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多行程裝置對逆交流平板薄膜模組中萃取效率之影響

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Effect of Multipass Arrangement on Solvent Extraction in Countercurrently Cross-Flow Rectangular Membrane Modules of Fixed Configuration

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Abstract— Solvent extraction through a countercurrently cross-flow multipass membrane module has been investigated both theoretically and experimentally. Considerable improvement of mass-transfer rate in a parallel-plate module of fixed dimensions is obtained if multipass, instead of single pass, are arranged in the phase (retentate phase or extract phase) where the mass transfer is dominated by its external film and not by the membrane resistance. Further, the performance in a countercurrently cross-flow device introduced in present study overcomes that in a cocurrently cross-flow one which we discussed in previous works.

Key Words : Liquid membrane extraction, Countercurrently cross flow, Multipass operation

INTRODUCTION

The performance of liquid solvent extraction through single-pass parallel-plate membrane modules has been discussed under parallel-flow (cocurrent and countercurrent flows) and cross-flow operations (Yeh and Hsu, 1999; Yeh and Chen, 2000a). Membrane solvent extraction is a dispersion-free separation process which can overcome the application limitations of conventional solvent extraction, such as flooding, intimate mixing, limitations on independent phase flow rate variations, requirement of density difference and inability to handle particulates (Kiani *et al.*, 1984; Lo and Baird, 1980; Prasad *et al.*, 1986).

Increasing fluid velocity in both phases of heat- or mass-transfer devices will increase the convective heat- or mass-transfer coefficient as well as the transport performance. The concept of increasing fluid velocity was applied to parallel-flow membrane devices by arranging multipass flow for liquid-liquid extraction (Yeh and Huang, 1995; Yeh and Chen, 2000b). Recently, the performance improvement by multipass operation in a cocurrently cross-flow membrane extractor was reported (Yeh and Chen, 2001). Instead of the cocurrently cross-flow operation, it is the purpose of this study to investigate the effect of multipass arrangement on the performance in a countercurrently cross-flow membrane extractor of fixed configuration.

THEORY

The mathematical performance for the countercurrently cross-flow operation employed in present study is more cumbersome than that for the cocurrently cross-flow operation applied in previous work because in present case, the solvent does not enter at the front end ($y = 0$) but at the other end ($y = B$). The assumptions made in present study are steady state, no chemical reaction, uniform concentrations, $C_a(x)$ and $C_b(y)$, uniform velocities across the flow channels, constant mass-transfer coefficients for specified concentrations and velocities and constant distribution coefficients.

Single pass arrangement

Figure 1 may serve to illustrate the structure and fluid flows in a single-pass cross-flow membrane extractor. This system consists of two perpendicular channels for the fluids a and b , respectively, which are separated by a microporous membrane sheet through which solute is extracted and transferred to its exposed surfaces. In the case that when fluids a and b are miscible, the pores of the membrane are filled with another fluid (fluid c) which is immiscible with these two fluids. The solute is extracted from phase a to phase c and then to phase b , or vice versa (Lo and Baird, 1980).

The expression for mass-transfer rate was derived in previous work (Yeh and Chen, 2000a).

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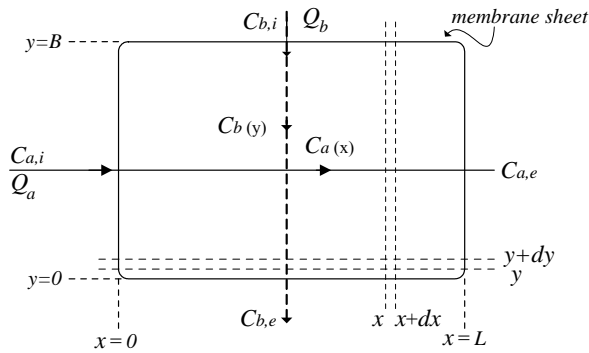


Fig. 1. Single-pass cross-flow flat-plate membrane extractor.

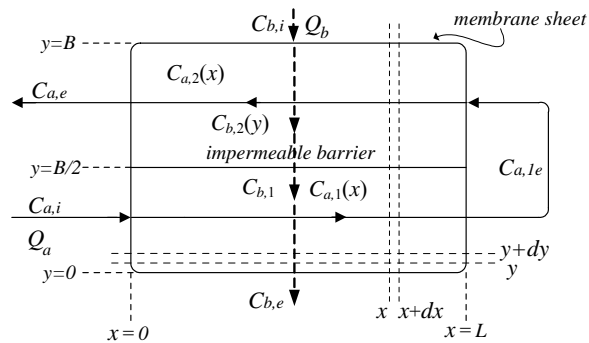


Fig. 2. Countercurrently cross-flow double-pass flat-plate membrane extractor.

$$W_1 = \frac{(Q_a/H_{ac})(H_{ac}C_{a,i} - H_{bc}C_{b,i})}{\left[1/(1-e^{-n})\right] + \left[\ell/(1-e^{-n\ell})\right] - (1/n)}, \quad (1)$$

where

$$\ell = \frac{Q_a H_{bc}}{Q_b H_{ac}}, \quad (2)$$

$$n = \frac{KBLH_{ac}}{Q_a}, \quad (3)$$

in which H_{ac} and H_{bc} are distribution coefficients of solute between two different phases, B and L are, respectively, the width and length of membrane sheet, Q_a and Q_b are the volumetric flow rates of phases a and b , while K denotes the overall mass-transfer coefficient. The rates of mass transfer predicted by Eq. (1) are in agreement with the experimental results (Yeh and Chen, 2000a).

