Computer Optimization of Reverse Osmosis Processing of Soy Whey with Cellulose Acetate Modules

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Synopsis

In processing full-fat soy flour to an acid-precipitated lipid protein concentrate a byproduct whey fraction results which, because of its high biological oxygen demand, represents a serious disposal problem. Processing of food waste streams by reverse osmosis has received considerable attention because of its low theoretical energy requirement since no phase change is involved. In a previous study at this Center a mathematical model was developed for the diffusive transport of soy lipid protein concentrate whey across cellulose acetate membranes. In this study, pumping energy and power costs combined with membrane life and replacement costs were incorporated into the original model to provide a basis for optimization from an energy cost standpoint. Computer-simulated runs were compared with experimental pilot-plant runs, and the agreement between predicted and actual results was quite good. Water flux rates were in the range of 3 to 7 gal/ft^2/day. Computer runs were used to optimize the processing of 100,000 gal/day of soy whey at 8000 ppm biological oxygen demand (BOD). Costs were at a minimum with a six-stage tapered flow primary reverse osmosis treatment over a porous cellulose acetate membrane, followed by a six-stage tapered flow reverse osmosis over a tighter membrane. BOD reduction was 94% at a cost of $5.45/1000 gal.

INTRODUCTION

The processing of full-fat soy flour to obtain an acid-precipitated lipid protein concentrate (LPC) curd[1] results in a whey byproduct which, because of its high biological oxygen demand (BOD), represents a serious disposal problem.

Membrane processing by reverse osmosis (RO) provides a new technology for treating this type of waste. A major advantage of reverse osmosis over evaporation or drying lies in the fact that no phase change is needed, thus lowering the energy required. The discovery by Reid and Breton[2] of cellulose acetate as an effective membrane material, together with the demonstration by Loeb and Sourirajan[3] of high-flux cellulose acetate membranes, initiated the recent interest in membrane science. Studies by the Office of Saline Water of the U.S. Department of the Interior in pursuit of techniques for the desalination of sea water[4-9] have further demonstrated that RO is an economically viable process. Because of these studies the basic physical relationships involved in reverse osmosis have been well documented,[10-13] and coefficients for soy LPC whey have been defined for the molecular (Fickian) diffusion mechanism observed.[14]

The objective of this study was to simulate the basic flow models by computer and, using this simulation, to determine the operating conditions that gave optimal results for BOD reduction and water removal at minimal operating (membrane replacement plus pumping) cost.

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ANALYTICAL METHODS

Feed solutions as well as the permeate and retentate streams from each run were analyzed for total solids by duplicate evaporations on a steam bath. Proximate analyses were conducted on spray-dried and freeze-dried samples. Moisture, crude fat, ash, and protein analyses were by Official AOCS Methods.[15] Nonprotein nitrogen was determined by the method of Becker et al.[16] Sugars were from the method of Black and Bagley.[17]

EQUIPMENT AND PROCEDURE

Equipment

The RO unit used in our experiments was the OSMO-3319 by Osmonics, Inc., Hopkins, Minn. The unit contains a cellulose acetate membrane, spiral wound to give a module with an effective area of 35 ft^2. The module was housed in a 4-in.-I.D. pressure vessel 2½ ft long. A staged centrifugal pump developed 185

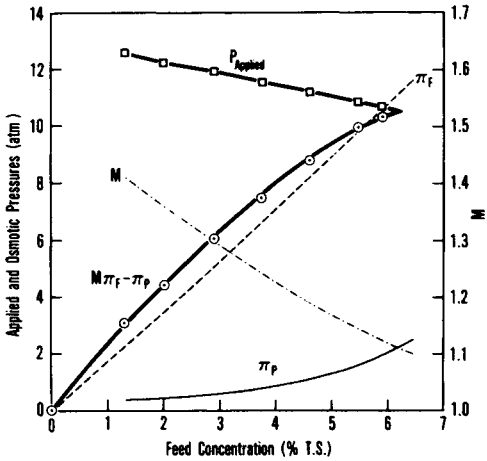

Fig. 1. Limiting concentration for low pressure RO unit (M-89, 25°C, P_i = 185 psig, (ΔP = 5 psig/stage, C = 1.34% T.S.).

