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The Effect of Concentration Level on the Gas Phase Absorption Coefficient

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A theoretical and experimental investigation has been made to determine the effect of the concentration level on the rate of mass transfer of a solute from a nontransfering or inert carrier gas in turbulent flow.

Diffusion theory predicts the gas phase mass transfer coefficient to be inversely proportional to the first power of the mean mole fraction of the nontransferring gas and to be a possible function of the rate and direction of mass transfer.

The results of an experimental study of the absorption of ammonia from mixtures with nitrogen into distilled water and aqueous ammonia solutions in a short wetted-wall column verified the predicted inverse relationship between the turbulent gas phase absorption coefficient and the first power of the mean inert mole fraction for inert mole fractions ranging from 0.068 to 0.934. The data also showed that for the range covered in this investigation there was no significant influence of the mass transfer rate on the product of the coefficient and the mean mole fraction of the inert carrier gas.

In chemical engineering operations such as gas absorption, evaporation into gas streams, and partial condensation the mass transfer in the gas phase may often be characterized as a process involving the diffusion of a solute through a nontransferring carrier gas between a dense boundary and a turbulent core. Heretofore, the effect of the concentration level on the transfer rate in this type of system has been taken into account in design by the use of the predictions of film theory based on the well-established effect of concentration level on mass transfer by molecular diffusion (*16*). Recent attempts to verify the predictions of film theory for evaporation into turbulent gas streams, however, have yielded conflicting results with respect to the correct relationship between the concentration level and the mass transfer coefficient. To obtain a better understanding of the phenomenon, this study applies the basic knowledge currently available on turbulent transfer processes and tests

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the theoretical analysis with experimental gas absorption data.

PREVIOUS EXPERIMENTAL WORK

A considerable number of experimental investigations of the gas phase mass transfer process have been carried out with the evaporation of pure liquids into turbulent gas streams. Those prior to 1952 have been summarized by Sherwood and Pigford (*13*) who concluded that in view of the lack of conclusive data on the effect of concentration level, the best procedure for correlation involved taking the $k_G p_{BM}$ product to be a constant in agreement with the predictions of molecular diffusion theory.

Several recent investigations of the effect of concentration level have been reported, all of which have used evaporating liquid systems. Cairns and Roper (*2*) studied humidification in a long wetted-wall column in which transfer rate data were obtained for steam mole fractions ranging from 0.03 to 0.85. Their experimental results did not agree with film theory predictions for either mass transfer or simultaneous heat and mass transfer (*3*). Their proposed correlation indicated k_G to vary inversely as the 0.83 power of p_{BM}.

The evaporation of carbon tetrachloride into air in a small wind tunnel was investigated by Westkaemper and White (*19*) over the inert mole fraction range of 0.3 to 1.0. An interrelation of the two independent variables, concentration level and gas flow rate, which occurred as a consequence of their experimental technique, precluded a unique correlation of the mass transfer coefficients.

Phillips (*10*) studied the rate of evaporation of water from falling films and from wetted cylinders into turbulent air streams at different total pressures and humidities. The data from the long wetted-wall column gave integrated values of the mass transfer rate. The coefficients were evaluated by an integral expression based on assumed heat and mass transfer mechanisms. The evaporation from the wetted cylinders, on the other hand, represented essentially point or differential transfer conditions. Heat transfer effects were taken into account by careful temperature measurements. The results from both these systems could be correlated on the basis of k_G being inversely proportional to p_{BM}.

From a study of the evaporation of carbon tetrachloride into air in a short column packed with ½-in. Raschig rings, Shulman and Delaney (*14*) and Shulman and Robinson (*15*) concluded that the mass transfer coefficient was inversely proportional to the two-third power of p_{BM}. The experimental data did not permit a conclusive empirical determination of the exponent, and the value two-third was chosen to correspond with the exponent on the diffusivity in the Schmidt number.

