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Mass Transfer and Pressure Drop in a Cocurrent Reciprocating Plate Extraction Column

A 5-cm-diameter reciprocating plate extraction column has been operated in cocurrent flow. The pressure drop for water flow, and the local and average mass transfer products for the system acetic acid/kerosene/aqueous sodium hydroxide have been measured. Under well-agitated conditions, the mass average transfer product is predictable by a model, which is also applicable to data reported earlier by Karr for a 2.54-cm-diameter column.

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SCOPE

Countercurrent liquid-liquid extraction may be performed in many types of column, among which the reciprocating perforated plate column developed by Karr (1959) is well known. Recently it was suggested (Karr, 1979) that the performance of the Karr column in cocurrent flow should be investigated. Although cocurrent extraction provides only one equilibrium stage, the throughput is not limited by flooding as in the case of

countercurrent flow. The lack of a flooding problem also permits the use of intense agitation with correspondingly high mass transfer rates, unless the excessive agitation leads to unduly slow settling of the phases.

The general objective of this study is to examine the characteristics of a 5-cm-diameter reciprocating plate extraction column in cocurrent flow. Because the flows through the column

are quite high, the pressure drop characteristics have been measured with water flowing at velocities up to 26 cm/s. The second part of the study has involved the measurement of mass transfer. The solute (acetic acid) is transferred from the dispersed phase (kerosene) to the aqueous continuous phase. Mass transfer was measured by the "indicator color change method" which removes the need for sampling and analysis of the column exit streams. Instead, a dilute solution of sodium hydroxide with a trace of phenolphthalein is fed as the aqueous phase. The location in the plate stack at which the red color of the alkaline phenolphthalein disappears corresponds to a known amount

of interfacial transfer of acetic acid. From this measurement it is possible to calculate the average and local mass transfer products.

A model has been developed to predict the average mass transfer product in the column from the system properties, frequency, amplitude, plate spacing, etc. The model assumes that drops are formed in the plate stack by an isotropic turbulence breakup mechanism; mass transfer coefficients are estimated from literature correlations for either (a) stagnant drops or (b) circulating drops.

CONCLUSIONS AND SIGNIFICANCE

The pressure drop in single-phase flow through the 1.46-m plate stack has been found to vary approximately with the square of velocity in the absence of plate motion. The reciprocation of the plates results in significant increases in the time-averaged pressure drop. These effects have been compared with a quasisteady-state model in which it is assumed that the frictional pressure drop at any moment corresponds to fully developed flow with an orifice coefficient of 0.60.

The mass transfer product (K_0a) for extraction of acetic acid from kerosene has been found to vary locally within the plate stack, tending to increase as the dispersion passes through it. The average value K_0a has been measured for a variety of conditions, viz., flow rates of each phase, plate reciprocation frequency, and plate spacing. The results have been compared

with a predictive model based on average energy dissipation rates, and agree reasonably well under well-agitated conditions, at frequencies in excess of 1.5 Hz. At low reciprocation frequencies, the model underpredicts the mass transfer performance. The model also agrees reasonably with earlier data of Karr (1983) obtained for the system acetic acid/o-xylene/water in a 2.54-cm column.

It is concluded that the isotropic turbulence model developed here is a useful interim approach to design for cocurrent flow. However, the observation that the local mass transfer products are varying within the plate stack points to the need for a more sophisticated model which could take account of the progressive breakup of the droplets as they pass through the column.

INTRODUCTION

The reciprocating plate column developed by Karr (1959) has been used extensively for countercurrent liquid-liquid extraction. The operating characteristics, design procedures and industrial applications of such columns have recently been reviewed by Karr (1980) and by Lo and Prochazka (1983).

In common with all other types of countercurrent extraction column, the Karr column cannot accept throughputs in excess of the limits dictated by the flooding condition. The flooding correlation for the Karr column (Baird et al., 1971) indicates that higher specific throughputs are possible than in most other column extractors; nevertheless, for most systems it is seldom possible for the superficial velocities of each phase to exceed 1.0 cm/s.

It has recently been suggested (Karr, 1983) that the possibilities of operating a reciprocating plate column in cocurrent flow should be explored. Karr (1983) operated a 2.54-cm column with the system water/acetic acid/o-xylene at superficial velocities up to 52.5 cm/s and obtained stage efficiencies which approached 100% at high frequencies and strokes of agitation. The main advantage of cocurrent over countercurrent flow is that the throughput can be greatly increased. Although of course only one equivalent stage at the most is available in a cocurrent column, this is sufficient for systems where the extraction factor is very high, or where extraction is accompanied by an irreversible chemical reaction.

The present work has been carried out using a 5-cm column in cocurrent flow. The objectives have been to obtain additional mass transfer data and to develop a model for the design of cocurrent reciprocating columns, based as far as possible upon first principles. A subsidiary objective has been to obtain data and a method of prediction for the frictional pressure drop when fluid flows through a stack of reciprocating perforated plates.

PRESSURE DROP

The absence of a flooding limitation in cocurrent flow allows the fluid velocities through the column to be greatly increased. Fric-

tional pressure drop, negligibly small under countercurrent conditions, becomes an important design factor for the pump sizes and hence for the capital and operating costs of cocurrent extraction.

The pressure drop measurements have been carried out using a single phase, namely water. It is anticipated that the pressure drops in two-phase emulsion flow would differ from the single-phase values only to the extent that the dispersion density differs from that of the single-phase fluid.