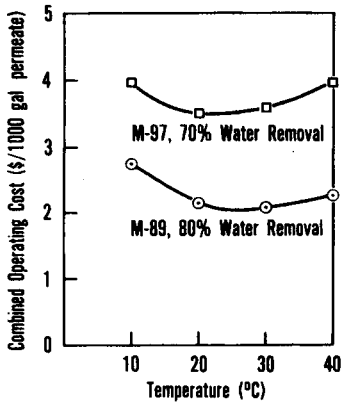

Fig. 2. Operating costs as a function of temperature (P_i = 185 psig, ΔP = 5 psi/stage, C = 1.34% T.S.).

TABLE I
Diffusion Model Equations for Soy LPC Whey

[1] $A_{M\text{-}89} = \dfrac{0.21727}{10^{0.00058P}} \exp\dfrac{-4754}{RT_a}$

[2] $A_{M\text{-}97} = 0.19724 \exp\dfrac{-5226}{RT_a}$

[3] $B_{M\text{-}89}\ (\times 10^6) = 15.22 + 0.038T^2 - 0.180CT$
[4] $B_{M\text{-}97}\ (\times 10^6) = 8.17 + 0.0022T^2 - 0.031CT$
[5] $r_{M\text{-}89} = 0.8786 - 0.0055C^2 - 0.000049T^2 + 0.00019CP$
[6] $r_{M\text{-}97} = 0.7862 - 0.0465C + 0.015T - 0.00026T^2 + 0.000208CP$
[7] $M_{M\text{-}89} = 1.03 + 860J_1$
[8] $M_{M\text{-}97} = 0.975 + 1620J_1$
[9] $J_{1\ M\text{-}89}\ (\times 10^5) = -5.578 + 0.186P + 0.4036C^2 + 0.0125T^2 - 0.1408CT - 0.02867CP + 0.00296TP$
[10] $J_{1\ M\text{-}97}\ (\times 10^5) = -3.769 + 0.090P + 0.1970C^2 + 0.0044T^2 - 0.0869CT - 0.01262CP + 0.00260TP$
[11] $\rho = 0.9996 + 0.0039C + 0.00007T + 0.00012C^2 - 0.0000064T^2 + 0.0000025TC$
[12] Same as [11][a]
[13] $\pi = 0.0059C\rho\ T_a\ (M_w = 139)$
[14] Same as [13][a]
[15] $r_{M\text{-}89}$ permeate on M-97 = 0.885
[16] $\pi = 0.0109C\rho\ T_a\ (M_w = 75)$
[17] Same as]16[[a]
[18] $F_{M\text{-}89,\ M\text{-}97}\ (\times 10^5) = 46.6 + 0.179P + 2.70 \times 10^6 J_1$
[19] $F_{M\text{-}89}$ permeate on M-97 $(\times 10^5) = -253.5 + 1.83P + 1.89 \times 10^6 J_1$
[20] Hydrolysis rate = $0.44 - 0.0304T + 0.00426T^2$ days^{-1} at pH 4.5
[21] Membrane life (days) = 2400/hydrolysis rate

[a] In this case C_P is to be substituted for C. Any errors due to the implied assumption of similar solutes in feed and permeate were found to be negligible.

psig pressure. Both tight and intermediate porosity modules were studied. The tight membrane was rated at 97% NaCl rejection and the intermediate, at 89%, with feed solution containing 1000 ppm NaCl pumped at 77°F and 400 psig. A cartridge prefilter completed the unit.

The feed tank was an agitated 30-gal stainless-steel jacketed kettle. Temperature was adjusted and controlled by flow through an Alfa-Laval plate heat exchanger serviced by hot water and a Borg-Warner brine chiller. Computer simulation was carried out on an IBM 1130 computing system.

Procedure

Pure water permeation rates at various operating temperatures and pressures were determined for the modules before running them on test solutions. This provided a bench mark for determining when the modules were adequately rejuvenated after a test run. Cleaning was discontinued when the modules regained 95% or more of their standard water rate.

