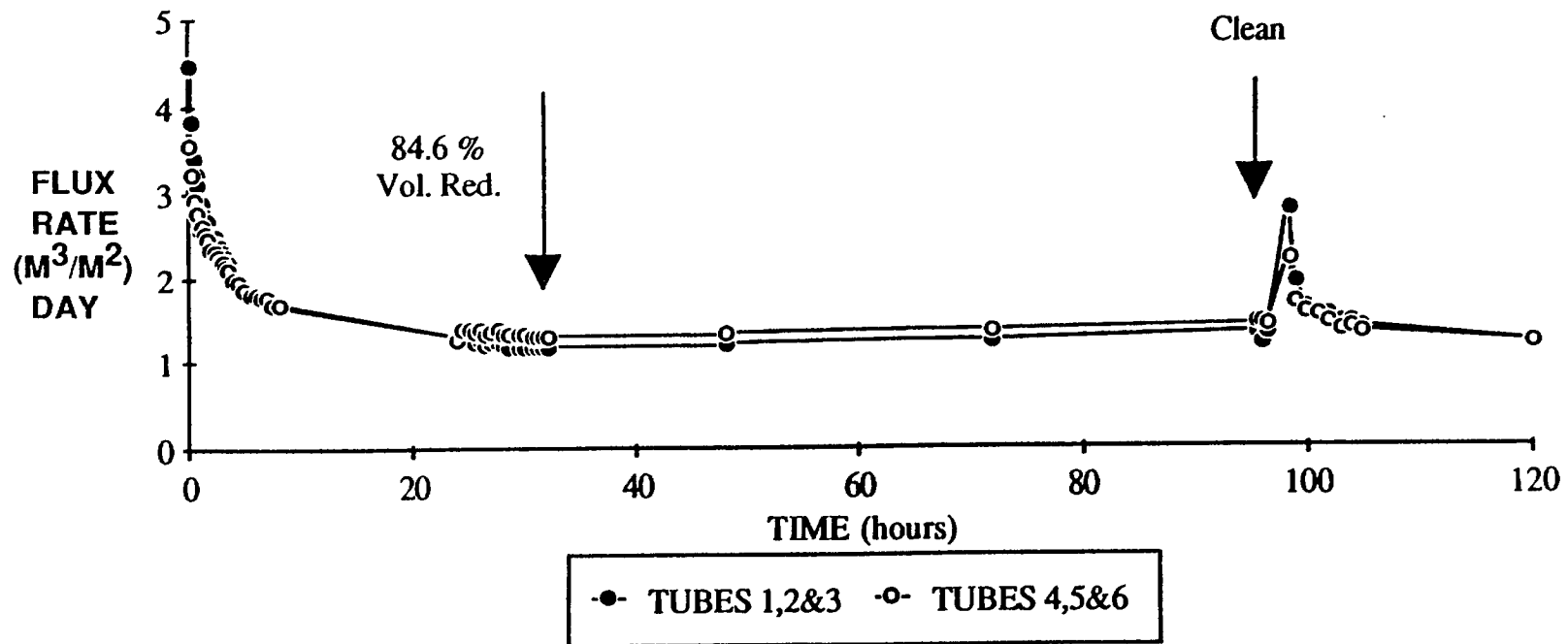


FIGURE 5

SITE E, IGF

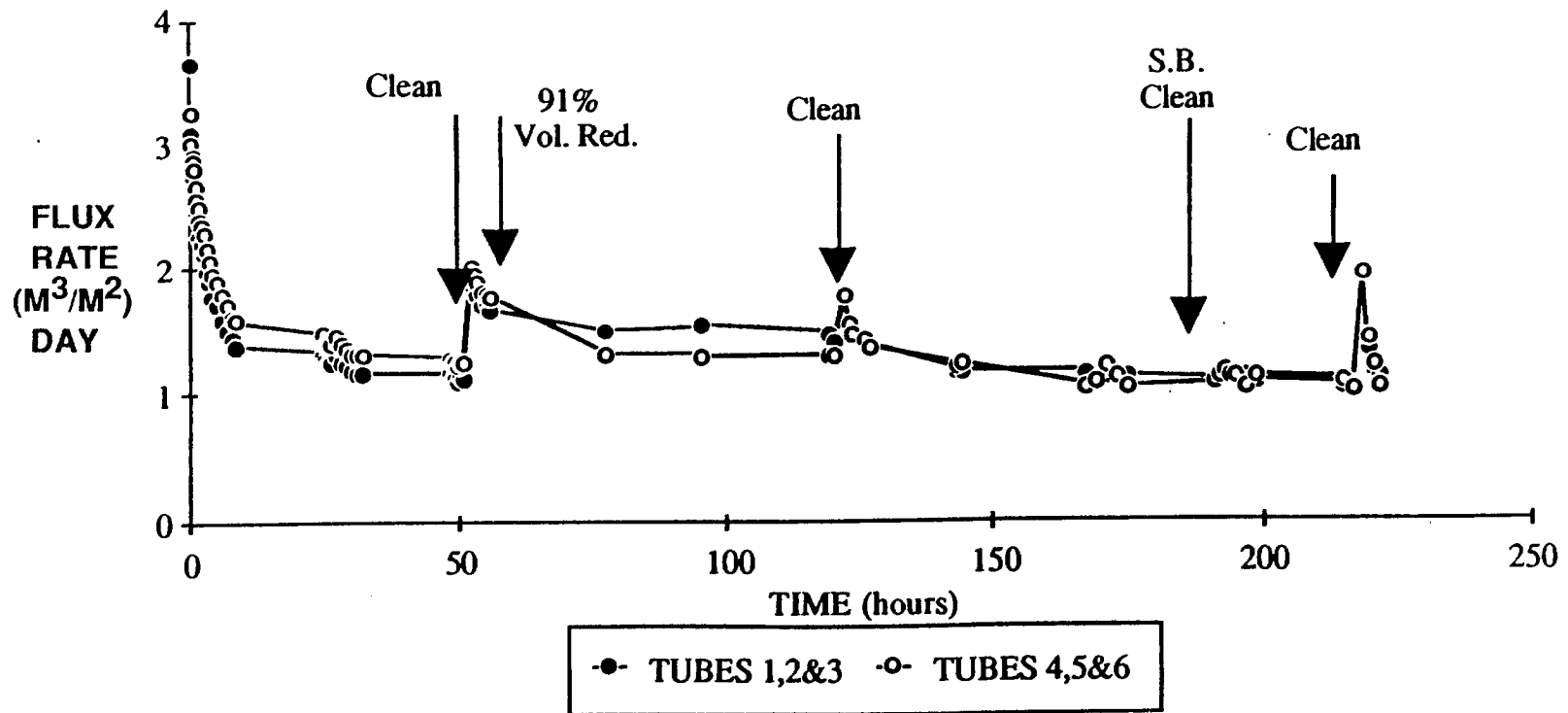