These previous studies have not presented a conclusive picture of the effect of the concentration level on mass transfer into turbulent gas streams. This situation may be the result of the inherent difficulties and assumptions due to simultaneous mass and heat transfer encountered in evaporation studies. For this reason an absorption system was chosen in this experimental investigation of the effect of concentration level. The apparatus used was a short wetted-wall column, which behaved as a differential absorber and permitted the study of point conditions. Furthermore, through proper choice of gas and liquid concentrations, the driving force, and hence the mass transfer rate, and the concentration level could be varied independently.

THEORETICAL ANALYSIS

In molecular diffusion in ideal gas mixtures the net rate of transfer of solute component A in the y direction with respect to fixed coordinates is expressed by the Stefan rate equation as the sum of a convective flux and a diffusion flux:

$$N_A = UC_T y_A - D_v C_T \frac{dy_A}{dy} \qquad (1)$$

The differential rate mechanism leading to Equation (1) is based for convenience on molecular diffusion because it leads to a value of the diffusivity D_v, that is essentially independent of the concentration in binary systems. In the transfer of solute through a nontransferring gas, a net convection velocity U established by the diffusion of the solute augments the diffusion flux $- D_v C_T \, dy_A/dy$ by the convective flux $U \, C_T \, y_A$ and leads to a particular solution of the general transfer equation. Integration of Equation (1) over the molecular diffusion path of length δ_L for one component diffusing at steady state gives

$$N_A = D_v C_T (y_{A_1} - y_{A_2}) / (y_{BM})(\delta_L) \qquad (2)$$

Equation (2) states that the transfer rate is proportional to the difference in the terminal values of the solute mole fraction across the diffusion path and inversely proportional to the mean concentration level y_{BM} of the nontransferring gas. Definition of the mass transfer coefficient as the transfer rate divided by the mole fraction driving force gives a coefficient that is found to be independent of total pressure and only slightly dependent upon temperature:

$$k_y \equiv N_A / (y_{A_1} - y_{A_2}) = D_v C_T / (y_{BM})(\delta_L) \qquad (3)$$

The mass transfer coefficient in the limit of low solute concentration k_{y^*} (the transfer coefficient for the case of negligible convective flux and hence also the coefficient for equimolal countercurrent diffusion) is then given by

$$k_{y^*} \equiv \lim_{y_A \to 0} k_y = D_v C_T / \delta_L \qquad (4)$$

The ratio of the two coefficients is equal to the inverse of the mean inert concentration term to the first power:

$$k_y / k_{y^*} = 1 / y_{BM} \qquad (5)$$

Experimentally, this relationship has been verified for at least three separate cases: molecular diffusion in the absence of forced convection, one-dimensional transfer postulated by film theory, and two-dimensional transfer from a flat plate into a laminar boundary layer as long as the transfer rates are relatively small (*8*). For the case of large concentration gradients, and hence large transfer rates, the ratio of the two coefficients as predicted by laminar boundary-layer theory deviates from the simple expression of Equation (5). This deviation is the result of the interaction of the diffusion and the momentum transfer processes and convection in the direction parallel to the plate.

In the treatment of mass transfer to or from a turbulent gas stream flowing adjacent to a dense boundary, film theory is frequently applied. Because of the hypothetical nature of the assumptions, however, it is difficult to judge the adequacy of the theory in actual applications. A somewhat more rigorous analysis can be made beginning with the turbulent diffusion equation based on time-averaged stream quantities and eddy transport coefficients. It is convenient to consider here the one-dimensional case because of the complexity of the more general case. Concentration gradients in the flow, or x, direction are assumed small compared with those in the direction of mass transfer y. The local rate equation based on time-averaged stream properties and the assumption that molecular and eddy transport processes are additive (*11*) is similar to Equation (1):

