Mass Transfer in a Horizontal Liquid-liquid Extraction Tube

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A study was made of the effect of physical properties and of geometrical and dynamical characteristics of several liquid-liquid extraction systems on the mass transfer coefficients. The two-component systems studied consisted of the following solvents with water: cyclohexanol, methyl ethyl ketone, furfural, normal butanol, and nitromethane. The transfer studies were made in a horizontal glass pipe with the phases flowing countercurrently at velocities from 1,000 to 15,000 lb./(hr.)(sq. ft.). The transfer of solvents into the water phase and of the water into the solvent phase was measured. The film coefficients of mass transfer for a solvent through the water film and for water through the solvent film were correlated by new dimensionless equations which include interfacial tension and diffusivity.

Since 1939 liquid-liquid extraction has assumed considerable importance as a unit operation of chemical engineering, particularly as an adjunct to distillation in the manufacture of organic chemicals and the processing of crude petroleum. Liquid-liquid extraction has proved to be valuable in the separation of azeotropic mixtures and in the concentration of chemicals from dilute solutions. For a long time liquid-liquid extraction was considered as the last resort for purification of chemicals, but with the introduction of more efficient equipment designs recently it has been winning favor. In all the processes mentioned in the literature (*16*) considerable progress has been made in the development of suitable equipment and the acquisition of the necessary operating techniques. However, this progress has been the result of trial and error in design and testing. More scientific design of extraction apparatus had to await development of a method for the correlation of the rate of extraction with the operating variables and the properties of the systems involved.

Quantitatively the mechanism of extraction and the dependence of extraction rates upon physical properties and dynamics involved have been assumed to be analogous to the similar, more extensively developed vapor-liquid operations. The major portions of the basic data on liquid-liquid contact equipment have been concerned with the performance of small-scale units with emphasis on the effect of the flow rates of both phases on the mass transfer coefficients. Expressed in terms of over-all mass transfer coefficients, *HTU* or over-all diffusional resistances rather than the individual values that are more fundamental and more widely useful for design purposes, the data are of value in the design of equipment for only those systems that have been investigated experimentally.

Few investigators (*3*, *8*, *10*) have attempted to separate the over-all coefficients into the individual resistances. They were able to do this only when certain restrictions were applied to their method of resolution. The restrictions set forth (*3*) require that the value of the distribution coefficient be unity and that either individual coefficient can be affected by both fluid rates but not both together. Invariably all have used the same method of resolution, whereby the individual resistances are added to give the over-all resistance and an over-all coefficient is obtained as a reciprocal of the latter. Early data indicate (*14*) that rates of flow of both phases affect the individual extraction coefficients of both films and that neither phase may be said to be controlling. In view of this, it would not be possible to break down the over-all coefficient into the individual coefficients.

It is the purpose of this investigation to study the properties affecting the film coefficients of mass transfer and to attempt to obtain a correlation of the properties of the system and of the operational variables which will enable one to predict approximately the values of the coefficients.

INDIVIDUAL TRANSFER COEFFICIENT

In 1934 Gilliland and Sherwood (*9*) proposed the following correlation for their results on the evaporation of water into air streams in a wetted-wall tower:

$$\frac{d}{x} = c\left(\frac{dG}{\mu}\right)^{n}\left(\frac{\mu}{\rho D}\right)^{m}$$

The ratio of the inside diameter of the column to the thickness of the stagnant layer was shown to be a function of the Reynolds and Schmidt numbers. The values of the constants, c, n, and m, were found to be 0.023, 0.83, and 0.44 respectively. This is apparently the earliest application of dimensional analysis to the study of film behavior during mass transfer.

By analogy therefore to relationships developed for heat transfer, Gilliland and Sherwood proposed the first correlation of the individual mass transfer coefficients with the physical properties of the fluids and the flow characteristics of the systems. In 1935 Fallah, Hunter and Nash (*7*) studied the transfer of phenol between water and kerosene in a wetted-wall column and found that the data for the kerosene film when kerosene was the core liquid could be correlated by the following equation:

$$\frac{kd}{D} = 0.94\left(\frac{dG}{\mu}\right)^{0.8}\left(\frac{\mu}{\rho D}\right)^{0.46}$$

