Membrane vacuum stripping process for volatile organics removal from water using an asymmetric PVDF (polyvinylidene fluoride) hollow fiber membrane module

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Abstract The vacuum membrane stripping process (VMS) for removing a VOC (trichloroethylene) from water was studied in a polyvinylidene fluoride (PVDF) hollow fiber membrane module containing 187 fibers with a length of 0.59 m. The porous PVDF asymmetric hollow fiber used was prepared by the wet phase inversion technique. The feed solution containing trichloroethylene (TCE) was passed through the lumen of the PVDF hollow fiber. Vapors of TCE and water were transported through the pores of the membrane into the shell side maintained at a sub-atmospheric pressure, and condensed in cold traps. The effects of down-stream pressure, solution temperature, feed concentration, and feed flow rate on TCE removal, TCE permeation flux, water permeation flux and TCE concentration of the permeated solution, as well as the mass transfer coefficient were investigated. The down stream pressure and solution temperature were identified as the major factors to control VOC removal and TCE concentration in the permeate stream. A small amount of the permeated solution with higher TCE concentration was obtained at the temperature of 50°C and pressure of 80 mmHg when about 90% TCE was removed from the water.

Keywords Hollow-fiber membranes; PVDF; vacuum stripping; VOC removal

Introduction Development of an efficient separation technique for the removal and recovery of volatile organic compounds (VOCs) from water is of great interest in water and wastewater treatment. The VOCs commonly found in water and wastewater are benzene, toluene, xylene, naphthalene, acetone and a wide range of chlorinated hydrocarbons (Bravo, 1994). Various technologies such as steam stripping, air stripping, carbon adsorption, chemical oxidation, and membrane systems have been applied for the removal of these VOCs from water (Walter and Smith, 1986). Of these technologies, air stripping and steam stripping technologies are most commonly used in industrial processes. Both technologies depend on the VOC volatility in water. The drawback of the air stripping process is that VOCs are only transferred from the liquid phase into air, which requires further treatment of the contaminated air. Steam stripping takes place at a higher temperature than air stripping. Since the volatility of VOCs is a strong function of temperature and high temperature enhances the removal of VOCs. Thus steam stripping is often more efficient and can achieve both high removal rate and low effluent concentration. The only waste stream generated is a small amount of very concentrated organic compounds. The major concerns are the energy intensity and precipitation of inorganic salts at elevated temperature (Bravo, 1994).

Conventional stripping processes are performed in stripping columns such as packed or plate columns. In order to affect efficient removal of VOC, a combination of porous membrane and the stripping process has been introduced. Semmens et al. (1989) and Zander et al. (1989) studied separation of VOCs from underground water by the combination of
porous polypropylene hollow fiber membrane and air stripping. The studies have shown that there are several advantages of membrane air-stripping over conventional packed tower air-stripping. These include larger interfacial area per unit volume, no loading and flooding limitations, reduction of dispersed phase back mixing, etc. However, the VOCs laden air still requires further treatment by carbon adsorption.

As a good VOC separation technique, high VOC removal efficiency and high VOC concentration in the collected solution are required. A small quantity of concentrated solution is easily treated and can be reused. Vacuum membrane stripping (VMS) or vacuum membrane distillation (VMD) is a good alternative for this application. This process is similar to membrane pervaporation. The difference is that the VMS process involves porous hydrophobic membrane, whereas the pervaporation process uses a non-porous dense membrane. In the VMS process, both the VOC and water vapor are transferred through the pores of a porous membrane under vacuum and are condensed in a cold-trap. Separation is thus based on the relative volatility of VOC with respect to water. The VMS process has several advantages over the steam stripping process. Firstly, because of operation at sub-atmospheric pressure, low temperature can be applied. Low temperature operation is beneficial to avoid possible fouling by precipitates and also energy saving. Moreover, the large membrane area available for vapor-liquid contact can greatly improve VOC removal efficiency. Lastly, scale-up is easy since a membrane process is modular.