FIGURE 6  
SITE H, FWKO

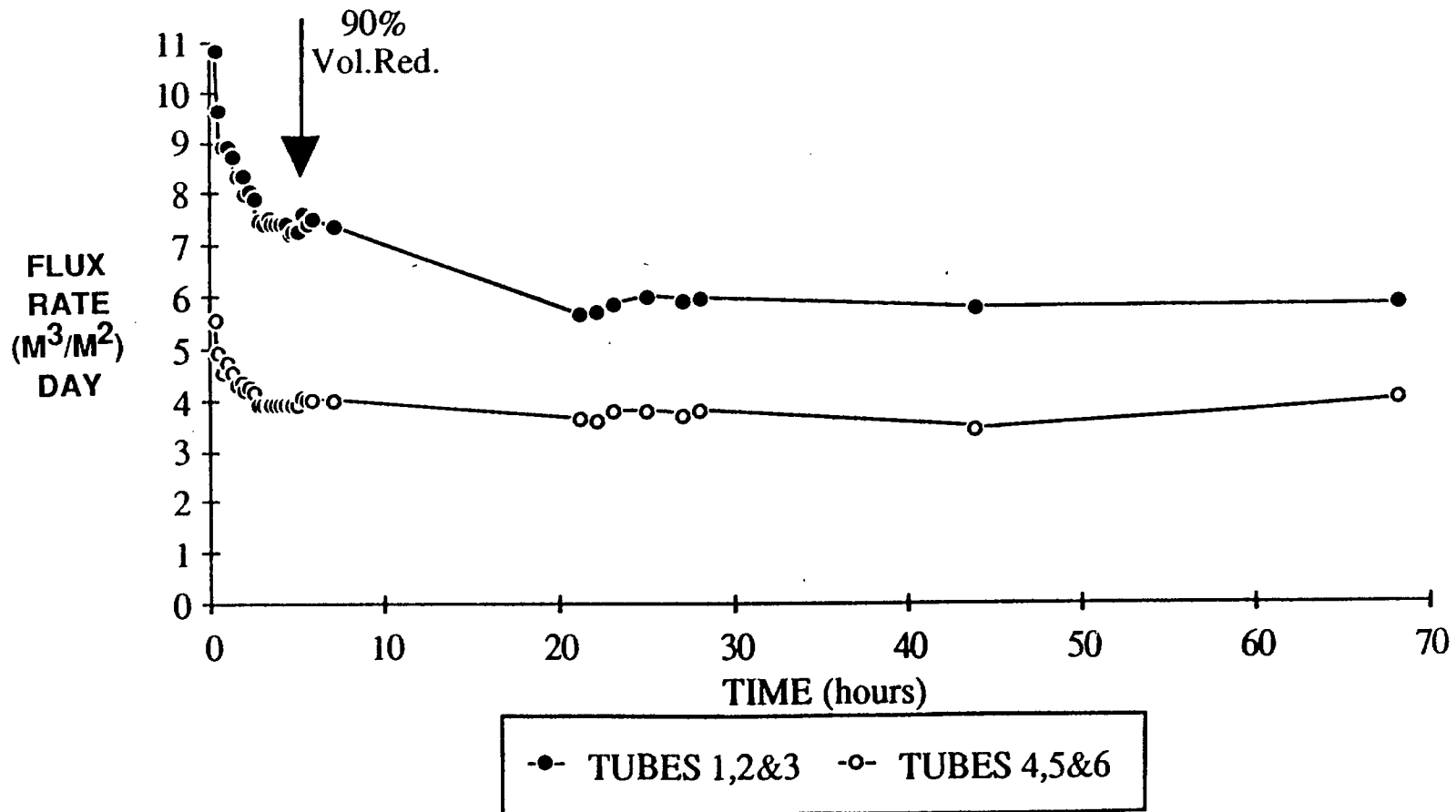


FIGURE 7  
SITE D, FWKO

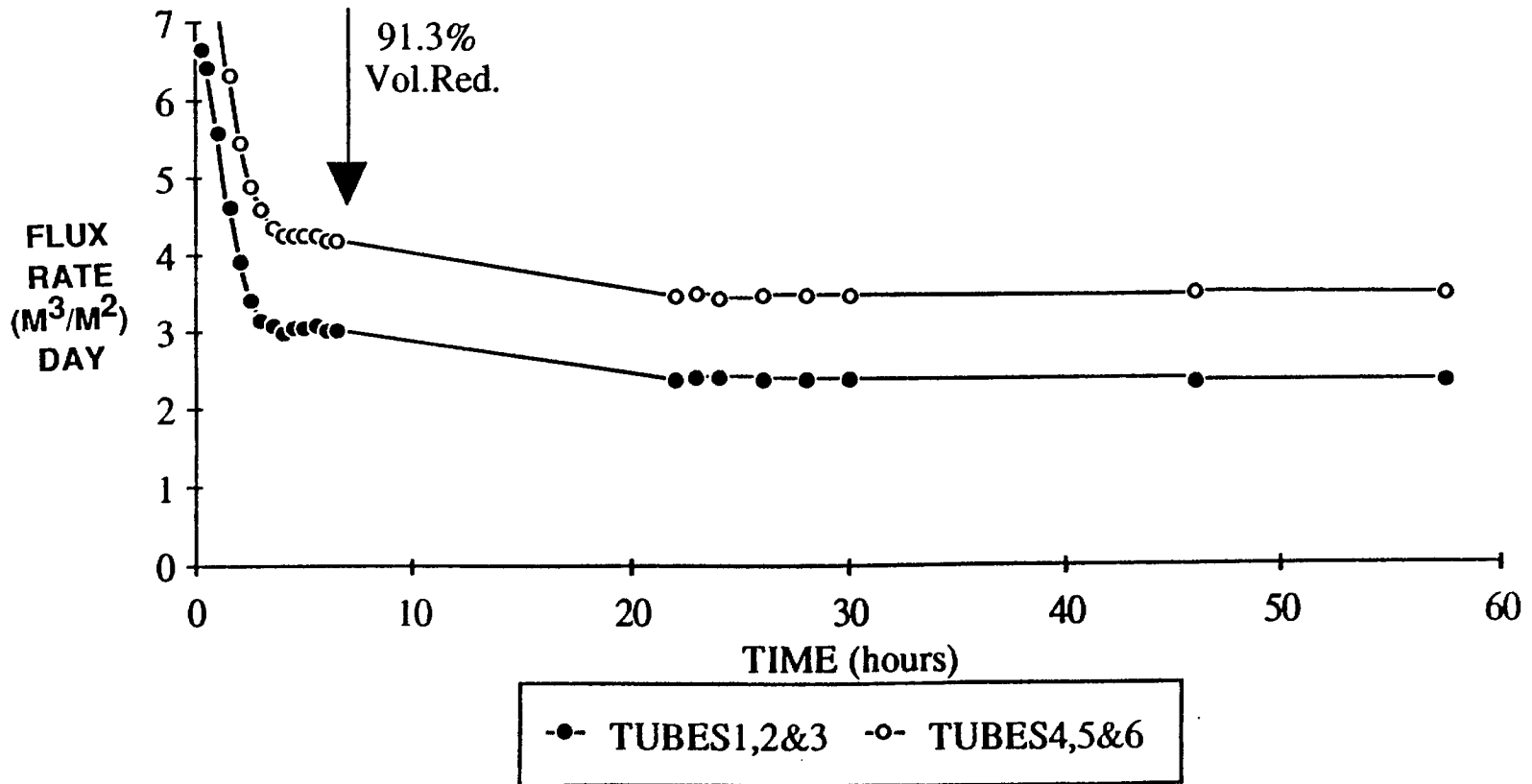




