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Publication Date

1967-11-01

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Proceedings of American Institute of
Chemical Engineers; Instituto de
Ingenieros Quimicos de Puerto Rico,
2nd Joint Meeting, Tampa, Florida,
May 19-22, 1968; for publication in
American Institute of Chemical Engineers
Journal

UCRL-17655
Preprint

UNIVERSITY OF CALIFORNIA
Lawrence Radiation Laboratory
Berkeley, California

AEC Contract No. W-7405-eng-48

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PART I: PURE LIQUIDS

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ABSTRACT

Normal pentane and isopentane were evaporated into nitrogen in laminar, cocurrent flow in a rectangular channel. Composition profiles within the gas phase were measured, and the evaporation rates were compared to theoretical predictions with satisfactory agreement. The interfacial mole fraction of the evaporating species in the gas phase ranged as high as 0.74. A solution is reported for the Leveque problem under high-flux conditions.

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INTRODUCTION

The rate of evaporation of a pure liquid into an adjacent laminar flowing gas stream can usually be predicted by the solution of a simplified form of the equation of convective diffusion written for the gaseous phase. A number of investigators^{5,17,18,etc.} have shown that if the vapor pressure of the fluid in question is sufficiently high, correction factors must be applied to the mass transfer coefficients which would be predicted by the simplified theory for the situation of low concentration level and low mass flux. A second problem which is encountered as the liquid volatility increases stems from the interaction of heat and mass transfer. As the mass transfer rate increases, the interfacial heat flux caused by the enthalpy of evaporation becomes significant. This heat flux brings about a reduction in the interfacial temperature of the liquid with a consequent lowering of the liquid vapor pressure at the interface. The result is a reduction in the mass transfer rate compared to that which would be predicted ignoring this effect.

The purpose of the present study has been to investigate the evaporation of a volatile liquid in an experimental system for which the fluid dynamics of both phases, as well as the low-flux and low-concentration-level transfer coefficients, are well known. Part II of this work deals with these phenomena for the situation of evaporation from liquid mixtures, where the resistance to mass transfer is divided between the liquid and gaseous phases.

Effect of Concentration Level upon Mass Transfer

In order to show the effect of the solute concentration level upon the interfacial mass transfer conveniently, let us first examine two forms of Fick's first law for a binary mixture of components A and B:

$$J_A^* = - c D_{AB} \nabla x_A \quad (1)$$

$$J_A^o = - (D_{AB} / \tilde{V}_A) \nabla \phi_A \quad (2)$$

The value of D_{AB} in the above two equations is the same, even if there are gradients in temperature and pressure, as long as $\tilde{V}_A$ remains constant and the substance to which Equation 2 is applied is incompressible and has a zero coefficient of expansion.¹²

Since the solution to most mass transfer problems requires the integration of a Fick's law expression similar to those given above, it is convenient to use Equation 1 for situations where the group $c D_{AB}$ does not vary throughout the fluid. This is found to be true for most gases at moderate pressures when temperature gradients are not large; therefore Equation 1 is the simplest form of Fick's law to use under these conditions. The conditions for Equations 1 and 2 to be equivalent are usually obeyed sufficiently closely in liquid solutions; therefore the application of Equation 2 is often found to be advantageous since a composition-dependent density does not affect that equation directly.

Both J_A^o and J_A^* are molar diffusion fluxes that are taken with respect to the fluid itself. In order to obtain a flux relative to stationary coordinates we may use the following equations:

$$N_A - x_A(N_A + N_B) = -c D_{AB} \nabla x_A \quad (3)$$

$$N_A - \phi_A(N_A + N_B \tilde{V}_B / \tilde{V}_A) = - (D_{AB} / \tilde{V}_A) \nabla \phi_A \quad (4)$$

Let us now consider Equation 3, which is useful for most binary gas mixtures at low pressure. If the value of $(N_A + N_B)$ is not equal to zero, then it is obvious from this equation that the value of the molar flux of component A relative to ^astationary coordinate, N_A , is dependent upon the concentration level of that component, x_A , as well as upon Δx_A . If Equation 3 is written at the gas-liquid interface, then the interfacial flux of A is concentration dependent; this is what is meant by the "concentration level effect." Since the value of x_A is always less than 1.0, we see that an increase in concentration level of the transferred species will result in an increase in total flux if $(N_A + N_B)$ is of the same sign as N_A ; otherwise an increase in concentration level will cause the total flux to fall.