In 1943 Brinsmade and Bliss (*3*) reported on the extraction of acetic acid in methyl isobutyl ketone with water in a wetted-wall tower. They separated the over-all film resistance into individual film resistances which were correlated by the following equations:

for the core fluid (ketone):

$$\frac{kd}{D_c} = 1.07\left(\frac{\mu\rho_c d}{\mu_c}\right)^{0.67}\left(\frac{\mu_c}{\rho_c D_c}\right)^{0.62}$$

and for the wall fluid (water):

$$\left(\frac{\mu_w^{\,2}}{\rho_w^{\,2} g}\right)^{1/3}\frac{k_w}{D_w} = 0.00135\left[\frac{4\Gamma(\rho_w - \rho_c)}{\mu_w\rho_w}\right]\left[\frac{\mu_w}{\rho_w D_w}\right]$$

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Table 1. Summary of Physical Properties for the Systems Used to Evaluate Correlation Equation

System	Diffusivities, (sq.ft./hr.) × 10^{-5}		Viscosities lb./(ft.)(hr.)		Interfacial tension lb./hr.^{2} × 10^5	Densities, lb./cu. ft.	
Water with	D_w	D_s	μ_w	μ_s	σ	ρ_w	ρ_s
Cyclohexanol	0.409	3.79	2.13	72.5	3.22	62.19	59.0
Nitromethane	2.01	6.45	2.13	14.7	1.63	62.19	70.7
n-Butanol	5.33	4.90	2.13	5.56	0.544	62.19	50.4
Furfural	8.32	4.73	2.13	3.56	1.345	62.19	72.0
Methyl ethyl ketone	32.3	4.57	2.13	0.92	0.286	62.19	50.1

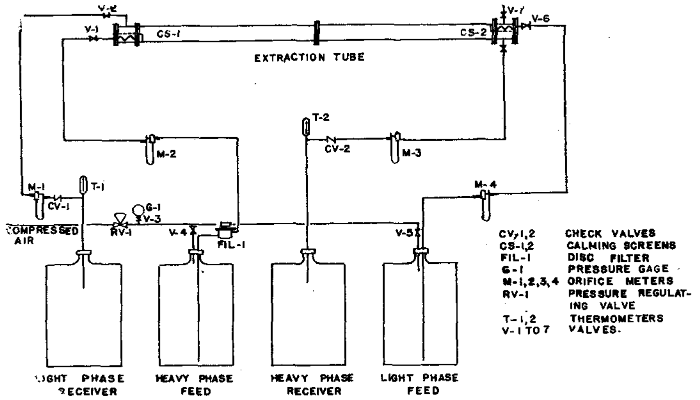

Fig. 1. Flow sheet for horizontal tube extractor.

Table 12. Summary of Constants for Correlation Equation

$$\frac{k_s d}{D_w} = A\left(\frac{d\sigma}{\mu_s D_w}\right)^a\left(\frac{\mu_s}{\mu_w}\right)^b\left(\frac{\mu_s}{\rho_s D_w}\right)^c\left(\frac{dG_s}{\mu_s}\right)^e\left(\frac{dG_w}{\mu_w}\right)^f$$

or

$$\frac{k_w d}{D_s} = A\left(\frac{d\sigma}{\mu_w D_s}\right)^a\left(\frac{\mu_w}{\mu_s}\right)^b\left(\frac{\mu_w}{\rho_w D_s}\right)^c\left(\frac{dG_s}{\mu_s}\right)^e\left(\frac{dG_w}{\mu_w}\right)^f$$

Mass transfer coefficients	A	a	b	c	e	f
	As originally calculated					
k_w	0.672	0.779	0.274	−1.47	−0.022	0.398
k_s	5.66 × 10^{17}	−1.07	−9.78	−4.03	0.923	−0.080
	As reevaluated					
k_w	0.552	0.670	0.254	−1.46	0.00	0.397
k_s	2.66 × 10^{17}	−1.012	−10.0	−4.17	0.927	0.00

Brinsmade and Bliss retain the same three dimensionless groups proposed by Fallah, Hunter, and Nash (*7*), revising them to fit the operating conditions.

The correlations reported show that the individual mass transfer coefficients have been generally recognized as a function of the Reynolds and Schmidt number (*10, 17*). However each relationship seems to vary the functions of the dimensionless group, possibly because of the differences in the nature of the systems and the condition of operation.