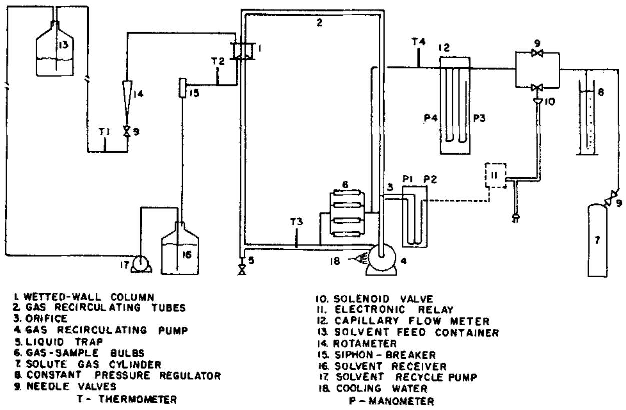

Fig. 1. Schematic diagram of the absorption apparatus.

$$N_A = U\, C_T\, y_A - (D_v + E_M)\, C_T \frac{\partial y_A}{\partial y} \tag{6}$$

In this case, $U = N_A/C_T$, so that Equation (6) can be reduced to

$$N_A = -\frac{(D_v + E_M)C_T}{(1 - y_A)} \frac{\partial y_A}{\partial y} \tag{6a}$$

At any point x along the direction of flow, Equation (6a) can be integrated across the diffusion path from $y = 0$ to $y = \delta_M$ which for steady state conditions at the boundary and in the bulk of the fluid gives Equation (6b):

$$(N_A)_{x=x} = \frac{(y_{A_1} - y_{A_2})}{y_{BM} \displaystyle\int_0^{\delta_M} \frac{dy}{(D_v + E_M)C_T}} \tag{6b}$$

The length average flux is then given by Equation (7):

$$(N_A)_{\text{avg}} = \frac{1}{h} \int_0^h \frac{(y_{A_1} - y_{A_2})}{y_{BM} \displaystyle\int_0^{\delta_M} \frac{dy}{(D_v + E_M)C_T}}\, dx \tag{7}$$

If the mole fractions of the solute in the free stream and at the gas-liquid interface do not vary appreciably over a segment h of the transfer apparatus, Equation (7) can be simplified to give Equation (8):

$$(N_A)_{\text{avg}} = \left[\frac{1}{h} \int_0^h \frac{dx}{\displaystyle\int_0^{\delta_M} \frac{dy}{(D_v + E_M)C_T}} \right] \frac{(y_{A_1} - y_{A_2})}{y_{BM}} \tag{8}$$

Combining Equation (8) with the definition of the mass transfer coefficient $k_y \equiv (N_A)_{\text{avg}}/(y_{A_1} - y_{A_2})$ one gets

$$k_y y_{BM} = \frac{1}{h} \int_0^h \frac{dx}{\displaystyle\int_0^{\delta_M} \frac{dy}{(D_v + E_M)C_T}} = k_{y^*} \tag{9}$$

where, as before, $k_{y^*} = \lim\limits_{y_A \to 0} k_y$. Thus the ratio of the mass transfer coefficients k_y/k_{y^*} for a turbulent boundary layer is predicted to be inversely proportional to the mean inert mole fraction as in the case of molecular diffusion Equation (5). It should be noted that in each case concentration level is introduced by considering the convective flux created by the net motion of the solute in the diffusion process.

The solutions to the fundamental transport equations used in boundary-layer theory do not support the expectation of equal exponents on the Schmidt and concentration level groups. In fact, these groups account for two distinctly different phenomena in the diffusion process. The Schmidt group in general accounts for the interrelation of the velocity and concentration gradients in the region in which molecular transfer processes prevail. The concentration level term, on the other hand, accounts for the resistance of the nontransferring gas to the net movement of the solute in both laminar and turbulent transfer. Laminar boundary-layer theory (*8*) for flow over a flat plate predicts that over the range of mass transfer rates normally encountered in chemical engineering operations the proper relationship is of the form $k_y y_{BM} \propto (N_{Sc})^{-2/3}$.

