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EFFECT OF FLUID PROPERTIES ON MASS TRANSFER IN THE GAS PHASE

E. J. Lynch and C. R. Wilke

April 2, 1953

Berkeley, California

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ABSTRACT

Water was vaporized into air, helium, and Freon-12 in turbulent gas flow over a wet-bulb thermometer and through a 1 ft. diameter tower packed with 1 in. Raschig rings.

The wet-dry-bulb measurements indicate that the gas film mass transfer coefficient varies with the Schmidt number to the -1/2 power for flow perpendicular to single cylinders.

Heights of a transfer unit, HTU, were measured in the packed tower for flow of the gases countercurrent to water over a range of gas and liquid rates. When compared at equal gas flow Reynolds numbers at constant liquid rate, HTU varied as the 0.9 power of the Schmidt group. When compared at equal values of

$$\rho u^2 (\rho = \text{gas density}, u = \text{velocity}),$$

HTU varied as the 0.47 power of the Schmidt group. By reference to the psychrometric study, the latter method of comparison of HTU's seems preferable, and indicates that further study of criteria for dimensional similitude in packed columns may be needed.

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INTRODUCTION

According to the two-film theory of Whitman (49), the resistance to mass transfer between a gas and liquid stream can be divided into two additive resistances which result from hypothetical films of stagnant gas and liquid on either side of the interface. The liquid film resistance in a packed column has been evaluated by Sherwood and Holloway (40) for the commonly used packings, but as yet a satisfactory correlation of gas film absorption coefficients has not been made.

It was the purpose of this investigation to clarify the role of the gas film resistance by determining the effect of the fluid properties on the rate of mass transfer in a packed tower. To do this, a twofold plan was followed. The most direct method was the actual measurement of mass transfer coefficients in a tower packed with 1 in. Raschig rings. To eliminate any influence from the liquid properties, it was decided to vaporize a single pure liquid into inert gases. The gases chosen were air, helium, and Freon-12. These gases, with water as the liquid, gave a thirtyfold variation in density, a twofold variation in viscosity, an eightfold variation in diffusivity, and a four-and-one-half fold variation in Schmidt number.

As the Schmidt number effect was of particular interest, the second method of approach utilized wet and dry-bulb thermometer measurements

to determine the psychrometric ratio. Since the Prandtl number of the gases is nearly constant, this method will give the variation of the mass transfer coefficient with the Schmidt number in evaporation from single cylinders. While it is not necessary that the same Schmidt number effect will hold for single cylinders and packed towers, the effect should be of similar magnitude and should serve to confirm the results obtained by the more direct method.

THEORY

Approximate analytical solutions for the rate of mass transfer from an interface to a fluid moving in forced convection in a straight circular pipe have been developed by various authors (5, 10, 33, 39) through the use of the analogy between heat, mass, and momentum transfer. These solutions have indicated that the mass transfer coefficient, k_G, for this case, is a function of the Reynolds number, $(du\rho/\mu)$, and the Schmidt number, $(\mu/\rho D)$. The Reynolds number ordinarily enters as a function replacing the friction factor. These solutions can be expressed generally by the equation:

$$k_G M_m p_f/G = \phi'(du\rho/\mu), (\mu/\rho D) \tag{1a}$$

where

ϕ' = some mathematical function

k_G = mass transfer coefficient - lb.moles/hr.-ft.^{2}

d = diameter of the tube in ft.

D = diffusion coefficient in ft.^{2}/sec.

u = velocity in ft./sec.

ρ = density in lbs.(m)/ft.^{3}

μ = viscosity in lbs.(m)/ft.-sec.

G = mass velocity of the gas in lbs.(m)/hr.-ft.^{2}

M_m = mean molecular weight in lbs.(m)/mole

p_f = film pressure factor (50). This term accounts for the presence and movement of the inert fluid.

This equation can be expressed alternately in terms of the transfer unit defined by Chilton and Colburn (8):

$$aHTU_G = \psi'(du\rho/\mu),\ (\mu/\rho D) \qquad (1b)$$

where HTU_G = the height of a transfer unit = $G/k_G a p_f M_m$-ft.,

and a = packing area/cu.ft. of packing- ft.^{2}/ft.^{3}.

These equations do not include the effect of the relative roughness on the mass transfer rate and consequently are limited to the particular pipe geometry for which the functions ϕ' and ψ' are experimentally evaluated.

Mass transfer in Packed Columns

Equations (1a) and (1b) have been frequently used to satisfy a need for some theoretical framework for the correlation of packed column data. However, this extension of the functional relationship developed for simple pipe flow has never been subjected to adequate experimental investigation to justify its general use. In attempting to correlate the results of this work, it was found that equations (1a) and (1b) were not satisfactory. Therefore an alternate method of correlation was developed which is based on the following qualitative analysis.

In the study of mass transfer rates from single cylinders placed perpendicular to the flow stream (29), it has been found that the bulk of the mass transfer taking place occurs in the vicinity of the stagnation point and in the region of the turbulent wake. These same two sections of the cylinder are of primary concern in the loss of momentum of the gas stream as manifested by the pressure drop. Consequently, it might be expected that the two phenomena, transfer of momentum and transfer of mass, are to a large extent functions of the same properties of the flow field, particularly for gases in which the diffusivities for mass and momentum are of similar magnitude. Pressure drop measurements made in flow past single cylinders (15) show that the drag coefficient becomes relatively constant at high values of the Reynolds number. Therefore, as a first approximation, the drag coefficient is independent of the Reynolds number and the pressure drop is dependent only upon the inertia of the gas stream, ρu_m^2. Since it is postulated above that the rate of mass transfer will depend on the same flow properties as the pressure drop, it may follow that the mass transfer coefficient will also be a function of the inertia force and not the Reynolds number at high flow rates. With this modification, equations (1a) and (1b) may be rewritten for application to single cylinders for conditions of high turbulence in the fluid:

$$k_G M_m p_f / G = \phi(\rho u_m^2), (\mu/\rho D) \qquad (2a)$$

$$aHTU_G = \Psi(\rho u_m^2), (\mu/\rho D) \qquad (2b)$$

where u_m = mean velocity of the flowing fluid in ft./sec. Pressure drop measurements made in packed absorption towers (36) have also shown very

little dependence on the Reynolds number over the usual operating range. Since a tower packed with Raschig rings may be considered in a broad sense as a group of cylinders, the extension of the above reasoning (equations 2a and 2b) to a packed tower would appear to follow.

Psychrometry

If the vapor content of a gas stream is known, the psychrometer can serve as a useful instrument for measuring the relative rates of heat and mass transfer from single cylinders. It has a distinct advantage over other methods for determining the dependence of k_G on Schmidt number in that no knowledge of the velocity term dependence is required. If an exact analogy can be drawn between the two processes of heat and mass transfer, the velocity term should enter into both rates in the same manner, and the ratio of the mass transfer coefficient to the heat transfer coefficient should be a unique function of the Schmidt and Prandtl numbers.

Colburn (10), and Chilton and Colburn (7) have developed equations for the correlation of heat and mass transfer data in terms of the j number, a quantity which is analogous to the friction factor in momentum transfer. According to this concept,

$$j_h = \left[h/C_p G\right] (C_p \mu/k)^{2/3} \tag{3}$$

$$j_d = \left[k_G M_m p_f/G\right] (\mu/\rho D)^{2/3} \tag{4}$$

where h = heat transfer coefficient - Btu./hr.ft.^{2} °F,

C_p = heat capacity of the gas - Btu./lb.(m) °F,

k = thermal conductivity - Btu./hr.ft.^{2} °F/ft.

If j_h is assumed to be equal to j_d, the psychrometric ratio is obtained:

$$(k_G M_m p_f/h)C_p = \left(\frac{\frac{C_p \mu}{k}}{\frac{\mu}{\rho D}} \right)^{2/3} \qquad (5)$$

It has been found (19) that $j_h = 1.08\ j_d$ for heat and mass transfer in packed beds and it might be expected that a similar relationship will hold for simultaneous heat and mass transfer from a wet-bulb thermometer.

In the above j-number analogy the value of k_G is determined directly from the wet-bulb data since it is multiplied by the term p_f, which corrects for variations in the partial pressure of the inert gas. The value of h, on the other hand, is the true heat transfer coefficient which would be obtained if there were no evaporation from the surface. To correct the wet-bulb data it is necessary to include a factor which will allow for the heat being carried away from the thermometer by the mass movement of the vapor molecules.

Consider a model as shown in Fig. 1 such that the transfer of heat and mass occurs only in the normal direction. With the surface temperature as a reference, the heat transferred to the surface, q_s, is equal to the heat entering the film, q_a, minus the heat being carried out by the vapor flow, $NC_m(t_a - t_s)$, or

$$q_s = q_a - NC_m(t_a - t_s) \qquad (6)$$

where N = number of moles diffusing per unit time per unit area - moles/hr.ft.^{2},

C_m = molar heat capacity of the vapor - Btu./lb.mole ^oF, and

t_a, t_s = gas and surface temperatures respectively - ^oF.

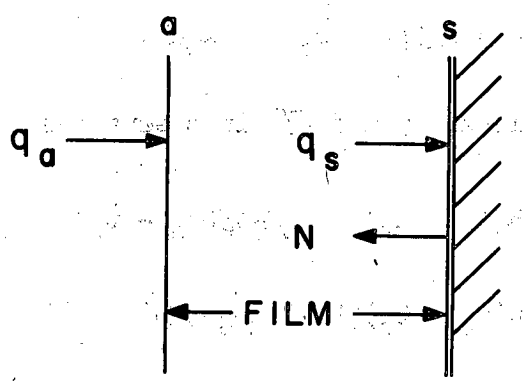

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Fig. 1. Schematic representation of heat and mass transfer taking place at a wet-bulb thermometer

If there were no flow of vapor, q_s would equal q_a.

$$q_s = q_a = h(t_a - t_s) \tag{7}$$

where h = heat transfer coefficient, assumed independent of the mass transfer movement.

The number of moles diffusing may be defined in terms of the mass transfer coefficient as

$$N = k_G(p_s - p_a) = k_G \Delta p. \tag{8}$$

Substituting from equations (8) and (7) into equation (6),

$$\begin{aligned} q_s &= h(t_a - t_s) - C_m k_G \Delta p(t_a - t_s) \\ &= h(t_a - t_s)(1 - C_m k_G \Delta p/h) \\ &= h(t_a - t_s)\gamma \end{aligned} \tag{9}$$

$$\text{where } \gamma = 1 - C_m k_G \Delta p/h. \tag{9a}$$

If a new pseudo-heat transfer coefficient, h', is defined as the total effective heat transfer coefficient including the convective transport of heat, then

$$h' = h\gamma \tag{10}$$

and the correct psychrometric ratio is:

$$\frac{k_G M_m p_f \gamma}{h'} C_p = \text{function } (\mu/\rho D), (C_p \mu/k). \tag{11}$$

EXPERIMENTAL

Physical and Thermodynamic Properties of Materials*

Viscosity

The viscosities of air, water vapor, and helium were taken from a review article by F. G. Keyes (27) in which the available literature data for several gases are collected and formulated by means of an empirical equation. The viscosity of Freon-12 was taken from a paper by Buddenberg and Wilke (6) on the measured viscosity of several gases. It was extended over the desired temperature range by the use of the equation of Hirschfelder, Bird, and Spotz (21).