In review of calculation methods for extraction, involving mass transfer in solid-liquid and gas-liquid systems, as well as in liquid-liquid systems, Pike (*13*) discusses the two-film theory. On the bases that diffusion is the controlling process and that the resistances to diffusion reside in two liquid films adjacent to the interface, he attempts to develop a consistent mathematical picture. From the literature he summarizes the effects of the Reynolds, Schmidt, end-effect, and Grashof numbers. He then presents the idea that some other factor is affecting the transport of matter in the liquid-liquid systems. This effect is ascribed to the presence of the mobile interface. "This factor is logically related to the physical properties of the two contacting phases. . . . Some day this mobility effect will be properly expressed in a function of the physical properties of the two phases. . . . This group has not yet been formulated. It seems to me that it would contain only the physical properties of the two adjacent films, such as densites, viscosities, and interfacial tension."

Bergelin, Lockhart and Brown (*2*) used a horizontal extraction tube to study the transfer of isopropyl alcohol between the tetrachloroethylene and water phases. They observed that the over-all transfer coefficient passed through a minimum when plotted against the velocity of the water phase, at constant velocity of the tetrachloroethylene phase, with the minimum located where the phase velocities were equal. This minimum did not occur when the transfer coefficients were plotted against the velocity of the tetrachloroethylene phase at constant water velocities. By photographs of the flow pattern when dye was injected into the stream, they showed that two types of currents were introduced, one caused by density changes as extraction proceeded and one by the effect of velocity of flow of one phase on the other. They observed that the direction of the color stream in the water phase was reversed and pulled along the interface in the same direction as the tetrachloroethylene phase. This portion of fluid must return at a high velocity in the upper portion of the tube and thus, they say, causes turbulence.

Colburn and Welsh (*4*), by using a two-component two-phase system of isobutanol and water in a packed tower, avoided the difficulty of evaluating film coefficients, for in this case the over-all coefficient of transfer of one component into the other is also the film coefficient for that transfer. They observed that the H. T. U. for transfer of the discontinuous phase was independent of the flow rate of the discontinuous phase but that this was not the case for transfer of the continuous phase as plotted against the flow rate of the continuous phase. Pike (*13*) claims that this system had too high a mutual solubility.

EXPERIMENTAL

The experimental approach to the problem may be summarized as follows. Using a horizontal tube to permit reasonably accurate evaluation of the interfacial area and using several two-component, two-phase systems, one can measure the value of the mass transfer coefficients of one component into the other phase at different mass velocities. By use of dimensional and statistical analysis, the correlation equation of the coefficient of mass transfer with the measured properties of the system was evaluated. Certain flow behavior and correlation behavior are discussed, and future studies are suggested.

Equipment and Procedure

The equipment used is shown schematically in Figure 1. The extraction tube is an 8-ft. length of 1½-in. Pyrex-glass tube. At each end of the extraction tube double-flanged spool pieces provide for attachment of the necessary feed inlets and return taps. Inlet and return lines of ¼-in. tubing are provided for the streams to and from the extraction tube.

The streams entering at each end of the column pass through a calming section of screen that covered the full cross-sectional area of the entering phase. The accurate cross-section area of each phase was 0.00614 sq. ft. and since the distance between ends of the separating partitions was 96 in., the interfacial area was exactly 1 sq. ft. An exception was the system *n*-butanol-water (Table 2), which was tested in a 2-in. tube 4 ft. long. For this system the phase cross sections were 0.00345 sq. ft.

In the experimental program material-transfer studies were made with two-component systems composed of solvent and water. In one series of experiments the water-flow rate was held as constant as possible and the solvent rate was varied from 1,000 to 15,000 lb./(hr.)(sq. ft.) in steps. In another series of determinations the water flow was varied over the same range while the flow rate of the solvent was held as constant as possible. The temperature of the fluids was kept at 80.6° to 84.2°F. (27.0° to 29.0°C.). The pressure required to force the liquids through the extractor was furnished by regulated, compressed air delivered to the top of the feed carboys at 9 lb./sq. in gauge pressure.

The interface between the phases was maintained at the center line of the entire length of the tube. A slight slope was necessary to keep the interface at the center of the tube over its full length, but this slope was not measured. After a time equivalent to six changes of the slowest moving phase through the tube, samples were taken simultaneously, over a short period of time, of both the solvent and water return streams.

Analysis of the samples taken was made by use of a Zeiss-Abbe refractometer. The refractive indexes of known solutions of the various solvents in water and of water in the various solvents were determined and are reported elsewhere (*15*). The procedure followed was that recommended by the American Society for Testing Materials (*1*).