FIGURE 8

FLUX RATE VS TIME  
SITE C, SKIM TANK

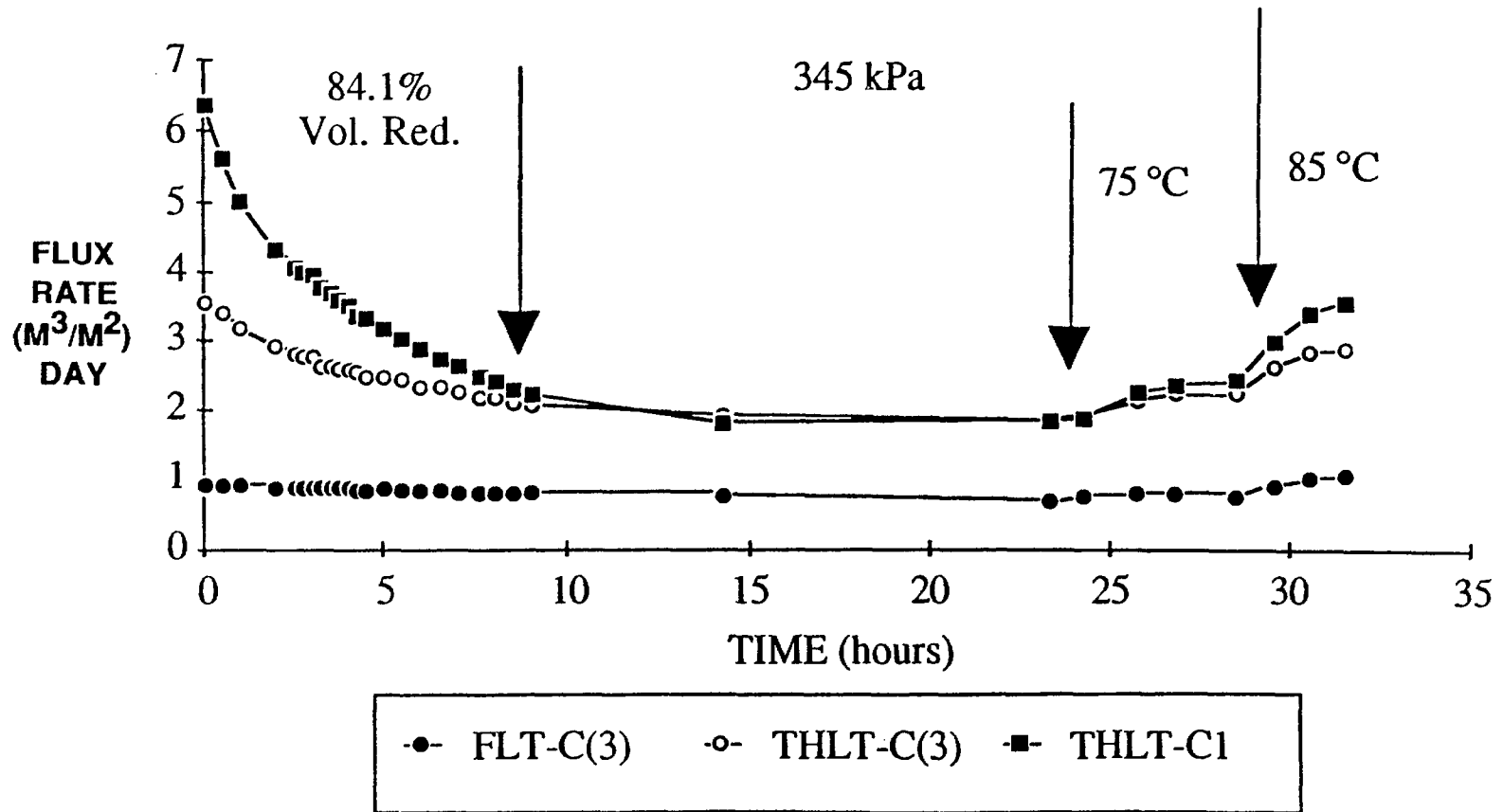


