



Permeability, permeance and selectivity: A preferred way of reporting pervaporation performance data

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ABSTRACT

A majority of the papers on pervaporation published in the *Journal of Membrane Science* (JMS) report results as raw data; that is, as fluxes (J_i) and separation factors (β_{ij}). These values are not only a function of the intrinsic properties of the membranes used, but also depend on the operating conditions of the experiments (feed concentration, permeate pressure, feed temperature): change the operating conditions and all the numbers change. This makes comparison of pervaporation data sets obtained under different operating conditions difficult. Furthermore, these values can lead to confusion in understanding the real trend of membrane intrinsic properties as operating conditions change. In this paper, the benefits of reporting the results in a normalized form, that is, as permeabilities (P_i), permeances (P_i/ℓ) and selectivities (α_{ij}) are described. Data from our laboratory and from two recent papers are used to illustrate these benefits.

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1. Background

Some time ago, one of us (JGW) wrote a letter to this publication advocating that pervaporation papers report membrane performance data as intrinsic, driving force normalized properties – permeability (P_i), permeance (P_i/ℓ) and selectivity (α_{ij}) – rather than as flux (J_i) and separation factor (β_{ij}) [1]. This proposal has not been widely accepted, and currently, a clear majority of the pervaporation papers appearing in JMS continue to use flux and separation factors. For example, there were 21 pervaporation papers in Volumes 325 through 334 of the JMS; only 5 reported selectivities and permeabilities or permeances. Among the papers that did report permeabilities or permeances, some used molar and others used mass-based terms. The other 16 reported their data as permeation fluxes and separation factors, permeate concentrations, or enrichment factors. This paper is another attempt at illustrating the benefits of driving force normalized properties.

Let's start with a few definitions. Pervaporation data are usually reported as membrane separation factors (β_{ij}) and fluxes (J_i). A separation factor (β_{ij}) is the ratio of the molar component concentrations in the fluids on either side of the membrane. Hence:

$$\beta_{ij} = \frac{c_{i\ell}/c_{j\ell}}{c_{io}/c_{jo}} \quad (1)$$

where subscript o represents the fluid at the feed surface of the membrane (point o) and subscript ℓ represents the fluid at the permeate surface of the membrane (point ℓ). Frequently, membrane separation factors are called membrane selectivities and are given the symbol α_{ij} instead of β_{ij} . This usage leads to confusion. A subgroup of authors report their data in terms of enrichments; that is, permeate concentration/feed concentration, which they also call membrane selectivity.

The membrane flux in pervaporation is usually given as a mass flux (J_i) with typical units $\text{kg}/\text{m}^2 \text{ h}$. In gas separation, flux is normally given as a molar flux (j_i) with units $\text{cm}^3(\text{STP})/\text{cm}^2 \text{ s}$. Ignoring simple numerical conversion terms (m^2 to cm^2 , h to s) this means:

$$J_i = j_i \frac{m_i}{v_i^G} \quad (2)$$

where v_i^G is the molar volume of gas i ($22.4 \text{ l}(\text{STP})/\text{mol}$) and m_i is the molecular weight of component i . The equations shown below use molar fluxes and molar concentration since it simplifies direct comparison of pervaporation data with the large body of gas permeation data.

The problem with reporting data in terms of fluxes and separation factors is that these values are not only a function of the intrinsic properties of the membranes used, but also depend on the operating conditions of the experiments (feed concentration, permeate pressure, feed temperature): change the operating conditions and all the numbers change. Therefore, using flux and separation factors makes comparison of pervaporation data sets obtained under different operating conditions difficult.

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A far more useful way to report pervaporation data is as membrane permeabilities (P_i), permeances (P_i/ℓ) and selectivities (α_{ij}) that are related to the intrinsic properties of the separation membranes. These terms are universally used to describe permeation through gas separation membranes using the standard solution-diffusion Eq. (2):

$$j_i = \frac{D_i K_i^G}{\ell} (p_{i_o} - p_{i_\ell}) = \frac{P_i^G}{\ell} (p_{i_o} - p_{i_\ell}) \quad (3)$$

where ℓ is the membrane thickness and D_i is the membrane diffusion coefficient (cm^2/s) of component i , p_{i_o} and p_{i_ℓ} are the partial pressures of component i on either side of the membrane (surfaces o and ℓ), K_i^G ($\text{cm}^3(\text{STP})/\text{cm}^3 \text{cmHg}$) is the sorption coefficient of component i , linking the partial pressure of the gas (p_i) to the concentration $c_{i(m)}$ in the membrane phase; hence:

$$c_{i(m)} = K_i^G p_i \quad (4)$$

It is implicit in the definitions shown in Eqs. (3) and (4) that concentration polarization effects are negligible. Eq. (3) is easily applied to gas separation experiments, since the partial pressures on the feed and permeate sides of the membrane are readily calculated from the measured total pressures and the feed and permeate concentrations.

