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페놀수지 생산공정에서 배출되는 반응성 폐수처리를 위한 중공사막 모듈 투과증발 공정모사

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Simulation of Pervaporation Process Through Hollow Fiber Module for Treatment of Reactive Waste Stream from a Phenolic Resin Manufacturing Process

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요 약: 반응성 페놀수지 폐액을 처리하기 위해 중공사막 모듈을 이용한 투과증발 막 탈수공정을 연구하였다. 이 공정의 거동을 예측하기 위한 모사모델을 확립하였고 여기에 사용되는 중요 기본 파라메타들을 평판형 막을 사용하여 직접 구하여 사용함으로써 공정모사의 정확성을 얻을 수가 있었다. 이들을 모사치와 중공사 투과증발 막으로 부터 직접 측정한 각 투과특성들을 비교한 결과 서로 잘 일치함을 보여 본 모사모델의 타당성을 입증하였다. 사용된 중공사막은 중공사 안쪽에 활성층이 도포되어 있으며 공급액은 중공사 내부로 공급하였다. 공급액의 막내에서의 흐름속도에 따라 온도분포가 결정되며 이에 따라 막 투과특성이 달라짐을 모사결과로부터 얻을 수가 있었다. 공급액 온도증가는 막을 통한 탈수 투과속도를 증가시킬 뿐 아니라 반응속도 증가로 인하여 물 생성속도도 증가시킴으로써 공급액 저장조 내의 수분 함량은 이들 상반된 공정들에 의해 결정이 됨을 보였다. 투과압력이 공급액 증기압보다 훨씬 작은 범위에서 증가할 경우 투과추진력인 공급액과 투과부의 투과물 활성도비 감소가 크지 않아 투과특성을 약간 저하시킨다. 그러나 투과압력이 공급액의 증기압에 접근할 경우 활성도비 감소가 현저하게 일어나 투과특성저하가 급격히 일어난다.

Abstract: For the treatment of reactive phenolic resin waste, a simulation model of pervaporative dehydration process has been developed through hollow fiber membrane module. Some of basic parameters were determined directly from dehydration of the waste liquid through a flat sheet membrane to get realistic values. The simulation model was verified by comparing the simulated values with experimental data obtained from hollow fiber membrane module. Hollow fiber membranes with active layer coated on inside fiber were used, and feed flew through inside hollow fiber. Feed flow rate affected membrane performances and reaction by providing a corresponding temperature distribution of feed along with fiber length. Feed temperature is also a crucial factor to determine dehydration and reaction behavior by two competing ways; increasing temperature increases permeation rate as well as water formation rate. Once the permeate pressure is well below the saturated vapor pressure of feed, permeate pressure had a slightly negative effect on permeation performance by slightly reducing driving force. As the pressure approached the vapor pressure of feed, dehydration performances declined considerably due to the activity ratio of feed and permeate.

Keywords: *pervaporation, simulation model, hollow fiber module, dehydration, phenolic resin*

1. Introduction

Some of waste streams discharged from a phenolic resin manufacturing process contain 20-26% phenol, 60-66% water and the rest part being balanced with several other chemicals. The waste streams are somewhat strong acidic solutions, of which pH is near 2. Because of high organic content in the waste, there is no conventional technology but burning-out of the solution. In this case, valuable chemicals cannot only be lost but

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also high disposal cost is required. Currently manufacturing companies are using these wastes to make a low value product to avoid the disposal cost of the wastes and barely achieve a breakeven. However, there is a potential to convert this stream into profitable by-product if the amount of water in the mixture is reduced to about 10-20%, for achieving the following two objectives; 1) Recovery of some of the valuable organic chemicals and 2) Safe discharge of treated wastewater.

A hassle to recover the organic chemicals is that the waste solutions have some reactivity even at room temperature, producing water. Thus, the recovery should be made at low temperature to depress the reaction. From this point of view, pervaporation process can be one of suitable processes for attaining the objectives. Actually, it is very advantageous of pervaporation that separation and permeation can be carried out effectively at low operating temperature. Previous work done as a preliminary step reveals that concentration of organic chemicals through dehydration is more efficient rather than separation through the permeation of organic components in the application because of much larger flux and greater separation efficiency of dehydration process.

A composite type of membrane is commonly used in industrial application, which is composed of an active layer on a support layer. In order to be applicable to the acidic solution described above, the active coating layer should be sustainable and also the support substrate is dimensional stable against the acidic circumstance. Two types of membrane modules, that is, plate-and frame module and hollow fiber module are available. Each one has its own advantages and disadvantages. The advantages of hollow fiber module over plate-and-frame are low module cost, flexibility in system design and easy installation. However, a serious shortcoming of hollow fiber membrane is a lack of robustness because of no reinforcement such as a non-woven fabric in membrane structure. Therefore, the range of operating conditions must be limited to low feed flow rate and low temperature. Slow feed flow can cause significant concentration and temperature profiles with location in membrane module, which affects negatively membrane performance. In hollow fiber module, each hollow fiber wall can divide the feed space into small compartments as many as fibers used in the module. Thus, feeding inside hollow fiber would provide uniform feed flow with a proper velocity each fiber with even at slow pumping, minimizing non-even flow occurring across the module and getting rid of any channeling of feed which would be possibly obtained in feeding outside hollow fiber. For the feeding inside hollow fiber, an active layer has to sit on inside surface of hollow fiber to permeate and separate the feed.