For the multistage experiments, feed solution was pumped to the module and the permeate and retentate streams were collected with the system operating in single pass mode. The outlet pressure of the retentate was observed, and this value was used as the applied pressure for the following run in which the retentate stream was fed to a clean module of the same type. This process was continued

TABLE II
Values Observed in Pilot Plant Reverse Osmosis vs Computer Predicted Values Feed: 772 lb Soy LPC Whey at 1.6% Total Solids (T.S.)

	M-89 Step 1 treatment Six-stage single pass						M-97 Step 2 treatment Six-stage single pass					
Stage	S-1	S-2	S-3	S-4	S-5	S-6	S-1	S-2	S-3	S-4	S-5	S-6
Permeate												
Observed lb	257	156	97	58	38	26	128	96	74	56	40	32
Predicted lb	261	156	92	53	29	16	141	103	75	54	39	27
Observed T.S. %	0.15	0.21	0.36	0.69	0.74	0.93	0.04	0.06	0.07	0.08	0.10	0.10
Predicted T.S. %	0.18	0.23	0.32	0.46	0.66	0.89	0.04	0.05	0.06	0.08	0.10	0.13
Retentate												
Observed lb	515	359	262	204	166	140	399	303	229	173	133	101
Predicted lb	511	355	264	211	182	167	386	283	208	154	115	88
Observed T.S. %	2.40	3.26	4.23	4.58	5.25	6.96	0.39	0.50	0.65	0.85	1.07	1.26
Predicted T.S. %	2.33	3.24	4.26	5.20	5.92	6.40	0.40	0.52	0.69	0.91	1.18	1.50

through successive stage until the driving force had diminished to a point where it could no longer produce adequate flux.

DISCUSSION

Mathematical Model

With the previously developed model,[14] the parameters A, B, and π for soy LPC whey were defined. For any given set of operating conditions (i.e., temperature, pressure, and feed concentration), the water flux rate and solute rejection could then be calculated. The incorporation of pumping energy requirements and power costs combined with membrane life and replacement costs gave the mathematical model a basis for optimization from a cost standpoint.

Computer Simulation

A computer program was used to simulate the diffusion (Fickian) flow model observed. In this model the solvent and solute fluxes can be related to the operating conditions in the following form:

$$J_1 = A[(P/14.7) - (M\pi_F - \pi_P)] \quad (1)$$

$$J_2 = B(C_{2F} - C_{2P}) \quad (2)$$

The applicability of the relatively simple reverse osmosis equations is probably due to the atypical nature of our soybean whey—unlike commercial soybean whey, the whey from our lipid protein concentrate (LPC) process is clear and essentially free of suspended solids. It also contains considerably less protein since our protein recovery in the curd is better than 95%. A has been shown to have an Arrhenius dependency on temperature, with membrane compaction due to pressure being observed in the 89 module[14] (Table I, eqs. [1] and [2]). B was dependent on feed concentration and temperature[14] (Table I, eqs. [3] and [4]), whereas the rejection was found to be a function of pressure as well as concentration and temperature[14] (Table I, eqs. [5] and [6]). These calculated rejection values were used to find the permeate concentration. M was found to be best expressed as a linear function of J_1 (Table I, eqs. [7] and [8]), and so at this point it was necessary to compute a rough value for J_1 based on the known operating conditions (Table I, eqs. [9] and [10]). Next, the feed and permeate solution

TABLE III
Comparative Energy Consumptions for Alternate Thermal Processes

Vacuum evaporation	kWhr/1000 gal water removed	BTU/1000 gal water removed	lb water removed/ 1000 BTU
Single effect	—	8,330,000	1.0
Triple effect	—	2,980,000	2.8
Spray drying	—	25,000,000	0.3
Reverse osmosis[a]			
89-module	25.2	258,000[b]	32
97-module	35.3	361,000[b]	23

[a] Water removal (75%) in six-stage tapered flow at 185 psig and 25°C.

[b] Based on conversion of thermal energy to electrical energy at 33% power efficiency.