Thermal conductivity

The thermal conductivity of air, water vapor, and helium was taken from the review article by Keyes (27) quoted above. The thermal conductivity of Freon-12 is that reported by Sherratt and Griffiths (38) over the range from 33.3° to 216.2° C.

Heat capacity, C_p

The heat capacity of air and water vapor was taken from the book by Keenan (26) on the thermodynamic properties of air. The heat capacity of Freon-12 was calculated from the work of Justi and Langer (25) who reported the heat capacity at constant volume. The heat capacity of helium is reported in Perry's Handbook (35) as constant for all temperatures.

Diffusivity

The diffusivities used were those measured by C. Y. Lee (28).

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*Properties of pure gases were modified for the presence of water vapor by established methods.

The measured values were corrected to the desired temperature by the equation of Hirschfelder, Bird, and Spotz (21).

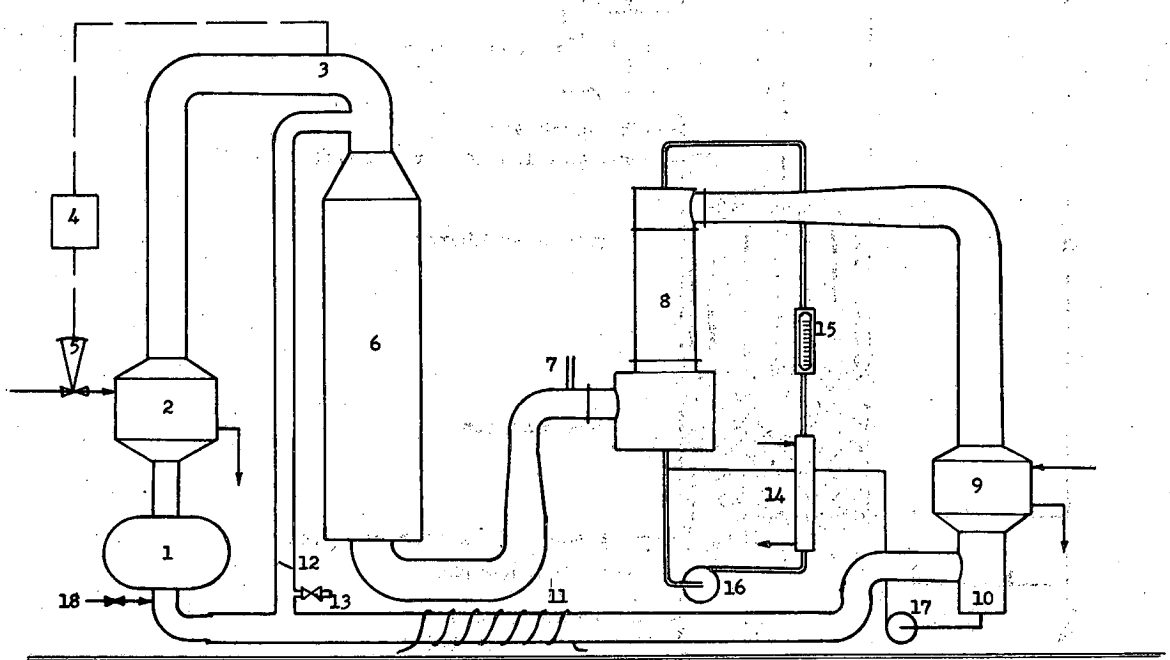
Gas specifications

The helium used was supplied by the U. S. Navy. It was 99.9% pure. The major impurity was water vapor. The Freon-12 used was supplied by Kinetic Chemicals. It was 99.95% pure. The major impurities are Freon-11 and Freon-13.

Equipment

The equipment is shown in Fig. 2. Gas circulation was provided by a positive displacement Roots-Connersville blower. Variation in gas rate was obtained by the use of a variable speed drive on the blower and a bypass on the circulation system. The gas flow rate was measured by means of a nozzle made to the specifications of Bean, Buckingham, and Murphy (3) which was placed in a 20 in. diameter section of the ducting. For low flow rates a calibrated orifice was placed over the end of the nozzle.

Gas entered the tower through a gas distributor which discharged through risers about 1 in. from the bottom of the packing. The gas was sampled in the distributor and its temperature measured at the top of one of the risers. The exit stream from the tower passed through a series of four mixing baffles in the exit duct before being sampled. Gas entering the tower was heated by steam coils. The steam supply was controlled by a Brown Electronik temperature recorder-controller. In general it was possible to keep the temperature fluctuations within $\pm 1/2^{\circ}$ F. at the control point. The metal of the flow cabinet and ducting provided a damping action which limited fluctuations to



1-Blower
2-Heater
3-Thermocouple
4-Temperature recorder-controller
5-Motor valve
6-Flow measuring chamber
7-Wet-dry bulb assembly
8-Absorption tower
9-Cooling coil
10-Condensate receiver
11-Superheating coil
12-Butterfly valve - by-pass control
13-Vent valve
14-Liquid heat exchanger
15-Liquid flowmeter
16-Liquid pump
17-Condensate return pump
18-Steam line for purging

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Fig. 2. General equipment layout

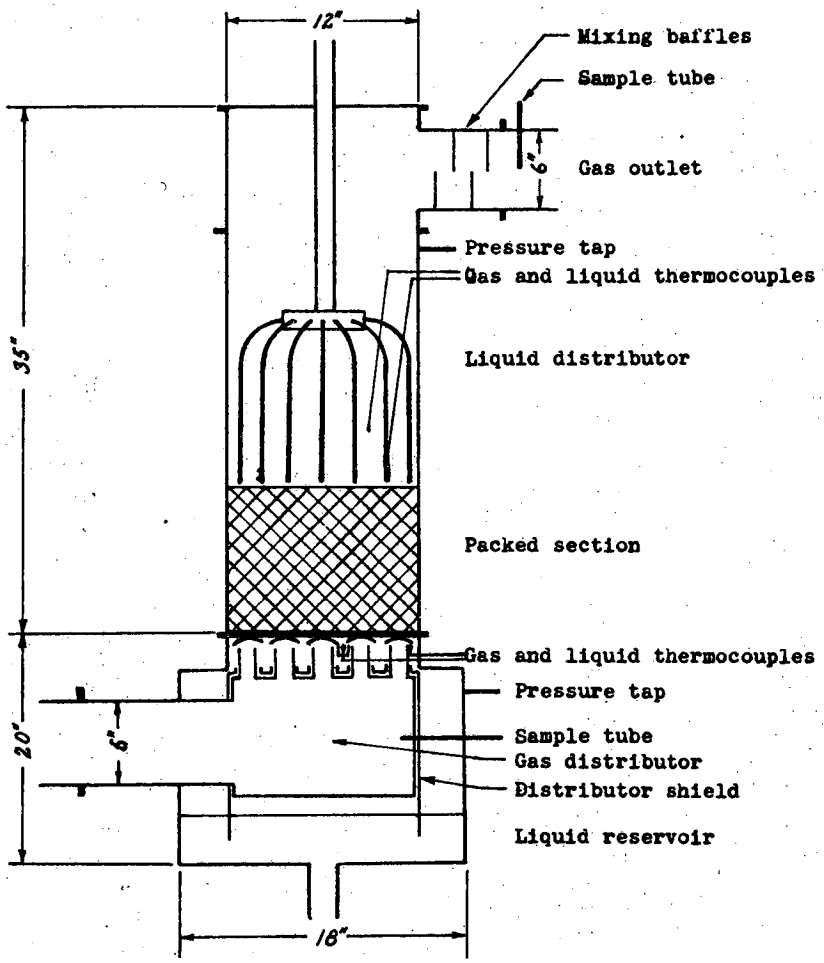

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Fig. 3. Absorption tower

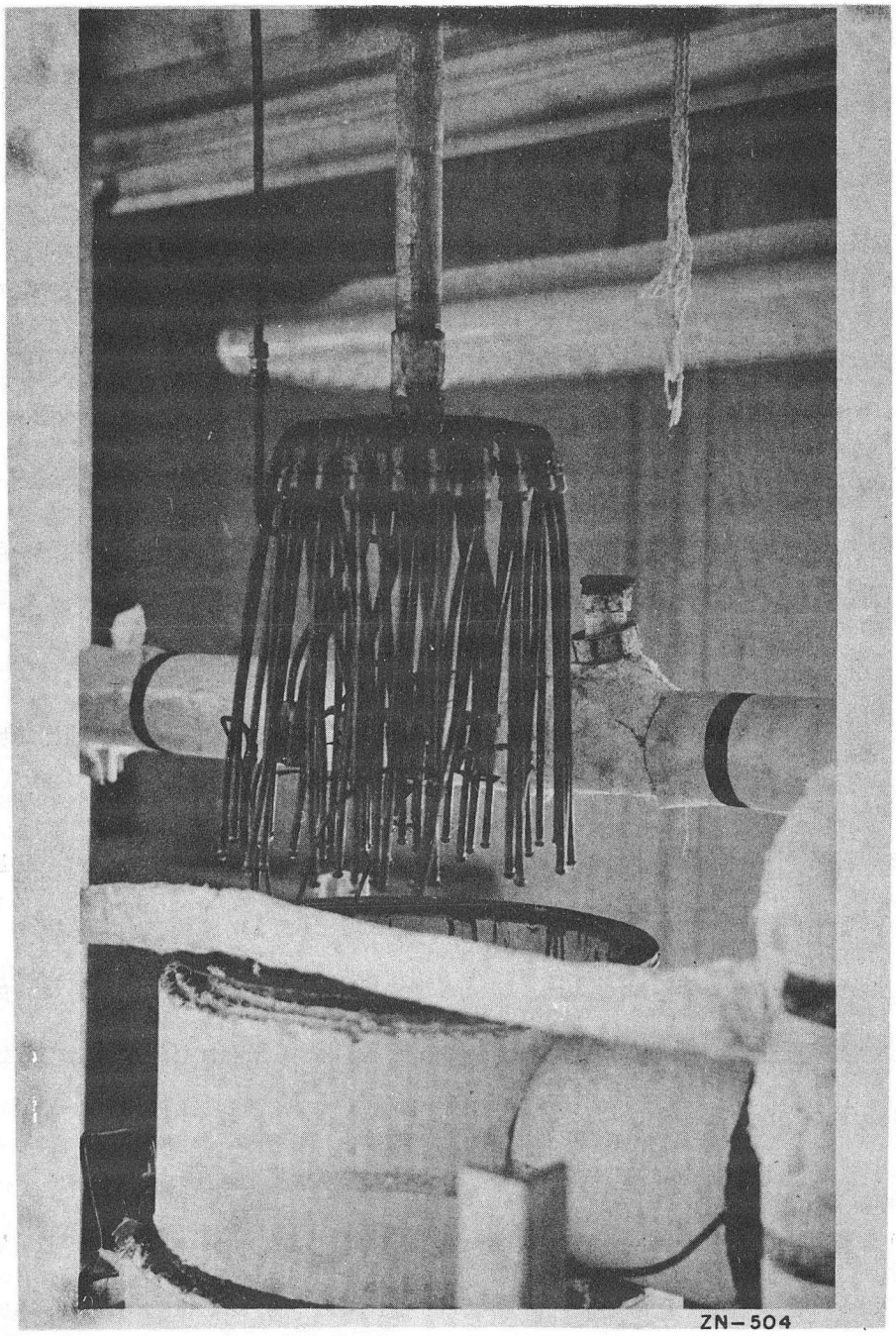

Fig. 4. Photograph of the liquid distributor

± 0.1°F at the tower inlet. The gas leaving the tower was dehumidified by a water-cooled coil before being recirculated.