In deriving the above results, it was assumed that gradients in the flow, or x, direction may be neglected and that the mass transfer rate is sufficiently small to avoid alternation of the turbulent transport characteristics as expressed by the total diffusivity. In the case of molecular diffusion in a laminar boundary layer these assumptions have been shown not to apply at high transfer rates, but it seems reasonable to expect them to be more applicable in a turbulent boundary layer, since the transfer conditions are more uniform and the shear stresses created by turbulence are relatively greater than the transport of momentum by the diffusing solute. In any event, both the tentative nature of turbulent diffusion theory and the inability to apply it rigorously to complex flow conditions require experimental investigation to be the final determinant of the functional relationship between the variables.

EXPERIMENTAL PROGRAM

Ammonia was absorbed into distilled water and aqueous ammonia solutions from various nitrogen-ammonia gas mixtures in a short wetted-wall column. Other variables, such as gas flow rate, liquid flow rate, temperature, and pressure, were held essentially constant.

The ammonia-nitrogen-water system offers several advantages for a study of this type. The Schmidt number in the nitrogen-ammonia gas phase is very nearly independent of the composition. The system gives a low liquid-phase resistance. Finally, the resistance in the liquid phase is simply one of molecular diffusion which is not complicated by the effect of chemical reaction except for the negligible amount involved

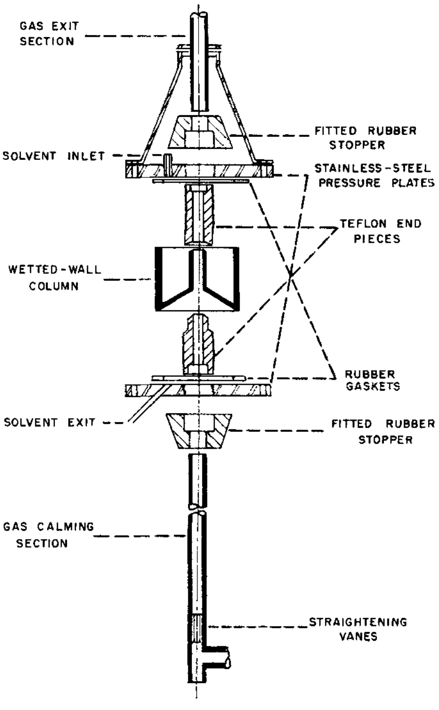

Fig. 2. Detail of the absorption apparatus (exploded view).

in ionization. The large solubility of ammonia in water is considered to be due to hydrogen bond formation (*4, 17*), and this is not believed to be rate controlling for absorption.

The experimental apparatus depicted schematically in Figure 1 can conveniently be divided into the absorption, the gas recirculation, the solvent feed, and the solute feed systems. The absorption section consisted of the short wetted-wall column provided with a 50-diam. entry and a 14-diam. exit tube to produce fully developed and reproducible gas flow conditions. The column was of the up-flow slot design reported by Vivian and Peaceman (*18*), having a diameter of 1.09 in. and producing a falling film of height 2.19 in. The entry and exit tubes of 32-mm. Pyrex glass were connected to the wetted-wall column with Teflon end pieces as shown in Figure 2, with the result that the passage for the gas flow was continuous and uniform in diameter throughout the absorption section. The wetted-wall column as a research tool offers the advantage of simple and reproducible flow patterns whose change with physical properties can be satisfactorily predicted.

The gas recirculation system was constructed of 1 × 1¼ in. clear plastic tubing to connect the gas recirculation blower with the ends of the absorption section. All joints in the plastic tubing were secured with a special plastic cement to eliminate leaks. The gas recirculation blower was a squirrel cage blower (10-in. diameter) mounted on the shaft of a totally enclosed motor which in turn was mounted on the blower housing. The motor was a 1/3 hp. end mounted, induction motor rotating at 3,400 rev./min. This motor-blower design could be adapted to provide a completely sealed unit which produced no detectable decomposition of the ammonia gas. The recirculating system was also provided with a 0.315-in. orifice for measuring the gas recirculation rate, with pressure taps and temperature wells, with ammonia make-up gas inlet, and with sample lines which were connected to the gas sample bulbs. The total pressure drop in the recirculation system was available to flush the gas sample bulbs.