FIGURE 9

FLUX RATE VS TIME  
SITE E, FWKO & HTS

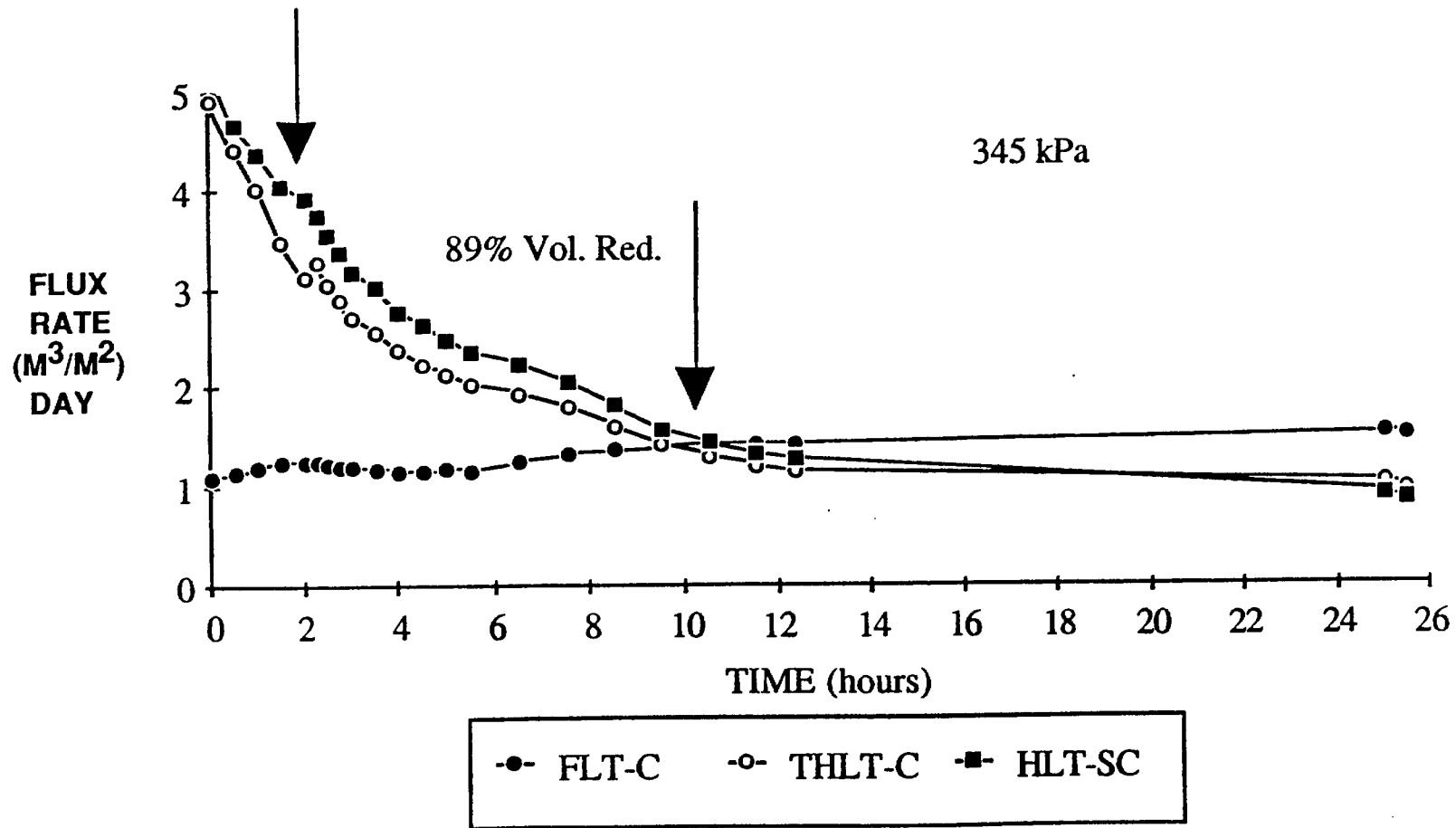


FIGURE 10

Yearly Costs for Conventional and Membrane Based Processes for Produced Water Treatment  
(5 year amortization of capital)