Some recent studies on VOC removal by the VMS process have been made. Bandini et al. has examined transport of various VOCs in a VMS process (Bandini et al., 1997, 1999) operated batchwise. In their research, the feed solution was passed through the membrane at sufficiently high flow rate, resulting in little change in VOC concentration in the inlet and outlet stream. The permeated vapors were condensed in a cold trap using liquid nitrogen. The VOCs tested include acetone, ethanol, isopropanol, ethylacetate, methylacetate and methytertbutyl ether. The VOCs concentration in the feed solution was varied from 0.5wt% to 9wt%. It has been shown that temperature, pressure, and feed flow rate will affect VOC concentration and amount of the recovered solution. High concentration of VOC in the recovered solution was obtained under optimal operating conditions.

This paper presents the results obtained from a study on the removal of trichloroethylene (TCE) in a continuous VMS system. Asymmetric polyvinylidene fluoride (PVDF) porous hollow fiber was used as the barrier separating the liquid and gas phase. A two-stage cold trap using ice-water in the first stage followed by liquid nitrogen in the second stage, were applied to condense the vapors. The effects of downstream pressure, feed solution temperature, its concentration and feed flow rate on the separation efficiency, permeate concentration and mass transfer coefficient have been studied and addressed.

**Experimental**

**Preparation of PVDF hollow fiber membrane**

The porous asymmetric PVDF hollow fiber membrane used in this study was prepared by the wet-phase inversion method from a polymer dope with the following composition: 20wt% PVDF (Keney-720), 74.9wt% dimethylacetamide, 3.7wt% LiCl and 1.4wt% H2O. Water was also used as both an internal and external coagulant. The hollow fiber was spun at the ambient conditions (25 ± 1°C, RH = 60%–65%). The detailed spinning procedure and the spinning conditions have been described elsewhere (Wang et al., 1995). The prepared hollow fiber was washed in tap water for one week. The water-wetted membrane was then immersed in ethanol for ten minutes and dried at the ambient conditions. The dried hollow fiber had an outer diameter of 0.735 mm and an inner diameter of 0.42 mm.
Characterization of hollow fiber

The structure of PVDF hollow fiber was characterised by the gas permeation method described elsewhere (Li et al., 1999). The nitrogen permeation rate was measured at different pressure differences (1–20 psi) and at room temperature (25 ± 1°C). By plotting the gas permeability versus the average pressure, intercept and slope were obtained which were used to determine the mean pore size (r) and the surface porosity (ε) over the effective pore length (L), ε/L. The average pore size, r and ε/L determined for the PVDF hollow fiber membrane were 5.04 × 10⁻⁸ m and 8.49 × 10⁻² m⁻¹, respectively. Based on these data, mass transfer coefficients in the membrane, K_M, at various temperatures were then calculated (Li et al., 1999). The calculated K_M at 30°C, 40°C, 50°C and 60°C are 0.0063 m/s, 0.0064 m/s, 0.0065 m/s and 0.0066 m/s, respectively. The nitrogen permeability of the membrane measured at 0.2 bar and 25 ± 1°C was 1.69 × 10⁻² cm³(STP)/cm².cmHg.s.

Module preparation

A normal shell-and-tube type hollow fiber module was fabricated. A stainless steel tube with inner diameter of 0.016 m acted as a shell which has two equal sized “Tees” connected at both ends. Both the tees were then connected to a reducer each on which the fibers were sealed using epoxy resin. The gaps between the fibers and the internal surface of the reducer were also bounded together with epoxy resin. The basic dimensions and other relevant information of the module are listed in Table 1.

