f = friction factor, defined by Equation (21)
$\mathbf{f}$ = external body force vector per unit mass
L = length of section of pipe considered in Section 4
N_1, N_2, N_3, N_4 = dimensionless groups defined by Equations (8), (11), and (16), respectively
p = pressure
P^* = defined by Equation (7)
r = radial cylindrical coordinate
R = characteristic length. Radius of tube in Section 4
t = time
$\mathbf{t}$ = stress tensor
t^*_{rz} = dimensionless physical component of stress tensor in cylindrical coordinates
$\mathbf{v}$ = velocity vector
V = characteristic scalar having the dimensions of velocity. The average axial component of velocity in Section 4
$x_i(\mathbf{x})$ = coordinates in an arbitrary, fixed, rectangular, Cartesian frame of reference
$X_i(\mathbf{X})$ = coordinates (in perhaps some other rectangular, Cartesian frame) of the material which we find around the position $\mathbf{x}$ at time t
z = axial cylindrical coordinate

Greek Letters

δ_{ij} = Kronecker delta
θ = cylindrical coordinate
μ = natural viscosity for a Noll simple fluid (7)
ρ = density
τ = natural time for a Noll simple fluid (7)
$\tau^{*-1}(N_3\tau^*)$ = dimensionally invariant (7) inverse of one of the three material functions that describe the behavior of a Noll simple fluid in a viscometric flow (2 to 5)
ϕ = potential in terms of which the body force vector per unit mass is defined, Equation (5)
ω = frequency of vibration in Equation (13)

Superscript

$*$ = a dimensionally invariant quantity. All lengths, velocities, times, and stresses are made dimensionless with respect to R, V, τ, and μ/τ, respectively
∇ = del operator

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Mass Transfer at High Mass Fluxes:

Part I. Evaporation at the Stagnation Point of a Cylinder

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An experimental method for measuring mass transfer rates at high mass fluxes is developed for the case of evaporation at the stagnation point of a cylinder. Evaporation of carbon tetrachloride at the stagnation point was measured and determined to be consistent with theoretical predictions. It was observed that water exhibits a significant interfacial resistance to evaporation at high transfer rates. Evaporation coefficients based on the discrepancy between experimental determinations and theoretical expectations were found to be a strong function of temperature and to be consistent with the majority of reported measurements made at lower temperatures. The experimental method should be useful for application to other configurations and flow conditions for studying mass transfer at high mass fluxes.

Considerable experimental and theoretical efforts have been directed to the study of mass transfer phenomena. Unfortunately, few experimental studies investigating the transport process at high concentration levels and at high transport rates have been reported. Cairns and Roper (*5*) studied simultaneous heat and mass transfer in a wetted wall column using the air-water system at conditions where high concentration levels were achieved over much of the apparatus. Westkaemper and White (*21*) investigated the effect of concentration level on the evaporation of carbon tetrachloride from a free liquid surface into air in turbulent flow. More recently, Schulman and Delaney (*19*) measured the evaporation of carbon tetrachloride into air in a packed column as a function of concentration

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level. Behrmann and Vivian (*4*) examined the effect of concentration level on the absorption of ammonia by water in a short wetted wall column. Unfortunately, these experimental studies were primarily directed toward delineating the effect of concentration level, and did not examine the influence of high mass transfer rates as such. Of more relevance to the latter question are the investigations of Johnson and Hartnett (*14*) for a cylinder and of Mickley et al. (*16*) for a flat plate configuration. In each case, heated air was injected through a porous surface into a lower temperature air stream. Presumably these investigations of the effect of high mass flux on heat transfer should be applicable to the mass transfer problem.

An experimental program was undertaken for the purpose of developing a reliable method by which the combined effects of high concentration levels and high mass transfer rates could be studied. The initial phase of this program (*15*), as reported in this paper, was devoted to the study of evaporation from the forward stagnation point of a cylinder. This system was chosen because the simplicity of its theoretical description provided a convenient means of evaluating the experimental procedure. During the course of the investigation, the influence of interfacial resistance on phase equilibria, particularly at high mass transfer rates, became apparent; accordingly, considerable emphasis was directed to this aspect of the problem.