The experimental arrangement (Figure 1) featured a static manometer tube arrangement for pressure drop measurement. Tap water at about 10°C was supplied through a flow control valve and a calibrated orifice flowmeter, entering the column near its base as shown. It passed upwards through a calming section of about 30 cm before entering the plate stack. This consisted of 1.46 m of perforated stainless-steel plates with 61% open area and perforation diameter 13.7 mm. The spacing between plates was 5.08 cm and the number of plates in the stack was 29.

Reciprocation was provided by an electric motor (0.56 kW) with a variable speed drive (0–400 rpm) coupled to an adjustable yoke by which the oscillation stroke could be set at values up to 4.5 cm.

The water leaves the top of the column by a simple overflow pipe as shown.

The time-average frictional pressure drop was measured in cm of water as the difference between the liquid level in the manometer tube (connected to the column beneath the plate stack) and the liquid level at the overflow.

Results in the Absence of Reciprocation

The flow through each stationary plate is comparable to flow through an orifice, although each plate in fact contains ten apertures; five complete 13.7-mm holes and five segments of holes between the plate and the column wall (Noh, 1981). The corresponding equation for frictional pressure drop in the absence of plate motion is:

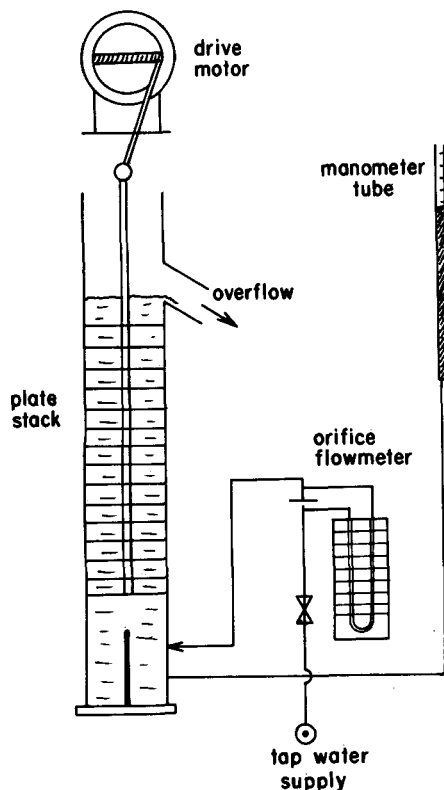


Figure 1. Schematic diagram of apparatus for measuring pressure drop.

$$\Delta P_f = \frac{n \rho_c u_c^2 (1 - \sigma^2)}{2 C_o^2 \sigma^2} \quad (1)$$

It is assumed here that no recovery of pressure occurs downstream of the plate (Jealous and Johnson, 1955).

The effect of superficial velocity upon measured pressure drop is shown in Figure 2 in comparison with a line representing Eq. 1 with the effective orifice coefficient $C_o = 0.60$. It was found that C_o varied between 0.42 at the lowest flow velocity and 0.75 at the highest. The Reynolds numbers (referred to hole diameter) ranged from 850 to 8,500. This variation of C_o with Re did not correspond with the effects given for standard single orifices (Perry, 1973) but this is not surprising as the perforated plate geometry is very different from that of a standard orifice.

Effect of Plate Reciprocation on Pressure Drop

The instantaneous displacement of each plate from its midpoint position is given by:

$$y = A' \sin \omega t \quad (2)$$

The amplitude A' is one half of the stroke, A . The plate velocity at time t is thus given by

$$\dot{y} = A' \omega \cos \omega t \quad (3)$$

The instantaneous velocity u_h of the water in the holes is estimated by a continuity equation.

$$\dot{y}(1 - \sigma) + u_h \sigma = u_c \quad (4)$$

This equation neglects to allow for "contraction" effects; these effects are included in the factor C_o which is added below.

The instantaneous pressure drop will include not only frictional effects but inertial effects because the flow is accelerating backwards and forwards through the holes. The time-averaged value of the inertial components is zero because the flow is not accelerating between one cycle and the next; only the frictional component of pressure drop will have a finite time-averaged value. By analogy with Eq. 1, we may write:

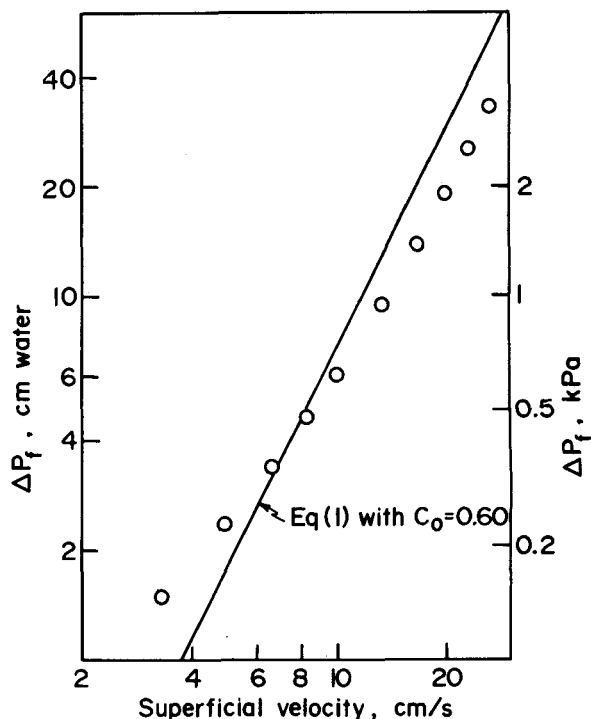


Figure 2. Effect of water velocity on pressure drop in the absence of agitation.

$$\Delta P_f = \frac{n \rho_c (u_h |u_h| - u_c^2)}{2 C_o^2} \quad (5)$$

The value of u_h is obtainable from Eq. 4 and the modulus term is included to permit a change of sign of the pressure drop if u_h is negative and its absolute value exceeds that of u_c . The orifice coefficient C_o in Eq. 5 represents an average effective value and is assumed to be independent of time.

