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Concentration of Dilute Acetone–Water Solutions Using Pervaporation

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Abstract

The separation of acetone–water mixtures by pervaporation has been studied. Four membranes were evaluated: a silicone composite (SC) membrane, a polydimethylsiloxane (PDMS) membrane, a polymethoxysiloxane (PMS), and a polyether-block-polyamide copolymer (PEBA) membrane. The silicone composite membrane exhibited a higher flux and selectivity than any of the other membranes studied. At a feed temperature of 50°C, a permeate-side pressure of 1 torr, and a feed concentration of 5.0%, the silicone composite membrane had a flux of 1.1 $kg/m^2 \cdot h$ and a selectivity of 50. The effects of temperature and permeate-side pressure on membrane transport were studied using the SC membrane. An increase in temperature increased the flux exponentially, but had little effect on selectivity. An analysis of the data shows that the trend agrees quite well with an Arrhenius-type relationship. As the permeate-side pressure increased, the flux decreased in a sigmoidal fashion over the range evaluated. Selectivity did not change significantly over the lower portion of the pressure range studied. The effect of feed concentration on flux and selectivity was also investigated.

INTRODUCTION

The separation of acetone–water mixtures by pervaporation has been studied. Pervaporation can be successfully used to recover various products from fermentation broths including ethanol, acetone, and butanol. One of our research goals is to integrate pervaporation with fermentation in an overall bioprocess production scheme. As a first step in that direction we have been investigating the pervaporative transport of various organics from aqueous systems. The separation of ethanol–water solutions and butanol–water solutions by pervaporation has been previously studied by our

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research group (*1–4*). This paper concentrates on the separation of acetone–water solutions.

Process Principles

Pervaporation (PV) selectively separates a liquid feed mixture, typically using a nonporous polymeric membrane. The separation is not based on relative volatilities like distillation, but on the relative permeation rates through the membrane. The prevailing model for PV is a solution-diffusion mechanism (*5*) which is common to many membrane processes. The transport theory and models have been presented elsewhere (*6–10*), and a summary of some important concepts is presented here. The permeating component of the feed goes into solution with the membrane at its surface and then diffuses through the membrane. A vacuum or sweeping gas is applied to the membrane on the permeate side. The permeating component desorbs from the membrane as a vapor and can be collected or released as desired. The chemical potential difference across the membrane (from the feed to permeate side) is the driving force for separation. The permeating component transports through the membrane because its partial pressure on the permeate side is lower than in the saturated vapor.

The permeability of a component can be expressed as a function of diffusivity and solubility in the polymer. Diffusivity and solubility are highly dependent on concentration, and there is significant interaction between the components in the mixture. Experimental studies are essential in determining separation performance and evaluating process parameters for scale-up and design.

The effectiveness of PV is measured by two parameters, flux and selectivity. Consider a binary mixture of components A and B. The flux is the rate of permeation per membrane area and can be expressed for the total permeate or for each specific component.

$$J_T = \text{total flux}$$

$$J_A = \text{flux of component A}$$

$$J_B = \text{flux of component B}$$

The flux has dimensions of mass/(area $\times$ time), $[M/L^2 \cdot t]$. Typical units would be $g/(cm^2 \cdot s)$ or $kg/(m^2 \cdot h)$, etc. The flux can be measured by knowing the mass of permeate collected, membrane area, and time of the experi-

mental run. The flux can also be defined by the phenomenological expression:

$$J_i = -L_i \frac{\Delta\mu_i}{l} \tag{1}$$

where L_i = phenomenological coefficient
$\Delta\mu_i$ = chemical potential driving force across membrane
l = membrane thickness

Selectivity is a measure of the membrane's separation efficiency. It is a ratio of the mass fractions of components A and B for the permeate and the feed.

$$\alpha_B^A = \frac{y_A/y_B}{x_A/x_B} \tag{2}$$

where y_A = mass fraction of component A in permeate
y_B = mass fraction of component B in permeate
x_A = mass fraction of component A in feed
x_B = mass fraction of component B in feed

The previous equation is for selective permeation of component A. A value greater than unity indicates the selective permeation of A over B, and a value less than unity indicates the selective permeation of B over A. The separation effectiveness is sometimes expressed by an enrichment factor, β. The enrichment factor is the ratio of a component's concentration in the permeate to its concentration in the feed.

$$\beta_A = y_A/x_A \tag{3}$$

APPLICATIONS

Pervaporation separations can be classified into three types. A type 1 separation is the removal of water from an aqueous/organic mixture. This type of separation uses water selective hydrophilic membranes to permeate water from the feed solution. Pervaporation is commercially used in the pharmaceutical industry for dehydration of solvents such as ethanol, isopropanol, and acetone. Organic solvents are used in this industry for many separations operations, including extraction, precipitation, crystallization, adsorption, ion exchange, and chromatography. Some water dissolves in

these organic solvents during use, and must be removed so that the solvents can be recovered and reused (*11*). Isopropanol and acetone are the two most widely dehydrated organic solvents according to Zenon Environmental, Inc. (*12*). Polyvinyl alcohol (PVA) membranes are one type commonly used for dehydration separations.

A type 2 separation is the permeation of organics from an aqueous/organic mixture. This type of separation uses hydrophobic/organophilic membranes, such as silicone-based polymers, to concentrate the organics in the permeate. One of the primary applications for this type of separation is the selective removal of organics from dilute biochemical processing streams. The removal and recovery of organic contaminants from wastewater streams has also been investigated (*13*).