EXPERIMENTAL APPARATUS AND PROCEDURE

A conventional recirculating wind tunnel of 6 in. x 6 in. cross section was employed for these studies. Flow could be regulated by speed control on the blower and by a damper valve arrangement permitting a range of control from once-through operation to complete recycle. Local air velocities were measured with a United Sensor pitot static tube whose pressure differential was determined, to an accuracy of ± 0.0002 in. of water, by a Flow Corporation micromanometer, model MM3. Nine 250-watt finstrip heaters controlled by a Leeds and Northrup unit were used to maintain the desired air temperatures within ± 0.2°F. up to temperatures of 190°F.

The test cylinder, Figure 1, consisted of a 1-5/16 in. diameter brass tube slotted to receive a ¾ in. × 6 in. strip of 400-mesh stainless steel screen. Care was taken to insure that the screen and the cylinder were electrically insulated from one another by interposing a film of epoxy resin. Six thermocouples, made of 36-gauge copper-constantan wire by arc welding, were soldered to the underside of the screen. Great pains were taken to minimize the size of the soldered connections with the result that the connections were less than 12 mils in diameter. The energy required for evaporation was provided by passing a.c. current along the length of the screen.

The system maintaining liquid feed to the test cylinder consisted of a liquid reservoir connected in parallel with a calibrated burette. Tygon and Teflon tubing were used for interconnections for the water and carbon tetrachloride experiments respectively. The reservoir was arranged so that its liquid level relative to the test cylinder could be adjusted to maintain the desired hydrostatic head. Evaporation rates were determined by measuring the time required for particular volumes of fluid to be drawn from the burette while the reservoir was isolated from the cylinder.

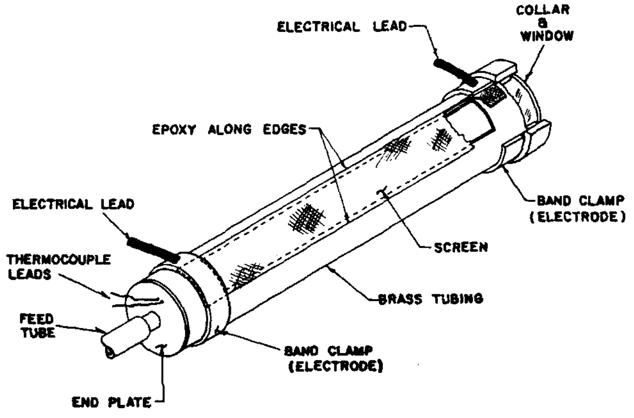

Fig. 1. Test cylinder.

The first step in preparing the system for an experimental run was to fill the reservoir with triple-distilled water or reagent grade carbon tetrachloride. The test liquid was brought to a boil with an immersion heater and allowed to degas for about 5 min. Subsequently, the liquid was allowed to flow into the test specimen positioned with the screen facing upward to permit the air to escape. The cylinder was then rotated 90 deg. in the upstream direction so that the centerline of the screen coincided with the forward stagnation point of the cylinder. The level of the reservoir was then lowered until it was about 1 in. below the screen and was maintained at this position manually throughout the experiment. The capillarity of the screen sufficed to keep air from penetrating into the cavity and also provided the suction required to replace the evaporating fluid. Preliminary tests indicated that the suction or negative hydrostatic head could be varied over a wide range without affecting either the evaporation rate or the surface temperature.

With the cylinder filled and in position, the blower was started and the speed was adjusted. The temperature controller and heating units were also put into operation. With the exception of several "adiabatic" runs made to obtain mass transfer coefficients at very low mass fluxes, all measurements were conducted under essentially isothermal conditions with the measured screen and air temperatures nearly equal. The term, *adiabatic* is somewhat of a misnomer in this application, since the nonevaporative brass surface reached a temperature close to that of the recirculating air and as such provided an additional energy source for the evaporating fluid. It is not surprising that the measured evaporative surface temperatures for these measurements were several degrees above the predicted wet bulb temperatures. Once steady state was reached, the screen, air and wet bulb temperatures, and the volumetric evaporation rates were measured.

RESULTS AND DISCUSSION

Treatment of the Experimental Data

The mass flux at the wall is given by

$$m_1 = -\rho_w D \frac{\partial w}{\partial y}\bigg|_w + \rho_{1w} v_w \qquad (1)$$

where the first term on the right is the diffusive flux and the second term is the convective flux.