From Eq. 4 it follows that

$$u_h = u_c (1 - V \cos \omega t) / \sigma \quad (6)$$

where

$$V = A' \omega (1 - \sigma) / u_c \quad (7)$$

If the parameter V (the maximum dimensionless velocity attained) is less than unity, the velocity u_h is always positive; at values of V greater than unity, the velocity u_h becomes negative for part of the cycle. Substitution of Eq. 6 in Eq. 5 gives

$$\Delta P_f = \frac{n \rho_c u_c^2}{2 C_o^2} \left[\frac{(1 - V \cos \omega t) |1 - V \cos \omega t|}{\sigma^2} - 1 \right] \quad (8)$$

The time-averaged pressure drop is obtained assuming C_o to be independent of time:

$$\overline{\Delta P_f} = \frac{n \rho_c u_c^2}{2 C_o^2} \left[\frac{(1 - V \cos \omega t) |1 - V \cos \omega t|}{\sigma^2} - 1 \right] \quad (9)$$

The above equation reduces to Eq. 1 as $V \rightarrow 0$. For small oscillations such that $V < 1$, the term $1 - V \cos \omega t$ is always positive and we can let

$$(1 - V \cos \omega t) |1 - V \cos \omega t| = (1 - V \cos \omega t)^2 \quad (10)$$

This can be shown to have a time-averaged value of $1 + V^2/2$. Substitution in the general Eq. 9 gives

$$\overline{\Delta P_f} = \frac{n \rho_c u_c^2}{2 C_o^2} \left[\left(\frac{1 - \sigma^2}{\sigma^2} \right) + \frac{V^2}{2 \sigma^2} \right] \quad (11)$$

(special case $V < 1$)

Thus a plot of $\overline{\Delta P_f}$ vs. the square of frequency should be linear provided there is no flow reversal in the holes.

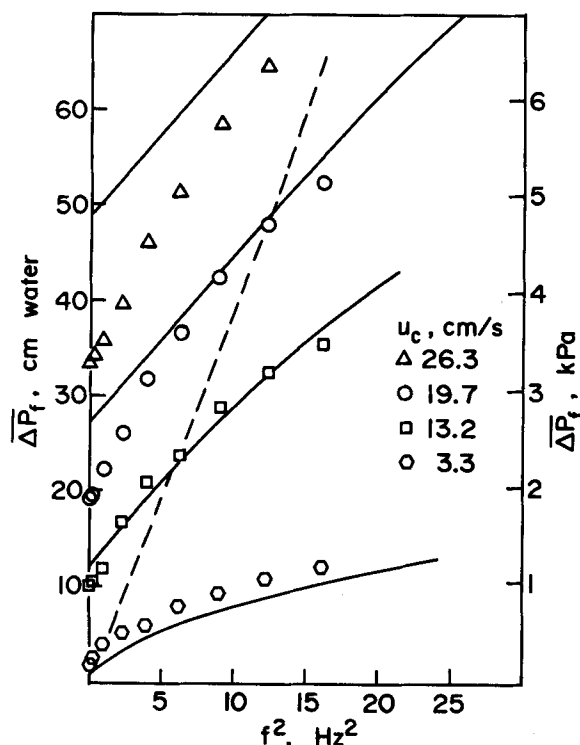


Figure 3. Effect of frequency on time-averaged pressure drop. — Eq. 9 assuming $C_o = 0.60$; - - - frequencies below which $V < 1$.

A second special case of Eq. 9 is that of $V \gg 1$; i.e., the oscillations are very large compared to the steady flow component. In this case it can be shown (Noh, 1981) that

$$\overline{\Delta P_f} = \frac{n \rho_c u_o^2}{2 C_o^2} \left[\left(\frac{4V}{\pi \sigma^2} \right) - 1 \right] \quad (12)$$

It was found that at high throughputs the column conditions generally corresponded to the case of $V < 1$, so the experimental data are presented (Figure 3) as a plot of ΔP_f vs. $(\text{frequency})^2$ as suggested by Eq. 11. The solid lines on the plot are general solutions (Noh, 1981) of Eq. 9 in which the orifice coefficient C_o has been taken to be 0.60. The dotted line in Figure 3 denotes the right hand boundary of the regime in which $V < 1$; i.e., Eq. 11 is applicable.

As seen already from Figure 2, the assumption $C_o = 0.60$ leads to an overestimate of ΔP_f at high throughputs and an underestimation at low throughputs in the absence of pulsation; this is also apparent from the predicted and experimental intercepts at $f = 0$ in Figure 3.

The effect of reciprocating motion is to increase the average Reynolds number for flow through the plate holes, leading to a more uniform value of C_o . Figure 3 indicates that at higher frequencies the choice of $C_o = 0.60$ gives reasonable predictions of the time-averaged pressure drop by means of Eq. 9.

From the practical designers' point of view, this section of the paper has shown that plate oscillation can increase the average pressure drop by a factor of 2 or more, compared with calculations based on flow through the stationary plate stack.