A type 3 separation involves the permeation of a particular organic from an anhydrous mixture. This type of separation is not in wide commercial use.

The primary application discussed in this paper is the selective removal of acetone from dilute mixtures similar to those occurring in fermentation operations. A review of the selective recovery of alcohols (including *n*-butanol, ethanol, and isopropanol) from fermentation broths by pervaporation has been published (*3*). Pervaporation has the potential to be used in the biochemical industry for product recovery as well as by-product concentration control. The efficiency of solvent fermentations can be significantly improved by a continuous removal of the product from active fermentation broths (*14*). The fermented solvents tend to inhibit microorganism productivity with increasing concentration. Therefore to increase fermentation efficiency, they should be removed from the fermentation broth, while the microorganisms should be kept inside. The separation process chosen should minimize thermal, chemical, and mechanical stress upon the microorganisms. This requirement makes pervaporation much more attractive than competitive processes such as reverse osmosis, distillation, and solvent extraction (*15*).

Acetone Separation from Biochemical Processes

Larrayoz and Puigjaner (*16*) studied butanol extraction through pervaporation in acetobutylic fermentation. The authors stated that substrate conversion during acetobutylic fermentation is limited by the concentration of butanol in the fermentation medium. Alternatives such as liquid–liquid extraction have been proposed, but most solvents used for extraction are toxic to bacteria. Fermentation coupled with pervaporation was compared to fermentation without pervaporation. Pervaporation was performed using air as the carrier gas and a silicone tube as the membrane. Fermentation with pervaporation was completed in 40 hours, while fermentation without

pervaporation required 48 hours. The butanol flux increased from 0.00442 to 0.01105 kg/(m^2·h) for butanol feed concentrations ranging from 1.4 to 1.75%. The acetone flux could not be obtained experimentally due to poor condenser efficiency.

The use of pervaporation in the continuous fermentation of acetone, butanol, and ethanol was investigated by Gudernatsch et al. (*14*). Composite hollow fiber membranes with an active layer of polydimethylsiloxane (PDMS) were coupled to fermentation operations. For the acetone–butanol–ethanol fermentation, the mass fraction of solvents in the fermentation broth was 0.0014. At a temperature of 37°C and a permeate pressure of 19 torr, pervaporation resulted in a total flux of 2 kg/(m^2·h), an acetone selectivity of 8.5, a butanol selectivity of 11.0, and an ethanol selectivity of 6.5.

Gudernatsch and coworkers (*15*) proposed a membrane separation process to allow a selective, continuous, and controllable removal of volatile bioproducts such as ethanol, acetone, and butanol from fermentation broths. The authors stated that pervaporation across highly permeable solvent-selective membranes seemed to be the most viable solution to this separation problem since it avoids the high mechanical, thermal, or chemical stresses often exerted upon the microorganisms by competitive processes such as reverse osmosis, distillation, or solvent extraction, and also holds the largest potential for simultaneous preconcentration of the product. The specific membrane selected was a composite hollow fiber type with the active surface inside. A porous support is coated with a solvent-selective layer of polydimethylsiloxane (PDMS). Characteristic results of vacuum pervaporation experiments obtained during 1 month of continuous fermentation of ethanol with *Saccharomyces cerevisiae* were presented to show the feasibility of the proposed separation process.

Masuda and coworkers (*17*) studied the pervaporation of organic liquid–water mixtures through substituted polyacetylene membranes. The membrane used for most of the experiments was a poly[1-(trimethylsilyl)-1-propyne] (PTMSP) membrane. The pervaporation of combinations of water and ethanol, acetonitrile, and acetone were evaluated. For the acetone–water separation using a PTMSP membrane, a flux of 2.17 kg/(m^2·h) and an acetone selectivity of 76 were observed at a process temperature of 30°C, a permeate pressure of 0.1 torr, and an acetone feed concentration of 10%.

Matsumara and Kataoka (*18*) studied the separation of dilute aqueous butanol and acetone solutions through an oleyl alcohol liquid membrane. This membrane was selected as the most suitable for separating volatile products resulting from acetone–butanol fermentation. For the acetone–water mixtures, feed solutions of less than 16% acetone were used. At a

feed concentration of 1% and a temperature of 30°C, the flux was approximately 0.04 kg/(m^2·h), and the selectivity was 160.

Ohya and coworkers (*19*) studied the concentration of acetone and *n*-butanol solutions by pervaporation using a porous polypropylene hollow fiber membrane. Acetone and *n*-butanol were concentrated in the permeation, but the separation factor of both was lower than their vapor–liquid equilibria. The permeation fluxes of acetone, *n*-butanol, and water increased exponentially with the increase of feed temperature and also increased with the decrease of permeate-side pressure. For a feed temperature of 30°C, a downstream pressure of 15 torr, and an acetone mole fraction of 0.0015 in the feed, the total permeation flux was approximately 1.1 kg/(m^2·h) and the mole fraction of acetone in the permeate was 0.075.