Double-pass arrangement

Figure 2 shows the schematic diagram of a countercurrently cross-flow double-pass device which is similar to the one in Fig. 1, except that an impermeate plate with negligible thickness is placed perpendicularly in the flow phase a to divide the channel into two subchannels (subchannels a_1 , and a_2) of same size. During extraction operation, the raffinate fluid (fluid a) with volume rate Q_a and inlet

concentration $C_{a,i}$, first enters subchannel a_1 at $x = 0$. After leaving from subchannel a_1 at $x = L$, the raffinate fluid then enters subchannel a_2 at $x = L$ and finally exits from subchannel a_2 at $x = 0$. The extract fluid (fluid b) with volume rate Q_b and inlet concentration $C_{b,i}$ enters channel b at $y = B$, then flows steadily through channel b and finally exists at $y = 0$. The total mass-transfer area of membrane surfaces, $S = BL = 2(B/2)L$, are the same for both single and double passes, as shown in Figs. 1 and 2.

Since the impermeable plate divides the raffinate phase into two regions, mass balances will also be taken for each of the two flow regions. By taking the mass balances through a differential area $dx \cdot dy$ in flow region 1, one obtains

$$-[Q_a/(B/2)] dy dC_{a,1} = K dx dy (H_{ac}C_{a,1} - H_{bc}C_{b,1}), \quad (4)$$

$$[Q_a/(B/2)] dy dC_{a,1} = [(Q_b/L) dx] dC_{b,1}. \quad (5)$$

Similarly, for flow region 2

$$[Q_a/(B/2)] dy dC_{a,2} = K dx dy (H_{ac}C_{a,2} - H_{bc}C_{b,2}), \quad (6)$$

$$[Q_a/(B/2)] dy dC_{a,2} = -[(Q_b/L) dx] dC_{b,2}. \quad (7)$$

The boundary conditions for solving $C_{a,1}$, $C_{a,2}$, $C_{b,1}$, and $C_{b,2}$ from Eqs. (4)-(7) are

$$\text{at } x = 0, \quad C_{a,1} = C_{a,i}; \quad (8)$$

$$\text{at } x = 0, \quad C_{a,2} = C_{a,e}; \quad (9)$$

$$\text{at } x = L, \quad C_{a,1} = C_{a,1e}; \quad (10)$$

$$\text{at } x = L, \quad C_{a,2} = C_{a,1e}; \quad (11)$$

$$\text{at } y = 0, \quad C_{b,1} = C_{b,e}; \quad (12)$$

$$\text{at } y = B/2, \quad C_{b,1} = C_{b,2e}; \quad (13)$$

$$\text{at } y = B/2, \quad C_{b,2} = C_{b,2e}; \quad (14)$$

$$\text{at } y = B, \quad C_{b,2} = C_{b,i}. \quad (15)$$

With the use of appropriate boundary conditions, $C_{a,1}(x)$ and $C_{b,1}(y)$ are solved from Eqs. (4) and (5), while $C_{a,2}(x)$ and $C_{b,2}(y)$ are obtained from Eqs. (6) and (7). Inspection of Eqs. (8)-(15) shows that the outlet concentrations, $C_{a,1e}$, $C_{a,e}$, $C_{b,2e}$, and $C_{b,e}$ were not specified in prior. Mathematically more relations are needed for determination of these values. They are

$$Q_a(C_{a,i} - C_{a,1e}) = Q_b(C_{b,e} - C_{b,2e}), \quad (16)$$

$$Q_a(C_{a,1e} - C_{a,e}) = Q_b(C_{b,2e} - C_{b,i}). \quad (17)$$

Once the outlet concentration of raffinate phase, $C_{a,e}$, is solved from Eqs. (4)-(17), the rate of mass

transfer is readily obtained. The mathematical treatment, which is cumbersome and will not be presented here, was done by following the procedure of Bowman *et al.* (1949) and may be referred to the Appendix for multipass operation. The result is

$$W_2 = Q_a(C_{a,i} - C_{a,e}) = -\xi_{a,e} Q_a [C_{a,i} - (H_{bc}/H_{ac})C_{b,i}], \quad (18)$$

where

$$-\xi_{a,e} = \left\{ 1 - (1 + \ell \xi_{a,e}) \left[-\frac{2}{n} + \ell - \frac{1}{1 - e^{n/2}} - \frac{\ell}{1 - e^{n\ell/2}} \right] \right. \\ \times \left. \left[\frac{2}{n} + \frac{\ell}{1 - e^{n\ell/2}} - \frac{1}{1 - e^{-n/2}} \right]^{-1} \right\} \\ \times \left[1 + \ell - \frac{2}{n} - \frac{1}{1 - e^{n/2}} - \frac{\ell}{1 - e^{n\ell/2}} \right]^{-1}. \quad (19)$$