TABLE IV
Comparison of Thermal and Electrical Energy Costs to Process 100,000 gal/day of Soy LPC Whey (1.6% Solids, 8000 ppm BOD)

	Step 1 RO treatment	Step 1 + Step 2 RO treatment	Step 1 + Step 2 + Step 3 RO treatment	Three-effect evaporation
Step 1—reverse osmosis to 6.3% Solids				
Total membrane area, ft^2	11,300	26,800	40,600	—
Membrane replacement cost ($/1000 gal)	0.70	1.74	2.66	—
Electric pumping energy (kWhr/1000 gal)		29.6	74.2	119
Pumping cost ($/1000 gal)	1.48	3.71	5.95	—
Steam cost ($/1000 gal)	—	—	—	8.03
Permeate production (gal/day)	78,400	75,000	74,500	74,600
BOD reduction (%)	81	94	99	100
Total cost $(1000 gal)	2.18	5.45	8.61	8.03
Step 2—Three-effect evaporation to 20% solids				
Steam consumption (lb/1000 gal evaporation)	2980	2980	2980	2980
Steam cost ($/1000 gal evaporated)	8.03	8.03	8.03	8.03
Water evaporated (gal/day)	13,000	17,000	17,600	17,400
Condensed steam (gal/day)	4,600	6,100	6,300	6,200
Step 3—spray drying 20% concentrate				
Water removed (gal/day)	7000	6400	6300	6400
Thermal energy cost ($/1000 gal)	50.0	50.0	50.0	50.0
Spray dried product (lb/day)	11,500	13,000	13,300	13,300
Thermal energy cost (¢/lb)	3.0	3.0	3.0	3.0

densities were calculated (Table I, eqs. [11] and [12]) to determine the solutions' osmotic pressures (Table I, eqs. [13] and [14]), which were based on the van't Hoff equation

$$\pi = C_s R T_a / MW \tag{3}$$

An approximate M_w of 139 was used as previously reported for soy LPC whey.[14] It was then possible to calculate exact solutions for J_1 and J_2 based on eqs. (1) and (2).

For permeate treatment, it was necessary to use modified equations for rejection and osmotic pressure values and an approximate M_w of 75 for permeate (Table I, eqs. [15]–[17]). In this case, J_2 was negligible and so the equations for B could be used or ignored. M values obtained from the rough J_1 values were similar in all cases, as were P values.

In order to scale up the microscopic level calculations to a macroscopic system, it was necessary to calculate the feed intake to a membrane so that it would be possible to determine what fraction of the original feed solution $J_1 + J_2$ represented. In this system the feed rate was found to be linearly related to the applied pressure and J_1 (Table I, eqs. [18] and [19]).

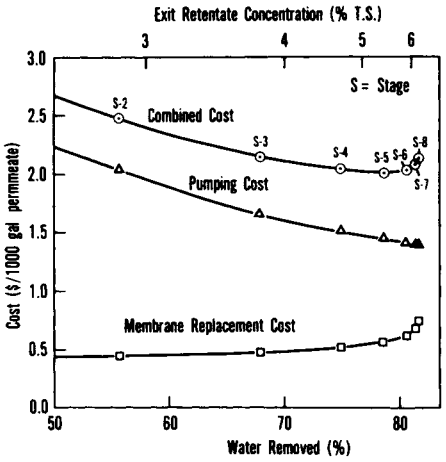

Fig. 3. Operating costs as a function of water removed (M-89, P_i = 185 psig, ΔP = 5 psi/stage, C = 1.34% T.S., T = 25°C).

To determine membrane life, equations were derived from the data of Vos et al.[18] on the hydrolysis of cellulose acetate (Table I, eqs. [20] and [21]). Based on manufacturer's data,[19] a maximum life of 1100 days was assumed. Membrane replacement cost was set at \$5/ft^2 (ref. 19) and a factory treating (100,000 gal/24 hr day) was assumed. Membrane cost for each stage was then found to be

$$\text{membrane cost (\$/1000 gal permeate)} = \frac{(5000)(\text{ft}^2 \text{ membrane area required})}{(\text{membrane life})(\text{flux in gal/day})} \quad (4)$$

Pumping cost, assuming electricity at 5¢/kWhr and consumption of 9.8 A at 110 V per 35 ft^2 module:

$$\text{pumping cost (\$/1000 gal permeate)} = \frac{(37)(\text{ft}^2 \text{ membrane area required})}{\text{flux in gal/day*}} \quad (5)$$

In each case the membrane area required was found by dividing the feed entering one 35 ft^2 module into the total plant feed and multiplying by 35 ft^2/module.