Experimental procedure

Feed solutions with different concentrations of TCE in water were prepared in 1,000 ml flasks. The mixture of TCE and water was homogenized under agitation for about a day. The set-up of experimental apparatus is schematically shown in Figure 1. The whole membrane module was immersed in a thermostat to control the temperature. The TCE laden water was first passed through a stainless steel coil in the thermostat to preheat the solution to the desired temperature prior to feeding into the lumen of the hollow fibers (tube side operation mode). The flow rate was controlled by a liquid mass flow controller (Brooks Instruments, USA). A vacuum system (Buchi Vacuum system B-178) was applied to control a preset sub-atmospheric pressure in the shell side. Water vapor and TCE vapor permeated through the membrane were first cooled down with an ice-cold trap and then with a liquid nitrogen cold-trap. Most of the water vapor was condensed in the ice-cold trap and the more volatile TCE was condensed in the liquid-nitrogen cold trap. For each experimental run, the weight of the condensed liquid was monitored continuously. The effluent TCE concentration was measured with a TOC-5000A (Shimadzu). For measuring VOC concentration of the condensed solution in the cold traps, the solution was first diluted prior to measurement using the TOC-5000A.

Table 1 Basic dimensions of the PVDF hollow fiber module

<table>
<thead>
<tr>
<th>Module i.d. (m)</th>
<th>1.6 × 10⁻²</th>
</tr>
</thead>
<tbody>
<tr>
<td>Module o.d. (m)</td>
<td>1.9 × 10⁻²</td>
</tr>
<tr>
<td>Fiber OD (m)</td>
<td>7.35 × 10⁻⁴</td>
</tr>
<tr>
<td>Fiber ID (m)</td>
<td>4.2 × 10⁻⁴</td>
</tr>
<tr>
<td>Module length (m)</td>
<td>0.59</td>
</tr>
<tr>
<td>Fiber length (m)</td>
<td>0.51</td>
</tr>
<tr>
<td>No. of fibers</td>
<td>187</td>
</tr>
<tr>
<td>Packing fraction</td>
<td>0.4</td>
</tr>
<tr>
<td>Surface/volume ratio</td>
<td>2147</td>
</tr>
<tr>
<td>Membrane area (m²)</td>
<td>2.2 × 10⁻¹</td>
</tr>
<tr>
<td>Membrane area (m²) (inner, m²)</td>
<td>1.26 × 10⁻¹</td>
</tr>
<tr>
<td>Lumen cross-section area</td>
<td>2.59 × 10⁻⁵</td>
</tr>
<tr>
<td>Lumen volume (m³)</td>
<td>1.32 × 10⁻⁵</td>
</tr>
</tbody>
</table>
The fluxes of permeated streams were determined from the amount of solution collected from both the condensers over the operating period (t):

\[ J_1 = \frac{W_1}{A_i} \]  
(1)

\[ J_2 = \frac{W_2}{A_i} \]  
(2)

Where \( J_1 \) and \( J_2 \) are permeated solution flux collected in the ice-cold trap (1) and liquid nitrogen cold trap (2), respectively, \( W_1 \) and \( W_2 \) are the amount of the condensed solution collected over time \( t \) in both the traps and \( A_i \) is inner surface area of the fibers. TCE permeation flux \( (J_v) \) and water permeation flux \( (J_w) \) can be determined as follows:

\[ J_v = J_1 C_1 + J_2 C_2 \]  
(3)

\[ J_w = J_1 + J_2 - J_v \]  
(4)

Where \( C_1 \) and \( C_2 \) are TCE weight concentrations of the solutions collected in the ice-water cold trap and liquid nitrogen cold trap, respectively. The average weight concentration on the permeate side was given by:

\[ C_p = \frac{J_1 C_1 + J_2 C_2}{J_1} \]  
(5)

The percent removal \( (R) \) of VOC is defined as:

\[ R = \frac{C_{in} - C_{out}}{C_{in}} \times 100 \% \]  
(6)