MASS TRANSFER

Preliminary experiments were carried out (Noh, 1981) on the extraction of acetic acid from kerosene (dispersed) by water in the cocurrently operated Karr column. The mass transfer performance was measured by sampling, separating and analyzing the phases leaving the column. The distribution coefficient m of acetic acid, defined as c_o/c_a , was found to be 0.0045. This was in close agreement with the data of Gordon and Reid (1922). Because of the low

value of m and because of the high stage efficiency of the column, the concentration of acid remaining in the exit kerosene phase was found to be extremely low. The uncertainties in analysis of such low concentrations led to uncertainties in estimating the mass transfer performance; these uncertainties may be compounded by the possible effects of interphase mass transfer during the phase separation of the samples.

It was, therefore, felt desirable to obtain a measure of mass transfer at intermediate points within the plate stack, without the need for sampling. This can be done by a modification of the indicator color change technique which was originally developed by Baird (1974) for measurement of axial mixing.

INDICATOR COLOR CHANGE METHOD

The principle of the method is that a fast irreversible reaction is allowed to occur between two components in the aqueous phase, typically an acid and a base. The zone at which the reaction occurs, and at which the concentrations of reactants are zero, may be seen if a trace of a suitable color-change indicator (e.g., phenolphthalein) is present. Because the reaction is very fast, its rate and hence the position of the color change are determined by mass transfer effects or hydrodynamic factors.

Consider a cocurrently operated contactor in which an acid transfers from the organic phase to the aqueous phase containing a free base. The acid concentration in the aqueous phase is maintained at negligibly small value because it undergoes a rapid reaction with a base. The base has negligible solubility in the organic phase compared to that in the aqueous phase. In the present work, the total throughputs ($u_o + u_a$) were at least 4 cm/s, and the axial dispersion coefficients can be estimated (Kim and Baird, 1976) to be in the order of $8 \text{ cm}^2/\text{s}$. Thus, the longitudinal Peclet number based on plate stack height is at least 75 and usually over 100 (Noh, 1981), so plug flow has been assumed. Thus the material balance for acid in the organic phase leads to a simplified differential equation:

$$u_o \frac{dc_o}{dz} + K_o a c_o = 0 \quad (13)$$

The overall mass transfer coefficient K_o is approximately equal to the organic film mass transfer coefficient if the acid is relatively very soluble in the aqueous phase and undergoes a very rapid reaction with base in that phase.

Integration of Eq. 13, with a boundary condition $c_o = c_{o1}$ at the column entry ($z = 0$) gives

$$c_o = c_{o1} \exp \left(- \frac{1}{u_o} \int_0^z K_o a dz \right) \quad (14)$$

This may also be written:

$$c_o = c_{o1} \exp(-NTU) \quad (14a)$$

It is assumed that one mole of nontransferring base reacts in the aqueous phase with each mole of acid transferred. If the base concentration is denoted by c_{ba} , the above stoichiometry dictates that

$$u_o(c_{o1} - c_o) = u_a(c_{ba1} - c_{ba}) \quad (15)$$

At the point in the column ($z = z_n$) at which the aqueous phase becomes neutral, it can be assumed that $c_{ba} = 0$ and it follows from Eq. 15 that

$$\frac{c_o}{c_{o1}} = 1 - \phi \quad (16)$$

where $\phi = (u_a c_{ba1}) / (u_o c_{o1})$

The factor ϕ is simply the ratio of moles of base fed in the aqueous phase to moles of acid fed in the organic phase; obviously ϕ must be equal to or less than 1.0 if neutrality is to be achieved within the column.

Substituting Eq. 16 in Eq. 14 leads to

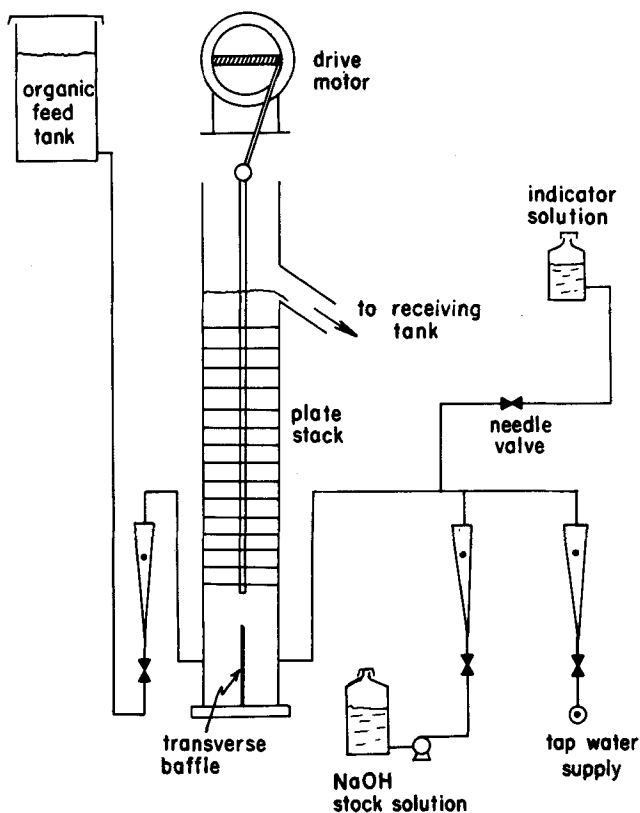


Figure 4. Schematic diagram of apparatus for measuring mass transfer product.

$$\frac{1}{u_o} \int_0^{z_n} K_o a dz = -\ln(1 - \phi) \quad (17)$$

If $K_o a$ is constant, then

$$z_n = -\frac{u_o}{K_o a} \ln(1 - \phi) \quad (18)$$

and

$$K_o a = -\frac{u_o}{z_n} \ln(1 - \phi) \quad (18a)$$

The organic phase superficial velocity u_o , and the factor ϕ , can be adjusted independently. The neutralization zone height z_n is measurable. Hence, Eq. 18a provides for calculation of the average mass transfer product from experimental data. In the event that $K_o a$ is a function of z , its local value can be found by measuring the effect of the stoichiometry ratio ϕ upon z_n . Differentiation of Eq. 17 at z_n gives

$$(K_o a)_{z_n} = -u_o \frac{d}{dz_n} [\ln(1 - \phi)] \quad (19)$$

APPARATUS AND PROCEDURES

A simplified schematic diagram of the apparatus is given in Figure 4.