Solvent was fed to the wetted-wall column by a gravity feed system. The rate was controlled by a needle valve and measured by a rotameter. To regulate the rate of solvent removal and maintain a constant liquid level in the exit slot of the wetted-wall column, a variable siphon was utilized. After the siphon breaker the liquid was accumulated in a solvent receiver from which it could be recycled to the overhead feed container if desired.

The solute feed system was designed for accurate control and measurement of the rate of addition of make-up ammonia to the recirculating system. A constant flow rate through the control needle valve was achieved by maintaining the pressure ahead of the valve constant by bubbling the excess gas from the cylinder through a constant head of liquid. In the start up of a run as well as between consecutive runs, the make-up ammonia gas was admitted by a relay actuated solenoid valve operating to keep the pressure in the recirculating system constant. The rate of addition of the make-up ammonia was measured with capillary flow meters.

The principle of measurement in these investigations was to determine the rate of absorption in the differential column by measuring the rate of addition of make-up ammonia to the circulating gas system. If there were no leaks or temperature change in the gas recirculation section, the only pressure change of gas in the section would result from absorption of ammonia in the falling film. Pure ammonia gas was added to the recirculation system at just the rate necessary to keep the system pressure constant. This was taken to be the rate of absorption. A comparison of the total volume of gas recirculation system (about 0.4 cu. ft.) with a low absorption rate (0.5 std. cu. ft./hr.) indicated it was necessary to maintain the pressure constant within 0.05 in. of carbon tetrachloride manometer fluid for 1 min. and to maintain the temperature constant with 0.1°C for 2 min. to reduce the error in the rate measurement due to each cause to 1%. These requirements were easily met, and the combined error from both sources was estimated to be less than 0.5%. The overall driving force for absorption was obtained from the difference in the mole fraction of ammonia in two gas samples: the first, taken with no detectable rate of absorption occurring in the column, contained the gas mixture in equilibrium with the solvent feed; the second, taken under steady state absorption conditions, gave the composition of the gas producing the measured absorption rate.

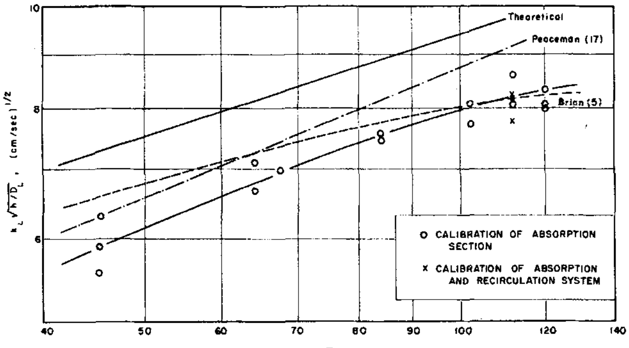

Fig. 3. Absorption of carbon dioxide into water in a short wetted-wall column.

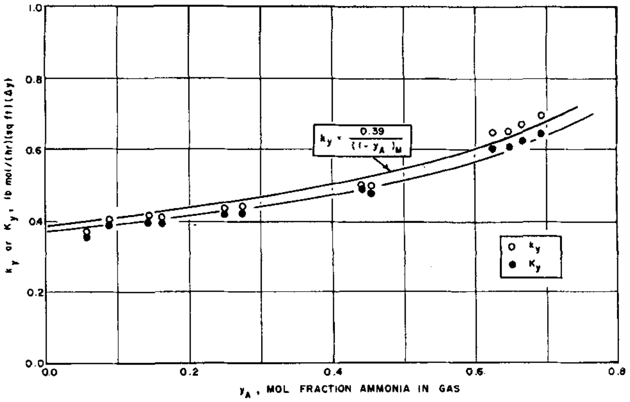

Fig. 4. Absorption of ammonia into distilled water in a short wetted-wall column.