Effect of Flux Level upon Mass Transfer

Whereas the effect of concentration level can be demonstrated by considering the magnitude of certain terms in the Fick's law expression, the effect of high flux or, equivalently, high concentration gradient enters in a more subtle manner. Usually the solution of the convective diffusion equation is carried out by first achieving a solution for the velocity profiles within the system, and then substituting these into the equation of convective diffusion; i.e., the flow equations and the mass transfer equations are solved independently and

sequentially. However, if the mass flux becomes sufficiently high, these equations can no longer be "decoupled," since a flux of mass also represents a physical flow of material and hence a velocity.

Thus a high flux mass transfer solution represents the solution to a physical problem where the mass flux of material has been considered as an additional velocity term, comparable in magnitude to the velocities present in the absence of mass transfer. This type of solution yields a different concentration profile shape from the low flux solution. On the other hand, the shape of the concentration profile is not altered by the concentration level effect; the concentration profile for a small concentration gradient at a high concentration level is similar to that for a small concentration gradient at a low concentration level.

In order to differentiate between the low and high flux solutions, it is convenient to define the local mass transfer coefficient as suggested by Bird, Stewart, and Lightfoot.²

$$k'_{x,loc} = \frac{N_{Ao} - x_{Ao} (N_{Ao} + N_{Bo})}{x_{Ao} - x_{A\infty}} \quad (5)$$

This form of mass transfer coefficient is useful for situations in which Equation 1 is the most convenient form of Fick's law, for example, for gas-phase mass transfer.

With the preceding general definition in hand, we can now define the mass transfer coefficient for the limit of low mass flux as

$$\begin{aligned} [\lim k'_{x,loc}] &= k_{x,loc} \\ (N_{Ao} + N_{Bo}) &\rightarrow 0 \end{aligned} \quad (6)$$

Since most of the existing mass transfer correlations are based on experiments performed under low flux conditions, it is convenient to predict the high flux performance by applying a numerical correction factor to the low flux results. For this purpose the following dimensionless variables are helpful:

$$R_{AB} = \frac{N_{Ao} + N_{Bo}}{k'_{x,loc}} = \frac{x_{Ao} - x_{A\infty}}{\frac{N_{Ao}}{N_{Ao} + N_{Bo}} - x_{Ao}} \quad (7)$$

$$\Phi_{AB} = \frac{N_{Ao} + N_{Bo}}{k_{x,loc}} \quad (8)$$

$$\theta_{AB} = k'_{x,loc} / k_{x,loc} \quad (9)$$

Both R_{AB} and Φ_{AB} are dimensionless expressions for the net interfacial mass flux. θ_{AB} is the correction factor which, when applied to a low flux expression for the mass transfer coefficient, yields the mass transfer coefficient at a high mass flux. Thus if one is able to obtain a solution to the high flux problem for a given flow situation, the results can be presented graphically in terms of any two of the above variables. Such solutions have been obtained for several frequently encountered flow geometries, and the graphical results can be found in Figures 21.7-2 and 21.7-3 in the text by Bird, Stewart and Lightfoot.²

Another complication that frequently arises at high flux levels is a variation of important physical properties across the concentration range between the interface and the bulk fluid.

However, this is not a necessary consequence of increased flux level, and therefore can be considered separately.

It should be noted that when the mass transfer coefficients are defined as in Equations 5 and 6, the concentration level effect is included; i.e., the value of $k'_{x,loc}$ is not a function of concentration level and varies only with changes in the value of $(\partial x_A / \partial y) / \Delta x_A$. An equation similar to 5 can also be written for other forms of Fick's law, such as that given by Equation 2.