Nguyen and Nobe (*20*) studied the extraction of dichloromethane, chloroform, bromoethane, acetone, and ethanol in dilute aqueous solutions by pervaporation using silicone tubular membranes. The authors noted that possible applications for the removal of organic solutes are the treatment of contaminated industrial and municipal water supplies and extraction of organics produced by fermentation. The solutes evaluated did not effect the permeation flux of water, and the presence of another organic solute did not influence the permeation of a given solute. The acetone component flux was approximately 0.011 kg(m^2·h) and the weight fraction of acetone in the permeate was 0.44 for a feed solution with an acetone weight fraction of 0.00134. For the same feed solution and a cellulose acetate membrane, the weight fraction of acetone in the permeate was 0.33, while for a microporous PTFE membrane the weight fraction of acetone in the permeate was 0.47. These values correspond to a sweeping gas (helium) flow rate of 0.095 cm^3/s and a temperature of 27°C.

Peirlot and Pons (*21*) examined the use of microporous hydrophobic membranes in a pervaporation system for elimination of inhibitory volatile metabolites produced in some solvent fermentations. Experiments were run without fermentation in order to study the effects of different operation parameters including liquid flow rate, gas flow rate, temperature, and membrane surface. Components included in the feed solutions for the experiments were ethanol, isopropanol, butanol, and acetone. A model was developed which can be used in the design of a pervaporation system for use with biological systems.

Other Acetone Separations

Karachevtsev and coworkers (*22*) studied pervaporation of ethanol–water and acetone–water mixtures using the organophilic composite membranes MDK and MDK-U (having block copolymers polyphenylsilsesquioxane/polydimethylsiloxane and polyurethane, respectively, as selec-

tive layers). These membranes withstand cyclic loads, permitting their use in the large-scale manufacturing of various separation modules, including spiral-wound and plate-and-frame configurations. The separation factor of the acetone–water mixture was higher for the MDK-U membrane than for the MDK membrane. At a downstream pressure of 2 torr, a temperature of 25°C, and a feed concentration of 2.5 wt% acetone, a total flux of 0.75 kg/(m^2·h) and a separation factor of 52 were observed for the MDK-U membrane.

Sakohara and coworkers (*23, 24*) used an acrylamide–acrylic acid copolymer gel to dewater an acetone–water mixture by pervaporation. The membrane was prepared in pores of a thin ceramic membrane of silica-alumina by copolymerizing acrylamide of primary monomer and *N,N'*-methylene-bis(acrylamide) as a crosslinking agent. At a feed temperature of 50°C, the flux of water was 6.17 kg/(m^2·h) and the separation factor reached about 2000 at a feed concentration of 95 mol% acetone.

Featherstone and Cox (*25*) studied the separation of acetone–water solutions by pervaporation using polypropylene membranes. The polypropylene and ethylene–propylene copolymer films used were permselective toward acetone at <75 mol% acetone in the feed, and permselective toward water at >80 mol% acetone in the feed. At a temperature of 84°C, a downstream pressure of 49 torr, and a 45% acetone feed, a flux of 0.11 kg/(m^2·h) and a separation factor of 2.5 were observed.

Bell and coworkers (*26*) measured the permeation rates of four different alcohols, water, and acetone in two types of polymers: polydimethylsiloxane (PDMS), a hydrophobic polymer, and cellulose, a hydrophilic polymer. Various models relating the molecular and chemical properties of the permeating components and the properties of the various polymers and membranes were developed and discussed.

Hauser and coworkers (*27*) measured the solubilities of organic–water mixtures, including water–ethanol, water–dioxane, and water–acetone, in polyvinyl alcohol using gas chromatographic and infrared spectroscopic analysis. These solubility measurements were then used in a model to calculate a predicted separation curve for water–ethanol mixtures.

Brun and coworkers (*28*) investigated pervaporation of dilute aqueous binary mixtures of benzene, chloroform, acetone, and ethanol through nitrile–butadiene and styrene–butadiene copolymers. Sorption of water and dilute aqueous solutions was investigated and a model relating the pervaporation results to equilibrium properties of the membranes was developed.

Bindal and Misra (*29*) compared the separation of two binary liquid systems (ethanol–water and methanol–acetone) by sorption to pervaporation. Wijmans and coworkers (*13*) studied the removal and recovery of

organic contaminants including chloroform, trichloroethane, ethyl acetate, acetone, and ethanol from aqueous streams by pervaporation.

EXPERIMENTAL MATERIALS AND METHODS

The pervaporation system used in the research is a bench-scale test unit shown in Fig. 1. The feed vessel is immersed in a temperature bath to keep the feed at constant temperature. The feed solution is continuously pumped through the membrane test cell and returned to the feed vessel. Two cold finger condensers immersed in liquid nitrogen baths are used to collect the permeate. A direct drive vacuum pump and a vacuum regulator are used to control the permeate-side pressure. A pressure meter and manometer are used to measure downstream pressure. Permeate flux is measured gravimetrically, and a refractometer is used to determine the feed and permeate concentrations.

The experimental studies utilized four different hydrophobic membranes and a hydrophilic polyvinyl alcohol (PVA) membrane. The first membrane studied was a hydrophobic/organophilic silicone composite (SC) membrane (Product Code: Type 1060) obtained from the GFT Division of Carbone USA Corp. in Boonton, New Jersey. The second membrane studied was a polydimethylsiloxane (PDMS) membrane originally obtained from the General Electric (GE) Corporation, Medical Systems Division in Schenectady, New York (now Mempro in Troy, New York). The final two hydrophobic membranes studied were a polymethoxysiloxane (PMS) membrane, also obtained from GFT, and a polyether-block-polyamide copolymer (PEBA) membrane. The hydrophilic PVA membrane was also obtained from GFT.