The mass transfer coefficient is generally defined in terms of the diffusive flux by

$$k = -\frac{D \dfrac{\partial w}{\partial y}\bigg|_w}{w_w - w_\infty} \qquad (2)$$

Substituting (2) and the relation $\rho_{1w} v_w = w_w m_1$ into (1)

$$k = \frac{m_1(1 - w_w)}{\rho_w (w_w - w_\infty)} = \frac{m_1}{\rho_w B} \qquad (3)$$

The evaporation rate (m) is obtained directly from the experimental data, while the mass transfer (or blowing) parameter B and the density of the fluid at the wall ρ_w are determined from a knowledge of the wall concentration. In evaluating the experimental results of this study, the concentration at the wall was obtained from the as measured surface temperature by assuming vapor-liquid equilibrium at the interface. It will subsequently be shown that this assumption is not valid in the case of water at high mass transfer rates.

Typical experimental and calculated results* for the evaporation of carbon tetrachloride and water are presented in Table 1. The results are shown in Figures 2 and 3 in the form of $N_{Sh}/(N_{Re})^{1/2}$ as a function of B.

The transport process for the experimental system under consideration can also be approached from a theoretical point of view. In brief, the momentum and diffusion laminar boundary layer equations are to be solved for the case of a constant properties fluid. Falkner and Skan (*9*) have considered this problem for an arbitrary geometry and have deduced a similarity variable by which the partial differential momentum and diffusion equations can be transformed into ordinary differential equations. The transformed equations have since been solved numerically by several investigators. For this work, the solutions obtained by Stewart and Prober (*20*) were used because of the wide range of Schmidt numbers covered in their paper.

In more detail, Equation (2) can be rewritten in dimensionless form as

$$k = -\frac{D}{l(x)}\frac{d\theta}{d\eta}\bigg|_w \tag{4}$$

where $\eta = y/l(x)$

and $l(x)$ as determined by Falkner and Skan is given by

$$l(x) = \frac{(2-\beta)\nu x^{1/2}}{U_\infty}$$

By the definition of the Sherwood number and noting that $\beta = 1$ for stagnation flow, one obtains

$$\frac{N_{Shx}}{(N_{Rex})^{1/2}} = -\frac{d\theta}{d\eta}\bigg|_w \tag{5}$$

In addition,

$$\frac{N_{Shx}}{(N_{Rex})^{1/2}} = \frac{kx}{D} \bigg/ \frac{U_\infty x}{\nu} \tag{6}$$

The velocity at the outer edge of the boundary layer in the neighborhood of the stagnation point of a cylinder may be written as

$$U_\infty = 2U_0 \sin\gamma$$

where γ is the angle subtended between the point of interest and the stagnation point and U_0 is the approach velocity. For small γ, the result can be further simplified to

$$U_\infty = \frac{4U_0x}{d}$$

by means of which (6) can be modified to

$$\frac{N_{Shx}}{(N_{Rex})^{1/2}} = \frac{1}{2}\frac{N_{Sh}}{(N_{Re})^{1/2}} \tag{7}$$

Combining (5) and (7), the desired result is obtained as

$$\frac{N_{Sh}}{(N_{Re})^{1/2}} = -2\frac{d\theta}{d\eta}\bigg|_w \tag{8}$$

It remains to show that the conditions under which the experimental measurements were conducted are compatible with the restrictions of the mathematical analysis, the results of which are to be employed for comparison. The transformation of Falkner and Skan requires that the system must be of wedge flow, and that the normal velocity v_w must be a prescribed function of the wedge angle. For the experimental system, the wedge flow requirement is satisfied at the stagnation point. For this case, v_w must be constant; this condition is satisfied by the constant evaporation rate characteristic of the test specimen. The one regard in which the experimental studies do not strictly conform is the constant fluid property requirement. In this connection, variations in property for the water studies are quite nominal with respect to the measurement of the mass transfer coefficient and can be disregarded. For the carbon tetrachloride case, property variations are significant and corrections must be made. The approximate method proposed by Hanna (*10*) was used for the purpose of converting the experimental vari-

TABLE 1. EXPERIMENTAL AND CALCULATED RESULTS

Run	Average surface temp., °F.	Air temp., °F.	Evap. rate,* m lb./(hr.)(sq. ft.)	W_w‡	U_o, ft./sec.	N_{Sc} (avg.)	k, ft./hr.	N_{Sh}	N_{Re} (avg.)	ϕ	$\frac{N_{Sh}}{\sqrt{N_{Re}}}$** (corrected)
					Carbon tetrachloride						
1	51.3	80.4	15.0	0.305	26.9	1.71	330	137.6	22,500	0.864	1.061
2	89.1	89.8	41.4	0.575	28.1	1.30	225	81.8	28,200	0.719	0.680
3	101.8	99.9	54.0	0.662	28.6	1.15	179	62.2	31,200	0.663	0.533
4	113.4	107.7	53.2	0.740	17.5	1.02	108	36.2	20,800	0.611	0.411