Chemicals

The chemicals used were technical-grade *n*-butyl alcohol, purified cyclohexanol, purified furfural, 99% methyl ethyl ketone, technical-grade nitromethane, and Blacksburg well water.

The required physical properties of these chemicals were measured and found to agree with published values. Solubility information between the solvents and water were measured and are reported elsewhere (*15*). Viscosities, densities, and interfacial tensions were measured (*12*) and plotted for use in correlation of the results obtained. The values of diffusion coefficients were estimated by Wilke's method (*19*). Table 1 is a summary of the values of the physical properties of the systems as used in this investigation.

Experimental Results

Mass transfer data and calculated results for the tests made are presented in Tables 2 through 6.* The quantities of each solvent and water entering and leaving the extraction tube were determined from the flow rates and compositions of the inlet and outlet streams. The quantities of solvent transferred into the water and of water transferred into the solvent phase were found by taking the difference between the quantities of each material in the inlet and outlet streams of the opposite phase. The concentration differences were calculated from the solubility limits and the experimentally determined composition of the entering and leaving streams (*18*). The average concentration differences are logarithmic when the ratios of these differences are more than 1.8.

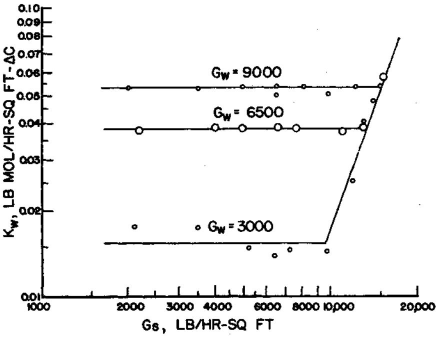

Fig. 2. Variation of K_W with mass velocity of the solvent phase, for the system furfural-water at 27°C.

Correlation of Results

Since previous attempts to correlate transfer coefficients with physical properties and with geometrical and dynamical characteristics of the systems have been criticized for omitting the interfacial tension and diffusivity, a new correlation equation was developed by dimensional analysis. This analysis resulted in an equation of six dimensionless groups, assembled in the following forms (*11*):

$$k_s d/D_w = A(d\sigma/\mu_s D_w)^a(\mu_w/\mu_s)^b \cdot(\mu_s/\rho_s D_w)^c(dG_s/\mu_s)^e(dG_w/\mu_w)^f$$

$$k_w d/D_s = A(d\sigma/\mu_w D_s)^a(\mu_w/\mu_s)^b \cdot(\mu_w/\rho_w D_s)^c(dG_s/\mu_s)^e(dG_w/\mu_w)^f$$

This set of dimensionless groups separates the effect of the phase mass velocities (Reynolds numbers), the viscosity ratio, and the interfacial tension. Evaluation of the constants enables one to evaluate each effect. This correlation equation reduces to that of Fallah, Hunter, and Nash (*7*) when the viscosity ratio and the interfacial-tension groups are assumed part of the constant.

Each dimensionless group for each system was evaluated, and these values with their logarithms are given in Tables 7 through 9* inclusive.

The correlation regression equations were written in the linear logarithmic form

$$\log kd/D = \log A + a \log (d\sigma/\mu D) + b \log (\mu_w/\mu_s) + c \log (\mu/\rho D) + e \log (dG_s/\mu_s) + f \log (dG_w/\mu_w)$$

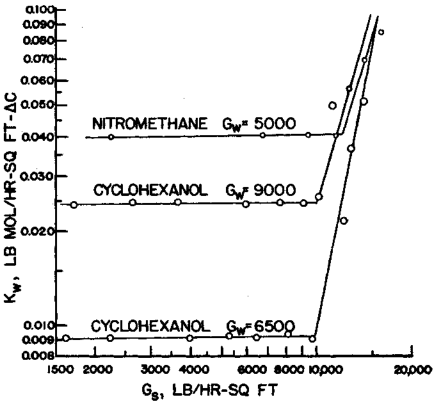

Fig. 3. Variation of K_W with mass velocity of the solvent phase, for systems nitromethane-water and cyclohexanol-water at 27°C.

By use of the method of least squares and the data of Tables 7 through 11*, which represent a total of 210 tests, a set of five equations was obtained for k_w and similarly for k_s. The constants in each set of equations were evaluated by the Abbreviated Doolittle Method, and are given in Table 12.