The organic feed was a solution of acetic acid in kerosene, with a concentration typically in the range 0.10 to 0.25 mol/L. It passed from an overhead reservoir (180L) via a flow control valve and a calibrated rotameter, to enter near the base of the column as shown.

The aqueous feed was a dilute solution of sodium hydroxide in water with a trace of phenolphthalein indicator. The solution was continuously prepared from metered flows of tap water and a stock solution (1.0 mol/L) of sodium hydroxide. The water flow greatly

TABLE 1. EXPERIMENTAL CONDITIONS*

Independent Variable: Caustic Stock Solution Feed Rate				
Series	u_A (cm/s)	u_o (cm/s)	f (Hz)	h (cm)
1	4.1-4.5	4.0	3.0	5.08
2	4.1-4.4	2.0	3.0	5.08
3	4.1-4.2	1.0	3.0	5.08
4	10.2-10.5	7.0	3.0	5.08
10	4.1-4.3	5.55	0	5.08
12	4.2-4.5	6.0	3.0	5.08
17	2.2-2.3	2.0	3.0	2.54
18	4.1-4.3	4.0	3.0	2.54
Independent Variable: Organic-Phase Velocity u_o				
Series	u_A (cm/s)	f (Hz)	h (cm)	
5	4.2	3.0	5.08	
9	5.5	0	5.08	
13	4.2	5.0	2.54	
14	4.1	1.0	2.54	
20	4.2	0	2.54	
Independent Variable: Frequency of Reciprocation, f				
Series	u_A (cm/s)	u_o (cm/s)	h (cm)	
7	4.3	3.0	5.08	
8	0.81	0.82	5.08	
15	4.1	2.3	2.54	
16	4.2	3.85	2.54	

* Reciprocation stroke A was constant in all tests, at 4.5 cm.

exceeded that of the stock solution with the dilution ratio typically in the order of 40. The flow of phenolphthalein solution was negligibly small in comparison with the flows of water and stock solution; a needle valve was used to regulate it so that the entering aqueous phase had a strong red color. Typically the indicator flow rate was 0.2 mL/s.

It is undesirable that the acid organic phase and the alkaline aqueous phase should mix intensely before entering the plate stack. Such mixing would lead to a large mass transfer "entry effect." Consequently, a transverse baffle was installed to minimize direct contact between the entering streams until they were about 20 cm beneath the plate stack. Under these conditions it was seen that the droplets remained very large until they entered the plate stack. The dispersion left the top of the plate stack by an overflow mechanism as shown.

The procedure was to start the column running with water only. The reciprocating drive was then adjusted to the desired frequency, and the caustic stock solution flow was started. The two flows were adjusted to give the desired values of u_A and C_{ab1} , and the indicator flow was started. Then the organic-phase flow was started at the desired value of u_o , with the concentration of acid c_{o1} being known from the analysis of the feed reservoir contents.

As soon as the acidic kerosene flow was started, the red color of the aqueous phase began to disappear at the upper end of the column. The color boundary generally reached a steady position within 1 minute, and this was then noted as the distance z_n from the foot of the plate stack. The observation of the color boundary was facilitated by the use of floodlight illumination and a white screen background. The precision of observation was about ± 2 cm.

Then without changing the organic flow rate or the tap water flow rate, the flows of caustic stock solution were adjusted so that the entering caustic concentration C_{ab1} was changed. Because of the high dilution ratio, the aqueous-phase flow rate remained almost constant. This allowed the ratio ϕ (Eq. 16) to be varied almost independently; the value of z_n at the new value of ϕ was noted as soon as the new steady state had been reached. The conditions for these tests are given in Table 1 (series 1-4, 10, 12, 17, 18).

Other series of tests were carried out in which the organic-phase velocity u_o was varied (series 5, 9, 13, 14 and 20) or the aqueous-phase velocity was varied (series). The agitation frequency f was

TABLE 2. SYSTEM PROPERTIES AT 10°C

Densities, kg·m ⁻³	
Tap Water	998
Kerosene	798
Viscosities, mPa·s	
Tap Water	1.01 ^a
Kerosene	3.0 ^a
Interfacial Tension, mN·m ⁻¹	
Tap Water/Kerosene	34.0 ^b
Molecular Diffusivity, m ² ·s ⁻¹	
Acetic Acid/Tridecane	0.99 × 10 ⁻⁹ ^c

^a From Perry and Chilton (1973).^b Measured by ring tensiometer.^c From Wilke and Chang (1955); tridecane is taken to be the pure substance whose properties approximate to those of kerosene.

also investigated as an independent variable (series 7, 8, 15, 16).

The plate spacing in the stack was either 5.08 or 2.54 cm, Table 1. The reciprocation stroke was not studied as a variable and was held constant at 4.5 cm. Experience has shown (Karr, 1980) that the stroke and frequency, as related through the plate velocity, have the same effect upon column performance.

The experiments were carried out at 10 ± 1°C. Relevant system properties are listed in Table 2.