Run	Average surface temp., °F.	Air temp., °F	Wet bulb temp., °F.	Evap. rate,* m lb./(hr.)(sq. ft.)	W_w‡	W_∞	U_o, ft./sec.	k, ft./hr.	N_{Sh}	N_{Re} (avg.)	$\frac{N_{Sh}}{\sqrt{N_{Re}}}$
						Water					
5	100.8	99.4	65.7	2.58	0.0424	0.0058	21.4	977	98.5	13,000	0.864
6	110.0	109.0	75.9	3.30	0.0560	0.0116	21.2	1040	101.8	12,570	0.909
7†	110.4	109.5	73.0	2.98	0.0550	0.0090	21.8	906	89.0	12,900	0.784
8	129.5	129.0	83.4	5.28	0.0991	0.0143	21.6	882	80.8	12,100	0.735
9†	129.7	129.7	79.0	5.16	0.0945	0.0099	22.1	865	79.5	12,400	0.715
10	155.2	154.3	90.9	9.40	0.2015	0.0169	22.0	705	59.5	11,450	0.557
11	151.0	153.6	90.8	9.05	0.1802	0.0170	29.0	773	66.1	15,300	0.535
12	169.2	168.2	95.7	12.4	0.2950	0.0196	22.0	592	47.9	11,200	0.453

* Surface area of the bare screen = 3.75 sq. in. Surface area of lens tissue = 4.50 sq. in.
† These data were obtained with lens tissue over screen.
‡ Assumes vapor-liquid equilibrium prevailed at the interface.
** Corrected for density variation by Equation (10).

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* Tables 1 and 2 have been deposited as document 8405 with the American Documentation Institute, Photoduplication Service, Library of Congress, Washington 25, D. C., and may be obtained for $1.25 for photoprints or 35-mm. microfilm. Table 1 summarizes the data for the evaporation of water. Table 2 summarizes the carbon tetrachloride results.

able density mass transfer coefficients to the corresponding constant density coefficients by the following expression

$$\phi = \frac{k_{\text{variable density}}}{k_{\text{constant density}}} = \frac{\ln \dfrac{M_w}{M_\infty}}{\dfrac{M_w}{M_\infty} - 1}$$

The values of the correction factor are listed in Table 1. It is to be noted that the maximum correction for density amounts to 61% at the highest value of B and is appreciably smaller at low B. At present there do not appear to be methods for correcting for the variation of viscosity. However, it is believed that this effect is less serious than that of density.

The theoretical curves shown in Figures 2 and 3 are based on the numerical results of Stewart and Prober and Equation (8). In the case of water, the curve shown corresponds to a $N_{Sc} = 0.7$. However, the Schmidt number is a strong function of B for the air-carbon tetrachloride system. In this instance the average film N_{Sc} is determined only by the evaporative surface temperature, since the concentration of carbon tetrachloride in the free stream was always zero. The theoretical curve was obtained by calculating the N_{Sc} for specific values of B; and obtaining the dimensionless gradient $\left.\dfrac{d\phi}{d\eta}\right|_w$ for the specific combination of B and N_{Sc} by crossplotting Stewart and Prober's results. Accordingly, the theoretical curve shown on Figure 2 is specific for carbon tetrachloride evaporating into pure air.

Analysis of the Carbon Tetrachloride Data

In Figure 2, the experimental and theoretical results for the evaporation of carbon tetrachloride based on interfacial equilibrium are compared. Although the agreement is, in general, satisfactory, with a deviation of generally less than 10%, the experimental data with but one exception fall slightly below the theoretical curve. Assessment of the reliability of the experimental data requires consideration of: (1) uncertainties between the actual and measured surface temperature, (2) the accuracy of the approximate density correction applied to the data, and (3) method employed to obtain an average film N_{Sc}.

In regard to the temperature uncertainty, a simple unidirectional heat conduction model was employed to estimate the equivalent thickness of a film of liquid carbon tetrachloride required to account for the discrepancy between the experimental and theoretical results. Upon applying the model (which reflects the heat load and conductivity of the fluid) to the data, an equivalent film thickness of approximately 0.75 mils was obtained over the entire range of the data. The uniformity of the calculated film thickness and the fact that it corresponds to approximately one-half the screen mesh opening suggests that a temperature drop between the thermocouples and the actual liquid surface is probably responsible for the discrepancy.