Multipass arrangement

Figure 3 illustrates the flows in a triple-pass countercurrently cross-flow membrane extractor with two impermeable barriers to divide the raffinate phase into three subchannels of same width, $B/3$. The effective surface areas of membrane sheet, $S = 3(B/3)L = 2(B/2)L = BL$, are the same in the triple-pass, double-pass and single-pass devices. The mathematical treatment of separation theory is also nearly the same as that in double-pass devices except that for triple-pass operation, six, instead of four, governing equations with more boundary conditions and overall mass balances are needed to describe the mass-transfer phenomena. They are

Governing equations:

$$- [Q_a/(B/3)] dy dC_{a,1} = K dx dy (H_{ac} C_{a,1} - H_{bc} C_{b,1}); \quad (20)$$

$$- [(Q_b/L) dx] dC_{b,1} = K dx dy (H_{ac} C_{a,1} - H_{bc} C_{b,1}); \quad (21)$$

$$[Q_a/(B/3)] dy dC_{a,2} = K dx dy (H_{ac} C_{a,2} - H_{bc} C_{b,2}); \quad (22)$$

$$- [(Q_b/L) dx] dC_{b,2} = K dx dy (H_{ac} C_{a,2} - H_{bc} C_{b,2}); \quad (23)$$

$$- [Q_a/(B/3)] dy dC_{a,3} = K dx dy (H_{ac} C_{a,3} - H_{bc} C_{b,3}); \quad (24)$$

$$- [(Q_b/L) dx] dC_{b,3} = K dx dy (H_{ac} C_{a,3} - H_{bc} C_{b,3}). \quad (25)$$

Boundary conditions:

$$x = 0, \quad C_{a,1} = C_{a,i}; \quad (26)$$

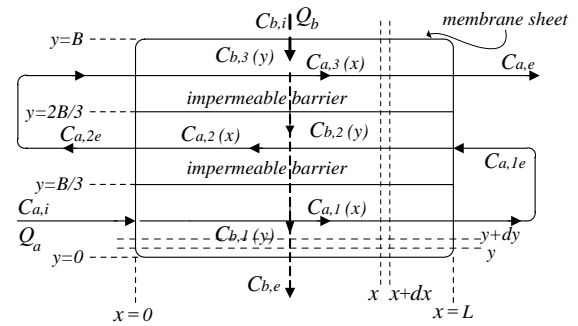


Fig. 3. Countercurrently cross-flow triple-pass flat-plate membrane extractor.

$$x = 0, \quad C_{a,2} = C_{a,2e}; \quad (27)$$

$$x = 0, \quad C_{a,3} = C_{a,2e}; \quad (28)$$

$$x = L, \quad C_{a,1} = C_{a,1e}; \quad (29)$$

$$x = L, \quad C_{a,2} = C_{a,1e}; \quad (30)$$

$$x = L, \quad C_{a,3} = C_{a,e}; \quad (31)$$

$$y = 0, \quad C_{b,1} = C_{b,e}; \quad (32)$$

$$y = B/3, \quad C_{b,1} = C_{b,2e}; \quad (33)$$

$$y = B/3, \quad C_{b,2} = C_{b,2e}; \quad (34)$$

$$y = 2B/3, \quad C_{b,2} = C_{b,3e}; \quad (35)$$

$$y = 2B/3, \quad C_{b,3} = C_{b,3e}; \quad (36)$$

$$y = B, \quad C_{b,3} = C_{b,i}. \quad (37)$$

Relation equations:

$$Q_a(C_{a,i} - C_{a,1e}) = Q_b(C_{b,e} - C_{b,2e}); \quad (38)$$

$$Q_a(C_{a,1e} - C_{a,2e}) = Q_b(C_{b,2e} - C_{b,3e}); \quad (39)$$

$$Q_a(C_{a,2e} - C_{a,e}) = Q_b(C_{b,3e} - C_{b,i}). \quad (40)$$

After solving $C_{a,1}(x)$ and $C_{b,1}(y)$, $C_{a,2}(x)$ and $C_{b,2}(y)$, and $C_{a,3}(x)$ and $C_{b,3}(y)$, respectively, from Eqs. (20) and (21), Eqs. (22) and (23), and Eqs. (24) and (25) coupled with the use of the appropriate boundary conditions listed in Eqs. (26)-(37), $C_{a,1e}$, $C_{a,2e}$, $C_{a,e}$, $C_{b,e}$, $C_{b,2e}$, and $C_{b,3e}$ are then determined with the use of Eqs. (38)-(40). The procedure for deriving the expression of mass-transfer rate though is not difficult and similar to that for a double-pass device but is more cumbersome, and is presented in the Appendix. The expression for calculating the total mass-transfer rate is.

$$W_3 = Q_a(C_{a,i} - C_{a,e}) = -\xi_{a,e} Q_a [C_{a,i} - (H_{bc}/H_{ac})C_{b,i}], \quad (41)$$

Table 1. Experimental results of double-pass cross-flow operations for water-acetic-methyl isobutyl ketone system with $Q_b = 0.125 \times 10^{-6} \text{ m}^3/\text{s}$ and $C_{b,i} = 0$.