PREVIOUS WORK

Colburn and Drew⁸ obtained a solution to the film model of mass transfer for high flux and high concentration level conditions, and developed the P_{BM} correction factor for the case where $N_{Bo} = 0$. This correction factor has frequently been included in mass transfer correlations. Colburn and Drew evaluated this correction factor for the evaporation of water into a turbulent air stream. Similar experiments concerning evaporation into turbulent gas flows were carried out by Cairns and Roper,⁵ by Westkaemper and White,²⁰ by Shulman and Delaney,¹⁷ and by Wasan and Wilke.¹⁹ These investigators arrived at various different conclusions regarding the appropriate correction factor to be applied. Vivian and Behrman¹⁸ surveyed the previous work and reported results of their own for the absorption and desorption of ammonia in a short-wetted-wall column. Their conclusion was that the P_{BM} correction factor is compatible with the available experimental data for mass transfer into a turbulent gas stream when $N_{Bo} = 0$.

There have been fewer studies of high flux and high concentration level mass transfer in laminar flows. Emanuel and Olander,⁴ Ranz and Dickson,¹⁶ and Mendelson and Yerazunis¹³ all examined situations to which laminar boundary layer theory should apply, and achieved varying degrees of success in matching their results to theoretical predictions.

The calculation of the effect of high flux level upon the mass transfer coefficients for a given system cannot be generalized in the manner used for the concentration level, since the high flux solution is dependent on the assumed fluid flow situation. The correction factors for various models do not differ greatly, however, at moderate flux levels.² At present, high flux solutions exist for three important flow geometries -- the film theory, for which the solution was first carried out by Lewis and Chang;¹¹ the penetration theory, for which the solution was obtained by Arnold;¹ and laminar boundary layer theory. A detailed account of the mathematical solution to each of the above problems has been given in the text by Bird, Stewart and Lightfoot.²

Unfortunately, none of the existing high flux solutions were directly applicable to the velocity profiles encountered in this work. Byers and King³ have shown that for short contact times, a flow geometry consisting of a finite interfacial velocity plus a linear velocity gradient away from the interface will closely approximate the exact, parabolic solution for the gaseous phase in cocurrent, gas-liquid channel flow. In view of this, a high flux solution was carried out for the case of a

linear velocity profile leading away from an immediate interface (the Leveque Model) in a semi-infinite fluid. This case and the penetration model represent the two extremes of the finite interfacial velocity-linear velocity gradient flow geometry.

Because of the complexity of the high flux Leveque problem, the equations were solved in finite difference form using the Crank-Nicholson six-point implicit formula.¹⁰ The results, when placed in the dimensionless form of θ_{AB} versus R_{AB} , are shown in Figure 1. The result for the laminar boundary layer model at a Schmidt number of 1.5 and that for the penetration model are also shown. As can be seen, the curve for the Leveque model differs very little from the results for the penetration model, and therefore either of these could be used to predict the high flux behavior for the gaseous phase of the experimental equipment, provided the contact times are not too long.

EXPERIMENTAL APPARATUS

The equipment used in this study was the same as that described by Byers and King⁴ in an earlier work. The gas-liquid contacting device (Figure 2) was a horizontal rectangular duct, with a large width-to-height ratio. The channel and the inlet and exit sections were made of Lexan polycarbonate. The gas and liquid phases passed through this channel in stratified, cocurrent, laminar flow with both the phases having equal depth. An entry section served to establish velocity profiles in each phase. The inside dimensions of the test section were: length = 18.0 in., width = 3.0 in. and height = 1.0 in. ($\frac{1}{2}$ in. in each phase). As can be seen in Figure 2, provision was made for sampling the

temperature and composition at several points along the exposure length, at $\frac{1}{2}$ in., 10 in. and 17.5 in. from the inlet. The temperature and composition probes were mounted on micrometer barrels that were inserted into the holes indicated in Figure 2. As a result these variables could be determined quite accurately as functions of vertical position.