It was confirmed in the preliminary work that the dehydrated liquid is reactive even at low temperature. To better manage this issue it is important to have some understanding of the reaction kinetics. Especially the dependency of the reaction on composition and temperature, which is associated with a change in reaction rate with increased level of dehydration. It is inevitable to find the best operating condition at which the reaction is minimized, the membrane performance is maximized and the product quality is maintained. For design of the pervaporation unit and process, it is also necessary to pick the optimum operating conditions that will provide a fine balance between membrane performance in terms of flux and separation efficiency and the service life [1]. The optimization of the process and finding of the best design factors can be achievable easily through the simulation of model, which can describe effectively the permeation and reaction behaviors in the dehydration. In this study, a simulation model of the pervaporation dehydration of the reactive waste stream was established based on the feeding inside hollow fiber in hollow fiber membrane module. A minimum number of pervaporations with flat sheet membrane were performed to get basic permeation data for the simulation model. To verify the established model, a comparison of the calculated values was made with experimental data that were obtained from hollow fiber membrane module. Here, hollow fiber

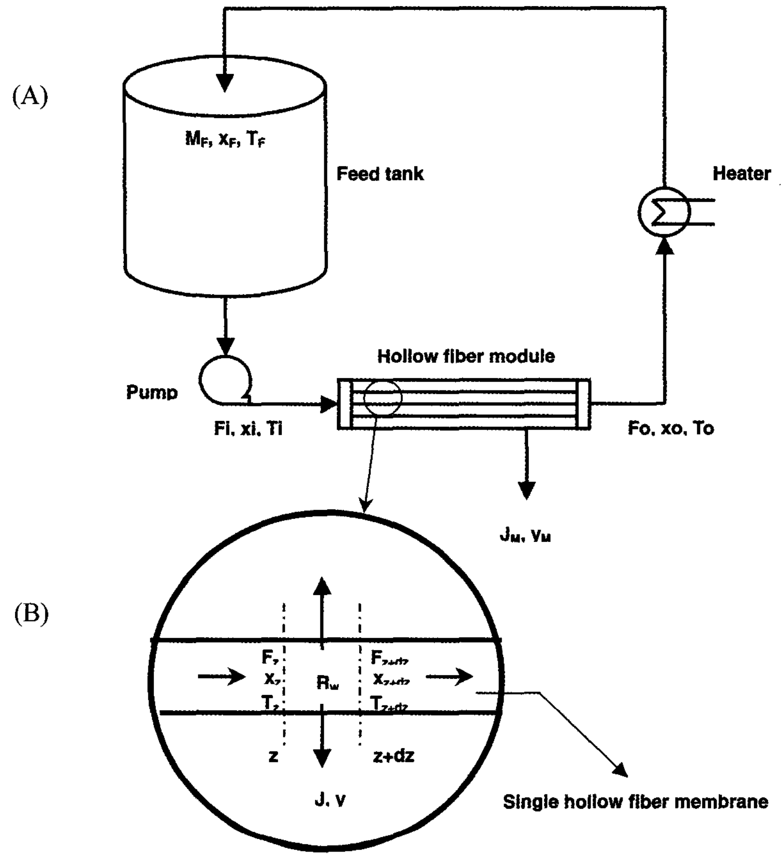

Fig. 1. (A) Schematical representation of pervaporation process with hollow fiber module and (B) principal balances across as differential element of volume in a hollow fiber.

membranes with an active layer on inside surface were used. Permeation behaviors at various operating conditions were discussed through analysis on the simulation results of the process.

2. Theory

Feed mixture circulates from the feed tank through the membrane module for a given period of time as described in Figure 1. During the circulation of the feed through the membrane module, selective permeation takes place through membranes in the module so that the feed amount in the tank, M_F, can not only decrease as much as the amount permeating through the membranes but also water content in the tank, x_F, decrease with time due to water-selective permeation through the membranes. Thus total mass change in the feed tank is balanced with the permeation amount through the membrane module for a differential time interval, dt, as follows;

$$dM_F = -J_M A_T dt \tag{1}$$

where J_M is an average flux through total membrane area, A_T, in the membrane module. The feed mixture has a reactivity to produce water. Thus, the amount of water in the feed tank that is a preferentially permeating component can be affected by competing effects; water permeation through the membranes and water formation by chemical reaction;