$C_{a,i}$ (kg mol/m ³)	$Q_a \times 10^6$ (m ³ /s)	$W_1^a \times 10^8$ (kg mol/s)	Cocurrent Flow ^b		Countercurrent Flow		
			$W_2 \times 10^8$ (kg mol/s)	I_2 (%)	$C_{a,e}$ (kg mol/m ³)	$W_2 \times 10^8$ (kg mol/s)	I_2 (%)
0.496	0.184	1.9670	1.9688	0.09	0.3845	2.0516	4.30
	0.433	2.1044	2.1607	2.68	0.4437	2.2646	7.61
	0.681	2.2064	2.2609	2.47	0.4615	2.3495	6.49
	0.929	2.2575	2.3411	3.70	0.4702	2.3968	6.17
	1.178	2.3560	2.3796	1.00	0.4753	2.4385	3.50
	1.426	2.4527	2.4527	0	0.4787	2.4667	0.57
2.02	0.184	6.1401	6.1456	0.09	1.6718	6.4069	4.35
	0.433	6.5037	6.7461	3.73	1.8548	7.1532	9.99
	0.681	6.9326	7.0892	2.26	1.9098	7.5046	8.25
	0.929	7.3205	7.3391	0.25	1.9368	7.7293	5.58
	1.178	7.5274	7.5392	0.16	1.9530	7.8926	4.85
	1.426	7.6719	7.6719	0	1.9637	8.0284	4.65

^a Yeh and Chen (2000a).^b Yeh and Chen (2001).

where

$$-\xi_{a,e} = \left\{ \left[(1 - e^{-n/3})^{-1} - \ell (1 - e^{n\ell/3})^{-1} + \ell - 1 - \frac{3}{n} \right] \times (\xi_{a,2e} + \xi_{a,1e}) - 1 \right\} \times \left[\frac{3}{n} + \ell (1 - e^{n\ell/3})^{-1} + (1 - e^{-n/3})^{-1} - \ell \right]^{-1}; \quad (42)$$

$$\xi_{a,1e} = (1 + \ell \xi_{a,e}) \left[\frac{3}{n} + \frac{\ell}{1 - e^{n\ell/3}} - \frac{1}{1 - e^{-n/3}} \right]^{-1}; \quad (43)$$

$$\xi_{a,2e} = [(1 - \ell)\xi_{a,1e} + \ell \xi_{a,e} + 1] \times \left[\frac{3}{n} + \frac{\ell}{1 - e^{n\ell/3}} + \frac{1}{1 - e^{-n/3}} \right]^{-1}. \quad (44)$$

EXPERIMENTAL

The chemicals, materials, dimensions of apparatus and experimental procedure are exactly the same as those employed and performed in previous work (Yeh and Chen, 2000b), except that in present experimental work, instead of cocurrent-flow operation, two countercurrently cross-flow multipass membrane extractors were employed. Experiments were carried out with the use of a membrane sheet ($L = B = 0.165 \text{ m}$) made of microporous polypropylene (Gelman Sciences, average pore size of $0.2 \mu\text{m}$, porosity of 70% and thickness of $178 \mu\text{m}$) as a permeable barrier to extract acetic acid (reagent ACS grade, Fisher) from aqueous solution by methyl isobutyl ketone (MIBK, reagent grade, Fisher). The membrane

sheet was inserted in parallel between two parallel plates of stainless steel, with same distance from them to divide the conduit into two channels (channels a and b , or phases a and b) of same height ($h = 1.9 \times 10^{-3} \text{ m}$). For double-pass extraction, an impermeable plate of negligible thickness was placed additionally in vertical to the upper plate and the membrane sheet at the center line of channel a to divide it into two subchannels (subchannels a_1 , and a_2) of height h and width ($B/2$), while for triple-pass operation two impermeable plates were placed to divide channel a into three subchannels of same width ($B/3$) and height h , as shown in Figs. 2 and 3.

RESULTS AND DISCUSSION

Experimental results

Many experimental data were obtained under various operating conditions ($Q_b = 0.125 \times 10^{-6} \sim 1.0 \times 10^{-6} \text{ m}^3/\text{s}$, $C_{b,i} = 0$, $Q_a = 0.184 \times 10^{-6} \sim 1.426 \times 10^{-6} \text{ m}^3/\text{s}$, $C_{a,i} = 0.496$ and 2.02 kg mol/m^3). Some of outlet concentrations, $C_{a,e}$ are listed in Tables 1 and 2 for double-pass and triple-pass operations, respectively, while the corresponding values of mass transfer rates, W_2 and W_3 , were calculated from Eq. (45)

$$W_N = Q_a(C_{a,i} - C_{a,e}), \quad N = 2, 3, \quad (45)$$

with the use of Tables 1 and 2, and the results are also listed in the tables.

The total mass-transfer resistance ($1/K$) may be considered as the sum of the three resistances (in phase a , within membrane, and in phase b) in series. However, there are lack of some physical properties of the system, such as membrane characteristic, and

Table 2. Experimental results of triple-pass cross-flow operations for water-acetic-methyl isobutyl ketone system with $Q_b = 0.125 \times 10^{-6} \text{ m}^3/\text{s}$ and $C_{b,i} = 0$.