**Mass transfer coefficient**

In a VMC process, VOC and water are first transported from the bulk of solution to the liquid-membrane interface where they evaporate and permeate through the pores of the membrane under the vacuum to the other side of the membrane or vapor-membrane interface, and then diffuse into the bulk of the vapor phase. In this transport process, three mass transfer resistances (liquid film, gas-filled membrane phase and vapor film) are encountered. In general, for a porous membrane, the resistances of gas-filled membrane phase and vapor film are small compared to the liquid film resistance and may be neglected (Das *et al.*, 1998). Under the steady state condition, the mass transport of VOC across the liquid film can be described as:

\[ Q_w \frac{dC_i}{dx} = K_L A_i (C_i - C) \]  
(7)
where \( Q_w \) is the volumetric flow rate of water solution (m\(^3\)/s); \( K_L \) is mass transfer coefficient (m/s); \( C^* \) is equilibrium concentration on the interface. Integrating Equation (7) with boundary conditions \( C = C_{in}, x = 0; \) and \( C = C_{out}, x = L, \) the mass transfer coefficient can be determined.

\[
K_L = \frac{Q_w}{A_v} \ln \left( \frac{C_{in} - C^*}{C_{out} - C^*} \right)
\]

(8)

For a solute at low concentration and low pressure, the VOC equilibrium concentration \( C^* \) is related to the pressure \( p, \) molar fraction \( x \) of VOC in the permeate side according to Henry’s law by:

\[
C^* = \frac{xp}{H}
\]

(9)

Based on reported Henry’s constant of TCE at the temperatures of 10°C to 30°C (Montgomery, 1996), the dependence of Henry’s constant, \( H \) with temperature could be expressed as:

\[
H = \exp \left( 7.891 - \frac{3714}{T} \right)
\]

(10)

where \( T \) is the temperature in K. The calculated Henry’s constants of TCE in water at the temperatures of 30°C, 40°C, 50°C and 60°C are 0.0128, 0.0173, 0.0232, and 0.0299 atm.m\(^3\)/mol, respectively.

**Results and discussion**

For the PVDF microporous membrane used in this study, the calculated mass transfer coefficient in the membrane is in the range from \( 6.3 \times 10^{-3} \) to \( 6.6 \times 10^{-3} \) m/s over the temperature range of 30–60°C. The mass transfer coefficient in the liquid film is usually in the range of \( 10^{-6}–10^{-7} \) m/s. Therefore, the influence of membrane resistance to VOC is negligible. The important process variables are downstream pressure, temperature, feed concentration and feed flow rate. The influence of each quantity has been investigated in detail and is presented below.

**Downstream pressure**

The effect of downstream pressure on TCE removal, TCE flux and water flux was first studied. The feed solution with the TCE concentration of 650 ppm was passed through the lumen side at a flow rate of 0.1 L/hr and temperature at 50°C. The downstream pressures were varied and maintained at 100 ± 3 mmHg, 80 ± 3 mmHg, 50 ± 3 mmHg and 1–3 mmHg, for each experimental run. The TCE removal was determined by measuring the outlet TCE concentration. Water permeation flux, TCE permeation flux and TCE concentration were determined by analyzing the condensed liquid in both the ice and liquid nitrogen cold traps. Figure 2 shows the TCE removal and the permeation fluxes of water and TCE. As depicted, the influence of the downstream pressure on TCE removal and TCE flux is not significant over the pressure range from 1mmHg to 100mmHg. The TCE removal only increased from 86% to 90% and the TCE permeation flux increased from 0.27 g/hr.m\(^2\) to 0.4 g/hr.m\(^2\) over this pressure range. However, the downstream pressure seriously affects the water permeation flux. At the pressure of 50 mmHg to 100 mmHg, water vapor permeation flux increased only slowly, but greatly increased if the pressure was further decreased to 1 mmHg. For aqueous solution containing low levels of VOC, the VOC transport is dominantly controlled by the liquid film resistance, while the water vapor transport is controlled by the membrane resistance. When the vacuum is much lower than the
water vapor pressure, the rapid increase in the driving force results in greater water vapor permeation.