The absolute reliability of the approximate density correction is difficult to ascertain for the case at hand. Comparison of the approximate correction with an exact high N_{Sc} solution (*11*) indicates that Hanna's factor may overestimate the coefficient by 10% at the higher carbon tetrachloride mass fluxes. The magnitude of the error at low N_{Sc} is not known at the present time.

There remains the question as to the basis on which to evaluate the Schmidt and Reynolds numbers. It was found by trial and error that the closest agreement between the experimental and theoretical results for carbon tetrachloride, over the entire range of the measurements, was obtained when these numbers were evaluated on average film properties.

In summary, the experimental method is confirmed by the agreement between measured and predicted mass transfer coefficients at high mass fluxes. The results support the conclusions of previous investigators (*22*) that carbon tetrachloride does not exhibit interfacial resistance, that is the evaporation coefficient is essentially unity. Finally, experimental evidence of the appropriateness of the approximate density corrections has been obtained.

Analysis of the Water Data

In contrast to the general agreement observed in the case of carbon tetrachloride, the experimental data for the evaporation of water fell significantly below theoretical expectations (Figure 3). In addition, the degree of departure from theoretical predictions becomes greater at the higher evaporation rates. Furthermore, correction of the surface temperature, using the simple conduction model and parameters described earlier, produces only a slight effect in the desired direction. Two possible explanations for the disparity present themselves: (1) contamination of the surface, and (2) interfacial resistance to mass transfer. Examination of the sequence in which the experiments were conducted reveals that the observed results would not be a consequence of accumulation of impurities as a function of prolonged evaporation. Furthermore, the contamination would have to occur at such rates and to exert such an effect as to produce evaporation results uniquely related to B. On these grounds, it was con-

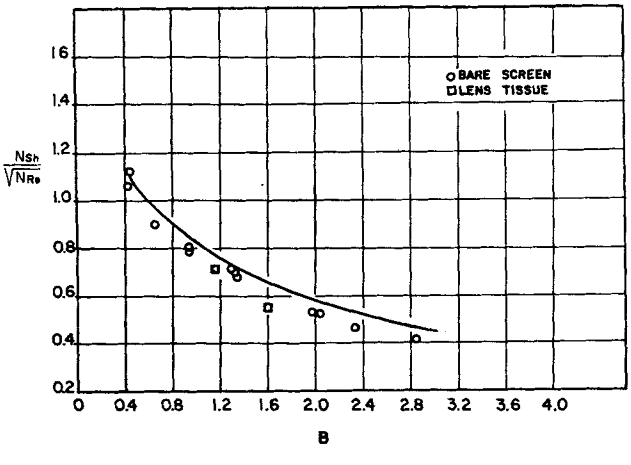

Fig. 2. Comparison of experimental mass transfer data with theory (*solid line*) for carbon tetrachloride.

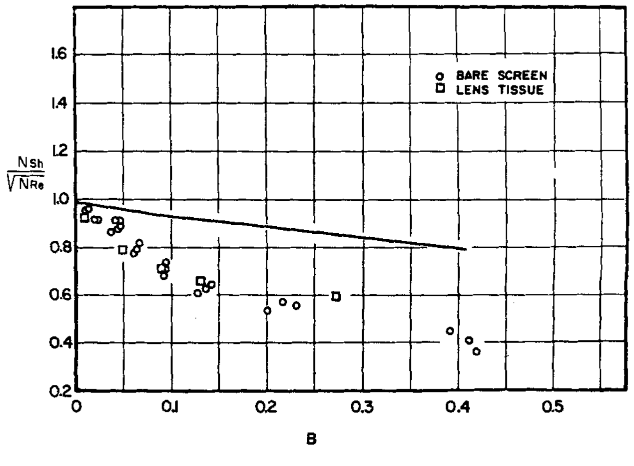

Fig. 3. Comparison of experimental mass transfer data with theory (*solid line*) for water.

cluded that surface contamination was not responsible for the observed results.