$C_{a,i}$ (kg mol/m ³)	$Q_a \times 10^6$ (m ³ /s)	$W_1^a \times 10^8$ (kg mol/s)	Cocurrent Flow ^b		Countercurrent Flow		
			$W_3 \times 10^8$ (kg mol/s)	I_3 (%)	$C_{a,e}$ (kg mol/m ³)	$W_3 \times 10^8$ (kg mol/s)	I_3 (%)
0.496	0.184	1.9670	2.0532	4.37	0.3745	2.2356	13.66
	0.433	2.1044	2.3858	13.37	0.4385	2.4897	18.31
	0.681	2.2064	2.5878	17.29	0.4569	2.6627	20.68
	0.929	2.2575	2.7034	19.75	0.4662	2.7684	22.63
	1.178	2.3560	2.7212	15.50	0.4725	2.7793	17.97
	1.426	2.4527	2.7664	12.79	0.4763	2.8092	14.54
2.02	0.184	6.1401	6.7123	9.32	1.6389	7.0122	14.20
	0.433	6.5037	7.9932	22.90	1.8311	8.1793	25.58
	0.681	6.9326	8.2333	18.76	1.8971	8.3694	20.73
	0.929	7.3205	8.4075	14.85	1.9283	8.5189	16.37
	1.178	7.5274	8.5758	13.93	1.9464	8.6700	15.18
	1.426	7.6719	8.6843	13.20	1.9586	8.7556	14.13

^a Yeh and Chen (2000a).

^b Yeh and Chen (2001).

we could not estimate K from the transport equations. Nevertheless, it was found in previous experimental results (Yeh and Chen, 2000b) that all values of overall mass-transfer coefficient K increase slightly with fluid velocities, v_a and v_b , within the velocity range of present interest, especially with v_b . Therefore, we assumed that K was independent of the velocity (v_b) in the organic phase. Accordingly, the following correlation equations were reached in previous work with the unit of v_a taken as cm/s.

$$K \times 10^4 = 7.256v_a^{0.104}, \text{ for } C_{a,i} = 0.496; \quad (46)$$

$$K \times 10^4 = 4.734v_a^{0.124}, \text{ for } C_{a,i} = 2.02. \quad (47)$$

The correlation predictions of mass-transfer rates, W_2 and W_3 , were calculated from Eqs. (18) and (41) with the use of Eqs. (47) and (48) as well as the following relations of v_a , v_b , ℓ , and n to Q_a and Q_b

$$v_a = \frac{Q_a}{h(B/N)} = \frac{Q_a \times 10^6}{(0.19)(16.5/N)}, \quad N = 2, 3, \quad (48)$$

$$v_b = \frac{Q_b}{hL} = \frac{Q_b \times 10^6}{(0.19)(16.5)}, \quad (49)$$

$$\ell = \frac{Q_a H_{bc}}{Q_b H_{ac}} = (Q_a/Q_b)(1/0.524), \quad (50)$$

$$n = \frac{KBLH_{ac}}{Q_a} = \frac{K(16.5)(16.5)(0.524)}{Q_a \times 10^6}, \quad (51)$$

where $H_{ac} = 0.524$ (Yeh and Huang, 1995) and $H_{bc} = 1$ for hydrophobic membrane used in present study. Some results of correlation prediction and experimental data for W_3 are plotted in Figs. 4 and 5 for comparison. Correlation qualitatively predictions

confirm with the experimental results. The reason why the experimental data are all larger than the theoretical values, may be due to the assumption that K was independent of v_b .

Effect of multipass operation on performance

The improvement of mass-transfer rate in membrane extractors with multipass operation is best illustrated by calculating the percentage increase in mass-transfer rate based on the performance of a single-pass device of same size

$$I_N = \frac{W_N - W_1}{W_1}, \quad N = 2, 3. \quad (52)$$

For the purpose of illustration, the experimental values of W_1 (Yeh and Chen, 2000a) as well as W_2 and W_3 obtained in previous work for cocurrent operations (Yeh and Chen, 2001) and those obtained in present study for countercurrent operations which are listed in Tables 1 and 2, were used to calculate I_2 and I_3 by Eq. (51). The results are also presented in the tables. It is seen from these tables that multipass operation improves the mass transfer rate. The enhancement increases with increasing pass number. It is also found that the improvements, I_2 and I_3 , increase with Q_a first and then decrease. This may be due to the fact that since the value of H_{ac} (0.524) in water-acetic acid-MIBK system is small and the mass-transfer resistance in phase a (multipass phase) is only scarcely predominant (Yeh and Huang, 1995; Yeh and Chen, 2000b). Accordingly, at rather higher inlet volume rate, Q_a , the fluid velocity is large enough even in single-pass devices, and though the effect of increasing fluid velocity on the performance

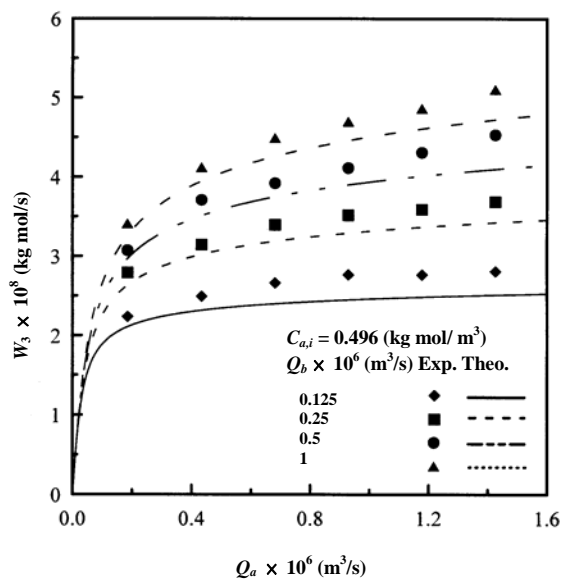


Fig. 4. W_3 vs. Q_a : $C_{a,i}=0.496$ kg mol/m³.