A schematic diagram of the experimental apparatus is given in Figure 3. The evaporating liquid was continuously circulated through a closed loop system, in which provisions were made for adjusting the temperature and flow rate and for adding liquid to the system. The gas was continuously withdrawn from cylinders through a pressure regulator, was passed through a heater for temperature control, and was then contacted with the liquid phase. After the exposure the two phases were separated by a thin metallic divider plate. Additional details concerning this apparatus are given elsewhere.^{4,7}

Important changes from the apparatus used by Byers and King were the following:

- (1) The use of Chromel-Constantan thermocouple temperature probes. These probes were constructed from 5-mil diameter wires with 0.5-mil Teflon coating, threaded through 23-mil diameter stainless steel tubing. The probe was bent so that the tip end was horizontal into the gas flow. The final 0.6 in. of the probe wires were left bare, extending beyond the stainless-steel support tube. The low conductivity metals chosen and the horizontal probe configuration

were necessary to overcome the problems of registering accurate temperatures in a low-velocity gas flow field.

- (2) The use of probes half-way along the test section, as well as at the exit.
- (3) A modification of the inlet calming section design to provide for better insulation of the inlet gas stream from the inlet liquid stream. Byers and King used a thin metallic divider plate separating the gas and liquid. In the present work the gas entered along a parabolic flow path which kept the inlet gas and inlet liquid physically far apart until near the end of the inlet divider. As a result, the final entry angle of the gas was 6° .
- (4) Windings of Nichrome wire placed around the gas entry section. A carefully controlled current passed through these wires could be used to make up for heat losses from the channel and maintain an adiabatic wall condition more closely.
- (5) The use of vacuum for gas sample withdrawal through the probes and from the bulk exit gas. This eliminated the need for running the channel under a slight positive pressure.
- (6) Installation of thermistor temperature monitors at various points on the walls of the channel and the entry section, and in the bulk fluids in the entry section.

In the present work nitrogen was employed as the gas-phase feed stream. Hydrocarbons were used for the liquid feed. n-Pentane and isopentane served as volatile liquid species,

while a linear paraffinic hydrocarbon mixture served as a non-volatile liquid species. This mixture, henceforth referred to as n-tridecane, was donated by the Atlantic Refining Co., and was in reality a still-cut from the Iso-sieve process composed of 16 mole % $n\text{-C}_{12}\text{H}_{26}$, 58% $n\text{-C}_{13}\text{H}_{28}$, 25% $n\text{-C}_{14}\text{H}_{30}$ and 1% $n\text{-C}_{15}\text{H}_{32}$.

Gas chromatography was used for gas and liquid-phase analyses. An Aerograph A-90-P2 chromatograph was employed with a column composed of silicone oil SE-30 on firebrick support.

LOW-FLUX HEAT AND MASS TRANSFER EXPERIMENTS

The previous work by Byers and King^{3,4} found good agreement between convective diffusion theory and experimental results for low-flux, low-concentration-level mass transfer. Evaporation rates for the cases of gas phase control and of resistances distributed between the gas and liquid phases were successfully analyzed.

In order to confirm the applicability of this approach for the case of heat transfer in the absence of mass transfer, a number of experimental runs were carried out in which heat was transferred from a warm nitrogen stream (approximately 32°C, inlet) to a nonvolatile liquid (n-tridecane) at approximately 20°C, inlet. Temperature measurements with the inlet probe indicated that heat transfer in the entering section amounted to the equivalent of an extra 1.5 inches of channel contact length. Under adiabatic wall conditions the experimental average heat transfer coefficients, which were obtained through integration of the entrance and exit temperature profiles, agreed

with the theoretically predicted coefficients quite well, with an average relative percent deviation of only 2.8%.⁷ Evaporation of n-pentane from n-tridecane into nitrogen at low solute concentration levels was also carried out, with close agreement between the experimental and theoretical mass transfer coefficients. Details of these heat and mass transfer measurements are given elsewhere.⁷