The potential role of interfacial resistance was deduced from the experimental results by applying the Hertz-Knudson equation (*1*) in the following form

$$m_1 = \alpha \sqrt{\frac{M_1}{2\pi RT}}\,(P_s - P_a) \qquad (9)$$

The evaporation coefficient α was determined using the experimentally measured evaporation rate, system temperature, and vapor pressure at the surface temperature, by forcing the actual partial pressure at the interface P_a to that value required to match the experimental and theoretical results. The calculated evaporation coefficients are compared in Figure 4 with results of other investigators as a function of temperature. The first impression gained from Figure 4 is the marked dependence of the evaporation coefficient on the temperature. In the low temperature range (32° to 90°F.) most of the data have been supplied by Alty and co-workers (*1, 2, 3*). Although these data are widely scattered, they support the temperature dependence observed in the present study. Also shown in Figure 4 is the measurement reported by Hickman (*12*), which was obtained using a high-velocity water jet. In his analysis, Hickman proposed that the actual value of the evaporation coefficient for the condition of his measurement may have been greater than the reported 0.25, perhaps approaching unity. However, he noted that experimental uncertainties did not permit this conclusion to be drawn. He explained the abnormally high value of α by the premise that the highly sheared surface and short exposure time precluded the accumulation of surface impurities and hindered the formation of an ordered surface structure at the molecular level. Hickman (*13*) subsequently studied evaporation of water from stagnant quiescent surfaces and concluded that the very small evaporation coefficients he noted qualitatively were due to the static state of the surface.

In the high temperature range (100° to 212°F.) the extensive set of data provided by Campbell (*6*) is in substantial agreement with the results of the present study. Of interest in this comparison is that Campbell's estimate of the evaporation coefficient was based on the evaporation of water from a flat plate into air in fully developed turbulent flow in a duct. It is to be noted that Campbell's experimental technique, namely evaporation from an electrically heated screen, was similar to that employed in this study. Nevertheless, the general agreement between evaporation coefficients as determined from measurements in markedly differing flow regimes would appear to support the results of this study.

In the present study it was not possible to control the evaporation rate independently of surface temperature. Accordingly, one cannot use these results to differentiate between the potential effects of temperature and mass flux. However, Campbell was able to regulate the concentration level in the recirculating air and thus separate the effects of these variables. His results indicate that the evaporation coefficient is not related to the mass flux over the range investigated experimentally.

Other data in the high temperature range are scarce. Pruger (*17*) measured the rate of evaporation of water from a flat horizontal surface utilizing two thermocouples, one situated above the liquid surface and a second below the surface. The position of the second thermocouple was determined microscopically, providing the data required for extrapolation to the actual surface temperature. Pruger observed that the liquid surface appeared to be superheated by several hundredths of a degree. His value of the evaporation coefficient of 0.02° at 212°F., deduced from the evaporation rate and the temperature rise, is high relative to the present results. The extreme accuracy requirements of his method, in combination with the uncertainties associated with the temperature extrapolation, tend to reduce the confidence with which the reported value is to be regarded. Silver and Simpson's studies (*18*) utilized the flashing of saturated water in a nozzle. Their reported values for α varied from 0.0006 to 0.008, indicating possible weaknesses in the simplified two-phase flow model used to analyze the data.

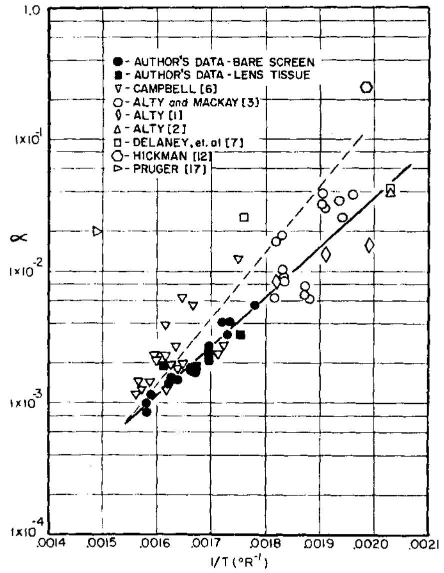

Fig. 4. Evaporation coefficient of water.