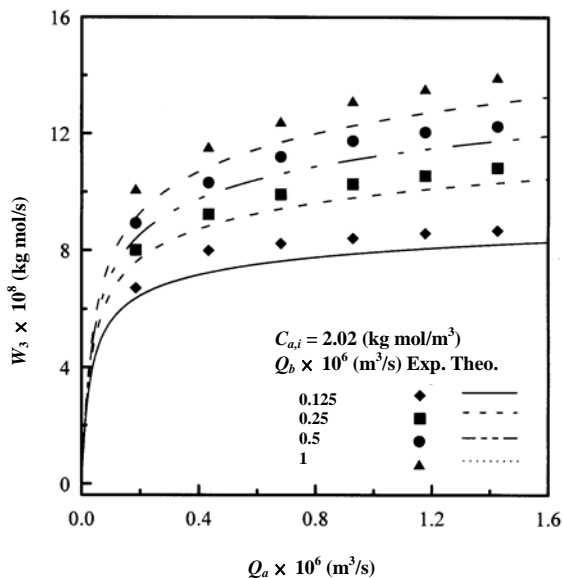


Fig. 5. W_3 vs. Q_a : $C_{a,i}=2.02$ kg mol/m³.

(W_N), as well as on K of water-acetic acid-MIBK system by applying multipass operation, is still favorable, the improvement (I_N) in performance based on a single-pass device of same size is declining because that W_1 increases with Q_a more rapidly than W_2 and W_3 do, as shown in the tables.

The power consumption in multipass cross-flow membrane extractors was already discussed in the previous work (Yeh and Chen, 2001), and the results are given in Table 3. Although the increase of power consumption due to multipass operation will rapidly increase with pass number, even the hydraulic dissipated powers of multipass operation are extremely small and thus, the power consumptions may be ignored.

Table 3. Power consumptions of multipass membrane extractors with $Q_b = 0.25 \times 10^{-6}$ m³/s (Yeh and Chen, 2001).

$Q_a \times 10^6$ (m ³ /s)	Power Consumption ($\times 10^8$ hp)		
	single pass	double pass	triple pass
0.184	0.048	0.196	0.452
0.433	0.268	1.096	2.522
0.681	0.663	2.712	6.239
0.929	1.234	5.047	11.612
1.178	1.984	8.115	18.669
1.426	2.908	11.894	27.364
1.600	3.661	21.497	34.450

Comparison of the performance between cocurrent and countercurrent operations

Tables 1 and 2 indicate that the performances of countercurrent-flow operation overcomes those of cocurrent-flow operation in both double-pass and triple-pass devices. This may be because that the mean concentration differences (driving force of mass transfer) between two phases in countercurrent-flow mass exchangers are larger than in cocurrent-flow one, like the same results in heat exchangers.

CONCLUSION

The performance of membrane extraction in countercurrently cross-flow multipass devices was investigated both theoretically and experimentally. The ordinary differential equations for solute concentration distributions were derived based on mass balances with the assumptions of uniform concentrations and velocities over the cross section of flow, and the concentration distributions as well as the outlet concentrations were then obtained by solving simultaneously the first-order equations. Once the outlet concentration is obtained, the overall mass-transfer rate and the improvement in separation may be calculated from the appropriate equations. The experimental works were carried out with a parallel conduit of stainless steel inserted with a membrane sheet made of microporous polypropylene to extract acetic acid from aqueous solution by methyl isobutyl ketone. Considerable improvement in performance is obtainable in present device if multipass, instead of single pass, is arranged. Further improvement is also achievable if countercurrent flow, instead of cocurrent flow, is operated.