An analysis of correlation coefficients between *kd/D* and the logarithms of the Reynolds numbers indicates that the effect of the Reynolds number of the solvent phase is insignificant with respect to k_w and of the water phase is with respect to k_s. That is, these groups do not contribute significantly to the calculation of the specified transfer coefficient. The Reynolds number of the solvent phase may be dropped from the equation for k_w and of the water phase for k_s. After these groups had been removed, the original constants were modified and are also presented in Table 12.

The statistic *F*, which in this case is the ratio of the variance due to regression to the error variance, was calculated for both regressions and found to be highly significant. It indicated that the type of equation used in this paper can explain a significant amount of variation. The multiple correlation coefficient between the measured and calculated values of the logarithms of $k_s d/D_w$ is 0.9066 and for the measured and calculated values of the logarithms of $k_w d/D_s$ is 0.539. The

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*Complete tabular data (Tables 2 through 11) may be obtained as document 5059 from the American Documentation Institute, Photoduplication Service, Library of Congress, Washington 25, D. C., for $5.00 for photoprints or $2.25 for 35-mm. microfilm.

*See footnote in column 2.

low value in the latter case signifies that the correlation is not very good and that some other factors which are not included in the equation may be affecting the values of $k_w d/D_s$.

Flow Characteristics in Horizontal Tube

In extraction processes, where moving partially miscible liquids are brought into contact for the purpose of transfer of a substance from one liquid to the other across the phase boundary, the diffusing substance must pass through various portions of the fluid that may be in viscous or turbulent flow. Strang, Hunter, and Nash (*16*) and Brinsmade and Bliss (*3*), for purposes of interpretation, concluded that for wetted-wall columns, the flow of the wall liquid is laminar and that of the core liquid turbulent. They offer no proof of these conclusions and since it is well known that mass transfer investigations involving the wetted-wall type of equipment have not in all instances given results that are in complete agreement, the postulates presented by Strang, Hunter, and Nash and by Brinsmade and Bliss, cannot serve as a basis for analyzing the patterns of fluid flows in the horizontal tube.

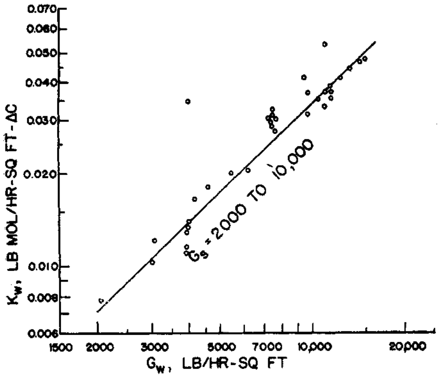

Fig. 4. Variation of K_W with mass velocity of the water phase, for the system methyl ethyl ketone-water at 27°C.

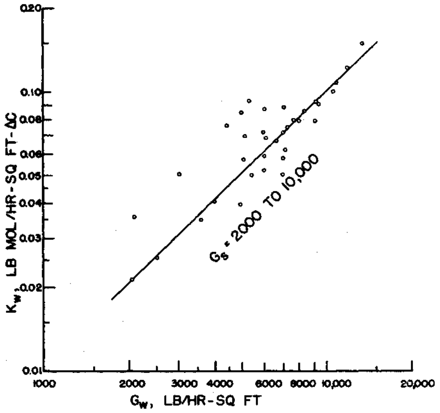

Fig. 5. Variation of K_W with mass velocity of the water phase, for the system nitromethane-water at 27°C.

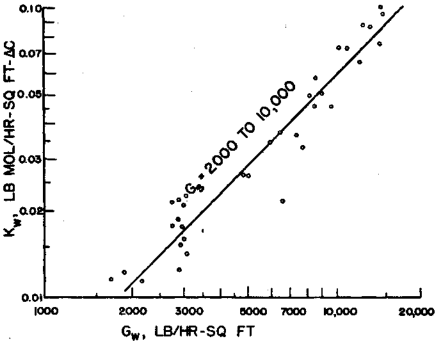

Fig. 6. Variation of K_W with mass velocity of the water phase, for the system cyclohexanol-water at 27°C.