A change of *water mass in tank*
= *water formation rate* − *water permeation rate*

$$d(M_F x_F) = R_w M_F dt - J_M y_M A_T dt \tag{2}$$

where R_w is a water formation rate per unit feed mass and y_M is a water content in permeate. Eq. (2) can be rewritten;

$$x_F dM_F + M_F dx_F = R_w M_F dt - J_M y_M A_T dt \tag{3}$$

Combining Eqs (1) and (3) gives

$$dx_F = \frac{R_w M_F + (x_F - y_M) J_M A_T}{M_F} dt \tag{4}$$

M_F, a feed amount in the tank at time t can be given by

$$M_F = (M_F)_0 - \int_0^t (A_T J_M) dt \tag{5}$$

$(M_F)_0$ is a initial feed amount and the integration term expresses a accumulated permeate amount for a permeating time t.

In order to calculate feed amount at a time t, membrane performance over the hollow fiber membrane module should be characterized with permeating time and location along with hollow fiber length. The permeation through single hollow fiber can be considered as continuous process described in Figure 1(B). Figure 1(B) presents schematically feed flow through a differential volume (dz) inside hollow fiber. When a driving force, activity gradient is created across membrane thickness, selective permeation takes place and then feed flow rate as well as feed composition changes through the differential volume. The feed temperature falls constantly because the heat of the evaporation of permeate is supplied from the feed side, resulting in decreasing flux[1,2]. Thus, three different balances over the differential volume are taken into account as follows;

$$dF = -JD_i m\pi dz \tag{6}$$

$$\begin{aligned} d(Fx) &= R_w m(D_i^2\pi/4)\rho dz - JymD_i\pi dz \\ &= (R_w \rho D_i/4 - Jy) mD_i\pi dz \end{aligned} \tag{7}$$

$$d(Fh_F) = -J\Delta h_v mD_i\pi dz \tag{8}$$

where F denotes feed flow rate at z, D_i inside diameter of hollow fiber, J total flux, m number of hollow fiber in the module, x the concentration of a selectively permeating component in feed, y the concentration of the component in permeate, h_F the enthalpy of feed flow, and the heat of the evaporation of permeate. Eqs (7) and (8) can be rewritten as follows, respectively,

$$xdF + Fdx = -(R_w \rho D_i/4 - Jy) mD_i\pi dz \tag{9}$$

$$FC_p dT = -J\Delta h_v mD_i\pi dz \tag{10}$$

where C_p is the heat capacity of feed liquid. From Eqs (6) and (9), the following equation can be obtained.

$$dx = \left[\frac{R_w \rho D_i}{4} + (x - y) J \right] \frac{mD_i\pi}{F} dz \tag{11}$$

At a position z in hollow fiber length,

$$F = Fi - mD_i\pi \int_0^z J dz \tag{12}$$

where Fi is a initial feed flow rate at the inlet of the module. The changes of feed flow rate, feed composition and feed temperature along with z direction can be determined if water formation rate, flux and permeate are expressed as functions of feed composition, feed temperature, and permeate pressure as follows, respectively,