The solution-diffusion model [2–4] shows that Eq. (3) also applies to pervaporation separations, where p_{i_o} and p_{i_ℓ} are the vapor pressures of component i on the feed and permeate sides of the membrane, and the permeability term (P_i^G) is the same as the gas permeability of the same membrane. Determining the permeate side vapor pressure in pervaporation is straightforward; the total permeate pressure and the permeate gas composition are all that is required. The problem is to calculate the feed side vapor pressure; conversion of the measured liquid phase composition and temperature into component vapor pressures is required. Today, the availability of computer simulation programs (ChemCad, HYSYS, ProSim) makes calculation of the partial vapor pressure of even complex liquid mixtures a trivial exercise, provided the correct equation of state is selected.

As one of the reviewers of this paper noted, the sentence above gives an apparent solution to the problem of calculating driving force (“computer simulation programs”) and the Achilles’ heel of such a solution (“provided the correct equation of state is selected”). Most computer process simulators do a good job of selecting the appropriate equation of state and the activity coefficients required for simple two-component mixtures. The vapor pressures calculated this way are normally within a few percent of the experimental values. However, obtaining reliable vapor pressures for multi-component mixtures is more difficult and may not be possible if one of the components is an electrolyte. Fortunately, the vast bulk of published data only involve two-component mixtures.

Researchers who do not have access to a process simulator can find experimental vapor pressure data and the predictions of several equations of state for many common binary mixtures in compilations such as the DECHEMA Vapor–Liquid Equilibrium Data Collection [5]. The appropriate equations and suggested numerical coefficients can then be incorporated into an MS Excel program to calculate vapor pressures from known liquid compositions at known temperatures.

One way to avoid calculating vapor pressures, used by some, is to recast Eq. (3) in terms of concentration.

$$j_i = \frac{P_i^G}{\ell} (\gamma_{i_o}^L x_{i_o}^L p_{i_o}^{\text{sat}} - p_{i_\ell}) \quad (5)$$

where $\gamma_{i_o}^L$ is the activity coefficient of component i in the feed liquid and $x_{i_o}^L$ is the mole fraction of the component i in the feed liquid.

$p_{i_o}^{\text{sat}}$ is the pure component i feed vapor pressure. Eq. (5) can then be written as

$$j_i = \frac{P_i^G}{\ell} (H_{i_o} x_{i_o}^L - p_{i_\ell}) \quad (6)$$

where H_{i_o} is the Henry’s Law coefficient and

$$H_{i_o} = \gamma_{i_o}^L p_{i_o}^{\text{sat}} \quad (7)$$

Eq. (6) can then be rewritten as

$$j_i = \frac{P_i^L}{\ell} \frac{x_{i_o}^L - p_{i_\ell}}{H_{i_o}} \quad (8)$$

where P_i^L is a concentration-based permeability with units $\text{mols}/\text{cm s}$.

At low permeate pressures, the term p_{i_ℓ}/H_{i_o} can often be neglected and so Eq. (8) becomes

$$j_i = \frac{P_i^L}{\ell} x_{i_o}^L \quad (9)$$

It follows from Eqs. (6) and (8) that

$$P_i^L = P_i^G H_{i_o} \quad (10)$$

This approach, leading to Eqs. (8)–(10), is a poor second best. The dependence of vapor pressure (driving force) on temperature and activity coefficients is now hidden in the terms P_i^L (definitely not constant). P_i^G , the gas phase permeability coefficient in Eq. (3), is only a modest function of temperature and concentration, but the liquid phase coefficient P_i^L is often a strong function of feed solution concentration and usually increases exponentially with temperature as the vapor pressure term now imbedded in P_i^L increases. The approach leading to Eqs. (8)–(10) should therefore be avoided.

The two main pervaporation membrane characterization terms we propose are therefore:

- Membrane permeability (P_i^G), a component flux normalized for membrane thickness and driving force and, as noted above in Eq. (3), given as

$$P_i^G = D_i K_i^G = j_i \frac{\ell}{p_{i_o} - p_{i_\ell}} \quad (11)$$

Permeability (P_i^G) is most commonly reported as Barrers ($1 \text{ Barrer} \equiv 1 \times 10^{-10} \text{ cm}^3(\text{STP})/\text{cm}^2 \text{ s cmHg}$). A variety of IUPAC-type units are also occasionally used – for example, $\text{m}^3(\text{STP})/\text{m}^2 \text{ s kPa}$ ($1 \text{ m}^3/\text{m}^2 \text{ s kPa} = 1.33 \times 10^{14} \text{ Barrer}$). Alternatively, when the membrane thickness is not known, membrane permeance (P_i^G/ℓ), a component flux normalized for driving force can be used; it is given as

$$\frac{P_i^G}{\ell} = \frac{D_i K_i^G}{\ell} = \frac{j_i}{p_{i_o} - p_{i_\ell}} \quad (12)$$

Permeance (P_i^G/ℓ) is most commonly reported as gas permeation units (gpu) ($1 \text{ gpu} \equiv 1 \times 10^{-6} \text{ cm}^3(\text{STP})/\text{cm}^2 \text{ s cmHg}$, ($1 \text{ m}^3/\text{m}^2 \text{ s kPa} = 1.33 \times 10^8 \text{ gpu}$).

- Membrane selectivity, defined as the ratio of the permeabilities or permeances of components i and j through the membrane:

$$\alpha_{ij} = \frac{P_i^G}{P_j^G} = \frac{P_i^G/\ell}{P_j^G/\ell} \quad (13)$$

In these definitions, the membrane flux is given in molar terms to allow direct comparison of pervaporation data with the larger body of gas permeation data (there are other advantages of using molar rather than mass coefficients, as discussed below).