The large section of the bottom of the tower served as a liquid reservoir. Liquid circulation was provided by a centrifugal pump. The flow rate was measured by a calibrated rotameter. The liquid entered the tower through a 44-point distributor which discharged with 1/2 in. of the top of the packing. Each of the tubes in the distributor delivered liquid at a rate which was within 10% of the average rate for a single tube.

The temperature of the liquid stream was controlled by passing it through a double pipe heat exchanger. The temperature of the liquid entering the tower was measured by a four-couple thermopile. The thermocouples were inserted into rubber sleeves which replaced the last 3 1/2 in. of four of the copper distributing tubes. The liquid temperature at the bottom of the packing was measured by a single thermocouple located in a small cup. The cup had a small drain hole in the bottom to ensure liquid circulation. In addition, the liquid temperature was measured by thermometers placed in the reservoir and in the line directly prior to its entry into the tower. These served solely as a check on the thermocouples.

The thermocouples used in determining the gas and liquid temperatures in the tower were copper-constantin and were calibrated against a Bureau of Standards thermometer. The potentials were read on a Leeds and Northrup K2 potentiometer. The inlet liquid temperature could be read to ± 0.01° C and the other temperatures to ± 0.04°C.

The absorption tower was 1 ft. in diameter and was packed by wet dumping with 1 in. carbon Raschig rings. The packing was supported by

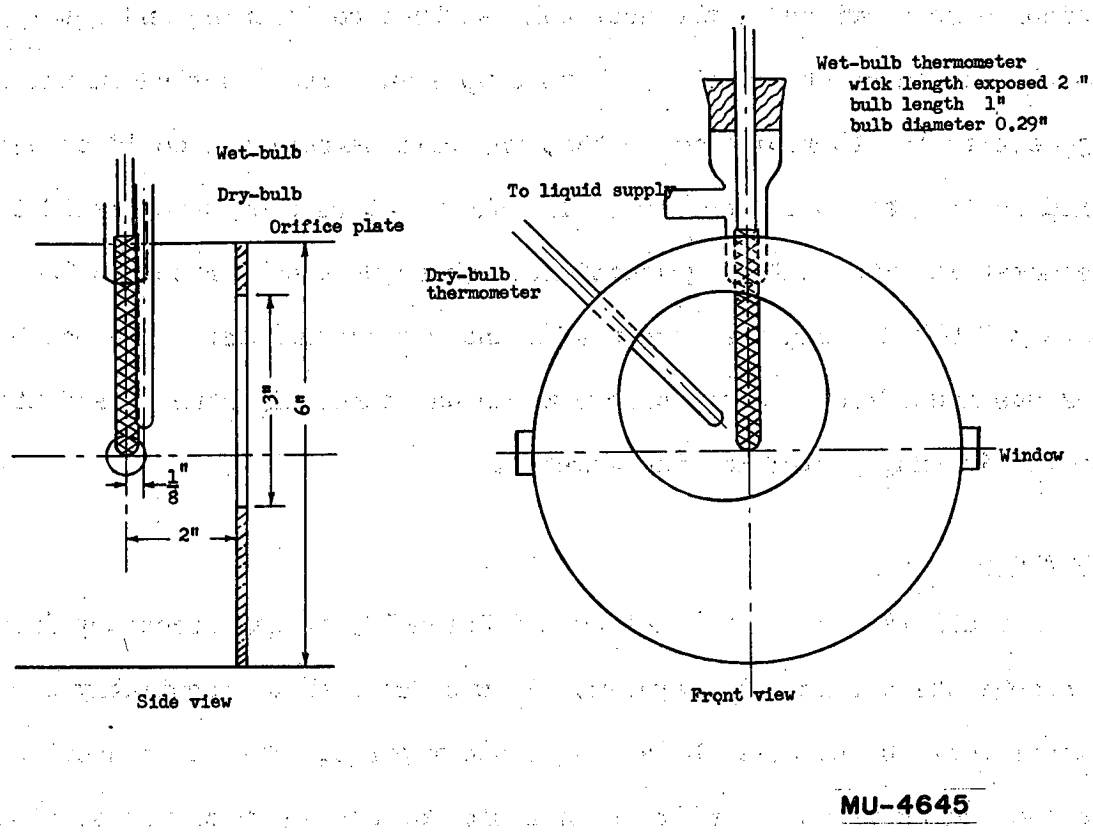

Fig. 5. Wet and dry-bulb thermometer arrangement

a 1 in. mesh steel screen which was in turn supported by a metal grid 1/8 in. thick. The tower was wrapped with a 1 in. thick glass wool insulation which was covered by two layers of corrugated asbestos paper.

The wet and dry-bulb thermometers for the psychrometric measurements were placed in the ducting immediately before the tower and were preceded by an orifice which ensured a uniform high velocity flow across the bulbs. Windows were provided in the duct wall so that positioning and wetting of the wick could be observed. A thermocouple was also inserted in the wall adjacent to the thermometers so that the wall temperature could be determined for radiation corrections. The thermometers used were precision thermometers with 0.1° C graduations. These were calibrated to the nearest 0.01° C against a Bureau of Standards thermometer. The wet-bulb wick was made from a cotton sleeve which was carefully sewn closed at one end to fit snugly over the thermometer.

Procedure

For all runs made with helium and Freon-12, it was necessary first to remove the air from the system. It was found that essentially complete removal could be effected by a single steam purge. The air concentration was checked by means of an Orsat analysis for oxygen which was sensitive to ± 0.05%. When purging was complete, condensed steam was removed from the lines and the necessary adjustments made to bring the equipment on stream. To prevent introduction of impurities into the gas stream, the equipment was operated with a slight positive pressure at the blower inlet.

After steady-state conditions were attained, gas samples were withdrawn simultaneously from the top and bottom of the tower by means of

aspirator bottles of known volume. The moisture content was determined gravimetrically by drawing the samples through tubes filled with Drierite. During the sampling period the temperature of the inlet liquid stream was carefully controlled. In general it was possible to operate with the inlet and exit liquid temperatures less than 0.5° C apart, and with a variation of the inlet temperature during the run of less than ± 0.05° C.

Immediately following sampling, water from an overhead supply was introduced to the wet-bulb wick and allowed to soak the cloth thoroughly. The water supply rate was then reduced until a pendant drop could be observed on the tip of the wick. The wet and dry-bulb temperatures were read. The water supply was stopped and the wet-bulb temperature checked for any further change. (Consecutive runs made at the same gas rate indicate that the moisture content of the gas stream remained essentially constant over long periods of time once steady-state conditions had been attained.)

All packed bed data were taken with a 7 1/2 in. depth of packing with the exception of the runs made with air at heights of 2 in. and 13 1/2 in. which were used to determine the amount of transfer taking place outside the packed section. All runs were made at a liquid rate of 1,575 lbs. (m)/hr.ft.^{2} except those made with air to check the effect of varying the liquid rate.

RESULTS

Psychrometric

The values of the psychrometric ratio obtained in this work for the three gases are shown plotted against the ratio Sc/Pr in Fig. 6. This graph also includes the results of Arnold (2), Bedingfield and Drew (4), Mark (30), and Dropkin (13). These data can best be approximated by a straight line having a slope of -1/2. This line satisfies a modified j number analogy if $j_h' = 1.10 j_d'$ where

$$j_h' = \frac{h}{C_p G} \left(\frac{C_p \mu}{k} \right)^{1/2} \tag{12}$$

$$j_d' = \frac{k_G M_m p_f}{G} \left(\frac{\mu}{\rho D} \right)^{1/2} . \tag{13}$$

This result is in good agreement with the results obtained from packed beds (19) where it has been found that the average ratio of j_h to j_d is 1.08.

Since the value of the Prandtl group is nearly constant for the results shown here, the heat transfer coefficient may also be considered constant. The curve as plotted then represents the variation of the mass transfer coefficient with Schmidt number for flow of gas perpendicular to single cylinders, i.e., k_G is proportional to $Sc^{-1/2}$.

Packed Column

The packed tower results are presented here in terms of the HTU instead of k_G. In order to calculate the HTU from the data, the following equation was used:

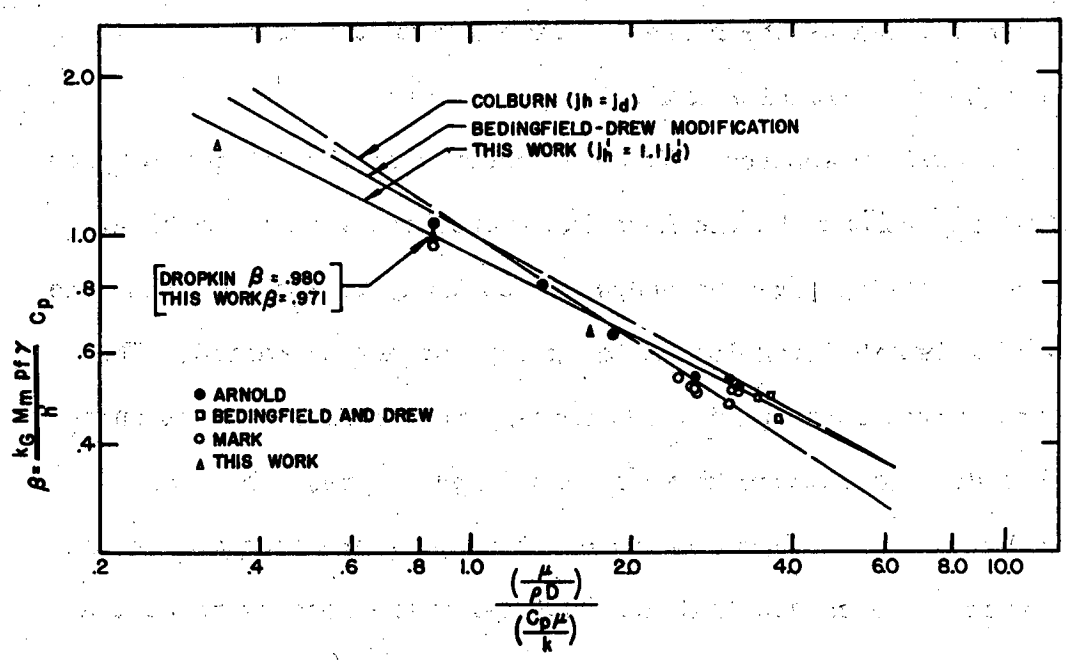

Fig. 6. The psychrometric ratio as a function of Sc/Pr

$$H_2 - H_1 = HTU_G \times \left[\left(\frac{y_f}{1 - y} \right)_{av.} \ln \left(\frac{y_1 - y_i}{y_2 - y_i} \right) \right] \qquad (14)$$

where y = mole fraction and $y_f = p_f/P$. The expression in brackets is the number of transfer units and its value can be computed from the data on the vapor concentrations at the inlet and exit of the tower and the temperature of the liquid stream. The height of the packed section, $H_2 - H_1$, when divided by the number of transfer units, gives the apparent height of a transfer unit.