$$R_w = r(x, T) \tag{13}$$

$$J = f(x, T, P) \tag{14}$$

$$y = g(x, T, P) \tag{15}$$

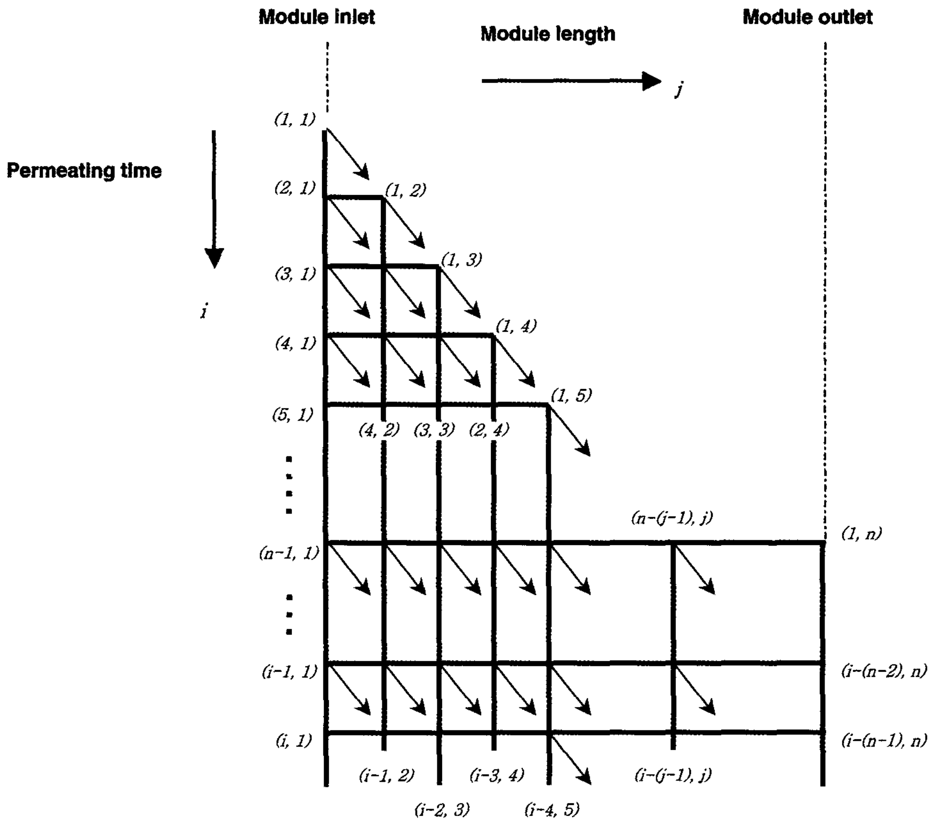

Fig. 2. Finite difference grid used in the numerical solution.

Hence, these three parameters can be also expressed as a function of location along with z direction.

Finite difference schemes are employed to get a numerical solution. This method involves dividing permeating time and hollow fiber length into finite elements, respectively. In continuous process, feed mixture flows through a sequence of finite element volumes to such an extent that the feed in an element volume closer to the outlet of the module has a longer residence time in the module at a given permeating time. Two-dimensional finite difference grid is applied to each to obtain a numerical solution as shown in Figure 2. Horizontal directional parameter j denotes the j^{th} element volume from the entrance of the module while vertical directional parameter i refers to feed introduced into the module at i^{th} time interval. Thus, a coordinate (i,j) in the grid indicates a feed that is located at the j^{th} element volume from the inlet of the module and introduced into the module at the i^{th} time interval. The differential forms of the parameters involved in the above equation, Eqs (10) and (11) can be transformed into the difference form of them, respectively, as follows;

$$T(i,j+1) = T(i,j) - \frac{J(i,j)\Delta h_v m D_i \pi}{F(i,j)C_p}\Delta z \qquad (16)$$

$$x(i,j+1) = x(i,j) + \left[\frac{R_w(i,j)\rho D_i}{4} + (x(i,j) - y(i,j))J(i,j)\right]\frac{m D_i \pi}{F(i,j)}\Delta z \qquad (17)$$

where $T(i,j)$, $F(i,j)$, $J(i,j)$, $x(i,j)$, and $y(i,j)$ are the respective parameters at the j^{th} element volume which is introduced into the module at the i^{th} time interval. The initial and boundary condition for each parameter can be given;

$x(1,1) = x_0$: initial water concentration in feed

$x(i, 1) = x_i$: water concentration to enter the membrane module

$T(1,1) = T(i,1) = T_F = T_i$: feed temperature in the tank

$F(1,1) = F(i,1)$: initial feed pumping speed

Now feed temperature, feed composition feed flow rate, flux and permeate composition at a location in the module and a permeating time can be determined by the numerical method. When feed enters into the module at the i^{th} time interval, the sequence of time interval in respective element volume can be given as $i-(j-1)$ with a position, j, along with membrane module. The permeation parameters at a coordinate (i,j) can be expressed as $R_w(i,j)$, $J(i,j)$ and $y(i,j)$ which are water formation rate, flux, and water content in permeate, respectively. In Eq. (5), $J_M A_T$ is a total permeation amount through the membrane module per unit time. Now the parameters in Eq. (5), that is, the total permeation amount and feed amount M_F can be calculated by using the finite difference scheme;

a. for $i < n$ that is a number of element volumes divided,

$$J_M(i)A_T = mD_i\pi \sum_{j=1}^{i} J(i-(j-1),j)\Delta z \tag{18}$$

$$M_F(i) = (M_F)_0 - mD_i\pi \sum_{i=1}^{i} \sum_{j=1}^{i} J(i-(j-1),j)\Delta z\Delta t \tag{19}$$

b. for $i \geq n$

$$J_M(i)A_T = mD_i\pi \sum_{j=1}^{n} J(i-(j-1),j)\Delta z \tag{20}$$

$$M_F(i) = (M_F)_0 - mD_i\pi \sum_{i=1}^{i} \sum_{j=1}^{n} J(i-(j-1),j)\Delta z\Delta t \tag{21}$$

Eq. (4) can be transformed into the difference form as follows;

$$x_F(k+1) = x_F(k) + \frac{R_w(k)M_F(k) + (x_F(k) - y_M(k))J_M(k)A_T}{M_F(k)}\Delta t \tag{22}$$

The initial condition for each parameter can be given;

$x(1) = x_0$: initial water concentration in feed

$M_F(1) = (M_F)_0$

$J(1) = f(x(1),\ T_F)$

$y(1) = g(x(1),\ T_F)$

3. Experimental

3.1. Membranes

The hollow fiber membrane used in this study was a commercial composite membrane, AzeoSep™-2100H (PetroSep Membrane Technologies Inc., Oakville, Ontario, Canada). Ionic polymers were coated on the inner surface of porous polyolefine hollow fiber. The support hollow fiber was dimensionally and chemically stable during the process but it is somehow vulnerable at elevated temperature. Thus, the hollow fiber composite membrane is for the application at a temperature lower than 60°C. The composite hollow fiber has 1.8 mm in inner diameter and 0.45 mm in thickness. The membrane module accommodates 18 fibers with 31 cm of effective membrane length and 315 cm^2 of total membrane area. The membrane is very effective for dehydration of acidic organic mixtures.