We will show that using permeabilities, permeances and selectivities decouples the variation in flux and separation factor due

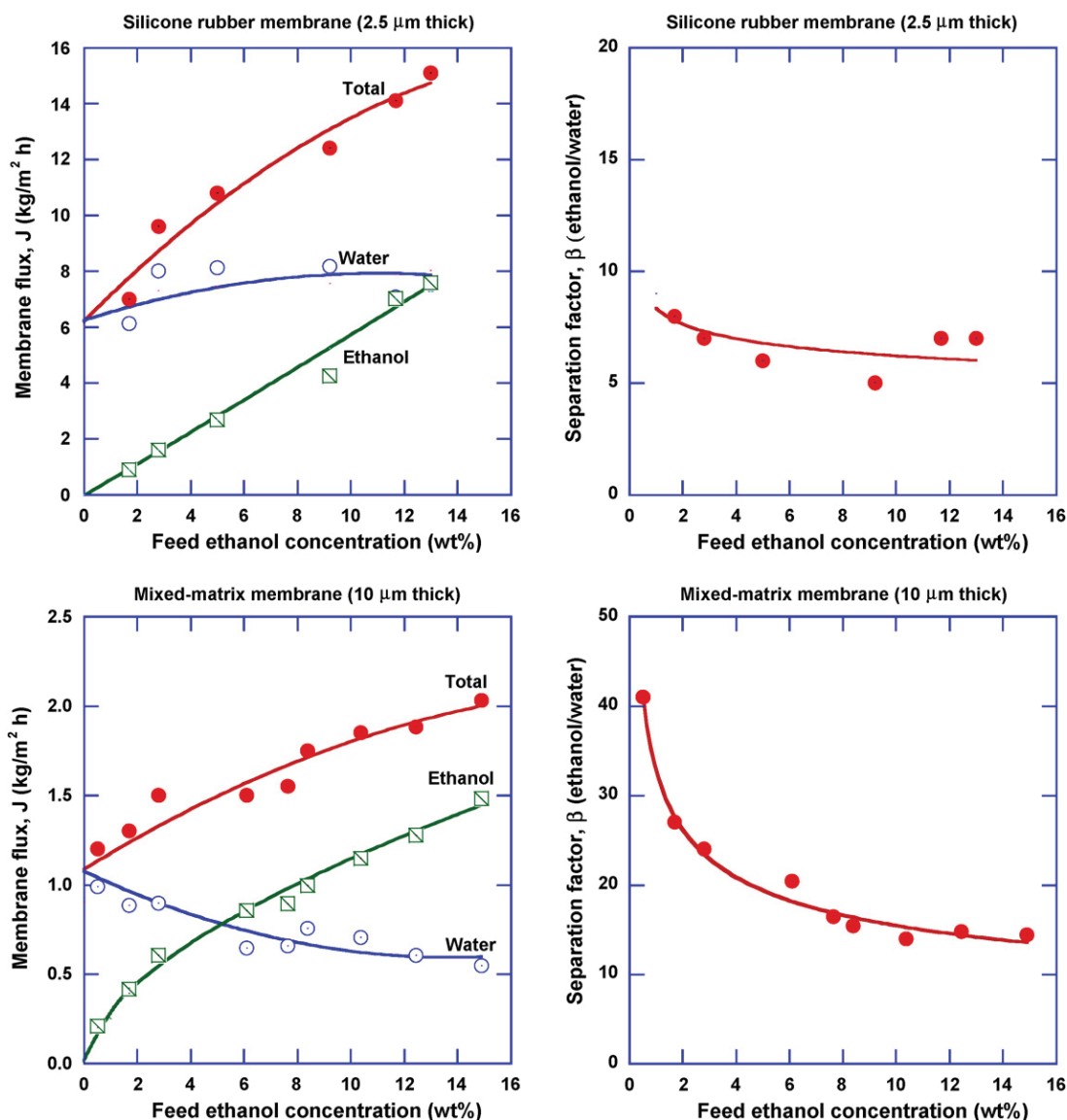


Fig. 1. Flux and ethanol/water separation factor data obtained with 2.5 μm thick pure silicone rubber membrane and 10 μm zeolite (60 wt%)-silicone rubber mixed-matrix membrane (flow through test cells, 75 °C, permeate pressure <5 Torr, stage-cut per cell of less than 0.2%).

to vapor-liquid equilibrium effects from true variations in the permeation properties of the membrane. The driving force normalized properties may not be completely constant and can change with feed concentrations and permeate pressures, but not nearly as much and not necessarily in the same direction as membrane flux and separation factors that are commonly reported.

2. Some sample calculations and discussion

2.1. Silicone rubber and silicone rubber-ZSM-5 zeolite mixed-matrix membranes

In this section, we will illustrate the benefits of using intrinsic membrane permeation coefficients rather than flux and separation factors. By way of example, some pervaporation data measured in our laboratory for the separation of ethanol from dilute aqueous solutions are shown in Fig. 1. Two types of membrane were used: a simple 2.5 μm thick supported silicone rubber membrane and a 10-μm thick mixed-matrix membrane consisting of 60 wt% ZSM-5 zeolite dispersed in a silicone rubber matrix made using the procedures described by Vane et al. [6] and teHennepe et al. [7].