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NOMENCLATURE

B	membrane width, m
$C_{a,1}, C_{a,2}, C_{a,3}$	solute concentration distribution in channel a_1 , in channel a_2 , in channel a_3 of raffinate phase, kg mol/m ³
$C_{a,1e}, C_{a,2e}, C_{a,e}$	solute concentration of channel a_1 at $x = L$, of channel a_2 at $x = 0$, and of channel a_3 at $x = L$, kg mol/m ³
$C_{a,e}, C_{b,e}$	outlet solute concentration in phase a , in phase b , kg mol/m ³
$C_{a,i}, C_{b,i}$	inlet solute concentration in phase a , in phase b , kg mol/m ³
$C_{a,jm}, C_{b,jm}$	mean values of $C_{a,j}$ and $C_{b,j}$, $j = 1, 2, 3$ kg mol/m ³
$C_{a,Nm}$	average concentration in channel a of N th pass, kg mol/m ³
$C_{b,1}, C_{b,2}, C_{b,3}$	solute concentration distribution within $0 \leq y \leq B/3$, $B/3 \leq y \leq 2B/3$, and $2B/3 \leq y \leq B$ of extract phase, kg mol/m ³
$C_{b,1e}, C_{b,2e}, C_{b,3e}$	solute concentration of phase b at $y=0$, $y=B/3$, and $y=2B/3$, kg mol/m ³
$C_{b,Nm}$	average concentration in channel b of N th pass, kg mol/m ³
D_a, D_b, D_c	solute diffusivity in phase a , phase b , and phase c , m ² /s
H_{ac}, H_{bc}	distribution coefficient between phases a and c and between phases b and c
h	height of flow channel, m
I_N	improvement of mass-transfer rate in N -pass device based on single-pass device, defined by Eq. (52)
K	overall mass transfer coefficient
L	length of membrane sheet, effect length of membrane extractor, m
ℓ, n	dimensionless group defined by Eq. (2) and Eq. (3)
N	pass number of flow in raffinate phase
Q_a, Q_b	volume flow rate in phase a , in phase b , m ³ /s
S	BL, total mass-transfer area of membrane sheet, m ²
v_a, v_b	fluid velocity in phase a defined by Eq. (48) and in phase b defined by Eq. (49), m/s
W_N	total mass-transfer rate in a N -pass membrane extractor, mol/s
x, y	rectangular coordinates, m

Greek symbols

$\xi_{a,e}, \xi_{a,1e}, \xi_{a,2e}$ dimensionless solute concentration,

defined by Eq. (A-7), Eq. (A-8), and Eq. (A-9)

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APPENDIX

Eqs. (42)-(44) are derived as follows:

Definitions

$$C_{a,1m} = \frac{1}{L} \int_0^L C_{a,1} dx, \quad (\text{A-1})$$

$$C_{a,2m} = \frac{1}{L} \int_0^L C_{a,2} dx, \quad (\text{A-2})$$



$$C_{a,3m} = \frac{1}{L} \int_0^L C_{a,3} dx, \quad (\text{A-3})$$

$$C_{b,1m} = \frac{3}{B} \int_0^{B/3} C_{b,1} dy, \quad (\text{A-4})$$

$$C_{b,2m} = \frac{3}{B} \int_{B/3}^{2B/3} C_{b,2} dy, \quad (\text{A-5})$$

$$C_{b,3m} = \frac{3}{B} \int_{2B/3}^B C_{b,3} dy, \quad (\text{A-6})$$

$$-\xi_{a,e} = \frac{H_{ac}(C_{a,i} - C_{a,e})}{H_{ac}C_{a,i} - H_{bc}C_{b,i}}, \quad (\text{A-7})$$

$$-\xi_{a,1e} = \frac{H_{ac}(C_{b,e} - C_{a,1e})}{H_{ac}C_{a,i} - H_{bc}C_{b,i}}, \quad (\text{A-8})$$

$$-\xi_{a,2e} = \frac{H_{ac}(C_{a,1e} - C_{a,2e})}{H_{bc}C_{a,i} - H_{bc}C_{b,i}}, \quad (\text{A-9})$$

$$-\ell\xi_{a,e} = \frac{H_{bc}(C_{b,e} - C_{b,i})}{H_{ac}C_{a,i} - H_{bc}C_{b,i}}, \quad (\text{A-10})$$

$$-\ell\xi_{a,1e} = \frac{H_{bc}(C_{b,e} - C_{b,2e})}{H_{ac}C_{a,i} - H_{bc}C_{b,i}}, \quad (\text{A-11})$$

$$-\ell\xi_{a,2e} = \frac{H_{bc}(C_{b,2e} - C_{b,3e})}{H_{ac}C_{a,i} - H_{bc}C_{b,i}}, \quad (\text{A-12})$$

Average concentrations

Integrating Eqs. (20), (22), and (24) from $y = 0$ to $y = B/3$, from $y = B/3$ to $2B/3$, and from $y = 2B/3$ to B , respectively, one has

$$-(3Q_a/B) dC_{a,1} = K(H_{ac}C_{a,1} - H_{bc}C_{b,1m}) dx; \quad (\text{A-13})$$

$$(3Q_a/B) dC_{a,2} = K(H_{ac}C_{a,2} - H_{bc}C_{b,2m}) dx; \quad (\text{A-14})$$

$$-(3Q_a/B) dC_{a,3} = K(H_{ac}C_{a,3} - H_{bc}C_{b,3m}) dx. \quad (\text{A-15})$$

Integrating of Eqs. (21), (23), and (25) from $x = 0$ to $x = L$ yields

$$-(Q_b/L) dC_{b,1} = K(H_{ac}C_{a,1m} - H_{bc}C_{b,1}) dy; \quad (\text{A-16})$$

$$-(Q_b/L) dC_{b,2} = K(H_{ac}C_{a,2m} - H_{bc}C_{b,2}) dy; \quad (\text{A-17})$$

$$-(Q_b/L) dC_{b,3} = K(H_{ac}C_{a,3m} - H_{bc}C_{b,3}) dy. \quad (\text{A-18})$$