Local Variations in Mass Transfer Product

When the caustic solution feed rate is varied with other parameters held constant, the effect of ϕ upon z_n is effectively obtained. Data from some of these tests are plotted in Figure 5 in the form suggested by Eq. 18. It will be seen that the slope of the plots is not constant; in general the curvature indicates that $K_o a$ is increasing as the dispersion moves up through the plate stack. It is also clear that the mass transfer effectiveness increases as the superficial velocities u_o and u_a are reduced. In the absence of plate agitation (test 10) the mass transfer is very poor, with neutralisation being correspondingly delayed.

Noh (1981) used a cubic-spline subroutine to analyse the slopes of the curves; the local mass transfer products $K_o a$ could thus be calculated by means of Eq. 19. Typical results are shown in Figure 6. While the data from test 2 show that $K_o a$ increases with z_n , the data in the absence of agitation (test 10) show an initial decline followed by a leveling-off. In the former case, it would be expected that the interfacial area a is increasing gradually as the dispersion rises through the plate stack. The apparent decline in $K_o a$ in the absence of agitation could be due to the relative importance in this case of the "entry effect," i.e., the existence of finite mass transfer beneath the plate stack ($z < 0$). The local $K_o a$ may also be affected slightly by the caustic concentration although this did not exceed 0.05 mol/L. Another possible factor is short-range mixing, although as noted earlier, the plate stack Peclet numbers were high enough to justify the plug flow assumption in most cases.

Model for Average Mass Transfer Product

The average mass transfer products $\overline{K_o a}$ calculated from the data of Figure 5 using Eq. 18 were less sensitive to z_n than the local values shown in Figure 6. It was, therefore, decided to develop a simplified model for $K_o a$ which could provide a basis of comparison with the data.

The specific interfacial area of a two-phase dispersion is given by

$$a = 6\epsilon/d_{32} \quad (20)$$

In cocurrent flow at high velocities, buoyancy-driven flows can be neglected, so the organic-phase holdup is assumed to be given by

$$\epsilon = \frac{u_o}{u_o + u_a} \quad (21)$$

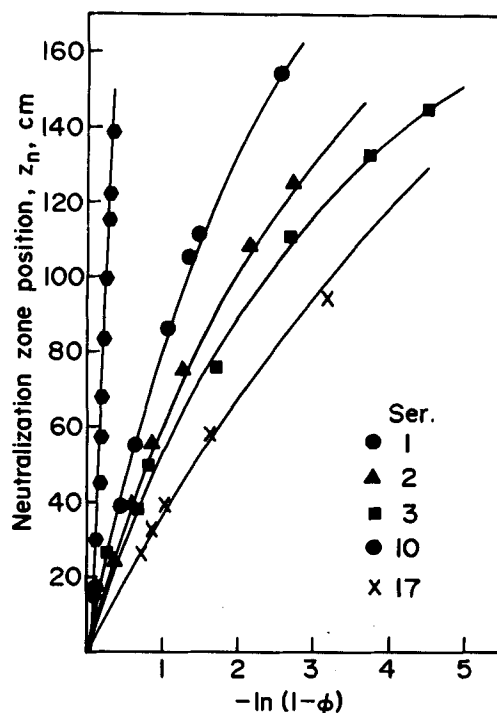


Figure 5. Effect of ϕ upon neutralisation zone position, z_n . Refer to Table 1 for data series parameters.

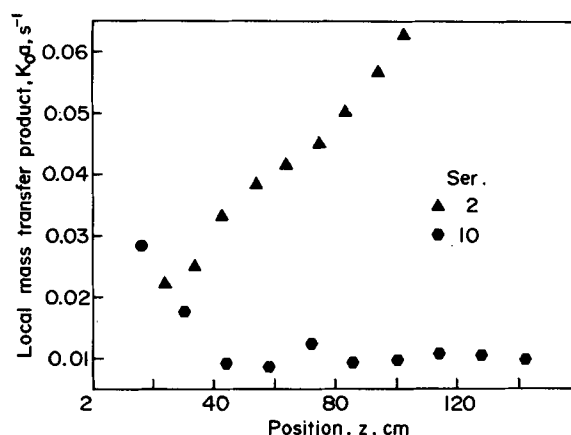


Figure 6. Variation of local mass transfer product with position in plate stack. Refer to Table 1 for data series parameters.

The Sauter mean drop diameter in reciprocating plate columns has been correlated by Baird and Lane (1973):

$$d_{32} = 0.36 \frac{\gamma^{0.6}}{\bar{\rho}^{0.2} \psi^{0.4}} \quad (22)$$

where $\bar{\rho}$ refers properly to the dispersion density:

$$\bar{\rho} = \epsilon \rho_o + (1 - \epsilon) \rho_a \quad (23)$$

The specific energy dissipation term in Eq. 22 contains two parts; one due to the effect of plate reciprocation and proportional to $(Af)^3$ (Hafez and Baird, 1978), and the second part due to the steady component of the flow and proportional to $(u_o + u_a)^3$.

$$\psi = \frac{\bar{\rho}}{h} \left(\frac{1 - \sigma^2}{C_o^2 \sigma^2} \right) \left[\frac{2\pi^2}{3} (Af)^3 + \frac{1}{2} (u_o + u_a)^3 \right] \quad (24)$$

In this work, ψ ranged between 250 and 50,000 W·m⁻³. Combination of Eqs. 20 through 24 results in the following expression for specific interfacial area.

$$a = \frac{6\bar{\rho}^{0.2}}{0.36\gamma^{0.6}} \left(\frac{u_o}{u_o + u_a} \right) \left\{ \frac{\bar{\rho}}{hC_o^2} \frac{(1-\sigma^2)}{\sigma^2} \left[\frac{2\pi^2}{3} (Af)^3 + \frac{1}{2} (u_o + u_a)^3 \right] \right\}^{0.4} \quad (25)$$

It must be noted that Eq. 22 for the droplet size is based on an assumed mechanism of droplet breakup by isotropic turbulence (Hinze, 1955). This concept has been found valid at high values of Af , but in the absence of reciprocation Eq. 25 should be treated with caution. In this case the energy dissipation tends to be concentrated immediately downstream of each plate and the turbulence is anisotropic.