Pressure

Higher pressures are known to increase the flux as predicted in eq. (1). However, this effect is not purely linear for two reasons. First, the polarization modulus M has been found to increase with flux rate and tends to lessen the benefit somewhat of increased operating pressure. Secondly, membrane compaction that reduced the flux rate was observed with the 89 module only. However, for this study energy costs were lowest at the maximum pump pressure (185 psig), and this was the pressure used in the first stage of a tapered flow system. The assumed pressure drop in the module was only about 5 psig, so the concentrate stream feeding the second stage was 180 psig, and so on down to a

* This flux rate includes permeates from all stages which follow those that the pump feeds into since there is no pumping following the first stage.

TABLE V
Comparative Energy Costs to Process 100,000 gal/day of Soy LPC Whey (1.6% Total Solids, 8000 ppm BOD)

Process	Unit operation	Water removed gal/day	Percent of total water removed	Energy cost[a] $/day
Step 1 RO treatment (81% BOD	Reverse osmosis	78,400	79.7	171[b]
removal) followed by three-effect	Three-effect evaporation	13,000	13.2	104
evaporation to 20% solids, then	Spray drying	7000	7.1	350
spray drying				625
Step 1 + Step 2 RO	Reverse osmosis	75,000	75.2	409[b]
treatment (94% BOD removal)	Three-effect evaporation	17,000	17.3	137
followed by 3-effect evap. to	Spray drying	6400	6.5	320
20% solids, then spray drying				865
Step 1 + Step 2 + Step 3	Reverse osmosis	74,500	75.7	641[b]
RO treatment (99% BOD removal)	Three-effect evaporation	17,600	17.9	141
followed by 3-effect evap. to	Spray drying	6300	6.4	315
20% solids, then spray drying				1098
Three-effect evaporation to 20%	Three-effect evaporation	92,000	93.5	739
solids (100% BOD removal)	Spray drying	6400	6.5	320
followed by spray drying				1059

[a] Based on electricity at $0.05/kWhr, steam at $2.70/1000 lb, and natural gas at $2.00/million BTU.
[b] Includes membrane replacement cost.

TABLE VI
Proximate Analyses of Soy LPC Whey Solution, Reverse Osmosis Permeates, and Spray-Dried Concentrate

Component	Soy LPC Whey		M-89 Permeate		M-97 Permeate		Spray-dried concentrate
	As is basis %	Dry basis %	As is basis %	Dry basis %	As is basis %	Dry basis %	%
Protein N-NPN) × 6.25	0.088	5.5	0.0003	0.1	—	—	5.7
Nonprotein N (NPN)	0.019	1.2	0.0027	0.9	—	—	1.2
Ash	0.429	26.8	0.1056	35.2	0.030	49.7	26.1
Crude fat	0.003	0.2	0.0006	0.2	—	—	0.2
Sugars							
Monomer	0.054	3.4	0.0474	15.8	0.012	20.6	2.9
Sucrose	0.352	22.0	0.0663	22.1	0.011	19.3	22.1
Raffinose	0.056	3.5	0.0090	3.0	—	—	3.6
Stachyose	0.301	18.8	0.0129	4.3	—	—	19.3
Others (by difference)	0.298	18.7	0.0552	18.4	0.007	10.4	18.9
	1.60	100.0	0.3000	100.0	0.060	100.0	100.0
BOD (ppm)	8000		1500		450		