Effect of Mass Transfer upon Interfacial Temperature

The high flux mass transfer results were obtained by vaporizing two pure fluids, n-pentane and isopentane, into nitrogen. Both of these liquids have boiling points that are only slightly above the run conditions. Consequently the mass flux levels were quite high and the effect of the latent heat of evaporation upon the temperature profiles could not be ignored. The following iterative calculational approach was found to yield an acceptable prediction of the experimentally observed decrease in the liquid interfacial temperature, under conditions where cellular convection cells did not arise in the liquid phase (see Part II). The first step in the procedure is to calculate the mass transfer rate assuming a constant interfacial temperature equal to the inlet liquid temperature. Thus if we employ the penetration model for the calculation of the gas-phase mass transfer coefficient (a conservative first approximation), then for the case of $N_{Bo} = 0$, $\Delta H_{vap} = \text{constant}$, we have

$$Q_{int} = k'_{x,loc} (x_{Ao} - x_{Ao}) \Delta H_{vap} \quad (10)$$

where

$$k'_{x,loc} = c \sqrt{\frac{D_{AB}}{\pi t}} \left(\frac{\theta_{AB}}{1 - x_{Ao}} \right) \quad (11)$$

The application of Fourier's law to the liquid interface, neglecting the small amount of heat transferred to the gas phase, yields

$$Q_{int} = - k_t \left. \frac{\partial T}{\partial y} \right|_{y=0} \quad (12)$$

If we also make the assumption of a constant liquid velocity in the region of the mass transfer interface, then penetration theory can be used to calculate the liquid heat transfer behavior. The x-direction velocity can be converted to t, the exposure time, by the relationship

$$x = U_{int} t \quad (13)$$

The preceding set of assumptions is now equivalent to the problem of a transient temperature penetration into a semi-infinite body, whose surface temperature gradient is given as a function of $t^{-\frac{1}{2}}$, i.e.,

$$k_t \frac{\partial T}{\partial y} = \beta_1 t^{-\frac{1}{2}} \quad (14)$$

The solution to this problem has been carried out by Carslaw and Jaeger⁶ with the assumption of a constant initial fluid temperature, T_∞ . Their solution can be manipulated to yield an expression for the interfacial temperature of the form

$$(T_{int} - T_\infty) = \frac{\beta_1 \alpha^{\frac{1}{2}} \Gamma(\frac{1}{2})}{k_t} \quad (15)$$

where α is the liquid-phase thermal diffusivity, and the value of β_1 can be determined from the combination of Equations 10, 11,

12, and 14. If Q_{int} is proportional to $t^{-\frac{1}{2}}$, T_{int} is independent of t .

Having determined the value of T_{int} , one can go to the vapor pressure curve of the fluid in question, obtain a new value of x_{Ao} corresponding to the new value of T_{int} , and repeat the above calculations. This iterative approach usually converged to a constant value of T_{int} after three or four iterations.

For the interpretation of high-flux evaporation data, interfacial temperatures obtained with the temperature probes were employed to obtain the equilibrium partial pressure of the evaporating species. It was found that the interfacial temperature would change by no more than 0.1 to 0.4°C along the channel. An average temperature was used to obtain x_{Ao} for converting the cup-mixing mole fraction to F_s .

High-Flux Mass Transfer Results

The high-flux-level mass transfer data were obtained in two ways, by the measurement of gas-phase concentration profiles and by the measurement of cup-mixing concentrations. The mole fraction of the volatile species was obtained by sampling a continuous flow from either the concentration probe or the exit stream, using the gas chromatograph. From three to six chromatograms were obtained for each experimental point, and an average value was calculated. The chromatograms for a single point usually gave between 5 and 10% spread.

Figure 4 represents an experimental concentration profile taken at the exit probe for a gas-phase interfacial mole fraction of 0.584. The experimental system was n-pentane evaporating into

pure nitrogen. The fraction saturation, F_s , is based upon the experimentally observed interfacial temperature according to the formula

$$F_s = x_A / x_{A0} \quad (16)$$

where x_{A0} is the equilibrium mole fraction at the experimental value of T_{int} .

Figure 5 is a comparison of several experimental concentration profiles taken at the same flow conditions, with varying values of interfacial concentration. The increase in over-all transfer rate and the decrease of k'_x as the value of x_{A0} is raised can easily be seen. For comparison the low-flux, low-concentration profile for the same flow conditions is given by the dashed curve.

In order to compare the experimentally obtained profiles with a theoretical approach, it was first necessary to integrate these profiles to obtain a cup-mixing concentration according to the equation

$$x_{A, \text{cup mix}} = \frac{\int_0^b u_x(y) x_A(y) dy}{b U_{x, \text{avg}}} \quad (17)$$

The experimental cup-mixing mole fraction was then converted to a fraction saturation by dividing it by the equilibrium mole fraction at the interfacial conditions. A theoretical value of the fraction saturation was obtained in the following manner.