Table 2 lists the permeation fluxes and TCE concentrations of the solution collected in both the ice-water cold trap and liquid nitrogen cold trap. At high vacuum pressures (80 mmHg and 100 mmHg), a small amount of the condensed solution with high TCE concentration was collected in both the traps. The TCE concentration for the solution collected in the liquid-nitrogen cold trap was 66%–76 wt%. It should be noted that the real TCE concentration could be higher than the data listed in Table 2. During dilution, some TCE may be lost. At high downstream pressure, a very small amount of the permeated solution was collected. For 100 g/hr feed solution, the total condensed solution collected in the both traps was only 0.08–0.1 g/hr when about 90% TCE was removed. This is beneficial from an environmental and economic point of view. When the vacuum pressure was reduced to 1 mmHg, the water vapor permeation rate was quickly increased. About 20 wt% feed solution was permeated and condensed in the ice-water cold trap. The permeated solution has quite low TCE concentration. Because heat transfer from the outer side is limited due to the vacuum in the shell side, the evaporation of a large amount of water results in a temperature decrease of the solution in the lumen. Thus, volatility of TCE in the solution is reduced. In addition, the low vacuum pressure provides a possibility to wet the membrane pores. Therefore, too low a vacuum pressure is not required for the removal of VOC with the membrane vacuum systems.

The mass transfer coefficient tends to decrease from $6.7 \times 10^{-7}$ m/s to $5.3 \times 10^{-7}$ m/s when the pressure is decreased from 100 mmHg to 1 mmHg (Table 2). At lower pressure,

Table 2 Permeation flux and TCE concentration of the permeated solution at different vacuum pressures

<table>
<thead>
<tr>
<th>Pressure (mmHg)</th>
<th>Ice water trap</th>
<th>Liquid nitrogen trap</th>
<th>$K_a$ (m/s) x $10^7$</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>$J_W$ (g/m²·hr)</td>
<td>$C_W$ (wt%)</td>
<td>$J_L$ (g/m²·hr)</td>
</tr>
<tr>
<td>100</td>
<td>0.3</td>
<td>1.79</td>
<td>0.33</td>
</tr>
<tr>
<td>80</td>
<td>1.62</td>
<td>0.44</td>
<td>0.33</td>
</tr>
<tr>
<td>50</td>
<td>27.8</td>
<td>0.174</td>
<td>0.48</td>
</tr>
<tr>
<td>1</td>
<td>112.9</td>
<td>0.03</td>
<td>0.87</td>
</tr>
</tbody>
</table>

Figure 2 TCE removal, water permeation flux and TCE permeation flux at different downstream pressures
more water is evaporated and solution temperature is decreased. The decrease in temperature of the solution is the major reason for decreasing mass transfer coefficient.

**Feed solution temperature**

The effect of solution temperature on the TCE removal performance was studied at a downstream pressure of 80mmHg and feed flow rate of 0.1 L/hr. The TCE concentration in the feed solution was 630–650 ppm. The temperature was varied from 30°C to 60°C. As shown in Figure 3 and Table 3, TCE removal and TCE permeation flux increased steadily with increasing temperature. The volatility and diffusivity of TCE in water increased with increasing temperature. Thus more TCE was removed with increasing temperature. For water, the permeation flux increases slowly over the temperature range from 30°C to 50°C. However, when temperature was increased from 50°C to 60°C, the water flux was quickly increased from 1.6 g/m² .hr to 75 g/m² .hr. At high temperature, more water evaporated and permeated through the membrane due to high water vapor pressure. The TCE concentrations of the solution collected in both the cold traps decreased with increasing temperature. Still, higher concentration of TCE was obtained at the liquid-nitrogen cold trap. Although an increase in temperature results in an increase of the TCE removal, more water passed through the membrane at high temperature. From the experimental results, the optimal temperature is around 50°C.