Fig. 1 presents the permeation data as mass fluxes and separation factors. Little can be gleaned from these plots. The zeolite mixed-matrix membranes are more selective than simple silicone rubber, as shown by the higher separation factors. On the other hand, membrane fluxes are lower, although this may be partially a membrane thickness effect.

The data from Fig. 1 are re-plotted in terms of membrane permeabilities and selectivities in Figs. 2 and 3.

This representation of the data is now far more enlightening:

- By normalizing for driving force, it becomes clear that the concentration dependence of flux in Fig. 1 is mostly due to changes in ethanol and water vapor pressure (driving force) with concentration. When this driving force effect is removed, the water and ethanol permeabilities of the pure silicone membrane are almost constant. The permeabilities of the mixed-matrix membranes are concentration dependent, even when the driving force contribution is removed. In particular, the permeability of ethanol is high at low ethanol concentration, but decreases to a plateau value at concentration above about 10% ethanol.

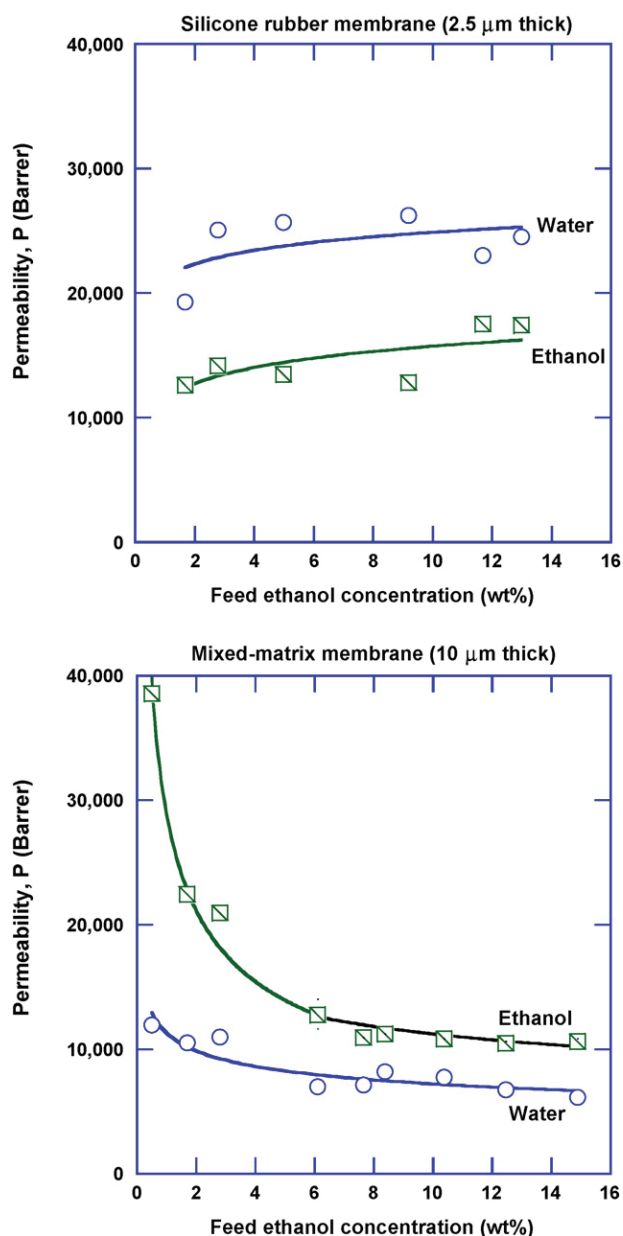


Fig. 2. The flux data of Fig. 1 reported as permeabilities (1 Barrer $\equiv 1 \times 10^{-10}$ cm³(STP) cm/cm² s cmHg) (equation of state NRTL).

- Pure silicone rubber membranes are water selective with a water permeability of about 25,000 Barrer and an ethanol permeability of 15,000 Barrer. These permeabilities are in line with the gas permeability data of similar silicone rubber membranes [8]. It follows that silicone rubber membranes are actually water selective. The apparent ethanol selectivity of the membrane shown by the separation factor curve in Fig. 1 is entirely a result of the higher volatility of ethanol compared to water. When the driving force effect is removed, ethanol has a lower permeability than water. Put another way, the separation achieved by a silicone rubber membrane for an ethanol–water mixture is actually lower than simple evaporation would produce.
- In contrast to the silicone rubber membrane, the mixed-matrix membrane is more permeable to ethanol than water. This membrane is therefore ethanol selective.
- Finally, the improved selectivity achieved by adding zeolite particles to a silicone rubber matrix is produced by a combination of increasing the permeability to ethanol and decreasing the per-