Fig. 7 shows the apparent HTU as a function of the group G/μ which is directly proportional to the Reynolds number for a given packing. Since only one packing size and shape was studied, it was felt that the inclusion of a length dimension in the group was not warranted. The peaks of the curves correspond approximately to the flow rate at which the pressure drop phenomenon known as "loading" occurs. This was determined by the simultaneous measurement of pressure drop across the packed section. The curves indicate that below the loading point the HTU can be represented as varying with the mass velocity to the 0.33 power, with the power decreasing thereafter as the flooding point is approached. It will be noted that there is a slight difference in the slope for the three curves, but the differences are insignificant in view of the precision of the data.

The more obvious difference among the three curves is the lateral separation of the point of maximum HTU. The vertical separation of the curves might be expected and explained on the basis of equation (1b) by the Schmidt number differences, but the lateral separation would lead one to suspect that the Reynolds number may not be the proper

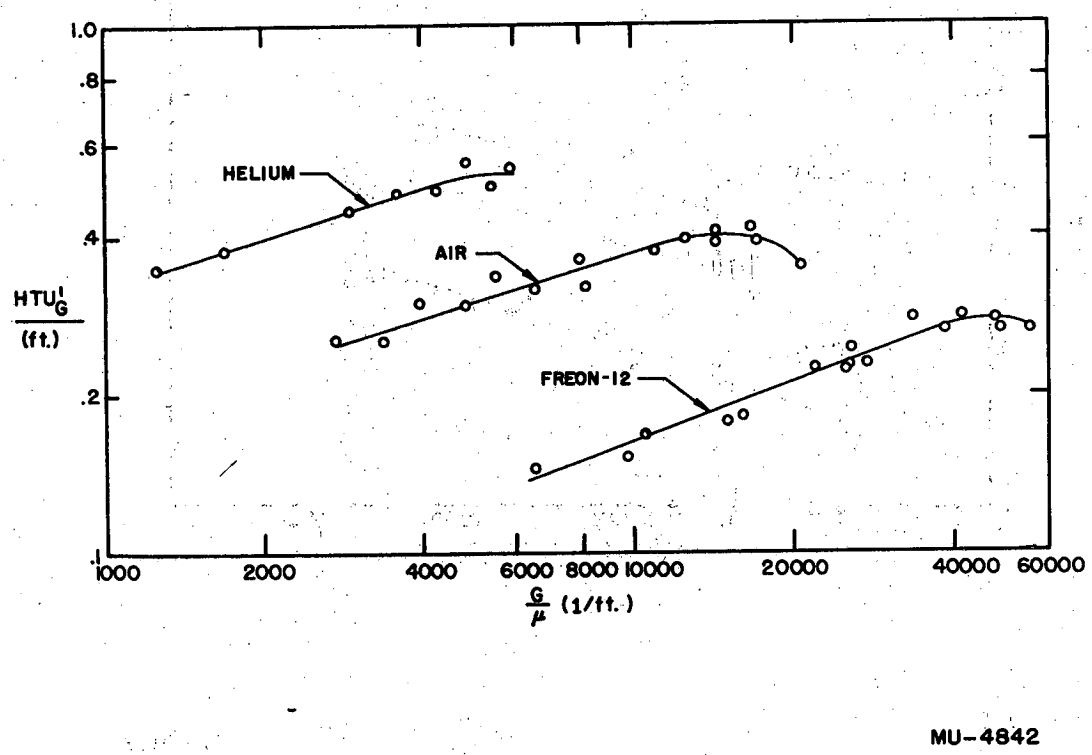

Fig. 7. HTU'_G as a function of G/ . L = 1575

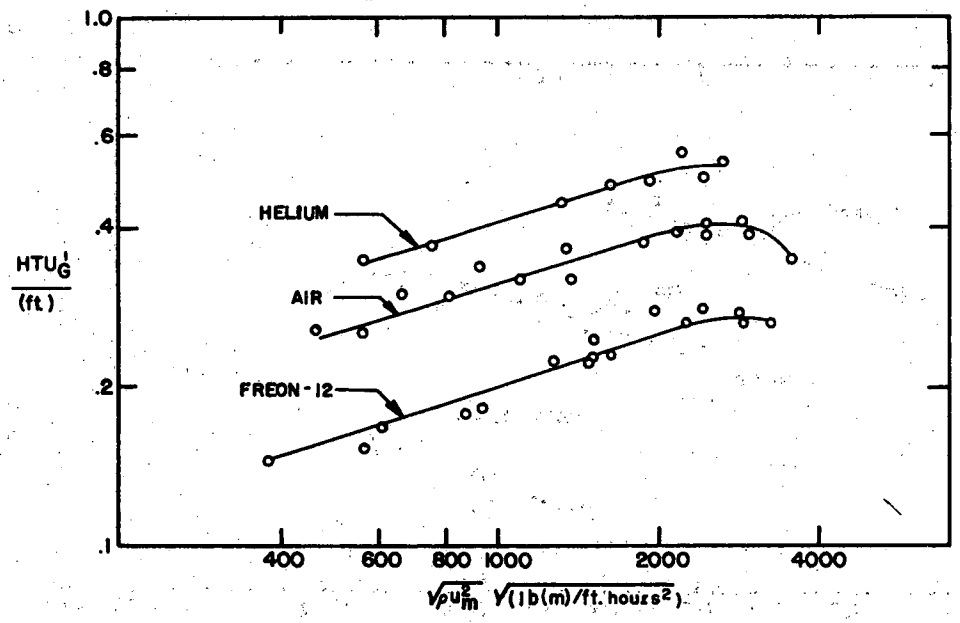

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Fig. 8. HTU'_G as a function of the inertia group. L = 1575

correlating modulus in this case. If, instead, the group suggested by equation (2b) is used, i.e., ρu_m^2, the curves of Fig. 8 are obtained. (The group ρu_m^2/length has the dimensional significance of inertia force per unit volume of fluid (47), and therefore the group ρu_m^2 will be designated the "inertia group.") The square root of the inertia group has been used in plotting in order to retain the first power dependence on the velocity that exists in Fig. 7. Similar usage in the correlation of flooding velocity (41) and pressure drop (36) data in packed columns would seem to justify the procedure. This group has also been used in the correlation of plate efficiencies (20) where it has been given the name F-factor. The curves of Fig. 8 are identical in shape with those of Fig. 7, but with the inertia group as the correlating modulus, the lateral separation has now disappeared and the peaks of all three curves appear at the same value of the inertia group within the limits of experimental error.

Cross plots of the HTU versus Schmidt number at a constant value of G/μ of 4,000 in Fig. 7 and at a constant value of $\sqrt{\rho u_m^2}$ of 1,000 in Fig. 8 are presented in Fig. 9. The cross plots indicate the Schmidt number function by which the HTU must be divided to cause a vertical alignment of the three curves of Fig. 7 and Fig. 8. The plot based on the Reynolds number would require this function to be $Sc^{0.9}$. This would indicate that the method of transfer in the tower was almost pure molecular diffusion, a conclusion which can hardly be justified in view of the known high turbulence level which exists in a packed tower at the flow rates studied. Correlation by the inertia group plot, on the other hand, requires an exponent of 0.47 on the Schmidt group to align the three curves.

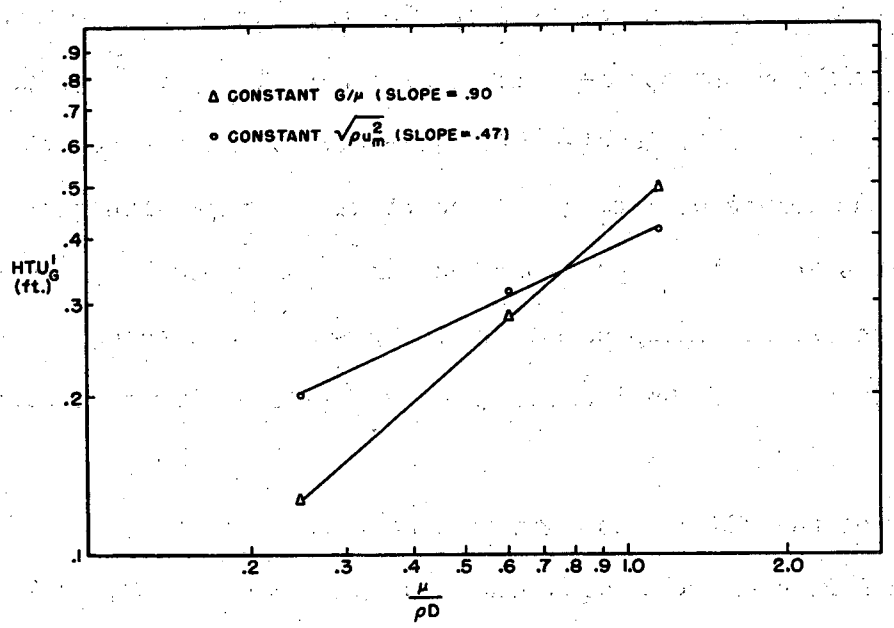

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Fig. 9. HTU_G' as a function of Schmidt number at constant values of G/ and inertia group. L = 1575

This exponent is in good agreement with that obtained in the psychrometric study and with some results obtained in the vaporization of liquids and organic solids in packed beds (9, 42). A simple power function of the Schmidt group appears to be satisfactory for correlation over the present range of system properties and experimental conditions.