In the start up procedure the entire solute feed and recirculation systems were purged of air with a stream of nitrogen. Nitrogen was preferred to air as the carrier gas to eliminate multicomponent effects. Then the wetted-wall column, which had previously been boiled in a detergent solution to facilitate wetting, was secured in place and the remainder of the absorption section flushed with nitrogen. The liquid flow and gas recirculation were then begun and a small driving force provided to prevent desorption during this period. When the steady state temperature had been reached, nitrogen was introduced through the solute feed system as ammonia was absorbed until the rate of absorption decreased to a negligible quantity. Two sample bulbs containing this gas mixture were then closed.

After solvent had been recycled to the overhead feed container, the driving force for absorption was established. When sufficient time had been allowed to attain steady state absorption conditions, the remaining experimental data and gas samples were taken. The liquid solvent was not recycled during this measurement period of a run. The runs made with distilled water required only the last steps above.

Analysis of the gas samples for ammonia involved titration with standard hydrochloric acid to a methyl-red end point. The standard acid was introduced into the gas sampling bulbs in such a way that the contents of the bulb were not exposed to the atmosphere until excess solution had been added. The samples were then back titrated with a standard ammonia solution.

RESULTS

Preliminary to the measurement of the gas-phase absorption coefficients, the wetted-wall column was calibrated to determine the liquid resistance to mass transfer

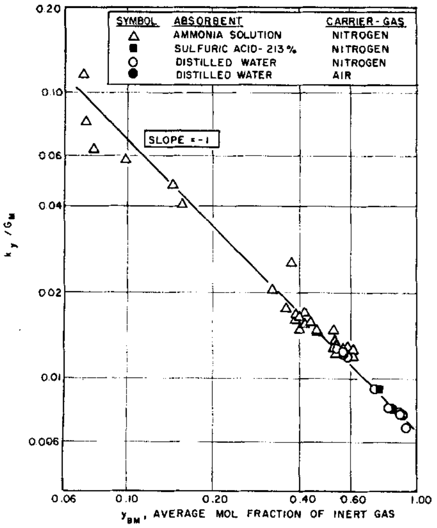

Fig. 5. Effect of inert gas concentration on absorption in a short wetted-wall column.

by absorbing carbon dioxide into distilled water. The results of the column calibration are presented in Figure 3 and are compared with data obtained in previous investigations with short wetted-wall columns (*6, 18*). Unsteady state diffusion theory predicts the factor $k_L\sqrt{h/D_L}$ to be a unique function of the liquid rate Γ, liquid viscosity, and density if the falling film is ideally laminar. The results indicate that the liquid resistance in this column was greater than that predicted theoretically but similar to that found in earlier investigations. The variation of the absorption coefficient with liquid rate, however, was that predicted by theory. The minor variations in resistance to absorption observed with different short wetted-wall columns are undoubtedly due to small entrance and exit slot effects which do not appear to affect the reproducibility for a given column. In Figure 3 the three runs indicated by crosses verified the technique used to measure the transfer rate in ammonia absorption. These runs, made at a gas Reynolds number of 5,350, were in good agreement with those taken with no gas flow. As a result of this study of the liquid-phase resistance in the apparatus, the highest liquid flow rate permitting smooth, laminar flow was chosen in order to suppress as much as possible the liquid resistance in the ammonia absorption.