As shown in Table 3, with increasing temperature from 30°C to 50°C, the permeated solution flux and TCE concentration are steadily increased. If the temperature increased

### Table 3 Permeation flux and TCE concentration of the permeated solution at different temperatures

<table>
<thead>
<tr>
<th>Temperature (°C)</th>
<th>Ice water trap</th>
<th>Liquid nitrogen trap</th>
<th>Liquid nitrogen trap</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>J₁ (g/m² .hr)</td>
<td>C₁ (wt%)</td>
<td>J₂ (g/m² .hr)</td>
</tr>
<tr>
<td>30</td>
<td>0.28</td>
<td>0.8</td>
<td>0.22</td>
</tr>
<tr>
<td>40</td>
<td>0.38</td>
<td>0.6</td>
<td>0.26</td>
</tr>
<tr>
<td>50</td>
<td>1.62</td>
<td>0.4</td>
<td>0.33</td>
</tr>
<tr>
<td>60</td>
<td>74.4</td>
<td>0.126</td>
<td>0.63</td>
</tr>
</tbody>
</table>

![Figure 3](https://iwaponline.com/ws/article-pdf/1/5-6/65/477301/65.pdf)
from 50°C to 60°C, the amount of the solution collected is quickly increased and low TCE concentration was obtained in the ice-water cold trap. As expected, the mass transfer coefficient increases with increasing temperature.

Studies on the influence of pressure and temperature indicate that the pressure and temperature are two important parameters in controlling VOC removal and the VOC concentration in the permeated stream. Low temperature and high downstream pressure are favorable for obtaining high concentration of VOC in the permeate stream, but can not achieve high VOC removal. Proper control of pressure and temperature is important. The choice of the pressure is closely related to the saturation vapor pressure of water. The water vapor pressures are 31.8, 55.3, 92.6, 149.4, 233.8 mmHg at the temperature of 30°C, 40°C, 50°C, 60°C and 70°C, respectively. It was suggested that the vacuum pressure used was 0.9 to 1.2 times the water vapor pressure (Bandini et al., 1997). Based on the water vapor pressure, the solution temperature should be less than 60°C. High pressure is not suitable to efficiently remove VOC. From the experimental results obtained in this study, vacuum pressure of 0.9 times of the water vapor pressure and temperature of around 50°C are optimum conditions for VOC removal from water.

**Effect of feed concentration**

The effect of the TCE concentration in the feed solution was examined at 50°C and a vacuum pressure of 80 mmHg. The feed concentrations were in the range from 124 ppm to 890 ppm. The TCE permeation flux increases with increasing feed concentration, while TCE removal and water permeation flux exhibited little influence as shown in Figure 4. An increase in the feed concentration results in increase in the TCE driving force, thus, more TCE permeated through the membrane. The feed TCE concentration is in the ppm range, the variation of the feed concentration does not affect the water concentration on the interface between membrane and water solution. Thus, water permeation flux had little change. About 90% TCE removal was obtained for the solutions with different feed concentrations. As shown in Table 4, higher TCE concentration in the permeated stream was obtained at a higher TCE concentration in the liquid feed. On increase in the feed concentration from 124 ppm to 890 ppm, the TCE concentration collected in the liquid-nitrogen cold-trap

![Figure 4](https://iwaponline.com/ws/article-pdf/1/5-6/65/477301/65.pdf)
increased from 50wt% to 85wt%. The permeation solution flux has little change. This further demonstrates that water permeation flux mainly depends on the vacuum pressure and temperature. With increasing feed concentration, the mass transfer coefficients are decreased.