As noted earlier, the overall mass transfer coefficient K_o can be taken as the organic-phase coefficient k_o because the aqueous-phase resistance is greatly reduced by the fast reaction of the acid. The mass transfer coefficient within a droplet depends strongly on whether the droplet is free to circulate or not. For stagnant droplets, Treybal (1963) has indicated that

$$k_o = \frac{2\pi^2}{3} \cdot \frac{D}{d} \quad (26)$$

Kronig and Brink (1950) derived the following expression for a circulating drop:

$$k_o = 17.9 \frac{D}{d} \quad (27)$$

In the present system, the droplets are expected to be small and stagnant at high agitation levels; at low agitation, the droplets will be larger and more likely to circulate. Accordingly, either Eq. 26 or 27 could apply, or the mass transfer coefficient k_o could be intermediate between the two predictions.

Comparison between Present Results and Model

A detailed comparison is given by Noh (1981) and it may be summarized by reference to Figure 7. This shows some typical data on the average mass transfer product $\bar{K}_o a$, plotted against the corresponding calculated values obtained by multiplying K_o (from Eq. 26 for stagnant droplets) and a (from Eq. 25). The full line represents agreement between this prediction and the observed values. The dashed line, which is raised above the full line by a factor of 2.56, represents the prediction for circulating drops using Eq. 27 instead of Eq. 26.

The data from test series 5, 7, 8, 15 and 16 all show a trend towards the stagnant droplet model at high agitation levels; the independent variables tested here were frequency, plate spacing and the superficial velocities of both phases. Some scatter of the points is not surprising in view of the averaging of the $\bar{K}_o a$ values which, as already noted, are varying with position in the plate stack.

In test 14, the agitation frequency is only 1.0 Hz and the data lie between the circulating drop prediction and the stagnant drop prediction. This is consistent with the greater proportion of circulating drops likely to exist in the dispersion at lower levels of plate

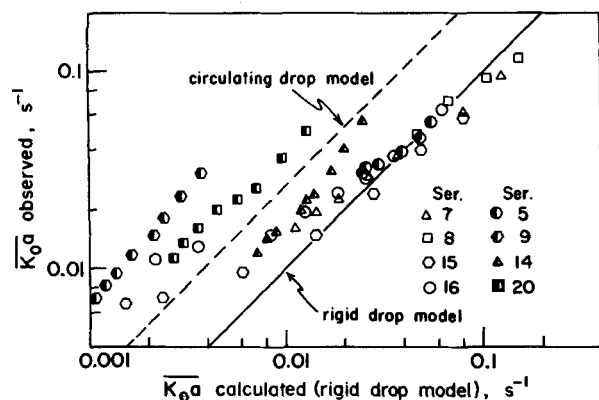


Figure 7. Comparison of $\bar{K}_o a$ observed (this work) with $\bar{K}_o a$ calculated. Refer to Table 1 for data series parameters.

agitation, corresponding to larger Sauter mean diameters (Eq. 22).

Results from two series of tests, viz., 9 and 20, lie above the circulating drop prediction. In these cases, no plate agitation was applied. Some data points from tests 15 and 16, at frequencies of less than 0.5 Hz, also lie in this region of Figure 7. Clearly the model is not applicable at very low (or zero) plate agitation and the observed mass transfer is significantly higher than predicted. It is suggested that this is because the droplet breakup mechanisms under such conditions are predominantly collision with the plates and dispersion in the immediate wake area behind the plates, leading to smaller drop sizes than predicted by Eq. 22.

Comparison between Karr's Data and Present Model

Karr (1983) operated a 2.54-cm diameter column in which o-xylene containing about 1% acetic acid was extracted cocurrently by water. The o-xylene (continuous) flow was kept at 16 times the water (dispersed) flow, such that the extraction factor was kept at approximately 2.2 throughout. Total superficial velocities could be increased to over 50 cm/s, i.e., considerably higher than in the present study. Because of the short contact times and the moderate value of the extraction factor, fairly accurate data on the column efficiency were obtainable by sampling and analysis of the exit stream.

Karr (1983) defined the column efficiency in terms of the acid concentrations in the organic phase.

$$E = \frac{c_{o1} - c_{o2}}{c_{o1} - c_{o2e}} \quad (28)$$

It can be shown that

$$\bar{K}_o a = - \frac{Fu_o}{(1+F)Z} \ln(1-E) \quad (29)$$

The experimental values of $\bar{K}_o a$ thus obtained from Karr's (1983) efficiency data may be compared with a model similar to the one developed above.