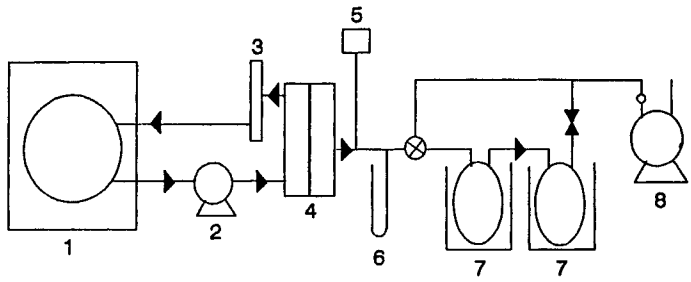

FIG. 1. Membrane pervaporation system process diagram. System components: temperature-controlled feed tank (1), feed pump (2), flowmeter (3), membrane cell (4), pressure meter (5), pressure manometer (6), permeate condensers (7), vacuum pump and pressure regulator (8).

The runs used to compare the four hydrophobic membranes were conducted at a benchmark condition which is a set of standard test conditions that has been used this past year in our laboratory for conducting preliminary screening tests on membranes. The benchmark process conditions are a feed temperature of 50°C, a feed flow rate of 1.5 L/min, a permeate side pressure of approximately 1 torr (mmHg), and a membrane area of 28.7 cm^2. Runs at the benchmark condition were repeated several times during the temperature and pressure studies to evaluate any change in membrane characteristics with increased processing time.

RESULTS AND DISCUSSION

A hydrophilic polyvinyl alcohol (PVA) membrane and a hydrophobic polymethoxy siloxane (PMS) membrane were tested at feed concentrations ranging from 0 to 100 wt% acetone. The resulting permeate concentrations were plotted on the same scale with acetone/water vapor–liquid equilibria data (Fig. 2). It can be seen that the PVA membrane is very selective for permeating water from acetone/water binary mixtures. In contrast, the PMS membrane selectively permeates acetone from the binary solutions. From comparison with the vapor–liquid equilibria data, it is evident that

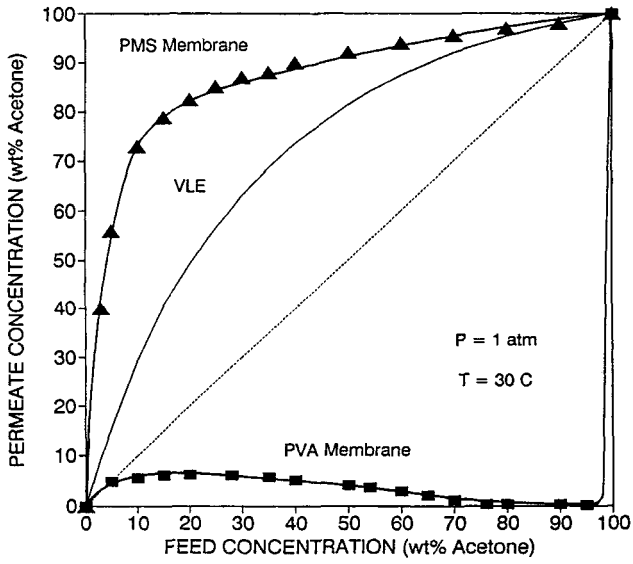

FIG. 2. Performance of PMS and PVA membranes for separation of acetone–water mixtures in comparison with thermodynamic vapor–liquid equilibria.

the pervaporation processes utilizing both membranes are much more selective than processes such as distillation in which the mechanism of separation is based on thermodynamic equilibria.

Four hydrophobic/organophilic membranes were tested experimentally at the benchmark condition and compared on the basis of flux and selectivity. The silicone composite (SC) membrane exhibited a higher flux than any of the other membranes studied. At an acetone feed concentration of 4.5% (weight), the SC membrane had a flux of 1.1 kg/(m^2·h) compared with fluxes ranging from 0.2 to 0.7 kg/(m^2·h) for the other three membranes studied (Fig. 3). The flux of the SC membrane increased from 0.36 to 1.1 kg/(m^2·h) as the acetone concentration in the feed increased from 0 to 4.5%.

The acetone selectivity of the SC membrane was high compared with that of the other membranes studied. At a feed concentration of 4.5% acetone, the acetone selectivity of the SC membrane was 50 compared to 55 for the PDMS membrane, 13 for the PMS membrane, and 3.7 for the

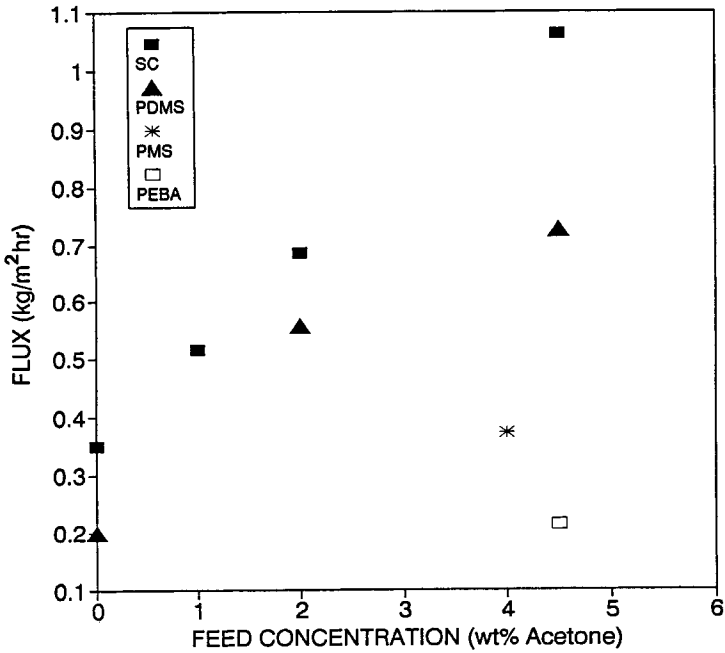

FIG. 3. Flux vs feed concentration at 50°C and 1 torr downstream pressure.