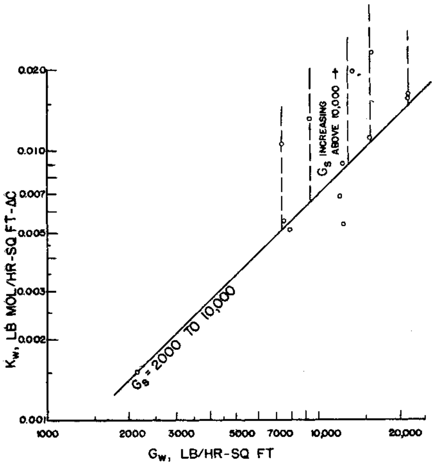

Fig. 7. Variation of K_W with mass velocity of the water phase, for the system *n*-butanol-water at 27°C.

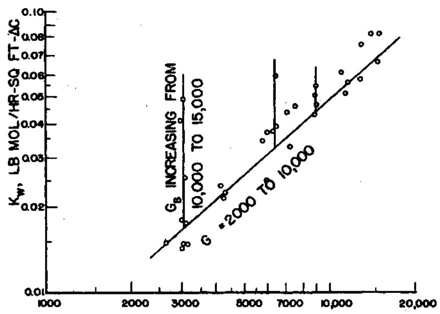

Fig. 8. Variation of K_W with mass velocity of the water phase, for the system furfural-water at 27°C.

Calculated values of the Reynolds number have been used to establish the character of the flow of each phase as either laminar or turbulent. From visual observation during countercurrent flow in a horizontal tube, it appears that the flow rate of one phase has an influence upon the degree of turbulence in the other phase. When the mass velocities of the two phases are equal in magnitude and opposite in direction, the velocity at the interface is close to zero. When the mass velocities are not equal in magnitude, a layer appears in the slower moving phase which apparently moves in the direction of the faster moving phase. The faster moving phase "shears off" a layer of the slower moving phase and causes it to flow in the direction of the faster moving phase. When the mass velocities of the two phases are equal in magnitude, the "plane of zero velocity" is at or near the interface between the two phases and as the velocity of one phase is increased this plane of zero velocity moves into the slower moving phase. In these studies, where a velocity ratio of 10 to 1 of the faster moving phase to the slower moving phase was the maximum, it was observed that the plane of zero velocity was as much as ¼ in. in the slower moving phase from the interface. The movement of the plane of zero velocity depends on the relative densities of the two phases and on the interfacial tension, which is a measure of the rigidity of the two-phase boundary. Thus with water and trichloroethane, the plane of zero velocity will move into the trichloroethane only a very small distance even at very rapid flow rates of the water phase.

Some visual observations of particles floating in the streams and of flow patterns observed with dyes introduced at various levels show the flow patterns very well. The character of flow in the region between the interface and the plane of zero velocity usually is laminar, but at extremely high velocities [above 11,000 lb./(hr.)(sq. ft.)] there is a rolling turbulence in this layer. This agrees with the observations of Bergelin, Lockhart, and Brown (*2*) and of Eargle (*6*).

These observations lead one to the conclusion that the transfer coefficient for one of the components into the opposite phase cannot be completely independent of the mass velocity (or Reynolds number) of the parent phase. At the lower velocities this is the observed behavior, but Figures 2 and 3 indicate that such is not the case at higher velocities where the turbulence introduced results in a sharp rise in the values of the transfer coefficients.

Behavior of Mass Transfer Coefficients

Since the systems involved are two-component, two-phase, in which one of the components is the transferred material, the over-all coefficient is equal to the individual coefficient; that is, $K_w = k_w$ and $K_s = k_s$.

When k_w is plotted against G_s at constant values of G_w, for any single system, as in Figures 2 and 3, one finds, after a region where k_w is a constant for a range of values of G_w, a region of sharp, linear increase. In the first region the transfer of solvent into the water phase is independent of the velocity of the solvent phase. This condition holds up to a particular value of G_s (which depends on G_w), when apparently the value of k_w becomes dependent on G_s and independent of G_w. This break is ascribed to the turbulence of the solvent phase agitating the interface film of the water phase.

It is interesting to note that the curves of k_w vs. G_s in the turbulent region [G_s greater than 11,000 lb./(hr.)(sq. ft.)] for all systems in which this break was observed fall almost together. No relation between Reynolds number and the break point could be found. The Reynolds number for cyclohexanol at the break point was 17.3; for nitromethane, 96.8; and for furfural at G_w: 3,000, 380; at G_w: 6,500, 469; and at G_w: 9,000, 540.