3.2. Pervaporation

Pervaprations with flat sheet membrane were carried out to produce permeation data, which would be used in the simulation model. Feed mixture was circulated from the feed tank through the membrane cell as described well elsewhere [1]. The membrane cell made of stainless steel holds a flat sheet membrane of 154 cm^2. Feed composition ranged from 66 wt.% to 10 wt.% water content. Feed temperatures were 30-60°C and permeate pressures were used from full vacuum to 60 torr. Pervaporative dehydrations of phenolic waste were also performed by using a hollow fiber membrane module to obtain real membrane performances and compare them with the theoretical values calculated from the established model. The schematic description of the pervaporation unit with hollow fiber module is depicted in Figure 1. The membrane module accommodates 18 fibers with 31 cm of effective membrane length and 315 cm^2 of total membrane area. Feed flow rates used were 60 kg/h. The feed and permeate were analyzed by using Karl-Fisher titration and gas chromatography with FID.

4. Results and Discussion

Fitting the permeation data of phenolic waste through flat sheet membrane produced the following dependencies of flux (J), permeate concentration (y) and reaction rate on feed composition, feed temperature and permeate pressure, respectively;

$$J = 1\times10^{8}\{8.2-18.5(2-x-x^{2})\}(1-6.37\times10^{-4}P^{1.517})\exp(-6337.9/T) \quad (23)$$

$$y = (0.92-0.12x)(1-5.59\times10^{-8}P^{2.817})(0.51+0.0043T-9.11\times10^{-6}T^{2}) \quad (24)$$

$$R_w = \mathrm{Exp}(44.97-14600/T-0.9\ln\frac{x}{100}) \quad (25)$$

where x and y are the water concentrations in feed and permeate, respectively, P the permeate pressure (torr) and T the feed temperature (K). The determined permeation functions were in agreement with the experimental values within ± 5% in the given range of operating condition. These functions were used in the simulation of the pervaporation dehydration processes. It is not possible to disclose the reaction mechanism and the list of the waste liquid composition because of the protection of related technologies.

First, to verify the established simulation model, the calculated values were compared with the real values measured from the hollow fiber module. Figure 3 exhibits the comparison of the simulated feed composition and total flux to the experimental data with permeating time. The measurements were repeated once more to see their reproducibility. As can be seen in the figure, the calculations have an excellent agreement with the real measurements although real

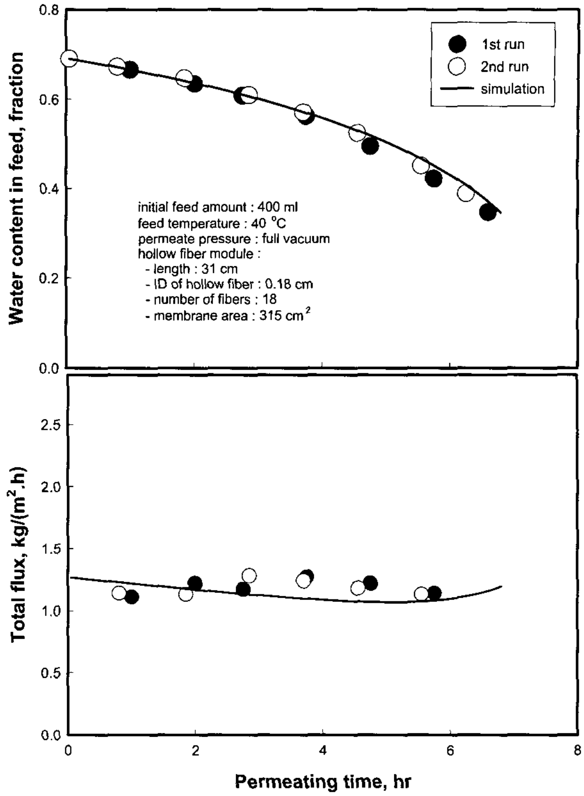

Fig. 3. A comparison of the simulated values with experimental data of water content in feed and flux through hollow fiber module at 40°C.