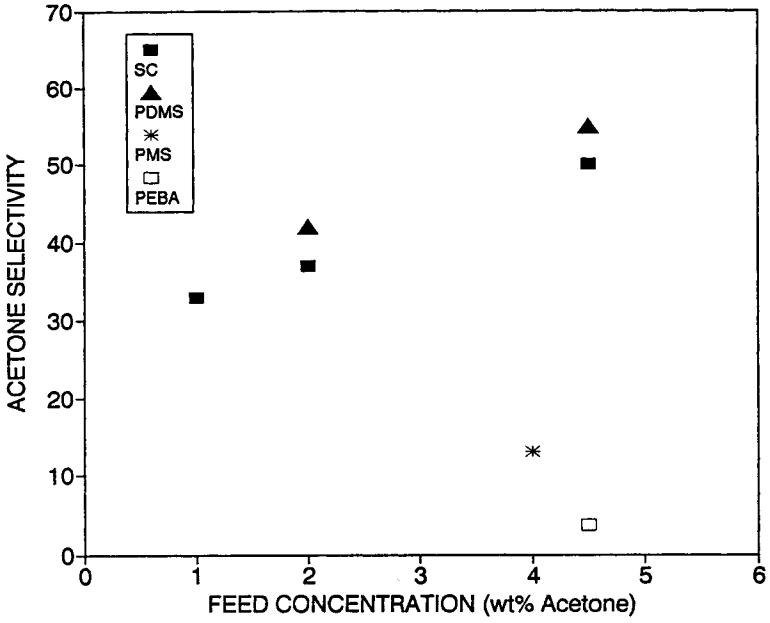

FIG. 4. Acetone selectivity vs feed concentration at 50°C and 1 torr downstream pressure.

PEBA membrane (Fig. 4). These selectivities correspond to permeate acetone concentrations of 70% for the SC membrane, 72% for the PDMS membrane, 35% for the PMS membrane, and 15% for the PEBA membrane (Fig. 5). The PDMS membrane had a slightly higher selectivity than the SC at feed concentrations of 2 and 4.5% acetone, but the SC membrane had a significantly greater flux in both cases. Therefore, the SC membrane was chosen as the best membrane for further studies. As the feed concentration increased from 1 to 4.5% acetone, the acetone selectivity of the SC membrane increased from 33 to 50.

At a feed concentration of 2%, the acetone selectivity was 37 using the SC membrane and 42 using the PDMS membrane. These selectivities are very good in comparison with those obtained through concentration of other dilute organic mixtures by pervaporation. The selectivities resulting from pervaporation using the PDMS membrane for feed concentrations of 2% ethanol, *t*-butanol, and *n*-butanol were 9, 20, and 28 respectively (*1, 2*). Clearly the acetone selectivity is superior to that of these other organics typically found in fermentation broths.

The SC membrane was utilized for studies to determine the effect of temperature and pressure on membrane performance. All temperature and

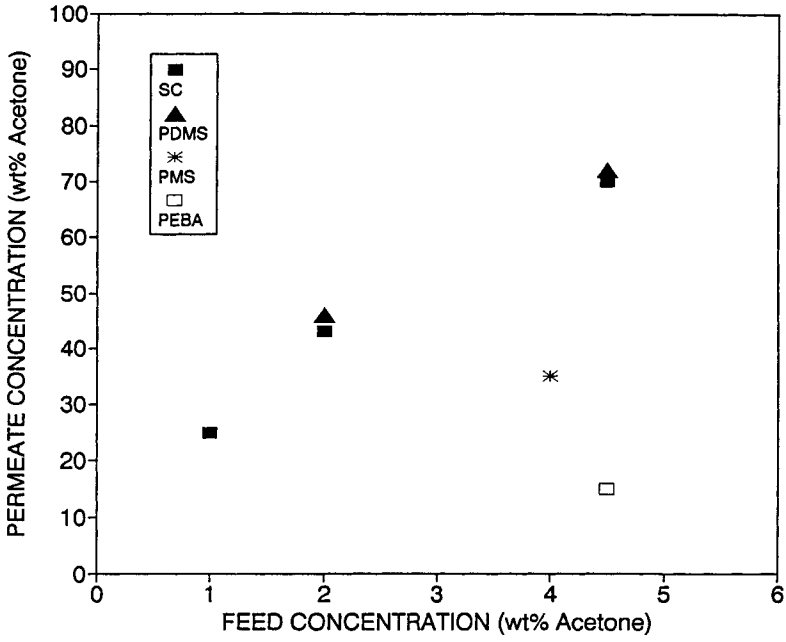

FIG. 5. Permeate concentration vs feed concentration at 50°C and 1 torr downstream pressure.

pressure studies were performed with a feed concentration of 2% acetone. Figure 6 shows the effect of temperature on total permeate flux and on the acetone and water component fluxes. Temperatures ranging from 20 to 68°C were studied. The optimum production rate was observed at the maximum temperature studied, 68°C. As the temperature increased from 20 to 68°C, the total flux increased from 0.2 to 1.3 kg/(m^2·h) and the acetone component flux increased from 0.1 to 0.5 kg/(m^2·h).