First a "low flux" value of the average fraction saturation was obtained based upon the steady-state flow conditions which would have occurred in the absence of mass transfer. For this purpose the numerical solution outlined by Byers and King³ was

used. The variation in gas viscosity due to composition and temperature was generally small and was ignored by presuming that the gas-phase velocity profile was parabolic, as would occur for constant viscosity. The fraction saturation is directly related to the average mass transfer coefficient based on the inlet driving force for the low-flux, low-concentration level case by the equations

$$k_{x,avg} x_{Ao} = N_{Ao,avg} \quad (18)$$

and

$$N_{Ao,avg} L = c x_{A,cup mix} U_{x,avg} b \quad (19)$$

$$F_{s,avg} = \frac{k_{x,avg} L}{c U_{x,avg} b} \quad (20)$$

By manipulation of the above two equations along with Equations 5, 9 and 17, we see that the correction factors which have been derived for the average mass transfer coefficient to account for flux level and interfacial concentration level are also applicable to the cup-mixing concentration.

Thus, the flux correction can be applied by multiplying the right-hand side of Equation 20 by θ_{AB} , and the concentration factor is applied by dividing by $(1 - x_{Ao})$. The appropriate value of θ_{AB} was obtained by first calculating R_{AB} for the experimental conditions, and then using Figure 1 to obtain θ_{AB} . Actually, due to the existence of a finite interfacial velocity, the correct value of θ_{AB} should lie somewhere between the penetration and Leveque solutions. Since the difference between the two is quite small, the penetration model was utilized in all calculations.

For the prediction of mass transfer coefficients, gas-phase diffusivities were computed by the approaches of Wilke and Lee²¹ and of Bird and Slattery.² The diffusivity of the N_2 -n-pentane system was taken as $0.0882 \text{ cm}^2/\text{sec}$ while that for the N_2 -isopentane system was taken as $0.0894 \text{ cm}^2/\text{sec}$, both at 20°C . The vapor pressures of n-pentane and isopentane were taken as 418.8 and 570.6 mm Hg, respectively, at 20°C . A tabulation of all experimental data and sources of physical properties is available elsewhere.⁷

In addition to the integration of the experimental concentration profiles, a number of cup-mixing concentrations were obtained directly by sampling the exit gas stream. The over-all experimental results are summarized in Table I, and in Figure 6. The gas flow rate was varied to give results at different Graetz numbers; however the liquid flow rate was held constant at 0.4 gal/min. The agreement between the calculational approach and the experimental results is quite good; however, there is some tendency for the experimental fraction saturation to be somewhat higher than the predicted values. This is not surprising, since a potentially important factor was ignored in the theoretical model. The experimental arrangement was one of confined flow; this meant that, as the liquid phase evaporated, it contributed a significant amount of material and therefore acceleration to the gas phase in the region near the interface. It is difficult to predict the exact influence of this acceleration upon the transfer coefficients; however, the qualitative effect should be to increase the mass transfer into the gas phase.

This increase in fraction saturation due to acceleration was calculated to be approximately 10% for $x_{A0} = 0.50$, if it is assumed that the velocity profile remains parabolic, i.e., that the acceleration is uniform across the gas phase and that variations in viscosity can be ignored. As can be seen from the experimental results given in Table I, a slight increase in the theoretical value of the fraction saturation would result in somewhat better agreement between the theoretical and experimental results.

If we assume that the concentration level correction is correctly given by dividing F_s by $(1 - x_{A0})$, then the experimental high-flux results can be used to test the flux-level correction factor, θ_{AB} . Figure 6 is a graph of θ_{AB} as a function of x_{A0} . The solid curve was obtained using the penetration high-flux solution, under the assumptions of $N_{B0} = 0$, and $x_{A\infty} = 0$. The experimental data points tend to be a little high; however, this is probably due to the previously discussed acceleration effect. The agreement is still fairly good, with the average relative error being +6.2%.