In assessing the reliability of the calculated evaporation coefficients, several factors must be taken into consideration. Foremost of these is the question of the theoretical curve with which the experimental mass transfer data are compared. As noted earlier the coefficients presented in Figure 4 for the air-water system are based on the theoretical curve for $N_{Sc} = 0.7$ rather than for the actual N_{Sc} of 0.6. In the light of the known effect of turbulent intensity on stagnation point heat transfer coefficients, it appears that extrapolation of the experimental data to zero mass flux provides the soundest method for establishing the theoretical curve. For these measurements the extrapolated value corresponds closely to the $N_{Sc} = 0.7$ asymptote. On the other hand it is possible that the experimental data are subject to some systematic error; in this case, an error of 5% would be sufficient to bring the experimental data into agreement with the theoretical asymptote for $N_{Sc} = 0.6$. Evaporation coefficients based on the $N_{Sc} = 0.6$ curve correspond to the dotted line shown in Figure 4. The effect is greatest ($\sim + 100\%$) at the lowest mass fluxes because of the extreme sensitivity of the calculation and amounts to only + 5% at the opposite extreme. It is not possible at this time to put forth arguments that can reduce the indicated uncertainty. However, it is worth noting that the potential error at high temperatures is small, and that evaporation coefficients calculated on the $N_{Sc} = 0.7$ basis are more consistent with the majority of values obtained by other investigators at lower temperatures.

The potential error associated with using the measured screen temperature, as indicated by the thermocouples, as the actual surface temperature can be estimated from an analysis of the carbon tetrachloride data. The simple unidirectional heat conduction model described earlier was used to estimate the effective thickness of the film resisting heat transfer at about 0.75 mils. As applied to the water experiments, this model indicates a maximum temperature error of 2°F. at the highest mass flux, run 10, and a corresponding increase in of 10%. At lower mass fluxes, the estimated temperature error is smaller but its effect on α is greater. By way of example, for run 5 the estimated temperature error is only 0.5°F. but the corresponding change in α is 20%. It is clear that the impact of experimental uncertainties is progressively more profound as evaporation rates and gradients become smaller. This may in large measure account for the extreme difficulties in replicating experimental values at low temperatures. On the other hand, the problem of measuring evaporation coefficients is eased at high temperatures where high mass fluxes can be achieved.

While the effect of variation in fluid density on the mass transfer coefficient is slight in the case of the water, $\phi = 1.01$ at low B and 1.03 at high B, the effect on the evaporation coefficient is of the same order as that of the temperature uncertainty but is in the opposite direction. At low B levels, the temperature and density factors essentially cancel; at high B, the net effect is to increase the coefficient by about 7%.

Finally, it is to be noted that the theoretical curve is for transport at the stagnation point, while the experimental data represent the average over an arc subtended by a 30 deg. angle on either side of the stagnation point. The experimental data should be increased between 2 to 3% for this reason. Applying this adjustment to the carbon tetrachloride data would reduce the required surface temperature correction for both fluids. Accordingly, the effect of the temperature error on the evaporation coefficients for water would be less than that described earlier. However, there would be a small residual effect due to the difference between average and local transfer rates at low temperatures.

The question of the nature of the surface from which the evaporation occurs was also considered. In this regard measurements made for both water and carbon tetrachloride, with and without a 1.5-mil thick lens tissue covering the screen, failed to reveal any significant effect on the results if the temperature drop through the tissue was taken into account.

Of the several uncertainties considered, the most significant by far is that related to the identification of the appropriate theoretical curve with which to compare the data. Despite this, however, the calculated evaporation coefficients as presented in Figure 4, are well grouped, show a strong correlation with temperature, and are consistent with the majority of measurements reported by other investigators. Finally, the consequences of the uncertainties are progressively less serious with increasing temperature and are estimated at less than 10% at the highest temperatures. The correlating line drawn on Figure 4 through the data points is given by the following equation

$$\alpha = 10.4 \times 10^{-10}\, e^{17,230/RT} \qquad (10)$$

Water is not unique in exhibiting a marked interfacial resistance at elevated evaporation rates. Wyllie (22) has summarized measurements for a number of materials and reports that polar and associated liquids tend to have low evaporation coefficients. The question as to the role played by the state of the liquid surface, (that is newly-formed as in the case of highly sheared surfaces or aged as in the case of static pool evaporation) remains unanswered. The relationship between the evaporation coefficient and temperature remains to be determined for materials in general. The success achieved in this study, by using a configuration amenable to theoretical analysis, suggests that experimental methods based on this approach can be developed to provide accurate and reliable data.

One final observation regarding the water evaporation measurements deserves some comment. It is to be noted (Table 1 and Figure 3) that the $N_{Sh}/\sqrt{N_{Re}}$ factor is a slight function of N_{Re}. For a given value of B, high N_{Re} data tend to lie below the low N_{Re} measurements. A similar tendency has been reported by Eckert (8) in connection with heat transfer coefficients in the neighborhood of the stagnation point. An explanation for this weak dependence on N_{Re} does not appear to be forthcoming from theoretical arguments.