The natural log (ln) of the flux was graphed versus the reciprocal absolute temperature, and an excellent linear correlation resulted, showing that the trend agrees quite well with an Arrhenius-type relationship (Fig. 7). This relationship held true for the actone and water component fluxes as well as the total flux. The predicted total and specific component fluxes are shown as solid lines on Figs. 6 and 7. The equations found for the total and component fluxes are:

$$J_T = 77020\, e^{-3760/T} \tag{4}$$

$$J_A = 20610\, e^{-3625/T} \tag{5}$$

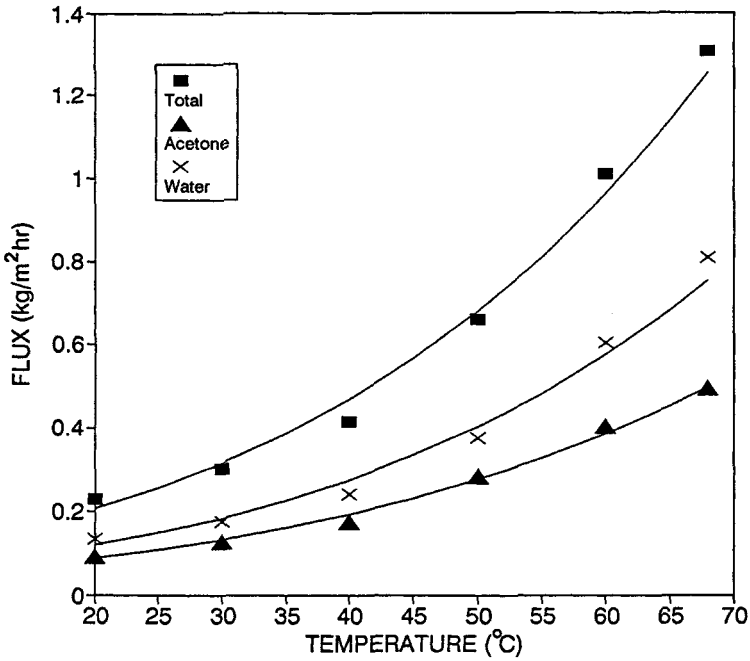

FIG. 6. Flux vs temperature for a 2 wt% acetone feed mixture and 1 torr downstream pressure.

$$J_W = 60160\, e^{-3849/T} \tag{6}$$

where J_T, J_A, and J_W are the total, acetone, and water component fluxes, and T is the absolute temperature in Kelvin. Activation energies of 7.2 and 7.6 kcal/mol were calculated from the acetone and water component flux and temperature data. The selectivity did not change significantly as temperature increased (Fig. 8).

The benchmark run was repeated after completion of the temperature studies, and the flux and selectivity results duplicated those obtained before the temperature studies. Therefore, one can conclude that the membrane is quite stable when operated under varying temperatures and extended processing times.

The effects of permeate-side pressure on membrane transport were also studied using the SC membrane. As the permeate-side pressure increased, the total flux decreased in a sigmoidal fashion (Fig. 9). As the downstream pressure was increased from 1 to 70 torr, the total flux decreased from 0.7 to 0.24 kg/(m²·h) while the acetone component flux decreased from 0.3

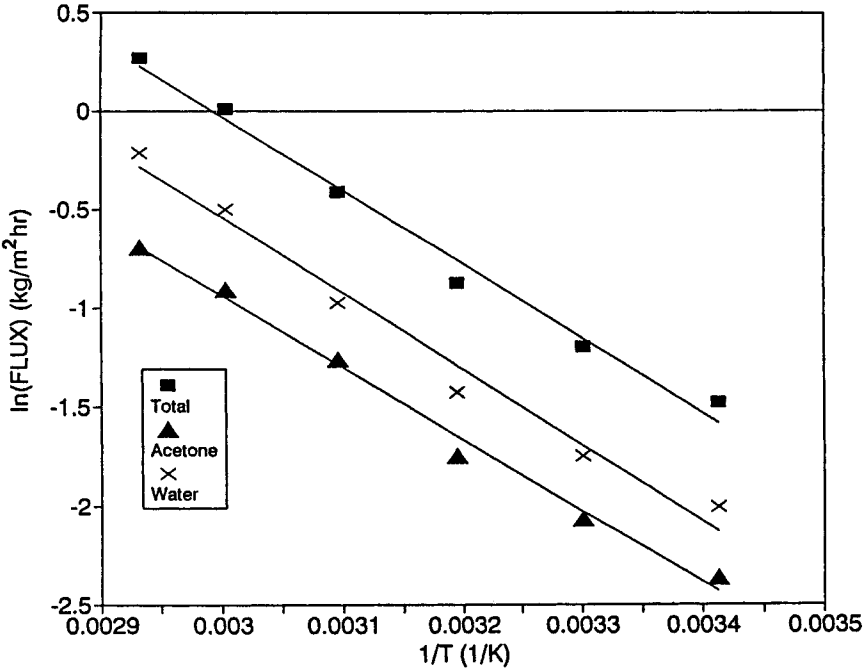

FIG. 7. Ln (flux) vs reciprocal temperature for a 2 wt% acetone feed mixture and 1 torr downstream pressure.