To obtain the true HTU, it is necessary to correct for "end effects," i.e., that amount of transfer taking place outside the packed section. In the system under study, the amount of transfer is dependent primarily on the area of liquid exposed to the gas, which in turn depends on the liquid distribution and the manner in which the liquid drips from the bottom of the packing. This should be reasonably independent of the gas being used so that the same end effect correction should apply to all three gases. If it can also be assumed that the effect of the gas properties on mass transfer is the same in the end section and in the packed section, then the end effect can be expressed in terms of an equivalent length of packing which should be constant for all gases and independent of the packed height. This procedure has been used by previous investigators (18, 44). If the end effect is constant, the apparent HTU values obtained at a given packed height need only be multiplied by a constant value to obtain the true HTU values. Therefore, the preceding analysis of the Schmidt number effect is not dependent on the use of true HTU values since all the data used were obtained at a single packed length.

It is possible to solve for the true HTU if two apparent values HTU'_{23} and HTU'_{24} are known for two packed lengths, Z_{23} and Z_{24}. Consider a tower as shown in Fig. 10 where the number of transfer units have been measured with packed lengths of Z_{23} and Z_{24} and with a constant end effect of the equivalent length Z_{12}.

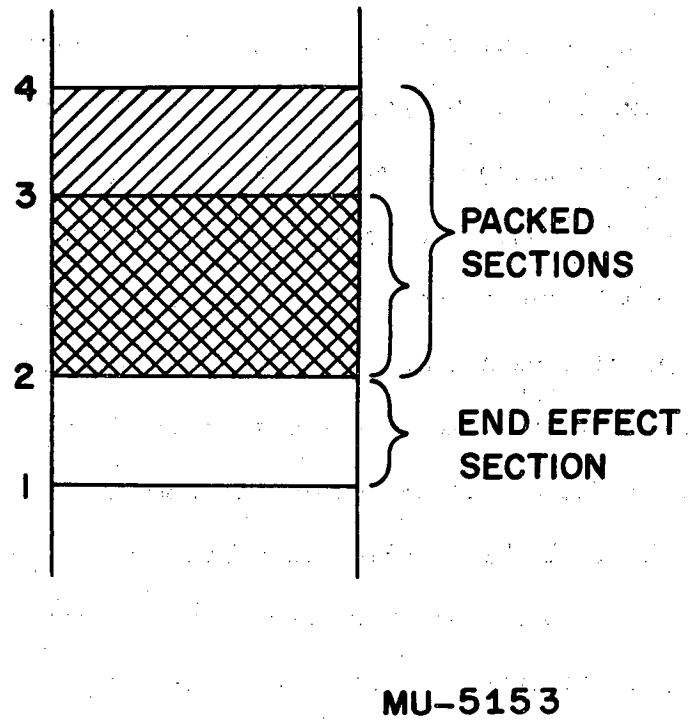

Fig. 10. Schematic representation of transfer sections in a packed column

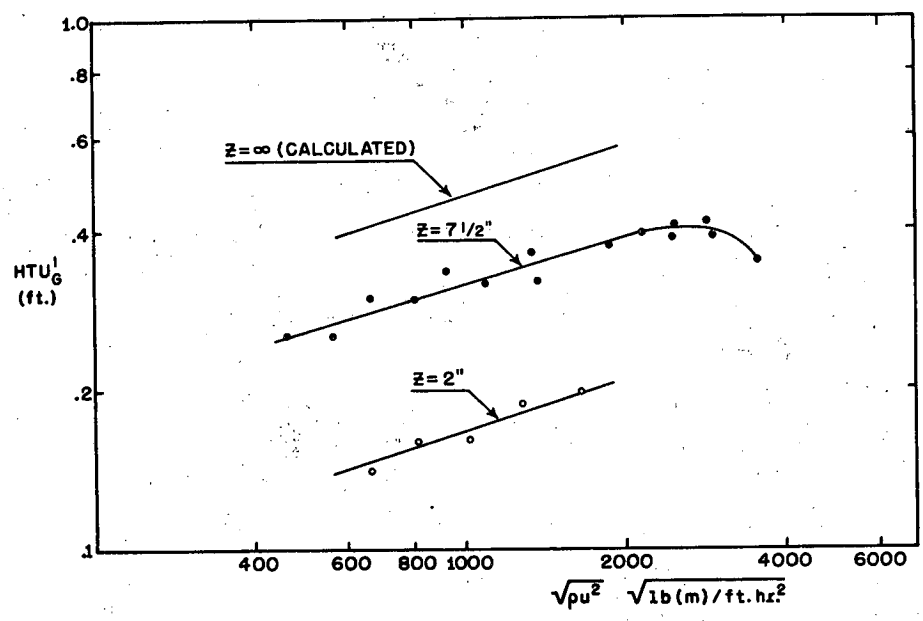

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Fig. 11 The apparent HTU as a function of packed height for vaporization of water into air. L = 1575

Then it is known that:

$$\text{(a)} \quad NTU_{14} = NTU_{12} + NTU_{24}$$

$$\text{(b)} \quad NTU_{13} = NTU_{12} + NTU_{23}$$

$$\text{(c)} \quad NTU_{23} = NTU_{24}(Z_{23}/Z_{24})$$

substituting (c) into (b),

$$\text{(d)} \quad NTU_{13} = NTU_{12} + NTU_{24}(Z_{23}/Z_{24})$$

subtracting (d) from (a) and rearranging,

$$\text{(e)} \quad NTU_{24} = \frac{NTU_{14} - NTU_{13}}{1 - Z_{23}/Z_{24}}$$

NTU_{24} is not a value that can be directly measured, but the corresponding value HTU_{24} is a true HTU. Substituting the values of the HTU for the NTU in equation (e), $NTU_{14} = Z_{24}/HTU'_{24}$; $NTU_{13} = Z_{23}/HTU'_{23}$; and $NTU_{24} = Z_{24}/HTU_{24}$,

$$\text{(f)} \quad Z_{24}/HTU_{24} = \frac{\dfrac{Z_{24}}{HTU_{24}} - \dfrac{Z_{23}}{HTU_{23}}}{1 - Z_{23}/Z_{24}} ,$$

rearranging,

$$HTU_{24} = \frac{HTU'_{24}\ HTU'_{23}\ (Z_{24} - Z_{23})}{Z_{24}\ HTU'_{23} - Z_{23}\ HTU'_{24}}$$

To evaluate the end effect in this work, bed lengths of 13 1/2 in., 7 1/2 in., and 2 in. were used. The 13 1/2 in. data scattered badly because the normal analytical errors were so greatly magnified by the approach to saturation at the top of the tower. Therefore these data were not used in determining the end effect. The equivalent packed height was found to be essentially independent of gas rate (see Fig. 11) with an average value of 3.51 in. This value was used to correct all of the data for end effects.

All of the HTU data taken at a constant liquid rate of 1,575 lb.(m)/hr.ft.^{2} may be correlated by the single curve of Fig. 12 which shows the effect of both gas velocity and gas properties on the gas film HTU. The average deviation of the data points from the line representing the mean is 4.25% with a maximum deviation of 12.5%.

The curves of Fig. 13 show the effect of the liquid rate on the HTU at several air rates. It is rather interesting that the HTU does not bear a simple functional relationship to the liquid rate. It might be expected that as the liquid rate was increased the HTU would decrease due to the increased wetting of the packing. This would explain the shape of the curves at liquid rates below 900 lbs. and above 1,500 lbs.(m)/hr.ft.^{2} The sudden reversal of the liquid rate dependence between these flow rates is difficult to explain on the basis of our present knowledge of flow distribution in a packed tower. One possibility is that the increased liquid flow over this range results in the blocking of some of the smaller flow passages, thereby preventing gas flow through these regions. This would effectively reduce the surface area available for transfer and result in higher values of the HTU.

It was observed that the average slope of the best straight line that could approximate each of the curves in Fig. 13 is approximately the negative value of the slope of the curve of the HTU versus gas rate (below the loading point.) This suggests that the group $\left(\sqrt{\rho u_m^2}\right)_G \Big/ \left(\sqrt{\rho u_m^2}\right)_L$ might serve as a parameter for general correlation as illustrated in Fig. 14. This modulus has been used extensively for the correlation of flooding data. The average deviation of the data from the line representing the mean is 6.2% and the maximum deviation is 24%. In view of the limited