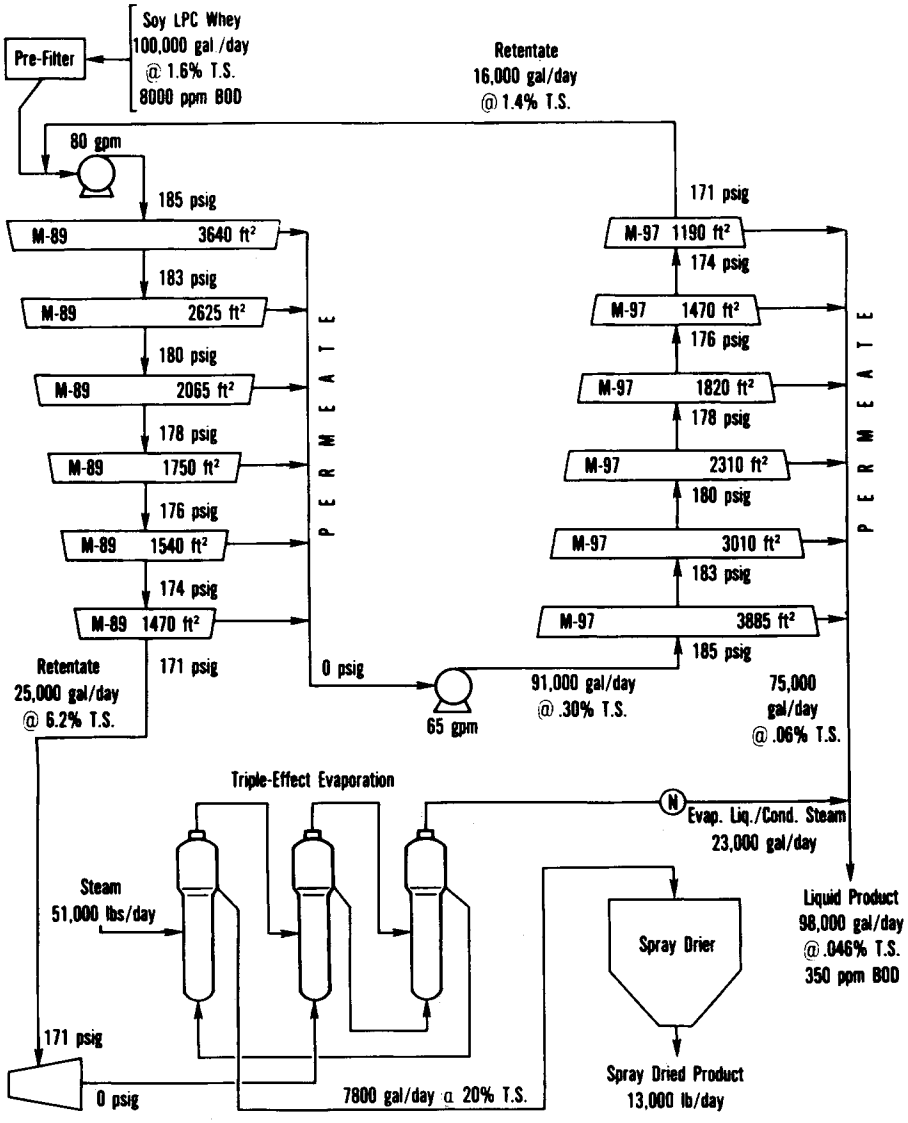

Fig. 4. Soy LPC whey treatment/recovery operation.

concentrate pressure of 160 psig feeding the sixth stage. With the osmotic pressure of the feed increasing with each passage through successive stages and the applied pressure dropping in 5 psig increments, the driving force ($P - M\pi_f + \pi_p$) is reduced to near zero at a solids concentration of 6.3% (Fig. 1).

Temperature

Water flux rate increases with temperature whereas membrane life decreases with temperature. At higher temperatures pumping costs are lower and less membrane area is required to process a given amount. However, the membranes must be replaced more frequently at higher temperatures. Combined pumping

and membrane replacement costs were found to be at a minimum at approximately 25°C (Fig. 2).

Membrane Selection

The combined pumping and membrane replacement cost was significantly lower for the more porous 89 module (Fig. 2). The solute rejection was somewhat lower while the water flux rate was much higher than with the 97 module.

Concentration

Pumping costs on a per gal basis decreased as successive stages were added, since energy was added only at the first stage. However, the addition of each successive stage increased the total membrane area and consequently the membrane replacement cost. A minimum combined pumping and membrane cost was found after six stages (Fig. 3).

Experimental vs. Computer Predicted Values

Soy LPC whey, 772 lb at 1.6% T.S. (total dissolved solids) was prepared in the pilot plant. The whey was processed in six single-pass stages over the 89 module. The permeates from these runs were combined and fed in six single-pass stages over the 97 module. The same runs were made by computer simulation. A modification in the program was necessary to accommodate observed pressure drops between stages of 2–3 psi, but the agreement between pilot plant and predicted results is quite good (Table II).