Figures 4 to 8 show the relation between k_w and G_w for separate systems. For values of G_s from 2,000 to 10,000 lb./(hr.)(sq. ft.) the values should fall along the line shown, but if G_s is more than 10,000 the value of k_w will fall above this line. Thus for furfural, Figure 8, if G_w is 3,000, the value of k_w will be about 0.015 lb./(hr.)(sq. ft.), unit concentration difference for G_s of 2,000 to 10,000. But at G_s of more than 10,000 the values of k_w fall on a vertical line, as shown. The correlation equation does not include these data, since any values of k_w for G_s above 10,000 were discarded before the correlation was made.

A first sight it may appear queer that the slope of the curve for k_w vs. G_w, at G_s from 2,000 to 10,000, is shown in Figures 4 to 8 as about 1.0; whereas, the method of least squares, for all the data for all systems, shows an exponent for the Reynolds number dG_w/μ_w of 0.4. If one should plot the data for all the systems on one graph, the curve through the points, as determined by the method of least squares, would have a slope of 0.4. As has been pointed out, this scattering of points results in poor correlation when the data for all systems are combined.

No such interesting curves could be obtained from the data collected for k_s plotted vs. G_w at constant G_s. The data was too scattered to permit location of any curves. However, statistically it was shown that k_s could be considered independent of the Reynolds number for the water phase. This means there should have been a horizontal portion to the curves. The scattered data is attributed to the higher transfer coefficients of water into solvent as compared with that for the solvent into water and to the necessity for more careful sampling and analysis.

LIMITATIONS OF THE METHODS

The equation proposed for calculation of mass transfer coefficients can at its present state of development be used only for calculations for mass transfer through the interface of a horizontal tube extractor. For any other type of extractor, no means of evaluating the area of the interface have been accepted. Furthermore, the length term in the Reynolds number, d, the diameter of the tube, must be interpreted and defined for other types of extractors. At present this method of attack is only one approach to the determination of the effect of various factors on the film coefficients.

The next step would be the application of the equation to extraction in the horizontal tube of three-component, two-phase systems, where a solute is transferred from one phase to the other. Then the equation could easily be applied to experiments using the wetted-wall towers. Finally the more difficult step will be to find a means of using the prediction equation with packed spray and baffle-type columns.

CONCLUSIONS

By use of a horizontal tube extractor and five two-phase, two-component systems, data have been collected for transfer of one component into the other. These data have been correlated as a series of dimensionless groups, related in exponential form. The constants for the equations have been determined. The data have been analyzed statistically. The value of the solvent-film coefficient for transfer of water into the solvent has been found to be independent of the mass velocity of the water phase between 2,000 and 10,000 lb./(hr.)(sq. ft.). The value of the water film coefficient for transfer of solvent into the water has been found to be independent of the mass velocity of the solvent phase between 2,000 and 10,000 lb./(hr.)(sq. ft.). Above a mass velocity of 10,000 lb./(hr.)(sq. ft.) the values of the logarithms of $k_s d/D_w$ and of $k_w d/D_s$ vary as a straight-line function of the logarithms of G_w and G_s, respectively, independently of the values of G_s or G_w, respectively.

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NOTATION

A = total interfacial area of contact, sq. ft.
A = constant
a = constant
b = constant
C = concentration of solute in solution, lb. moles/cu. ft.
c = constant
D = diffusivity, sq. ft./hr.
d = diameter of extraction tube, or inside diameter of tower, ft.
e = constant
f = constant
G = mass rate of flow, lb./(hr.)(sq. ft.)
g = acceleration of gravity, ft./hr.^{2}
K = over-all mass transfer coefficient, lb. moles/(hr.)(sq. ft. of unit concentration difference)
k = film coefficient of mass transfer, lb. moles/(hr.)(sq. ft. of unit concentration difference)
m = constant
N = moles solute transferred, lb. moles
n = constant
u = velocity, ft./hr.
x = thickness of film, ft.

Subscripts

c = core fluid
s = solvent phase
w = wall fluid or water phase

Greek Letters

Δ = finite change or difference
θ = time elapsed, hr.
σ = interfacial tension, lb./hr.^{2}
μ = viscosity, lb./(ft.)(hr.)
ρ = density, lb./cu. ft.
Γ = weight rate of flow per unit periphery, lb./(hr.)(ft.)

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