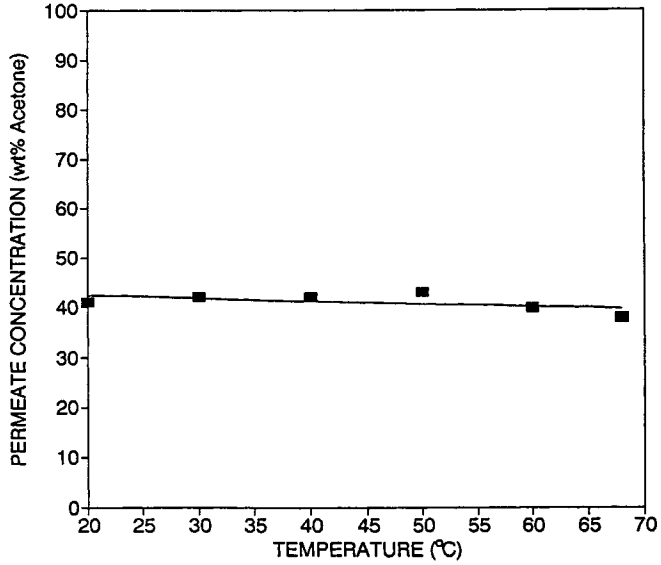

FIG. 8. Permeate concentration vs temperature for a 2 wt% acetone feed mixture and 1 torr downstream pressure.

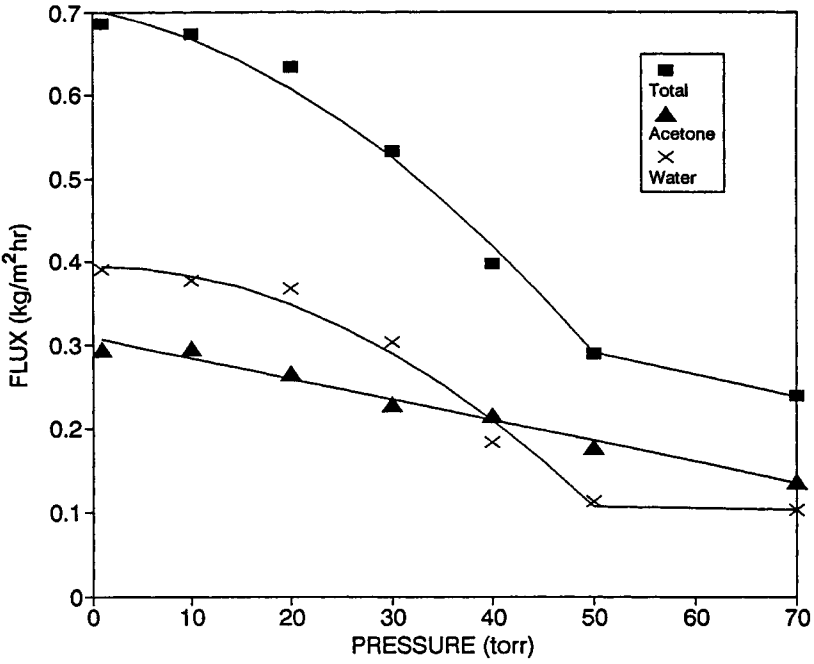

FIG. 9. Flux vs permeate side pressure for a 2 wt% acetone feed mixture at 50°C.

to 0.14 kg/(m²·h) and the water component flux decreased from 0.39 to 0.1 kg/(m²·h). The flux did not change significantly below a permeate-side pressure of 10 torr. The water component flux decreased more quickly than the acetone component flux, causing the acetone selectivity to increase with higher permeate-side pressure (Fig. 10). The selectivity stayed relatively constant below a downstream pressure of 30 torr.

The effect of pressure on the water component flux was modeled using a slight variation to the equations developed by Yoshikawa and coworkers (*30*):

$$J_W = A(P_2 - P_*) + B(P_*^2 - P_3^2) \qquad \text{when } P_3 \leq P_* \tag{7}$$

$$J_W = A(P_2 - P_3) \qquad \text{when } P_3 > P_* \tag{8}$$

where J_W[kg/(m²·h)] is the water component flux, A and B [kg/(m²·h·torr)] are constants, and P_2, P_*, and P_3 [torr] are the upstream pressure, saturation vapor pressure, and downstream pressure, respectively. An inflection point occurs at the saturation vapor pressure, where the flux versus permeate-