**Feed flow rate**

At a temperature of 50°C and feed concentration of 600–620 ppm, the influence of the feed flow rates was studied under a vacuum pressure of 80 mmHg. The flow rate was in the range of 0.1 L/hr to 0.5 L/hr. Figure 5 illustrates the effect of feed flow rate on TCE removal, TCE flux and water flux. With increasing flow rate from 0.1 L/hr to 0.5 L/hr, TCE removal slightly decreased from 89 to 85. Water permeation flux and TCE permeation flux increased. As discussed above, the evaporation of water in the fiber will reduce the solution temperature. With increasing feed flow rate, the temperature difference between the inlet and the outlet became less. The water vapor flux increase is mainly caused by the temperature. Two factors contribute to the TCE flux increase, e.g. concentration gradient and temperature gradient. The higher feed flow rate results in the TCE driving force along the

<table>
<thead>
<tr>
<th>Concentration (ppm)</th>
<th>Ice water trap</th>
<th>Liquid nitrogen trap</th>
<th>K_V (m/s) x 10^7</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>J_v (g/m².hr)</td>
<td>C_v (wt%)</td>
<td>J_w (g/m².hr)</td>
</tr>
<tr>
<td>124</td>
<td>1.59</td>
<td>0.39</td>
<td>0.25</td>
</tr>
<tr>
<td>340</td>
<td>1.60</td>
<td>0.50</td>
<td>0.30</td>
</tr>
<tr>
<td>650</td>
<td>1.53</td>
<td>0.44</td>
<td>0.26</td>
</tr>
<tr>
<td>890</td>
<td>1.53</td>
<td>0.68</td>
<td>0.27</td>
</tr>
</tbody>
</table>

*Figure 5* TCE removal, water permeation flux and TCE permeation flux at different feed solution flow rates
hollow fiber. On the other hand, with increasing feed flow rate, the temperature difference along the lumen of the fiber becomes less. This improves the TCE permeation. Higher feed flow rate leads to a higher TCE concentration in the condensed liquid as listed in Table 5. The permeated solution flux was also increased on increasing the feed flow rate. With increasing solution flow rate, the mass transfer coefficient quickly increased from $6.02 \times 10^{-7}$ m/s to $24.2 \times 10^{-7}$ m/s.

**Conclusions**

The removal and recovery of trichloroethylene (TCE) from water using a PVDF hollow fiber membrane vacuum stripping process was studied. This process is efficient for the removal and recovery of VOC from water by controlling at the optimal downstream pressure and solution temperature. The downstream pressure and solution temperature are two major factors in the membrane vacuum stripping process. A very small amount of permeated solution (~0.1 wt% of the feed solution) with high TCE concentration can be obtained at a downstream pressure of 80 mmHg and at a temperature of 50°C. Further increase in temperature and reduction in pressure yielded little improvement in the TCE removal, but greatly increased the amount of condensed solution. Two stages of cooling condensation are beneficial to obtain high VOC concentration in the permeated solution. The 60–85 wt% TCE concentration in the permeate stream was obtained from the feed concentration of 100–1000 ppm with TCE removal of up to 90%. The water permeation flux is much less sensitive to changes in the feed solution concentration and feed flow rate compared to pressure and temperature. The influence of temperature, pressure and feed concentration on the mass transfer coefficient is much less than that of the feed flow rate. The working pressure and temperature suitable for a more convenient process are approximately 80 mmHg and 50°C.

**Symbols**

- $A$ = inner surface area of hollow fiber membrane, $m^2$
- $C_1$ = concentration of permeated solution collected in ice-cold trap, (wt%)
- $C_2$ = concentration of permeated solution collected in liquid nitrogen-cold trap, (wt%)
- $C_\text{in}$ = concentration of feed solution, (ppm)
- $C_\text{out}$ = concentration of outlet solution, (ppm)
- $C^*$ = equilibrium concentration on the liquid/membrane interface, (ppm)
- $H$ = Henry’s constant, (atm.m$^3$/mol)
- $K_L$ = mass transfer coefficient, (m/s)
- $J_v$ = VOC permeation flux, (g/m$^2$.hr)
- $J_w$ = water permeation flux, (g/m$^2$.hr)
- $J_1$ = permeated solution flux collected in ice-cold trap, (g/m$^2$.hr)
- $J_2$ = permeated solution flux collected in liquid nitrogen cold trap, (g/m$^2$.hr)
- $R$ = removal of VOC, (%)
Acknowledgements
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References