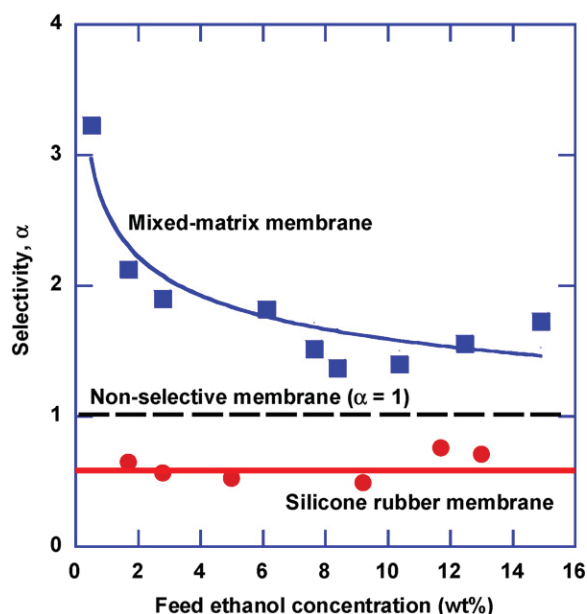


Fig. 3. Ethanol/water separation factor data of Fig. 1 reported as ethanol/water selectivities.

meability to water. It appears that ethanol passes through the zeolite particles faster than through the silicone polymer matrix and hence the ethanol permeabilities are enhanced. Water cannot permeate through the zeolite particles and so these particles increase the tortuous path taken by the permeating water. The result is a decrease in water permeability.

The reduced permeability and selectivity of ethanol in the mixed-matrix membranes at high concentrations may be due to a crowding (saturation) effect of ethanol diffusion through the zeolite pores.

A line marked $\alpha = 1$ is also shown in Fig. 3. This line represents the performance of a membrane with no intrinsic membrane selectivity. Such a membrane would achieve the same separation as simple evaporation of a portion of the liquid into the vapor phase. This membrane would have a separation factor of ~ 8 – 14 at a feed liquid ethanol concentration of 0–15 wt%, but the separation obtained with such a non-selective membrane (for example, a microporous film) is entirely due to the different relative volatilities of ethanol and water as they evaporate from the liquid. The membrane itself makes no contribution to the separation.

As a final aside, some JMS pervaporation papers cited earlier report pervaporation permeabilities and permeances in mass terms. It follows from Eq. (2) that selectivities calculated this way will be different from the mole-based values in Eq. (13) by a term:

$$(\alpha_{ij})_{\text{mass}} = (\alpha_{ij})_{\text{molar}} \frac{m_i}{m_j} \quad (14)$$

For the case of ethanol (i) from water (j), this means that a non-selective membrane has a molar selectivity of 1.0, but a mass-based selectivity of 2.6 ($1 \times 46/18$). For an isopropyl alcohol–water mixture, a non-selective membrane again has a molar selectivity of 1.0, but the mass-based selectivity is now 3.3 ($1 \times 60/18$). This type of odd result is another reason for preferring molar-based permeabilities and selectivities.

2.2. Ethanol dehydration membranes

In the data set shown in Figs. 1–3, separating the contributions of driving force and intrinsic membrane properties revealed that

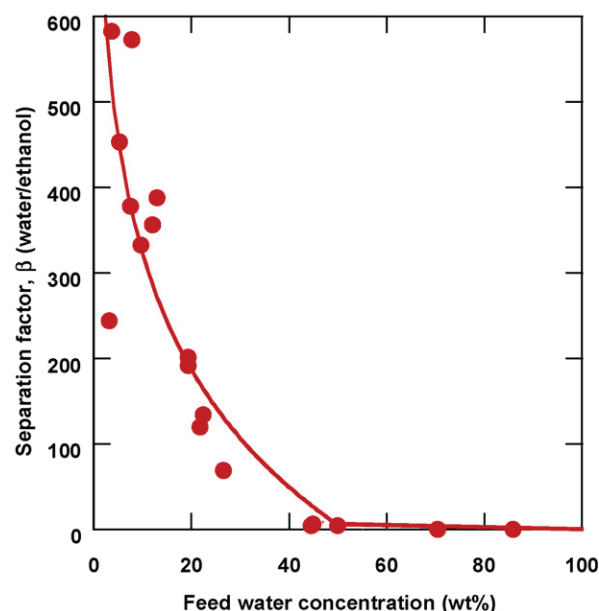
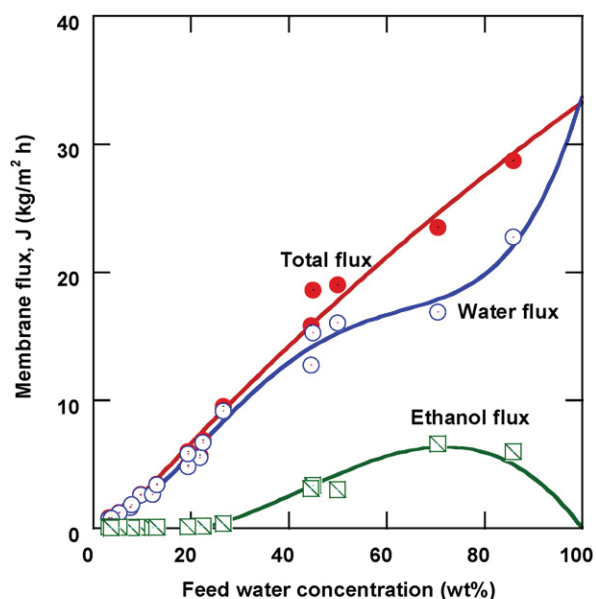


Fig. 4. Dehydration of ethanol/water solutions with a commercial cellulose ester membrane (75°C, permeate pressure <5 Torr), reported as membrane flux and water/ethanol separation factor data.

operating conditions changed the driving force but not the membrane. This behavior is not always found. Figs. 4 and 5 show results also measured with ethanol/water mixtures, but this time with a cellulose ester composite membrane that is very selective for water but whose permeability changes with water concentration.