Alternate Processes

The removal of water by thermal processes such as single and triple effect evaporation and spray drying is shown in Table III. Even though electric energy is much more costly than thermal energy, the order of magnitude of energy consumption of phase change processes over membrane processes is enormous. The energy unit costs for this study were a steam cost of $2.70/1000 lb, natural gas at $2.00/million BTU, and an electric power cost of $0.05/kWhr.

RESULTS

Preliminary runs were made by computer simulation, treating each variable independently to determine the operating conditions and mode of operation where the combined pumping and membrane replacement costs were lowest. The best operating conditions were found at 25°C and 185 psig. Optimum water removal by RO was found to be approximately 75% with a solids content of 6.3% in the retentate. Economics dictated that the 89 module with its higher flux rates be used in the Step 1 treatment to remove the majority of the solids, followed by the 97 module for the Step 2 and Step 3 treatments of the 89-M permeate. The balance of the water (ca. 25%) would be removed by conventional evaporation to a solids content of 20% (further concentration results in viscosity problems) followed by spray drying.

Computer simulation runs were made as follows: (a) Step 1 RO treatment on 89 module. A BOD reduction of 81% was predicted at a combined pumping and membrane replacement cost of $2.18/1000 gal (Table IV). (b) Step 1 RO treatment on the 89 module, followed by Step 2 RO treatment of the 89 permeate on the 97 module. This mode of operation predicted a BOD reduction of 94% at a combined cost of $5.45/1000 gal. (c) Step 1 RO treatment on the 89 module, followed by Step 2 RO treatment of the 89 permeate followed by Step 3 RO treatment of the 97 permeate. This method predicted a BOD reduction of 99% at a cost of $8.61/1000 gal. This treatment was seen to be slightly more costly than straight triple-effect evaporation (Table IV).

The total energy and membrane replacement costs to process 100,000 gal soy LPC whey by the various methods were calculated (Table V). Step 1 RO treatment on the 89 module followed by Step 2 treatment on the 97 module (Fig. 4) resulted in a 94% BOD reduction at a cost that was almost $200.00/day less than triple-effect evaporation. If the spray-dried product could be sold for as little as 3¢/lb it would more than offset the cost of spray drying. The spray-dried product has a high carbohydrate and mineral content. If the permeate was recycled to process, an additional savings of approximately $50.00/day could be realized assuming a fresh water cost of approximately $0.50/1000 gal. (Table VI). The costs observed here seem favorable when compared to alternative water removal methods[20–33] and are comparable to other figures arrived at for membrane processes in general.[20,26,27,34–41]

Analyses were made by L. T. Black, J. D. Glover, F. B. Alaksiewicz, and K. M. Rentfro. Pilot-plant equipment was operated by R. L. Brown. The mention of firm names or trade products does not imply that they are endorsed or recommended by the U.S. Department of Agriculture over other firms or similar products not mentioned.

APPENDIX: LIST OF SYMBOLS

A	Solvent permeability coefficient	g/cm^2 sec atm
B	Solute permeability coefficient	cm/sec
C	Solute concentration in the feed	% total solids
C_P	Solute concentration in the permeate	% total solids
C_S	Solute concentration	g/liter
C_{2F}	Solute concentration in the feed	g/cm^3
C_{2P}	Solute concentration in the permeate	g/cm^3
F	Feed flow rate to membrane	g/cm^2 sec
J_1	Solvent flux rate	g/cm^2 sec
J_2	Solute flux rate	g/cm^2 sec
M	Polarization modulus	...
M_w	Molecular weight of solute	g/mole
P	Applied hydraulic pressure	psig
r	Solute rejection	...
R	Gas constant = 0.082 l-atm/mole-K = 1.987 cal/mole-K	...
T	Feed solution temperature	°C
T_a	Absolute temperature of feed solution	K
ρ	Density of feed solution	g/cm^3
π	Osmotic pressure	atm
π_F	Osmotic pressure of feed solution	atm
π_P	Osmotic pressure of permeate	atm

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