CONCLUSIONS

An effective experimental method for studying transport processes at high mass transfer rates was developed. The evaporation of carbon tetrachloride at the stagnation point of a cylinder normal to the main motion of the fluid was measured and determined to be consistent with theoretical predictions. It was observed that water exhibited a significant interfacial resistance at high evaporation rates. Evaporation coefficients based on the discrepancy between experimental determinations and theoretical expectations were found to be a strong function of temperature and to be consistent with the majority of reported measurements made at lower temperatures.

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NOTATION

B = $(w_w - w_\infty)/(1 - w_w)$
D = diffusion coefficient, sq. ft./hr.
d = cylinder diameter
k = mass transfer coefficient, ft./hr.
$l(x)$ = $(2 - \beta)\, \nu x/U_\infty{}^{1/2}$, ft.
M = molecular weight, lb.-mole
m = evaporation rate, lb./sq. ft.-hr.
N_{Rex} = $\dfrac{U_\infty x}{\nu}$, Reynolds number based on x and local velocity
N_{Re} = $\dfrac{U_o d}{\nu}$, Reynolds number based on diameter and approach velocity
N_{Sc} = $\dfrac{\mu}{\rho D}$, Schmidt number
N_{Shx} = $\dfrac{k x}{D}$, Sherwood number based on x
N_{Sh} = $\dfrac{k d}{D}$, Sherwood number based on diameter
R = gas constant
T = absolute temperature °R.
U = velocity, ft./hr.
v = velocity component normal to surface, ft./hr.
w = concentration, mass fraction
x = distance along the surface measured from the stagnation point, ft.
y = distance measured normal from the surface, ft.

Greek Letters

α = evaporation coefficient, dimensionless
β = constant defined by the wedge angle $(\pi\beta)$ in wedge flow
η = $y/l(x)$, transformation variable
θ = $(w - w_\infty)/(w_w - w_\infty)$, dimensionless concentration

μ = dynamic viscosity, lb./ft.-hr.
ν = kinematic viscosity, sq. ft./hr.
ρ = density, lb./cu. ft.
ϕ = k variable density/k constant density = $\left(\ln \frac{M_w}{M_\infty}\right) \Big/ \left(\frac{M_w}{M_\infty} - 1\right)$

Subscripts

1 = diffusing component
w = evaluated at wall
o = approach conditions
∞ = evaluated at outer edge of boundary layer

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Performance of a Packed Distillation Column with Electrically Heated Packing

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Electric current was applied across the coke packing of a glass column, distilling carbon tetrachloride-benzene at total reflux. Heat generated by electricity boiled the liquid on the packing surface. Bubbles rising across the liquid film reduced its resistance to mass transfer, improving the column efficiency.

Theoretical analysis and experimental results are presented.

Attempts to improve the efficiency of a packed distillation column have been directed mostly to the development of new packing materials and to methods providing better distribution of the liquid. The literature contains numerous recommendations as to the best types of packing elements and distributors to use. Little or no work seems to have been directed toward inducing turbulence to the liquid layer on the packing surface. If a suitable method were devised in this respect, it would help reduce the liquid film resistance and improve the efficiency. The results of Furnas and Taylor (*6*) on a water-ethanol system and Yoshida et al. (*12*) on a benzene-toluene system and using ordinary packed columns at various reflux ratios, show that in the rectifying section of a packed column, the rate of mass transfer is controlled by the liquid film resistance. This conclusion is supported further by the results of Hands and Whitt (*7*) and Edye (*4*) on columns packed with small rings. It is, therefore, reasonable to assume that the mixing of the liquid layer should improve the efficiency, particularly in the rectifying section.

The simplest method to induce turbulence to the liquid in a packed column is to increase its flow rate. The net effect, however, is rather small, as can be seen from the empirical expressions for the liquid film coefficient presented by Brown and Rosenberg (*1*) and Morris (*11*) for design purposes.

An interesting paper by Kafarov (*8*) describes another method by which a violent mixing of the liquid is achieved by *emulsification* through the penetration of bubbles of vapor. This condition is said to have provided, at the same time, a fast equalization of liquid concentration, an even distribution, and a longer residence time, all of which helped to raise the efficiency. Unfortunately, the operating limits for such a column appear to be narrow.