Fig. 4 shows a plot of membrane flux and separation factor as a function of feed water concentration. At a low water concentration in the feed, the membrane is very selective and the separation factor is high, but the flux of water and ethanol through the membrane is low. As the water concentration in the feed is increased, both the ethanol and water fluxes increase and the membrane separation factor falls. At a water concentration of about 70 wt%, the ethanol flux reaches a maximum value. At higher water concentrations, the ethanol flux declines, going to zero ethanol flux at zero ethanol concentration in the feed. The water flux continues to increase sharply, reaching about 34 kg/m² h at 100 wt% water. Something is going on, probably related to plasticization of the membrane by water, but

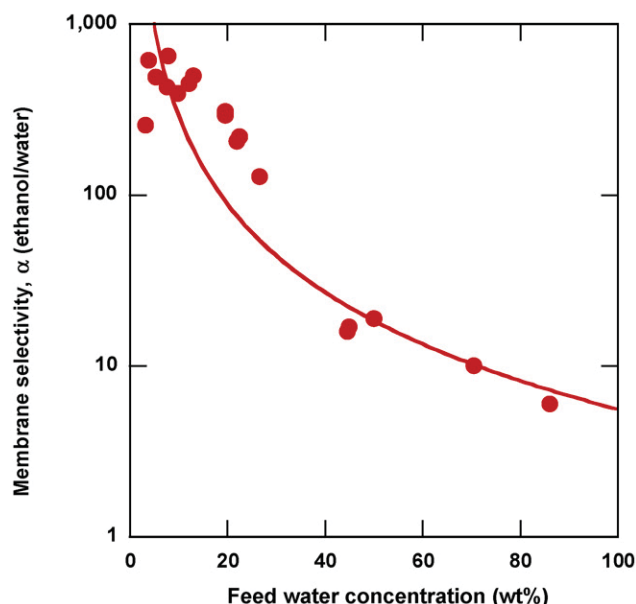
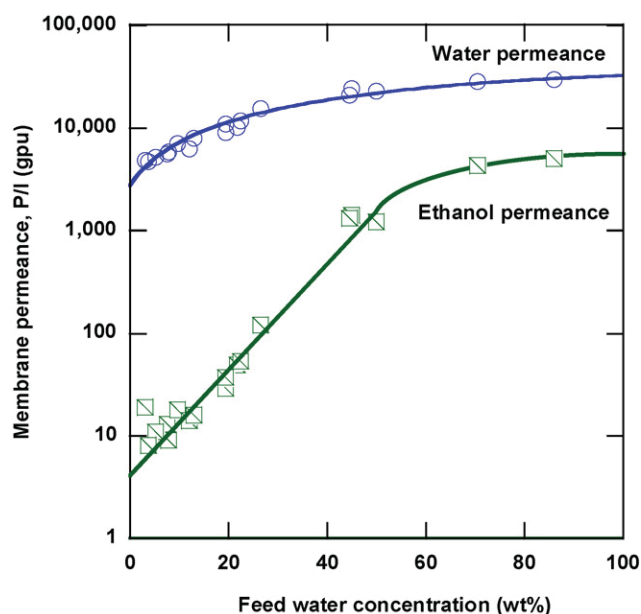


Fig. 5. The permeation data of Fig. 4 re-plotted as permeances ($1 \text{ gpu} \equiv 1 \times 10^{-6} \text{ cm}^3(\text{STP})/\text{cm}^2 \text{ s cmHg}$) and membrane water/ethanol selectivities (equation of state NRTL).

it is hard to sort this out from the membrane flux and separation factor data as presented in Fig. 4.

The same data are re-plotted as permeances and selectivities versus feed concentration in Fig. 5. The situation is now clearer. At low water concentrations in the feed, the membrane is extremely selective; for example, at a water concentration of less than 5 wt%, the membrane has a water permeance of about 5000 gpu and the ethanol permeance is below 10 gpu. As the water concentration increases, the ethanol and water permeances both increase. This increase is probably due to swelling of the membrane by water, leading to plasticization. Plasticization increases the water permeance from 5000 gpu at less than 5 wt% water to 32,000 gpu at 100 wt% water, but has a much larger effect on ethanol permeance. The corresponding ethanol permeance increases from less than 10 gpu at less than 5 wt% water to more than 6000 gpu at 100 wt% water, an almost 600-fold increase. As a consequence,