fluxes are relatively smaller than the calculated due to unsteady-state conditions accomplished during the incipient permeation. The water concentration in permeate ranged 98.3−96.0 wt.% during the dehydration. According to these findings, it should be noted that the established model is of practical value to predict the pervaporative dehydration process. Now, let's examine the permeation and separation behavior of the reactive phenolic resin waste through the simulation of the process with the hollow fiber membrane module. The simulation was based on the treatment of 5 kg of waste feed and 3 stages of hollow fiber membrane modules through which feed mixture flows sequentially. Each module is 0.42 long and involves 71 fibers, having an effective membrane area of 0.5 m^2. The feed mixture is circulated from the feed tank through the hollow fiber membrane module until target dehydration is achieved; dehydration from 66 wt.% of

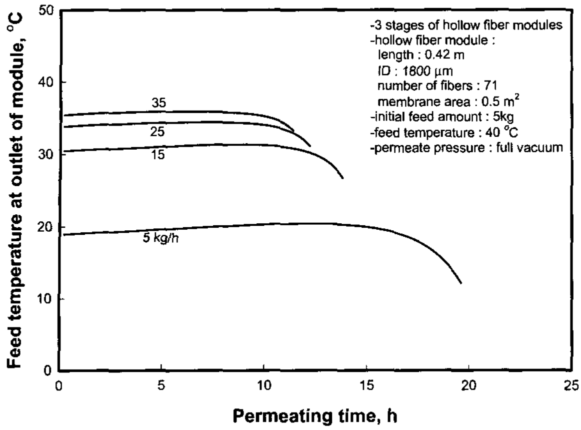

Fig. 4. Simulated feed temperature at outlet of module with permeating time at different feed flow rates at 40°C.

water content to 10 wt.%.

Actually, feed flow rate through hollow fibers is related to the temperature distribution of feed along with fiber length; the slower the feed flows, the more the temperature gradient is developed. It is because when feed flows slower, it undergoes more separation and permeation so that feed should supply more heat to evaporate the permeating liquid into permeate, and thereby feed temperature decreases more as it flows along with hollow fiber. Figure `4 shows the feed temperature at the outlet of module with permeating time at different flow rates. As discussed just before, the feed temperature at the exit site of the module decreases at slower feed flow. It is very interesting to note that the temperature increases slightly with permeating time and then decreases rapidly. The slight increase in the temperature is associated with more concentrating organic in the retentate leading to less permeation and less heat needed for the evaporation. As the retentate gets concentrated, chemical reaction proceeds faster and produces more water as described in Eq. (25) expressing reaction rate above. Figure 3 illustrates that flux decreases with permeating time and then increases again due to producing significantly water by the chemical reaction. Thus more evaporation energy should be supplied from the feed and thereby the temperature of feed eventually decreases. Figure 5 presents water concentration in feed, cumulative amount

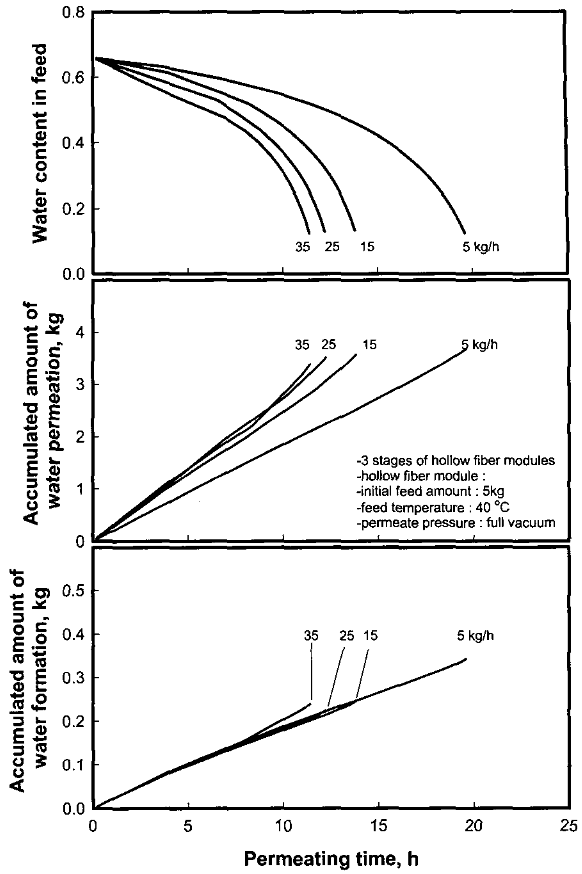

Fig. 5. Plots of permeation performance and water formation rate with time at different feed flow rates at 40°C.