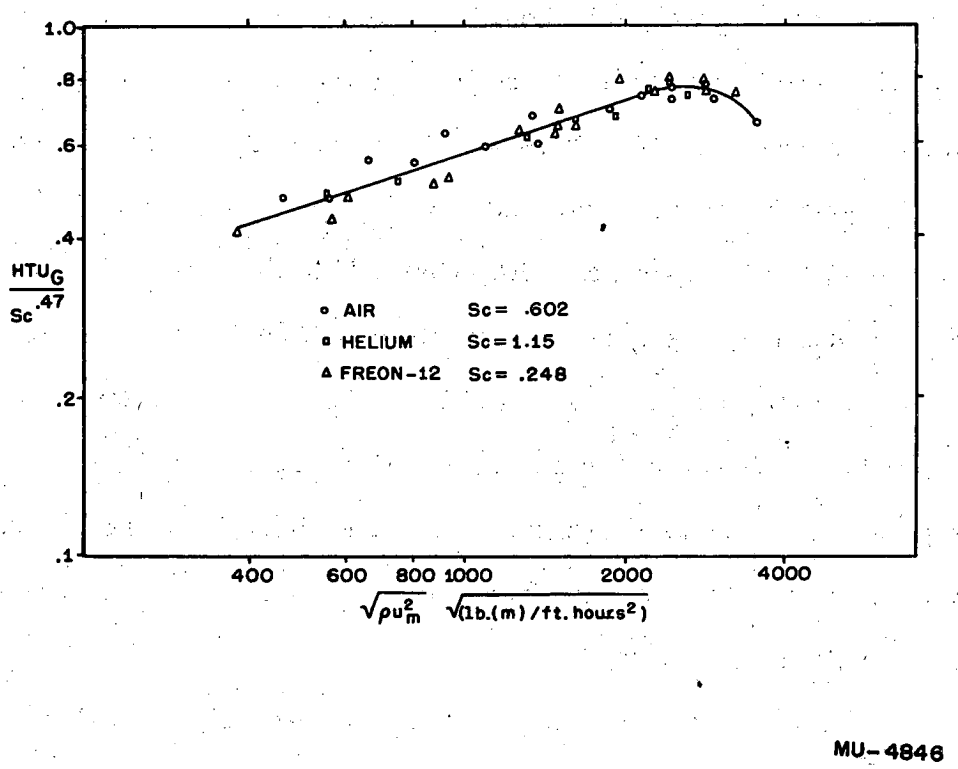

Fig. 12. HTU_G as a function of the inertia group and Schmidt number. L = 1575

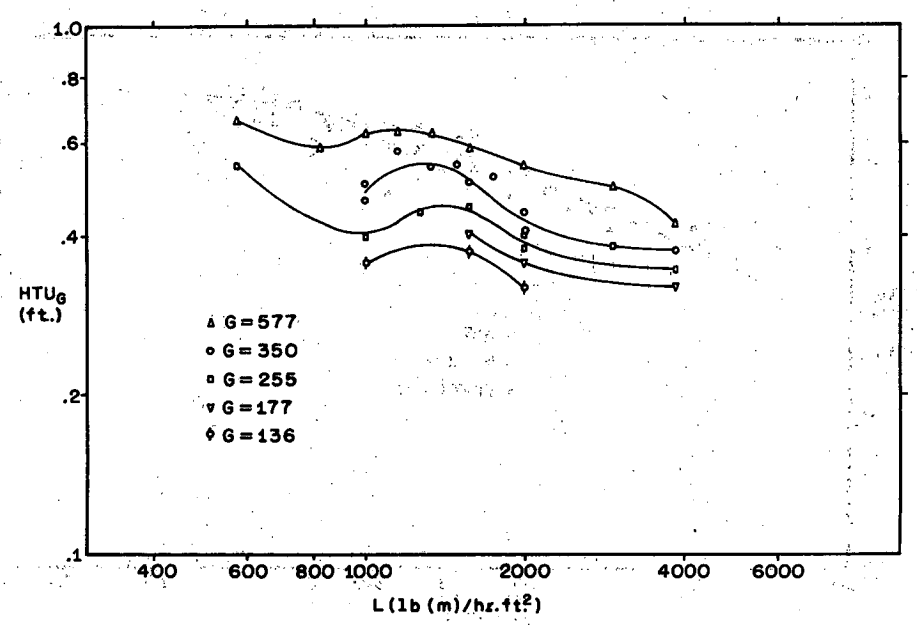

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Fig. 13. HTU_G as a function of the liquid rate

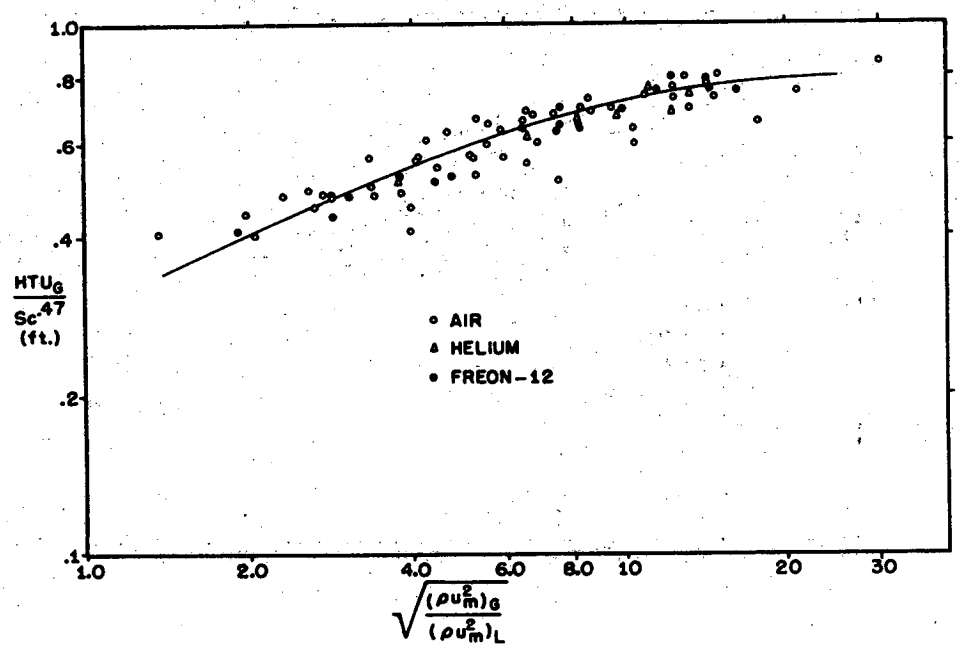

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Fig. 14. HTU_G as a function of the flooding modulus

scope of the liquid variables covered in this work, it is not possible to attach a final significance to this correlation. The influence of liquid density could enter in other ways, and the effect of other properties such as viscosity and surface tension remains to be considered.

DISCUSSION

In spite of the large volume of previous work on the subject, no completely reliable data exist for the resistance to mass transfer in the gas phase in a packed tower. From the numerous studies which have been made, a number of works have been selected to show the variation among investigators that has resulted even though they all used the same packing (1 in. Raschig rings), the same inert carrier gas (air), and directed the experiments specifically toward the determination of gas film data. The results plotted on the graph, Fig. 15, for gas film HTU have been corrected for Schmidt number effect to the system studied in this work (air-water) assuming the HTU to be proportional to $Sc^{1/2}$. No correction was made for slight variations in packing properties since many of the investigators did not report these quantities. A list of the studies is included here. For a more complete description, reference should be made to the original papers.

Dwyer and Dodge (14) - absorption of NH_3 in water,

Fellinger (18) - absorption of NH_3 in water and H_2SO_4,

Houston and Walker (22) - absorption of NH_3, acetone, methanol and ethanol in water,

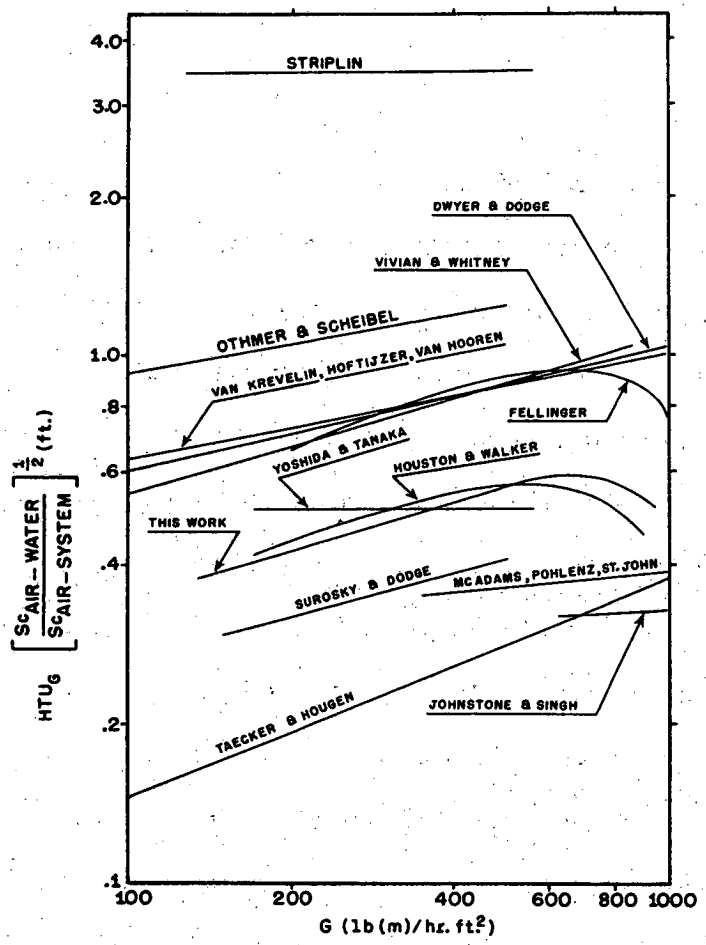

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Fig. 15. HTU_G as a function of gas rate for various investigations

Johnstone and Singh (24) - absorption of SO_2 in NaOH,

McAdams, Pohlenz, and St. John (31) - vaporization of water,

Othmer and Scheibel (34) - absorption of acetone in water,

Striplin (43) - absorption of water in phosphoric acid,

Surosky and Dodge (44) - vaporization of water, methanol, benzene, and n-ethyl butyrate,

Taecker and Hougen (45) - vaporization of water from porous solids,

Vivian and Whitney (48) - absorption of SO_2 in water,

Yoshida and Tanaka (51) - vaporization of water.

Van Krevelin, Hoftijzer, and Van Hooren (46) reviewed a number of gas film mass transfer studies to obtain in a general equation. Since their equation is based largely on the absorption of NH_3 in water, it gives results on the order of those found by Dwyer and Dodge, Vivian and Whitney, and Fellinger. The curve of Taecker and Hougen pertains to systems without liquid flow and is not intended to be comparable to the other systems.

It will be noted that the technique of absorbing NH_3 and SO_2 in water has yielded much higher values of the HTU than either the vaporization of liquids or the absorption of NH_3 and SO_2 in strong chemical solutions (absorption followed by a rapid chemical reaction.) Fellinger found that absorption of NH_3 in H_2SO_4 gave HTU's which were 1.5 to 2.5 times smaller than those for the absorption in water. These acid data scattered somewhat and were not published, but these results would seem to indicate that the nature of the liquid phase resistance to mass transfer in the NH_3-water system may have been improperly evaluated.

An alternate explanation of the discrepancy is the existence of an interfacial resistance postulated by some authors (11, 12, 16). Resistance to mass transfer usually has been considered to exist only in a gas film and a liquid film with a condition of equilibrium existing at the interface between them (49). However, it is possible that the large mass transport in one direction might result in an increase in the concentration of vapor molecules at the interface if the accommodation coefficient were very different from unity. Thus, the actual partial pressure at the interface could be much higher than the equilibrium value and this would result in a lower transfer rate from that which would be expected. Only fundamental experiments on the rate of NH_3 absorption in water will determine whether either or both of these factors is involved in the wide difference in results obtained from the two types of experiments.

The discrepancy between the various studies based on the vaporization of water is probably the result of the analytical difficulties involved in this type of investigation. Saturation of the carrier gas with water occurs very rapidly so that it is difficult to determine the driving forces for mass transfer at the tower exit with the required precision. To alleviate this condition, it is necessary to use very short packed sections. This introduces large relative errors in the measurement of the bed length and also magnifies the effect of the transfer which takes place outside of the packed section.

Although there is a wide variation in the absolute value of the HTU, most of the investigators indicate that it will vary with approximately the 0.3 power of the velocity at flow rates below the loading

point. There is also the growing realization that this portion of the curve is not exactly a straight line and that above the loading point the value of the HTU undergoes a marked decrease for reasons which are not yet entirely clear. The data of Sherwood and Holloway (40) show that the liquid film HTU is independent of the gas rate until the loading point is reached. At the loading point there is a rapid improvement in the rate of mass transfer. This would indicate that at this point there is a marked change in the liquid flow pattern which results in an increased contact area. It is not known whether this change is simply an increase in the wetted area of packing caused by increased gas pressure drop, a rippling of the liquid surface, or a more severe disintegration of the liquid streams, but it seems reasonable to assume that the effect is also responsible for the improved mass transfer rates in the gas film. Fig. 15 gives some indication of the importance of the area effect. The curve of Taecker and Hougen was obtained by using completely wetted porous packing without liquid flowing. The values of the HTU that were obtained by this procedure are approximately one-half those obtained in this work. This also suggests that there would be a great dependence of the gas film HTU on the liquid being used since the fraction of the total available area that is wetted will depend on the characteristics of the wetting fluid.