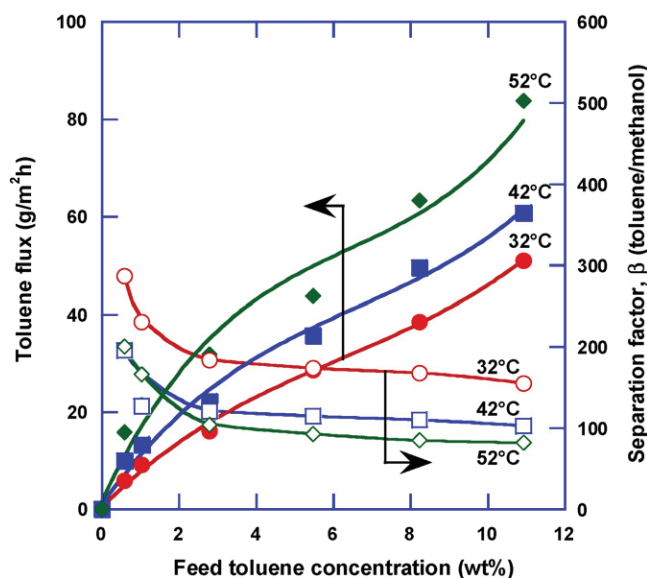


Fig. 6. The separation of toluene from toluene/methanol mixtures using crosslinked natural rubber membranes. Data of Ray and Ray [9], presented as flux and separation factor data. Permeate pressure = 1 mmHg; feed temperatures range from 32 to 52 °C, as shown.

the water/ethanol selectivity falls to a selectivity of only 5 at very high water concentrations. Data obtained with similar hydrophilic membranes have shown that some membranes even become ethanol selective at very high water concentrations.

The data in Fig. 5 also highlight the very high permeances of these water permeable pervaporation membranes. In the concentration range where these membranes are usually used, 0–20 wt% water, the membrane has a water permeance of 5000 to 10,000 gpu. This permeance is almost 1000 times higher than the oxygen permeance of membranes used to separate oxygen from air, and about 50 times the CO₂ permeance of membranes used to separate CO₂ from natural gas.

2.3. The effect of operating conditions

Some data illustrating the effect of temperature and feed concentration on pervaporation membrane performance are shown in the form originally reported by Ray et al. in Fig. 6 [9]. This data set shows pervaporation of toluene/methanol mixtures through a crosslinked natural rubber membrane. The membrane appears to be quite selective, with separation factors ranging from 50 to 200, depending on the feed toluene concentration and temperature. The membrane flux increases almost linearly with concentration. As the feed temperature increases, fluxes increase, but the separation factor falls. Removal of small amounts of toluene from methanol is not an important chemical separation commercially, but for an engineer faced with this separation problem, these results would be encouraging. The toluene/methanol mixture has an azeotrope at about 30 wt% toluene. This means the vapor–liquid equilibrium curves from 0 to 30% toluene are very close, so separation by distillation is not easy. Ray's data, on the other hand, show pervaporation can produce better than 95% toluene in the permeate in a single separation step.

The same data set is transformed into membrane permeance and selectivity data in Fig. 7. A number of additional conclusions can now be drawn. First, the increase in flux with increasing temperature is completely due to the increase in vapor pressure driving force. As the permeance curves in Fig. 7 show, the intrinsic permeance actually decreases slightly as temperature increases. Diffusivity almost always increases with increasing temperature,

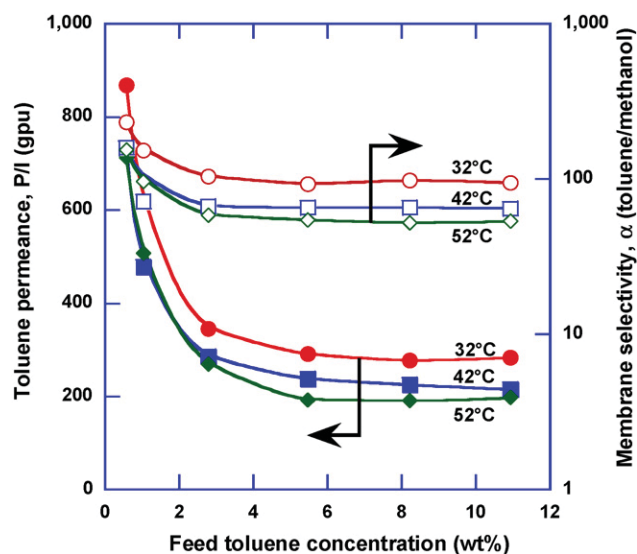


Fig. 7. The permeation data of Ray and Ray [9] shown in Fig. 6, re-plotted as toluene permeances (1 gpu $\equiv 1 \times 10^{-6}$ cm³(STP)/cm² s cmHg) and toluene/methanol selectivities (equation of state NRTL).

while sorption usually decreases. This indicates that in these membranes, the change in sorption outweighs the change in diffusivity. On the other hand, the decrease in membrane selectivity as the temperature increases is in line with expectations. But even at 52 °C, the membrane still has a toluene/methanol selectivity of 50. Since toluene is a substantially larger molecule than methanol, the high toluene/methanol selectivity must be due in large part to the high preferential sorption of the hydrophobic toluene compared to the hydrophilic methanol.