of water permeation and cumulative amount of water formation with time at different feed flow rates and 40°C. At higher feed flow rate, dehydration is faster because of higher feed temperature in the membrane module. In the plots of the cumulative amount of water formation, the slope of each line represents the formation rate of water by chemical reaction and its end denotes a point to reach 10 wt.% of water content in feed by dehydration. During all the period of permeation, water formation rates are almost equal regardless of feed flow rate except the permeation system with 35 kg/h of which reaction proceeds at the same rate and then faster after about 8 h than the other permeation systems with slower feed flow. It should be noted that the accumulated amount of water formation decreases to a minimum value at 25 kg/h and then increases with increasing feed flow rate for the dehydration to reach 10 wt.% of water content in feed. The parabolic trend of the water formation can be caused by the .competing effects; 1) the effect of reaction time and 2) the effect of reaction temperature. At fast feed flow in the module, average feed temperature is high as shown in Figure 4 while permeation time required is short due to fast dehydration, and *vice versa*. Either higher temperature or longer permeation time that is equivalent with reaction time increases the amount of product. At high feed flow, water formation is dominated by the temperature effect, whereas long reaction time is more likely to affect significantly to produce water at slow feed flow. Now that 25 kg/h of flow rate does not only produces minimum water formation but also is not much different from 35 kg/h of flow rate in the water content in feed and the cumulative amount of water permeation with time, 25 kg/h must be optimum flow rate.

Figure 6 demonstrates the effect of feed temperature on the permeation and reaction. As feed temperature increases, water permeation progresses faster. So, the feed is more concentrated with time, as shown in this figure. At 60°C, it is very interesting to note that water content in feed is increasing with time rather than decreasing at the beginning stage of dehydration. It is because the formation rate of water is larger than that its depletion rate at this temperature, resulting in more water accumulated in the tank, indicating that the reaction requires higher activation energy than the permeation does. Higher feed temperature produced more water and the dehydrated liquid is reacted to greater extent, as shown in the figure. Thus the formation of more water would devalue the resulting dehydrated liquid as reusable material. Looking at the plot of accumulated amount of water formation, the accumulated amount increases exponentially with increasing temperature. At 50°C, total amount of produced water is about 0.1 kg, which is equivalent with 80% organic species reacted, showing the reaction is gone too much. At 40°C, 0.045 kg of water is produced, reacted only 35% of organic species, that is, more than 50% of organics remain unreacted. At temperature lower than 40°C, permeation rate of water is too small while a temperature higher than 50°C reaction proceeds too much during the permeation. Therefore, taking into consideration membrane performance and reaction progress, 40°C seems to be optimum.

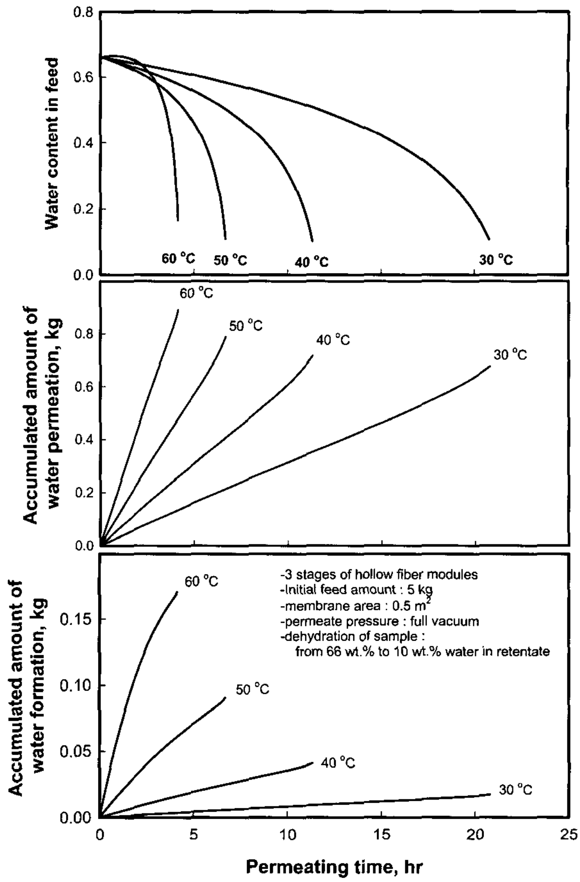

Fig. 6. Plots of permeation performance and water formation rate with time at different feed temperature at 25 kg/h of feed flow rate.

Figure 7 shows the variation of feed temperature against permeating time at different permeate pressure. When permeate pressure is higher, permeation rate is smaller due to smaller driving force to the permeation, so that the temperature drop by the permeation can be smaller. The saturated vapor pressure of pure water is 55 torr at 40°C and the calculated vapor pressure of initial feed is 66 torr. All the permeate pressures used are lower than the feed vapor pressure, that is, permeation taking place across membrane at the permeate pressures. However the effect of permeate pressure is

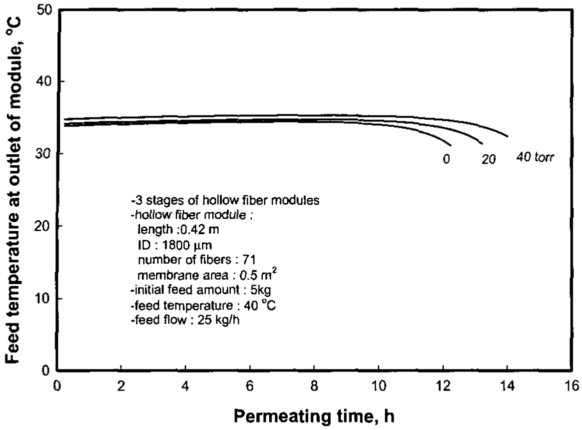

Fig. 7. Simulated feed temperature at outlet of module with permeating time at different permeate pressures at 40°C.