The dependence of the gas film HTU on the liquid rate as determined by several investigators is shown in Fig. 16. All of the curves show a decrease in the HTU with increase in liquid rate but the dependence on liquid varies widely. Dwyer and Dodge, Yoshida and Tanaka, and Vivian and Whitney indicate that the HTU varies as approximately $L^{-0.2}$ over

the entire range of liquid rates measured. The curve of McAdams, Pohlenz, and St. John is taken from their heat transfer results since their mass transfer results scattered excessively. The data of Surosky and Dodge scatter too much to warrant the assumption that the HTU is independent of liquid rate above a value of 1,200. Actually their data might be fitted as well by either a sloping line or an S-shaped curve. The curve of Fellinger based on liquid rates of 500, 1,500, and 4,500 shows the greatest dependence on liquid rate, the HTU varying with $L^{-0.7}$ at low flow rates and changing gradually to $L^{-0.3}$ at high rates. The shape of the curve is almost identical with that suggested by Van Krevelin, Hoftijzer, and Van Hooren. Fellinger's three values could also be fitted by a curve of the type found in this work. It is difficult to draw a general conclusion from these studies or to say whether they offer either supporting or contradictory evidence for the unique liquid rate dependence found by the authors.

The number of works devoted to the determination of the Schmidt number effect is considerably more limited. Houston and Walker (22) absorbed NH_3, acetone, methanol, and ethanol from air into water and found that the HTU was proportional to the diffusivity to the -2/3 power. Scheibel and Othmer (37) absorbed acetone and methyl ethyl ketone from air into water and obtained data which indicated that the HTU was proportional to the diffusivity to the -1/2 power. However, in trying to correlate their data with the data of other investigators (14) on the absorption of NH_3, they were led to the final conclusion that the correct exponent on the diffusivity was -1. The results of both Houston and Walker and Scheibel and Othmer have possible sources

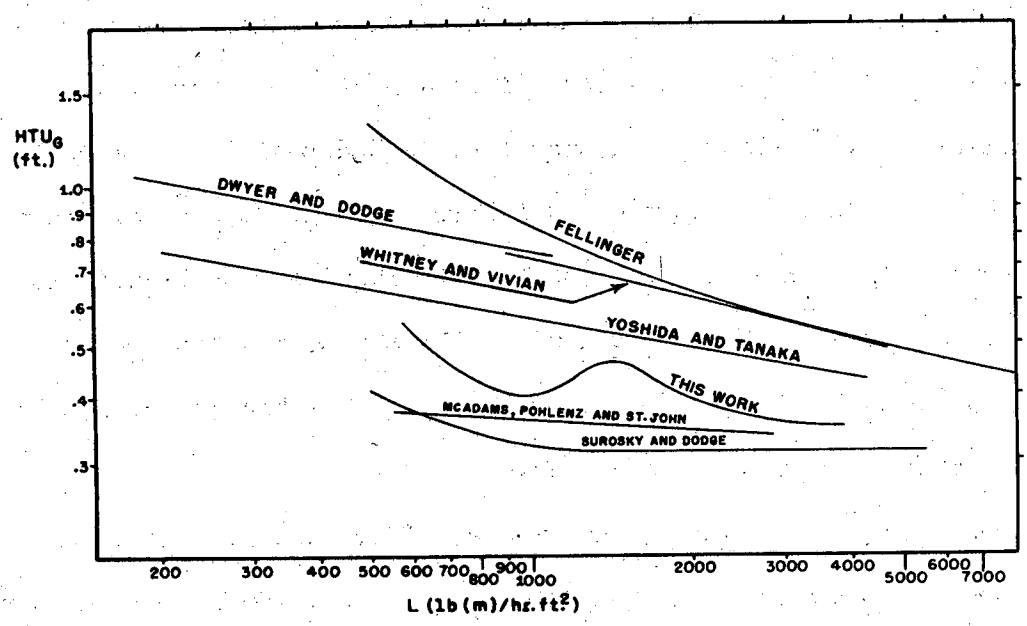

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Fig. 16. HTU_G as a function of liquid rate for various investigations. G = 200

of uncertainty in that it is necessary to correct for the liquid film resistance which is present. In addition, the variation in diffusivity is small.

These difficulties can be avoided by the technique of vaporizing pure liquids into air. Mehta and Parekh (32) vaporized water, methanol, benzene, and toluene and found that the HTU was proportional to $D^{-0.17}$. Surosky and Dodge (44) used water, methanol, benzene, and ethyl butyrate as the liquids and found that their data could be correlated using HTU proportional to $D^{-0.15}$. The data of Simkin (42) and Chrisney (9) on the vaporization of solids and liquids from packed beds in the absence of liquid flow indicate that the HTU is proportional to $D^{-0.36}$. This suggests that in the vaporization of pure liquids with liquid flow there are complications being introduced by the manner in which the various liquids wet and flow over the packing. The possible error from this source has been discussed previously.

It is possible to avoid the above difficulties by the technique used in this work of using a single liquid for vaporization and varying the Schmidt number through the use of different carrier gases. However, it is necessary to ascertain that the various gases do not have different effects on the liquid flow pattern and effective mass transfer surface. If the gas does cause a change in the liquid flow, it should do so primarily by the pressure it exerts. Elgin and Weiss (17) and Jesser and Elgin (23) have found that the liquid hold-up in a packed tower is independent of the gas flow rate at constant liquid rate up to the flooding point. In addition, it was found in this work that the pressure drop across the packing was the same for all three gases at the same value of

the inertia group. Therefore, with this group as the correlating modulus the relative mass transfer rates should be independent of the influence of the pressure drop on flow pattern and effective surface.

TABLE I

Summary of Wet-bulb Data for Various Investigations*

Investigator and System	Pr	Sc	Sc/Pr	$\frac{k_G M_m P r \gamma}{h'} C_p$
Mark				
air-benzene	.702	1.71	2.44	.532
-chlorobenzene	.704	2.17	3.08	.502
-toluene	.704	1.86	2.64	.511
-carbon tetrachloride	.700	1.87	2.67	.499
-ethylene tetrachloride	.704	2.15	3.05	.474
-ethyl acetate	.709	1.83	2.58	.514
-water	.704	.602	.855	.955
Arnold				
air-water	.704	.602	.855	1.049
-methanol	.706	.970	1.37	.800
-propanol	.705	1.30	1.85	.642
-toluene	.704	1.86	2.64	.537
-chlorobenzene	.704	2.17	3.08	.531
-m-xylene	.704	2.24	3.18	.514
Bedingfield and Drew				
air-naphthalene	.704	2.56	3.64	.494
-p-dichlorobenzene	.704	2.22	3.15	.509
-camphor	.704	2.65	3.76	.448
-p-dibromobenzene	.704	2.41	3.42	.487
Dropkin				
air-water	.704	.602	.855	.980

TABLE I (Continued)

Summary of Wet-bulb Data for Various Investigations*

Investigator and System	Pr	Sc	Sc/Pr	$\frac{k_G M_m p_f \gamma}{h'} C_p$
<u>This Work</u>				
air-water	.702	.602	.858	.971
helium-water	.687	1.15	1.673	.650
Freon 12-water	.740	.248	.335	1.482

* The physical properties used in the Schmidt and Prandtl numbers were evaluated at 25°C. Diffusivities for the systems studied by Bedingfield and Drew were recalculated by the method of Arnold (2).

TABLE II

Experimental and Calculated Results for Wet and Dry Bulb Thermometer Measurements

Exp. No.	Dry Bulb	Wet Bulb	Wall	G	p_f	γ	$\frac{k_G M_m p_f \gamma}{h'}$
5	50.93	29.49	48.67	11,604	.967	.985	3.99
6	51.91	29.98	49.25	13,271	.965	.986	4.09
7	51.33	30.27	49.01	15,010	.965	.986	4.00
8	51.66	30.72	48.83	17,460	.963	.986	3.99
1H	50.04	30.06	46.80	3,521	.965	.985	.562
2H	50.75	30.95	49.72	4,974	.964	.984	.527
3H	50.29	31.32	49.48	6,175	.962	.985	.519
4H	51.02	30.69	49.40	4,246	.965	.984	.521
5H	50.23	31.37	49.14	5,631	.962	.985	.511
6H	50.11	31.79	49.19	6,672	.961	.986	.530
14F	51.33	32.17	48.54	20,090	.962	.981	11.26
15F	48.49	31.33	46.38	27,780	.964	.983	10.69
16F	48.19	31.77	46.30	35,180	.963	.982	10.08
17F	48.20	30.74	45.86	18,608	.966	.981	10.59

All temperatures in °C.

G = gas flow rate over the thermometers - lb.(m)/hr.-ft.^{2}.

p_f = film pressure factor - atmospheres.

γ = heat transfer factor.

TABLE III*

Experimental and Calculated Results for Vaporization in Packed Towers

Exp. No.	L	G	T_{AB}	T_{WT}	T_{WB}	y_{iT}	y_{iB}
1	1,575	357.2	34.66	29.00	28.70	.03981	.03912
2	1,575	246.6	32.79	29.01	28.88	.03986	.03955
3	1,575	177.6	31.23	28.88	28.86	.03938	.03950
4	1,575	123.2	30.76	29.16	29.43	.04022	.04084
5	1,575	496.9	39.84	30.87	31.02	.04435	.04469
6	1,575	571.0	40.97	31.30	31.44	.04540	.04573
7	1,575	648.2	40.39	31.45	31.54	.04574	.04594
8	1,575	757.2	40.39	31.51	31.46	.04583	.04564
9	1,575	292.9	34.14	28.69	28.62	.03907	.03891
10	1,575	215.5	32.48	28.57	28.15	.03882	.03788
11	1,575	150.9	31.10	28.59	28.83	.03888	.03942
12	1,575	366.1	35.45	29.06	29.09	.03988	.03994
13	1,575	648.2	40.81	30.87	30.89	.04437	.04438
14	1,575	775.4	40.86	31.41	31.28	.04567	.04527
15	1,575	934.9	41.10	31.66	31.54	.04619	.04578
16	1,010	135.9	29.72	27.62	27.79	.03691	.03728
17	1,995	135.9	29.61	27.74	27.42	.03717	.03648
18	1,995	176.7	30.39	28.00	27.58	.03773	.03685
19	3,850	176.7	30.50	28.05	28.13	.03784	.03802
20	1,010	255.3	32.71	28.52	28.52	.03887	.03887

* See nomenclature at end of table for units.