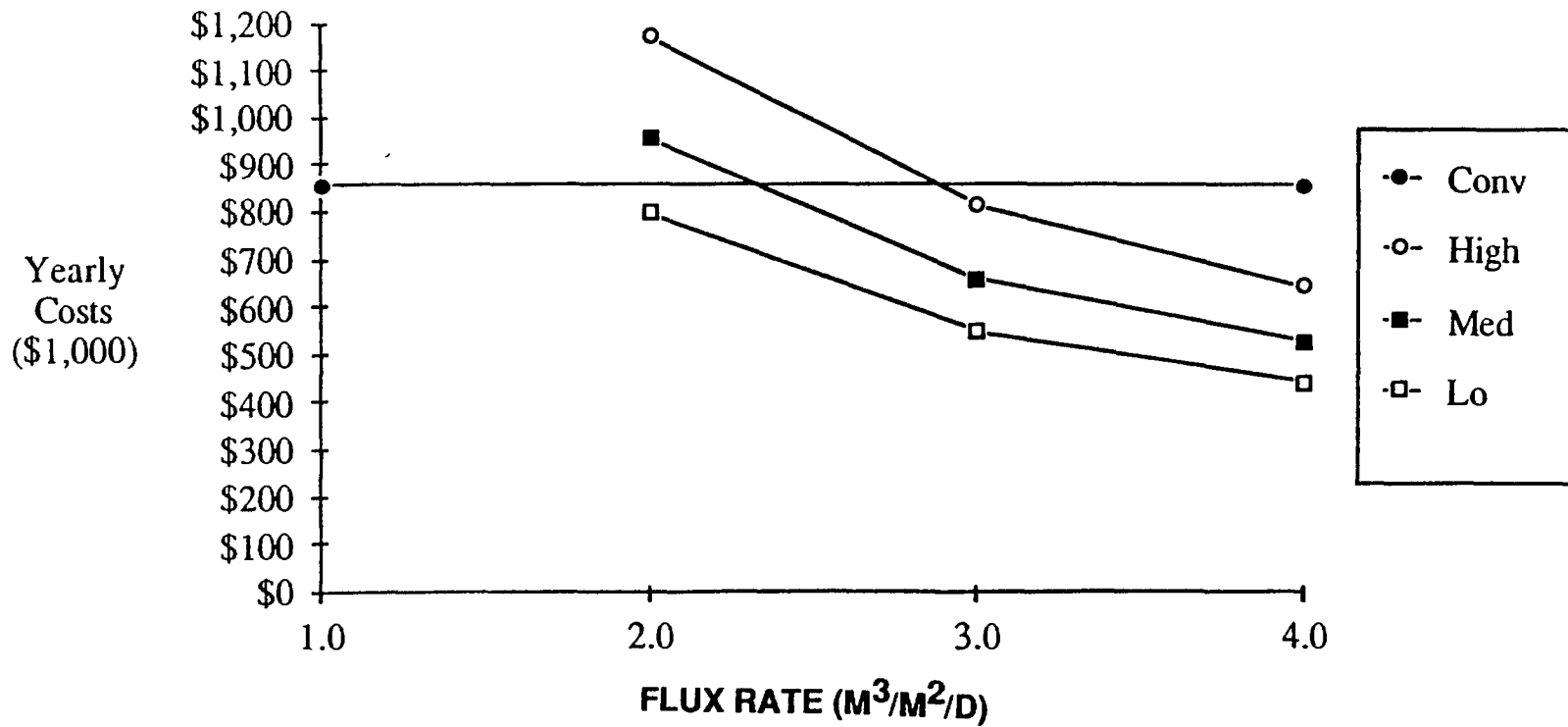


FIGURE 11

Yearly Costs for Conventional and Membrane Based Processes for Produced Water Treatment  
(10 year amortization of capital)