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NOMENCLATURE

b	channel half width (cm)
c	total concentration (gmoles/cm ³)
D_{AB}	diffusion coefficient in the binary system A-B (cm ² /sec)
F_s	fraction saturation of the gas phase
ΔH_{vap}	heat of vaporization (cal/gmole)
J_A^o	molar flux of species A relative to the volume average velocity
J_A^*	molar flux of species A relative to the molar average velocity
k	mass transfer coefficient at low flux conditions (moles/cm ² -sec)
k'	mass transfer coefficient (applicable at high flux conditions) (moles/cm ² -sec)
k_t	thermal conductivity (cal/cm-sec-°C)
L	exposure length (cm)
N_A	molar flux of component A relative to stationary coordinates (moles/cm ² -sec)
P_{BM}	log mean pressure of component B
Q_{int}	interfacial heat flux (cal/cm ² -sec)
R_{AB}	dimensionless flux ratio
T	temperature (°C)
t	time (sec)
$U_{x,avg}$	average gas velocity in the x-direction (cm/sec)
U_{int}	interfacial velocity in the x-direction (cm/sec)
u_x	velocity in the x-direction (cm/sec)

$\tilde{V}$	partial molal volume
x	horizontal distance variable (cm)
x_A	mole fraction of Component A, gas phase
y	vertical distance from the bottom of the channel (cm)
z_{Ao}	arbitrary interfacial concentration (dimensionless)
α	thermal diffusivity (cm^2/sec)
β_1	variable used in heat flux solution, $c \Delta H_{\text{vap}} \theta_{AB} \sqrt{\frac{D_{AB}}{\pi}} \left(\frac{x_{Ao} - x_{A\infty}}{1 - x_{Ao}} \right)$
Γ	gamma function
Δ	indicates the difference between two quantities
θ_{AB}	dimensionless flux correction factor
ϕ_A	volume fraction of Component A
Φ_{AB}	dimensionless flux ratio
∇	vector differential operator

Subscripts

A,B	refers to the components A, B; etc.
AB	refers to the binary system composed of components A and B
avg	average value of a quantity
int	evaluated at the interfacial position
loc	a local or point quantity
o	evaluated at the interfacial position
x	based upon mole fractions
ϕ	based upon volume fractions
∞	quantity evaluated at a large distance from the gas-liquid interface, equivalent to inlet conditions.

Table I
High Flux Experimental Results

Graetz No. $(D_{AB}L/U_{x,av}b^2)$	x_{Ao}	Experimental F_s	Theoretical* F_s	Percent error
<u>Evaporation of n-Pentane into N₂</u>				
0.0454	0.403	0.248	0.235	+ 5.6
0.0454	0.418	0.278	0.239	+14.0
0.0454	0.505	0.291	0.248	+14.8
0.0842	0.410	0.377	0.353	+ 6.4
0.0842	0.580	0.380	0.382	- 0.5
0.0842	0.580	0.427	0.382	+10.5
0.1510	0.439	0.535	0.481	+10.1
0.1510	0.580	0.591	0.525	+11.2
0.0810	0.418	0.357	0.338	+ 5.3
<u>Evaporation of Isopentane into N₂</u>				
0.0956	0.560	0.402	0.400	+ 0.2
0.0956	0.655	0.435	0.421	+ 3.2
0.1368	0.560	0.490	0.483	+ 1.4
0.1368	0.681	0.488	0.516	- 5.7
0.1368	0.740	0.509	0.536	- 5.3
Average percent error = + 5.1				

*Based upon velocity profile which would occur in the absence of mass transfer.

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Figure 1. Effect of high mass flux on the mass transfer coefficient for the Leveque model (dashed curve).

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Figure 2. Schematic of channel for gas-liquid contact, showing dimensions and probe locations.

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Figure 3. Experimental apparatus.

Figure 4. Exit concentration profile for Run No. 240. $x_{A0} = 0.584$.

System: n-Pentane ~ Nitrogen.

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Figure 5. Exit concentration profile for various interfacial mole fractions.

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Figure 6. Experimental values of high flux correction factors for the evaporation of n-pentane and iso-pentane into nitrogen.

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