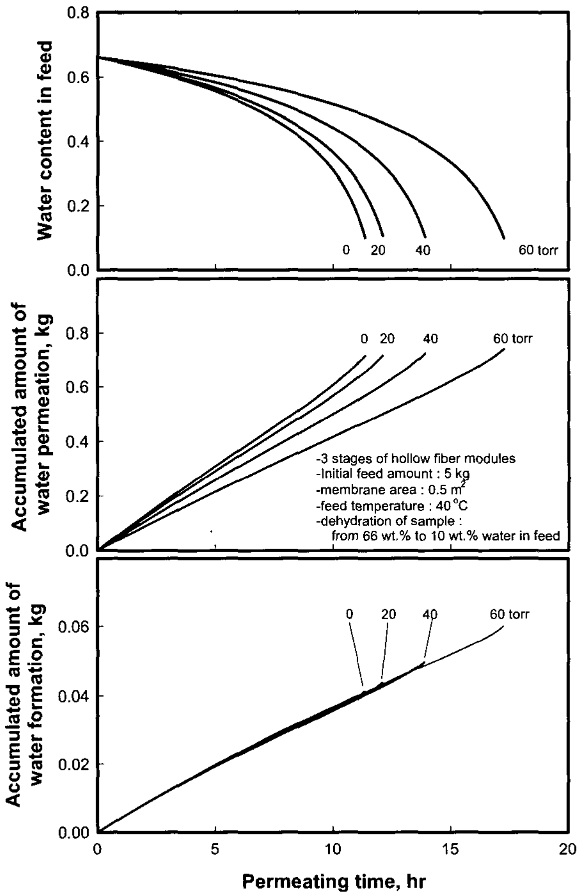

Fig. 8. Plots of permeation performance and water formation rate with time at different permeate pressures at 40°C.

not as significant as the other effects. Figure 8 shows the effect of permeate pressure on permeation and reaction. Up to 40 torr of permeate pressure, water permeation performance and water formation by reaction do not change so much with permeate pressure. However as the permeate pressure approach the vapor pressure of the feed, dehydration rate decreases dramatically and the accumulated amount of water formation also increases as much. Obviously, it is attributed to a significant decrease in the activity ratio of feed to permeate, which is a driving force for the permeation. Especially since separation factor is dependent on the permeate pressure, optimum permeate pressure should be compromised in terms of separation efficiency required and capital cost for the pumping and condensing system.

4. Conclusions

A simulation model of pervaporative hollow fiber module has been established for the dehydration of phenolic resin waste that has a reactivity to produce water. Some of permeation parameters and reaction parameters used in the model were determined directly from the dehydration of the waste through a flat sheet membrane. The simulation model was verified by comparing the simulated values with experimental data obtained from the dehydration of the waste through hollow fiber module. Hollow fiber membranes with active layer coated on inside fiber were used, and feed flew through inside hollow fiber. Feed flow rate affected membrane performances and reaction by providing a corresponding temperature distribution of feed along with fiber length. 25 kg/h of feed flow rate yields minimum water formation and reasonable dehydration rate. Feed temperature is also a crucial factor to determine dehydration and reaction behavior by two competing ways; increasing temperature increases permeation rate as well as water formation rate. Thus, optimum temperature was found to be 40°C at which the reaction is minimized, the membrane performance is maximized and the product quality is maintained. Once the permeate pressure is below the saturated vapor pressure of feed, permeate pressure affected negatively by reducing driving force. As the pressure approached the vapor pressure of feed, dehydration performances declined considerably due to the activity ratio of feed and permeate. The established model can not only be used as design tool but also provide a tool for the techno-economical analysis of pervaporation dehydration for the treatment of the reactive phenolic resin waste.

List of Symbols

A_T : effective membrane area in module (m^2)
C_p : heat capacity of feed (kcal/(kg·°C))
D_i : inner diameter of hollow fiber (m)
dz : differential length of hollow fiber (m)
F : feed flow rate (kg/h)
h_F : enthalpy of feed (kcal/mol)
Δh_V : heat of the evaporation of permeant (kcal/mol)
J : flux (kg/(m^2·h))
m : number of fibers in unit module (ea)
M : feed mass in feed tank at a time t (kg)
R_W : rate of water formation (kg/(kg·h))
t : permeating time (h)
T : feed temperature (K)
x : concentration of a selectively permeating component in feed (wt.%)
y : concentration of a selectively permeating component in permeate (wt.%)
ρ : density of feed (kg/m^3)

Subscripts

F : feed
M : membrane
T : total system
0 : initial state

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