Applying the solution-diffusion model and Eq. (3) to this data also shows the importance of permeate pressure in these experiments. In the Ray paper, the permeate pressure is described as “(an) adjustable downstream pressure that was maintained at 1 mmHg.” Usually, a permeate pressure this low means that the permeate vapor pressure term ($p_{i,p}$) in Eq. (3) can be ignored, but in these experiments, the feed toluene concentrations are low and the feed temperature is also as low as 32 °C. Under these conditions, the toluene vapor pressure of the feed is only 0.8–1.5 mmHg. Since the membrane is very selective, the permeate toluene concentration ranges from 60 to 95 mol%. This means the toluene vapor pressure on the permeate side of the membrane is 0.6–0.95 mmHg, which is 50–80% of the feed vapor pressure. The vapor pressure of the permeate has to be taken into account in Eq. (3), and very precise control of the permeate pressure is required to obtain reliable data. This effect is most noticeable at low toluene concentrations and may be a partial explanation for the curious two-to-threefold increase in permeance and selectivity at low toluene concentrations. It is hard to come up with another logical reason for this result.

Clearly, these insights can only be elucidated when the pervaporation results are presented in the normalized form.

2.4. Comparing different membranes

As a final example, we introduce Table 1, taken from a paper by Svang-Ariyaskul et al. [10]. The authors were trying to compare the properties of several pervaporation membranes used to dehydrate isopropyl alcohol (IPA). The table illustrates the difficulty of comparing data from different sources. The temperature of the feed solutions varied from 30 to 80 °C, equivalent to a tenfold difference in vapor pressure driving force. The water concentrations varied

Table 1

Comparison of pervaporation performance for the dehydration of isopropanol. In all cases, the permeate pressure is assumed to be a hard vacuum [10].

Membrane	% IPA in the feed	Temperature (°C)	Flux (g/m ² h)	% Water in permeate	Permeance (P/ℓ) _{H₂O}	Selectivity $\alpha_{H_2O/IPA}$	References
1. Crosslinked CS ^a on PSf	90	80	2733	82.0	3600	37	Lee et al. [11]
2. Silica ^{a,b}	92	70	1900	85.6	4400	57	Gallego-Lizon et al. [12]
3. Crosslinked PVA ^{a,b}	91	70	1350	95.7	3300	192	Gallego-Lizon et al. [12]
4. CSPVA 7525	90	60	644	>99.999	2400	>10,000	Svang-Ariyaskul et al. [10]
5. Two-ply CS and sodium alginate	90	60	554	99.6	2000	2000	Moon et al. [13]
6. Crosslinked CS	90	30	400	97.5	6500	290	Nawawi et al. [14]
7. PANi doped with PAA	85	60	180	90.0	490	55	Lee et al. [11]
8. CS/HEC	90	60	175	>99.9	640	6400	Jiraratananon et al. [15]
9. Asymmetric PVA	90	70	100	98.4	240	500	Yu et al. [16]

CS: chitosan; PVA: polyvinyl alcohol; PSf: polysulfones; HEC: hydroxyethylcellulose; PANi: polyaniline; PAA: polyacrylic acid. Note that numbers appearing in Table 1 are not the exact numbers from others' studies; they are approximations.

^a Composite membranes.

^b Commercially available membranes from Sulzer Chemtech, Germany.

from 8 to 15%, an additional 1.5-fold difference in vapor pressure driving force. This large range of operating conditions makes a comparison of the data based on flux and separation factor data difficult. The problem is compounded by the range of membrane thickness – from thin composite membranes to 100 μ m films.

The first four columns of Table 1 show the data as originally reported. The last two columns show the membrane permeances and selectivities calculated from the reported numbers and Eqs. (11) and (12). The first three membranes listed are composite membranes probably having selective layer thicknesses in the 2–5 μ m range. The last six membranes in Table 1 are membrane films with thickness on the order of 50–100 μ m.

Based solely on the raw data of membrane flux reported by the authors, one would think the first three composite membranes are more water permeable than the other six membrane films, since their fluxes are larger. However, this is probably only a thickness effect. Now let us check the water permeance data shown in column 6 of Table 1. The first three composite membranes have water permeance between 3000 to 5000 gpu. Membrane selectivities range from 37 to 192 – on the low end for commercial membranes used for this separation. But in the permeance comparison, the chitosan (CS) membrane (#6) is worth noting. This relatively thick membrane actually has a higher permeance than the composite membranes (#1–3), even though the flux values for these membranes are four-to-sevenfold higher. This is because the chitosan data were obtained at 30 °C, while the composite membrane data were obtained at 70–80 °C. When the difference in vapor pressure driving force is taken into account, the chitosan membrane has the higher permeance. An operating temperature of 30 °C is a long way from the normal 90–110 °C operating range of industrial plants for this separation. Nonetheless, the chitosan membrane is probably worth another look, especially if made as a thinner composite membrane. Again, this insightful information only comes to light when normalized permeance is reported.

3. Conclusion

Membrane gas separation data are routinely reported as membrane permeabilities, permeances and selectivities. This procedure allows data from divergent sources to be easily compared under widely varying experimental conditions. The study of pervaporation got off on the wrong foot, and most pervaporation data are

reported only as raw data; that is, fluxes and separation factors. These terms are not normalized for driving force and so change when the conditions of the experiment change. This usage significantly reduces the value of the data and is an unnecessary barrier to development and information exchange in the field.

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