Integrating Eqs. (A-13), (A-14), and (A-15) from $x = 0$ to $x = L$, one obtains

$$H_{bc}C_{b,1m} = H_{ac}C_{a,i} + H_{ac}(C_{a,1e} - C_{a,i}) / (1 - e^{-n/3}); \quad (\text{A-19})$$

$$H_{bc}C_{b,2m} = H_{ac}C_{a,2e} + H_{ac}(C_{a,1e} - C_{a,2e}) / (1 - e^{-n/3}); \quad (\text{A-20})$$

$$H_{bc}C_{b,3m} = H_{ac}C_{a,2e} + H_{ac}(C_{a,e} - C_{a,2e}) / (1 - e^{-n/3}). \quad (\text{A-21})$$

Integrating Eqs. (A-16), (A-17), and (A-18) from $y = 0$ to $y = B/3$, from $y = B/3$ to $2B/3$ and from $y = 2B/3$ to $y = B$, respectively, results in

$$H_{ac}C_{a,1m} = H_{bc}C_{b,e} + H_{bc}(C_{b,2e} - C_{b,e}) / (1 - e^{-n/3}); \quad (\text{A-22})$$

$$H_{ac}C_{a,2m} = H_{bc}C_{b,2e} + H_{bc}(C_{b,3e} - C_{b,2e}) / (1 - e^{-n/3}); \quad (\text{A-23})$$

$$H_{ac}C_{a,3m} = H_{bc}C_{b,3e} + H_{bc}(C_{b,i} - C_{b,3e}) / (1 - e^{-n/3}). \quad (\text{A-24})$$

Average concentration difference

Integrating Eq. (A-13) from $x = 0$ to $x = L$, we have

$$H_{ac}C_{a,1} - H_{bc}C_{b,1m} = (H_{ac}C_{a,i} - H_{bc}C_{b,1m})e^{-(nx/3L)}. \quad (\text{A-25})$$

Integrating again from $x = 0$ to $x = L$, we obtain

$$H_{ac}C_{a,1m} - H_{bc}C_{b,1m} = (3/n)(H_{ac}C_{a,i} - H_{bc}C_{b,1m})(1 - e^{-n/3}). \quad (\text{A-26})$$

Substitution of Eq. (A-19) into Eq. (A-26) to eliminate $(H_{ac}C_{a,i} - H_{bc}C_{b,1m})$ yields

$$H_{ac}C_{a,1m} - H_{bc}C_{b,1m} = (3/n)H_{ac}(C_{a,i} - C_{a,1e}). \quad (\text{A-27})$$

Actually, Eq. (A-27) also represents the mass balance for the first pass and this can be more evident if we rearrange it with the use of the definition of n in Eq. (3), i.e.

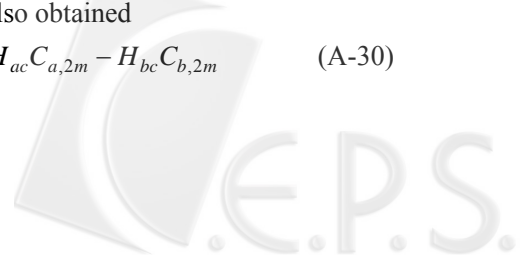
$$Q_a(C_{a,i} - C_{a,1e}) = KL(B/3)(\Delta C)_{1m}, \quad (\text{A-28})$$

where, $(\Delta C)_{1m}$ denote the average concentration difference for the first pass,

$$(\Delta C)_{1m} = H_{ac}C_{a,1m} - H_{bc}C_{b,1m}. \quad (\text{A-29})$$

Accordingly, by referring to Eq. (A-27) to take mass balances for second and third passes, the following expressions are also obtained

$$(\Delta C)_{2m} = H_{ac}C_{a,2m} - H_{bc}C_{b,2m} \quad (\text{A-30})$$



$$= (3/n)H_{ac}(C_{a,1e} - C_{a,2e}), \quad (A-31)$$

$$(\Delta C)_{3m} = H_{ac}C_{a,3m} - H_{bc}C_{b,3m} \quad (A-32)$$

$$= (3/n)H_{ac}(C_{a,2e} - C_{a,e}) . \quad (A-33)$$

vided by $(H_{ac}C_{a,i} - H_{bc}C_{b,i})$ with the use of Eqs. (A-7)-(A-14), one obtains

$$(-3/n)\xi_{a,1e} = \frac{e^{\xi_{a,1e}}}{1 - e^{(n\ell/3)}} - \frac{\xi_{a,1e}}{1 - e^{-n/3}} - 1. \quad (A-34)$$

After rearrangement, Eq. (A-34) becomes Eq. (43). Similarly, substituting Eqs. (A-20), (A-23), (A-21) and (A-24) into Eqs. (A-31) and (A-33) to eliminate $C_{a,2m}$, $C_{b,2m}$, $C_{a,3m}$, and $C_{b,3m}$, respectively, and then rearranging, one has Eqs. (44) and (42).

Results

Substituting Eqs. (A-19) and (A-22) into Eq. (A-27) to eliminate $C_{a,1m}$ and $C_{b,1m}$, and then di-

多行程裝置對逆交流平板薄膜模組中萃取效率之影響

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摘 要

本文以理論與實驗方法，探討逆交流平板薄膜模組中多行程操作對萃取效率之影響，結果發現：若阻力非在薄膜內時，在流體相中設置多行程裝置，其效率遠比大小相同的單行程薄膜萃取器高，而逆交流操作又比順交流者為佳。