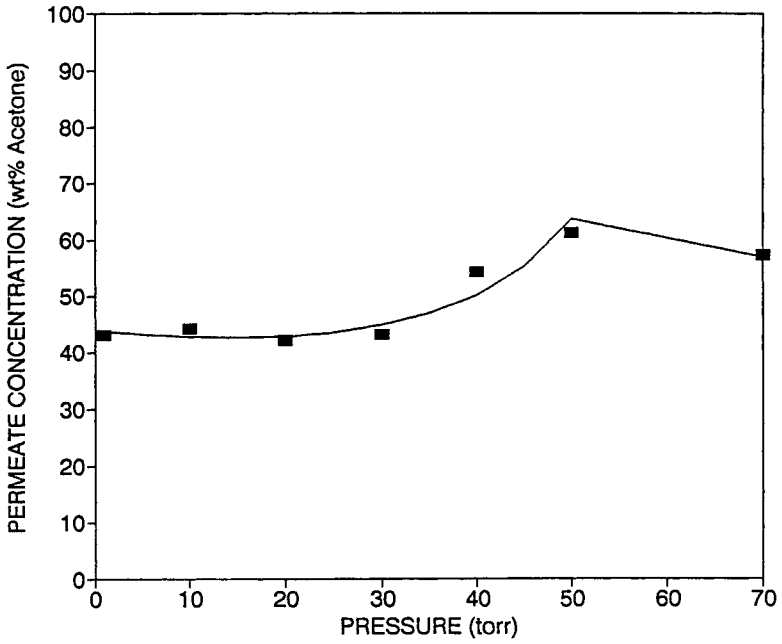

FIG. 10. Permeate concentration vs permeate side pressure for a 2 wt% acetone feed mixture at 50°C.

side pressure data show a sudden change from the parabolic form of the first equation to the linear form of the second equation. The water component data fit this relationship quite well, with A, B, and P_s set to 1.5×10^{-4} kg/(m^2·h·torr), 1.15×10^{-4} kg/(m^2·h·torr), and 50 torr, respectively. The saturation vapor pressures of water at temperatures of 50, 40, and 30°C calculated using the Antoine vapor pressure equation and associated constants were 92.3, 55.3, and 31.9 torr, respectively. Since all pressure studies were performed with a feed temperature of 50°C, the discrepancy in the water-vapor pressure can be explained in two ways. The tubing between the constant temperature bath and the membrane is not insulated; therefore, the temperature in the membrane could be lower than 50°C, causing the vapor pressure of water in the bulk system to be less than 92.3 torr. Second, an interaction between the water and the membrane would cause the vapor pressure of water in the membrane (50 torr) to be different than that in the bulk system.

The acetone component flux and pressure data did not fit the relationship described above. The flux was nearly linear over the entire range of pres-

sures studied:

$$J_A = 0.3092 - 0.0025P_3 \quad (9)$$

At a temperature of 50°C, the saturation vapor pressure of acetone calculated using the Antoine equation is 615 torr. Therefore, it is not surprising that over the pressure range studied (1 to 70 torr), the acetone flux did not drop sharply.

It can be predicted that the flux will continue to fall as the pressure is increased above 70 torr and the limiting pressure is approached. The limiting pressure, approximately 615 torr for this system, is the pressure at which no further transport is possible because the permeate would no longer be a vapor (dew-point pressure of the permeate mixture). Since the acetone has a much greater vapor pressure than water, the limiting pressure is essentially the vapor pressure of acetone. The benchmark run was again duplicated after completion of the pressure studies, and the membrane proved to be stable when operated under varying pressures.

An estimation of system process stream compositions and flow rates for the silicone composite (SC) membrane at operating conditions of 50°C and 1 torr is given in Fig. 11. The estimation is based on the experimental performance of the SC membrane. This block diagram sets the retentate composition to whatever desired levels are needed, and the flow rates are

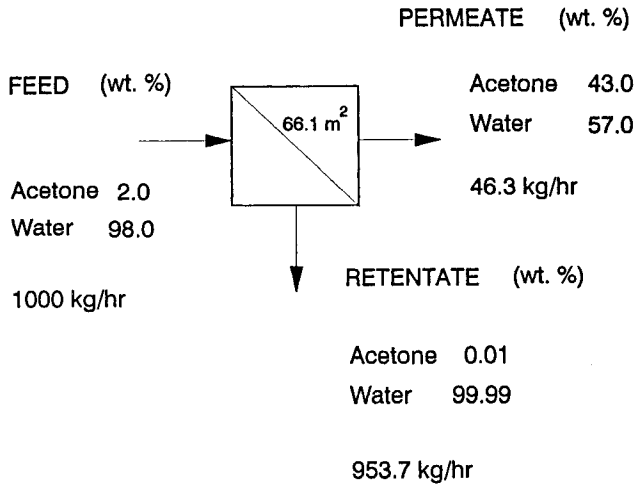

FIG. 11. An estimation of system process stream compositions and flow rates for the SC membrane.

calculated from an overall material balance. For this example a retentate acetone concentration of 0.01% was used with a feed flow rate of 1000 kg/h. This gives an indication of what the process stream compositions would be when translating the membrane laboratory test data on permeability to typical process flows in actual commercial operation.

CONCLUSIONS

Pervaporation is an effective type of membrane separation with great application potential for the biochemical and pharmaceutical industry. Pervaporation coupled with fermentation not only concentrates the fermentation products but also increases fermentation efficiency by reducing product inhibition. In addition, pervaporation minimizes the thermal, chemical, and mechanical stresses often exerted upon the microorganisms by competitive separation processes.

The concentration of dilute acetone solutions has been studied using four different membranes. The silicone composite (SC) membrane obtained from GFT was found to be the most effective in terms of flux and selectivity. This membrane was used to study the effects of process parameters, including feed temperature, permeate-side pressure, and feed concentration, on the pervaporative transport. As the feed temperature increased, the flux increased exponentially while the selectivity remained constant. An increase in permeate-side pressure decreased the flux in a sigmoidal fashion. The optimum production rate occurred at low permeate-side pressure and high feed temperature.

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