Because the distribution coefficient of the solute was highly in favor of water, it was assumed that the main mass transfer resistance lay in the organic phase, i.e., $K_o \approx k_o$. In this case the organic phase is continuous. Rigid sphere behavior is assumed and the well-substantiated correlation of Levins and Glastonbury (1972) is adopted:

$$Sh_o = \frac{k_o d_{32}}{D} = 2 + 0.47 \left[\frac{d_{32}^{4/3} \nu^{1/3}}{\nu} \left(\frac{D_s}{D_T} \right)^{0.28} \right]^{0.62} Sc^{0.36} \quad (30)$$

The molecular diffusivity D of acetic acid in o-xylene is taken as $2.06 \times 10^{-9} \text{ m}^2 \cdot \text{s}^{-1}$ (Wilke and Chang, 1955), and the impeller to tank diameter ratio (D_s/D_T) is effectively taken to be unity. The Sauter mean droplet size is calculated by means of Eq. 22 using the interfacial tension of $25 \text{ mN} \cdot \text{m}^{-1}$ reported by Karr (1983).

$\bar{K}_o a$ estimated from the value of k_o (Eq. 30) and that of a (Eq. 25) is shown on the abscissa of Figure 8 with Karr's data, processed via Eq. 29, on the ordinate. The range of values of $\bar{K}_o a$ obtained by Karr (1983) is significantly higher than that of the present work because of the higher flow velocities and agitation frequencies (generally 6.67 Hz) employed. Karr's data generally lie somewhat higher than the rigid-drop model prediction, perhaps because of droplet circulation or oscillation, but the order of magnitude is correct.

The present model for $\bar{K}_o a$ assumes that the properties of the dispersion, e.g., drop-size distribution, are independent of height. Thus although the model appears to be quite useful in predicting the column performance, it can only be regarded as an interim approach in view of the observed variation of $\bar{K}_o a$ with position in the plate stack (Figure 6). It was observed at low frequencies of agitation that the drop size continually decreased as the dispersion progressed through the column. A similar trend at high frequencies is expected, but this could not be visually observed because of the fineness of the dispersion. Further work, both experimental and theoretical, is needed on the local variation of $\bar{K}_o a$; in particular it would be desirable to measure the drop size distribution at different points in the plate stack.

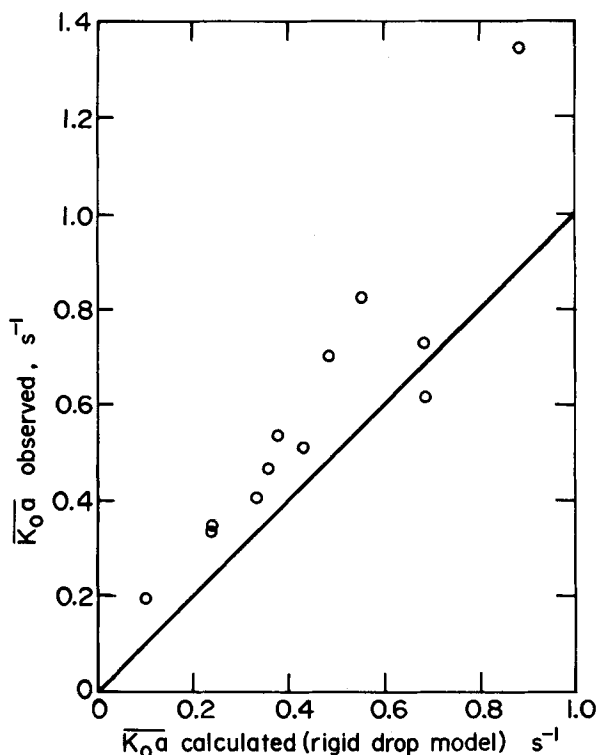


Figure 8. Comparison of $K_o a$ observed (Karr, 1983) with $K_o a$ calculated, assuming $F = 2.2$.

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NOTATION

A	= stroke, m
A'	= amplitude $\left(= \frac{1}{2} \text{ stroke} \right)$, m
a	= specific interfacial area, $\text{m}^2\text{-m}^{-3}$
C_o	= orifice coefficient
c	= concentration, mol m^{-3}
d	= drop diameter, m
d_{32}	= Sauter mean diameter, m
D	= molecular diffusivity, $\text{m}^2\text{-s}^{-1}$
D_s	= impeller diameter, m
D_T	= tank diameter, m
E	= stage efficiency
F	= extraction factor $= mu_o/u_a$
f	= frequency, Hz
h	= plate spacing, m
K	= overall mass transfer coefficient, m-s^{-1}
k	= film mass transfer coefficient, m-s^{-1}
m	= distribution coefficient $= C_{oe}/C_{ae}$
NTU	= number of transfer units
n	= number of plates
ΔP_f	= frictional pressure drop, Pa
Sc	= Schmidt number $= D/\nu$
Sh	= Sherwood number, Eq. 30
u	= superficial velocity, m-s^{-1}
u_h	= velocity of water through holes, m-s^{-1}
V	= dimensionless reciprocation velocity, Eq. 7

y	= plate displacement, m
\dot{y}	= plate velocity, m-s^{-1}
Z	= height of plate stack, m
z	= distance from base of plate stack, m
z_n	= value of z at which aqueous phase is neutral, m

Greek Letters

γ	= interfacial tension, N-m^{-1}
ϵ	= holdup of organic phase
ϵ_v	= power dissipation rate per unit mass, W-kg^{-1}
ν	= kinematic viscosity, $\text{m}^2\text{-s}^{-1}$
ρ	= density, kg-m^{-3}
σ	= fractional open area of plate
ϕ	= ratio of mols caustic fed to mols acid fed
ψ	= power dissipation rate per unit volume, W-m^{-3}
ω	= angular frequency $= 2\pi f$, s^{-1}

Subscripts

a	= aqueous phase
b	= base
c	= continuous phase
d	= dispersed phase
e	= equilibrium
o	= organic phase
1	= bottom of plate stack
2	= top of plate stack

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