The results for absorption of ammonia into distilled water are shown in Figure 4. The overall and gas-phase coefficients, which differ by only 4 to 6%, are plotted as a function of the overall driving force y_A. The temperature range was 25° to 27°C. The gas Reynolds number varied from 3,150 to 3,320, and the data shown have been reduced to a Reynolds number of 3,200 with experimental data on the effect of Reynolds number obtained in the apparatus and reported elsewhere (*1*). Hence, the only significant variables are the concentration level in the gas phase and the mass transfer rate. These data correlate well with the relationship $k_y y_{BM}$ = (constant) on which the upper line in Figure 4 is based, but they do not show the separate effects of concentration level and transfer rate.

The results of the absorption of ammonia into distilled water and aqueous ammonia solutions are shown in Figure 5, where a form of the Stanton number for mass transfer k_y/G_M is plotted against the mean inert mole fraction y_{BM}. The data indicate that the gas-phase absorption coefficient is inversely proportional to the first power of the mean inert mole fraction. The average deviation from the line representing $k_y Y_{BM}$ = (constant) is 7%, and from the statistical analysis of the data the slope of the line n in Figure 5 is estimated to be $-0.95 > n > -1.05$ within the 95% confidence limits. Because of the narrow range of flow rates used, these Stanton numbers are virtually independent of the gas rate. The average gas Reynolds number was 3,200. Essentially the same results were obtained with nitrogen or air used as carrier gas. The agreement of the data of the run with sulfuric acid absorbent (negligible liquid resistance) and runs with distilled water verifies the manner in which the liquid resistance was taken into account for absorption into distilled water.

The gas temperature varied only slightly in this investigation (23.0° to 27.2°C.), and this variation did not cause any consistent trend of the gas-phase coefficient. The maximum variation of the Schmidt number to the two-third power was only 5%, and allowance for this variation does not significantly affect the results shown in Figure 5.

In correlating the data in Figure 5, it was assumed that the mass transfer rate was not a significant variable. In Figure 6, modified Stanton numbers $k_y Y_{BM}/G_M$ calculated from the absorption data are plotted against the mass transfer rate. The horizontal line is derived from the average value for $k_y Y_{BM}/G_M$. It is evident from Figure 6 that within the precision of the data the mass flux is not a significant variable in this case.

CALCULATION OF THE GAS-PHASE COEFFICIENTS

The experimental data gave directly the overall absorption coefficients. The gas-phase coefficients were calculated by subtracting from the overall resistance the liquid resistance, as determined from data in which the gas-phase resistance was negligible. Allowance was made for temperature effects in the liquid due to the heat of absorption. The equilibrium data used were those of Sherwood (*12*) and Morgan and Maass (*9*). In the range of mean inert mole fractions from 0.4 to 0.6 the liquid resistance accounted for about 5% of the total resistance for water as absorbent and for 20 to 35% of the total resistance for ammonia solutions as absorbent. The inert concentration was taken to be that of nitrogen alone, since molecular diffusion theory predicts that the small amount of water vapor is swept along with little resistance to the tranfser of ammonia.

DISCUSSION

The theoretical analysis leading to Equation (9) and the experimental results indicate the importance of the net transfer in turbulent diffusion as well as in molecular diffusion. For the combined processes of molecular and eddy diffusion in the net transfer of one component, the gas-phase absorption coefficient is inversely proportional to the first power of the mean inert concentration. Fur-

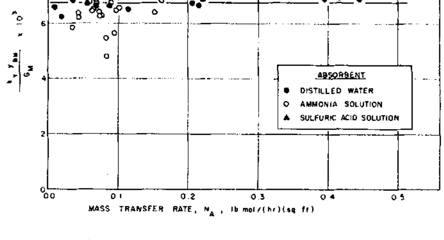

Fig. 6. Effect of mass transfer rate on absorption in a short-wall column.

thermore, the results appear to justify the assumptions regarding uniform conditions along the short column and a turbulence pattern which is independent of the mass transfer flux in the range covered in these experiments.

The findings of previous investigators permit the extension of these results. Gilliland and Sherwood (*7*) and Zabban and Dodge (*20*) found the gas-phase coefficient k_{y^*} to be independent of the total pressure in such widely different apparatus as the wetted-wall column and the packed column. As a consequence, the product $k_y y_{BM}$ can be expected to be independent of the total pressure. It can also be expected that the product $k_y y_{BM}$ is reasonably independent of temperature for given gas flow and diffusion characteristics. These considerations substantiate the previously suggested form for the *J*-factor correlation (*13*):

$$J = \left(\frac{k_y y_{BM}}{G_M} \right) (N_{Sc})^{2/3} = \psi(N_{Re})$$

The nature of the gas-phase boundary layer in a wetted-wall column is relatively simple compared with that found in most industrial contacting equipment. It would be expected, however, that the same relationship between the gas-phase coefficient and the concentration level would apply to more complicated gas flow patterns. The nature of the boundary layer and the turbulence level which determine E_M in Equation (9) determine in effect the value of the coefficient k_{y^*}, but for given flow conditions the product $k_y y_{BM}$ would still be expected to equal k_{y^*}. A variable in addition to the Reynolds and Schmidt numbers which could have an effect on the relation $k_y y_{BM} = k_{y^*}$ in industrial contacting equipment is the tranfer flux, as points of flow transition and separation are sensitive to the direction and rate of the cross-flow velocity. Hence, the rate of mass transfer may be a more important variable in commerical equipment than in the wetted-wall column. Such an effect, however, would be a mass transfer rate effect on turbulence patterns and not a concentration effect.

CONCLUSIONS

In the design of chemical engineering contacting equipment in which a solute is transferred through a nontransferring gas (evaporation, partial condensation, and absorption), the gas-phase coefficient can be expected to be inversely proportional to the first power of the mean concentration of the nontransferring gas. Apparently, the relationship for turbulent transfer is similar to that which has long been known for molecular diffusion. The effect of the absorption rate upon the gas-phase coefficient was not significant in this investigation.

ACKNOWLEDGMENT

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NOTATION

C_T = total gas concentration, lb.-moles/cu. ft.
D_v = diffusivity of solute in the vapor, sq. ft./hr.
D_L = diffusivity of solute in the liquid, sq. ft./hr.
E_M = eddy diffusivity in the gas phase, sq. ft./hr.
G_M = molar gas rate through column, lb.-moles/ (hr.) (sq. ft.)
K_y = overall gas-phase mass transfer coefficient, lb.-moles/(hr.) (sq. ft.)
N_A = mass transfer rate, lb.-moles/(hr.) (sq. ft.)
P = pressure, atm.
U = mole-average velocity, ft./hr.
V = mass-average velocity, ft./hr.
d = diameter of column, ft.
h = height of column, ft.
k = mass transfer coefficient, lb.-moles/(hr.) (sq. ft.) (driving force)
k_c = based on concentration driving force
k_G = based on partial pressure driving force
k_y = based on mole fraction driving force
k_{y^*} = based on mole fraction driving force, zero solute concentration
k_L = liquid-phase coefficient, cm./sec.
p_{BM} = logarithmic mean inert partial pressure, atm.
x,y = space coordinates, ft.
y_A = mole fraction of solute in the vapor
y_{BM} = logarithmic mean of the inert mole fraction
J factor = $N_{St} \times N_{Sc}^{2/3}$
N_{Re} Reynolds number = $dV\rho/\mu$
N_{Sc} Schmidt number = $\mu/\rho D_v$
N_{St} Stanton number = $k_y y_{BM}/G_M$

Greek Letters

Γ = liquid flow rate, g./(cm.) (sec.)
δ_L = length of diffusion path, ft.
δ_M = boundary-layer thickness, ft.
μ = molecular viscosity, lb./(ft.) (hr.)
ρ = gas density, lb./cu. ft.

Subscripts

1 = gas-liquid interface
2 = gas entering wetted-wall column

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