TABLE III (Continued)

Experimental and Calculated Results for Vaporization in Packed Towers

Exp. No.	L	G	T_{AB}	T_{WT}	T_{WB}	y_{iT}	y_{iB}
21	1,995	255.3	32.58	28.37	28.62	.03853	.03909
22	3,850	255.3	32.24	28.34	28.44	.03846	.03868
23	1,010	354.3	34.87	28.85	28.80	.03963	.03950
24	1,995	354.3	34.64	28.82	28.88	.03956	.03969
25	3,850	354.3	34.48	28.80	28.88	.03951	.03968
26	1,150	331.5	34.98	28.31	27.87	.03822	.03725
27	1,340	330.6	34.87	28.34	27.92	.03829	.03736
28	1,500	329.7	34.61	28.62	28.10	.03892	.03775
29	1,750	328.7	34.61	28.64	28.41	.03896	.03849
30	1,010	350.0	36.12	29.17	28.70	.04020	.03912
31	1,995	350.0	34.92	29.07	28.94	.03997	.03966
32	1,010	350.0	36.17	29.33	29.27	.04054	.04042
33	1,280	254.8	32.97	29.06	28.88	.04002	.03960
34	1,995	254.8	32.66	28.84	28.41	.03951	.03854
35	2,935	254.8	33.48	28.82	28.70	.03947	.03919
36	575	254.8	33.52	29.40	29.20	.04081	.04034
37	1,150	577.2	39.69	30.95	30.94	.04456	.04451
38	1,340	577.2	39.56	30.94	30.89	.04453	.04435
39	1,995	577.2	39.56	30.86	30.78	.04407	.04385
40	2,935	577.2	38.78	30.95	30.86	.04456	.04430
41	3,850	577.2	38.57	30.93	30.73	.04451	.04396

TABLE III (Continued)

Experimental and Calculated Results for Vaporization in Packed Towers

Exp. No.	L	G	T_{AB}	T_{WT}	T_{WB}	y_{iT}	y_{iB}
42	820	577.2	40.21	31.04	30.73	.04479	.04398
43	575	577.2	40.73	31.17	30.70	.04512	.04390
44	1,010	577.2	38.88	30.41	29.98	.04313	.04206
45	1,575	281.5	30.65	27.88	27.35	.03733	.03618
46	1,575	225.8	30.24	27.90	27.32	.03739	.03614
47	1,575	181.1	29.53	28.12	27.58	.03788	.03670
48	1,575	349.7	32.71	28.49	28.15	.03877	.03800
49	1,575	209.2	31.57	28.15	27.45	.03796	.03643
50	1,575	217.6	32.01	29.27	28.96	.04052	.03980
51	1,575	272.1	33.31	29.56	29.56	.04117	.04117
52	1,575	177.5	31.70	29.37	29.53	.04075	.04113
53	1,575	339.5	36.51	31.39	31.10	.04579	.04504
54	1,575	436.4	40.21	32.26	32.24	.04807	.04802
1H	1,575	137.5	38.78	29.07	28.94	.03990	.03959
2H	1,575	202.1	42.53	29.93	29.79	.04181	.04146
3H	1,575	255.5	42.24	30.36	30.24	.04277	.04244
4H	1,575	169.7	41.64	29.62	29.72	.04106	.04129
5H	1,575	231.3	41.75	30.53	30.19	.04315	.04230
6H	1,575	277.6	42.37	31.11	30.84	.04451	.04379
7H	1,575	79.03	34.77	28.70	28.02	.03898	.03747
8H	1,575	58.85	33.36	28.49	27.84	.03851	.03708

TABLE III (Continued)

Experimental and Calculated Results for Vaporization in Packed Towers

Exp. No.	L	G	T_{AB}	T_{WT}	T_{WB}	y_{iT}	y_{iB}
1F	1,575	1,056	41.30	33.95	34.08	.05260	.05284
2F	1,575	1,301	41.43	33.88	34.05	.05238	.05273
3F	1,575	1,514	41.59	34.16	34.18	.05314	.05302
4F	1,575	1,745	42.16	34.39	34.37	.05375	.05349
5F	1,575	807.8	40.26	33.65	33.79	.05177	.05204
6F	1,575	796.7	38.52	33.25	33.58	.05060	.05142
7F	1,575	505.1	38.91	32.55	32.78	.04875	.04926
8F	1,575	306.1	37.50	32.31	31.27	.04813	.04528
9F	1,575	203.8	35.76	32.31	32.67	.04813	.04898
10F	1,575	472.6	35.65	33.00	33.21	.04981	.05040
11F	1,575	329.4	34.74	33.48	33.57	.05111	.05137
12F	1,575	688.0	37.81	33.91	34.04	.05234	.05271
13F	1,575	872.9	40.21	34.20	34.59	.05328	.05451
14F	1,575	874.2	43.00	35.09	35.90	.05592	.05844
15F	1,575	1,216	41.10	34.06	34.12	.05292	.05306
16F	1,575	1,545	40.73	33.98	33.96	.05263	.05251
17F	1,575	808.3	39.51	33.76	33.83	.05207	.05225

TABLE III (Continued)

Experimental and Calculated Results for Vaporization in Packed Towers

Exp. No.	y_T	y_B	NTU	HTU'	HTU	HTU/Sc$^{.47}$	G/μ	$\sqrt{\rho}\, u_m$
1	.03686	.02257	1.725	.362	.532	.675	7,996	1,340
2	.03718	.02240	1.856	.337	.495	.628	5,529	925.6
3	.03709	.02117	2.080	.300	.441	.560	3,978	663.6
4	.03849	.02153	2.453	.255	.375	.476	2,764	461.4
5	.04092	.02639	1.675	.373	.548	.695	11,057	1,877
6	.04168	.02759	1.585	.394	.579	.735	12,686	2,158
7	.04196	.02851	1.529	.409	.601	.763	14,404	2,448
8	.04230	.02968	1.509	.414	.609	.773	16,822	2,858
9	.03690	.02354	1.958	.319	.469	.595	6,559	1,099
10	.03701	.02305	2.104	.297	.437	.555	4,832	807.7
11	.03750	.02303	2.475	.253	.372	.472	3,385	566.2
12	.03772	.02477	1.950	.321	.472	.599	8,190	1,374
13	.04110	.02805	1.609	.388	.571	.725	14,417	2,450
14	.04255	.02980	1.601	.390	.573	.727	17,227	2,930
15	.04377	.03134	1.787	.350	.515	.654	20,780	3,531
16	.03586	.02347	2.577	.243	.354	.453	3,060	508.9
17	.03643	.02351	2.864	.218	.320	.406	3,060	508.9
18	.03670	.02322	2.583	.242	.356	.452	3,974	662.5
19	.03702	.02354	2.872	.218	.320	.406	3,974	662.5
20	.03734	.02363	2.300	.272	.400	.508	5,724	958.7
21	.03718	.02387	2.423	.258	.379	.481	5,724	958.7

TABLE III (Continued)

Experimental and Calculated Results for Vaporization in Packed Towers

Exp. No.	y_T	y_B	NTU	HTU'	HTU	HTU/Sc$^{.47}$	G/μ	$\sqrt{\rho u_m}$
22	.03739	.02365	2.643	.236	.347	.440	5,724	958.7
23	.03771	.02479	2.037	.319	.469	.595	7,940	1,333
24	.03799	.02491	2.243	.279	.410	.520	7,940	1,333
25	.03823	.02502	2.438	.257	.378	.480	7,940	1,333
26	.03546	.02352	1.605	.389	.572	.726	7,429	1,243
27	.03583	.02362	1.720	.363	.534	.678	7,409	1,240
28	.03632	.02354	1.699	.368	.541	.687	7,389	1,237
29	.03652	.02378	1.797	.348	.512	.650	7,366	1,233
30	.03786	.02448	1.834	.341	.501	.636	7,844	1,317
31	.03809	.02465	2.078	.301	.442	.561	7,844	1,317
32	.03839	.02487	1.965	.318	.467	.593	7,844	1,317
33	.03805	.02404	2.067	.302	.444	.563	5,712	956.8
34	.03797	.02394	2.250	.278	.409	.519	5,712	956.8
35	.03805	.02360	2.397	.261	.384	.487	5,712	956.8
36	.03770	.02357	1.685	.371	.545	.692	5,712	956.8
37	.04075	.02817	1.456	.429	.631	.801	12,824	2,182
38	.04083	.02831	1.467	.426	.626	.794	12,824	2,182
39	.04123	.02835	1.697	.368	.541	.687	12,824	2,182
40	.04207	.02859	1.842	.339	.498	.632	12,824	2,182
41	.04276	.02869	2.167	.288	.423	.537	12,824	2,182
42	.04155	.02855	1.561	.400	.588	.746	12,824	2,182

TABLE III (Continued)

Experimental and Calculated Results for Vaporization in Packed Towers

Exp. No.	y_T	y_B	NTU	HTU'	HTU	HTU/Sc$^{.47}$	G/μ	$\sqrt{\rho}\, u_m$
43	.04121	.02837	1.380	.453	.666	.845	12,824	2,182
44	.03966	.02704	1.466	.426	.626	.794	12,824	2,182
45	.03686	.02897	2.731	.416	.524	.665	6,350	1,053
46	.03691	.02667	2.983	.380	.478	.607	5,090	844.4
47	.03764	.02602	3.796	.299	.376	.477	4,083	677.3
48	.03832	.02586	3.296	.344	.433	.549	7,871	1,312
49	.03784	.02634	4.433	.256	.322	.409	4,714	783.5
50	.03510	.02440	1.044	.160	.442	.561	4,868	817.4
51	.03552	.02537	1.028	.162	.448	.569	6,022	1,023
52	.03569	.02461	1.183	.141	.390	.495	3,980	666.5
53	.03709	.02393	.886	.188	.520	.660	7,543	1,282
54	.03957	.02830	.842	.198	.547	.694	9,670	1,654
1H	.03691	.02749	1.398	.447	.657	.615	2,923	1,312
2H	.03819	.02851	1.275	.490	.720	.674	4,283	1,928
3H	.03924	.03010	1.252	.499	.734	.687	5,420	2,431
4H	.03729	.02754	1.294	.483	.710	.665	3,598	1,621
5H	.03916	.02998	1.128	.554	.814	.762	4,904	2,201
6H	.04081	.03193	1.165	.536	.788	.738	5,891	2,634
7H	.03661	.02472	1.683	.371	.545	.510	1,683	754.1
8H	.03627	.02359	1.796	.348	.512	.479	1,252	562.1
1F	.04986	.02712	2.240	.279	.410	.790	33,750	1,966

TABLE III (Continued)

Experimental and Calculated Results for Vaporization in Packed Towers

Exp. No.	y_T	y_B	NTU	HTU'	HTU	HTU/Sc$^{.47}$	G/μ	$\sqrt{\rho}u_m$
2F	.04942	.02531	2.230	.281	.413	.796	41,420	2,424
3F	.04998	.02634	2.250	.277	.407	.784	48,200	2,820
4F	.05128	.02706	2.370	.264	.388	.748	55,430	3,257
5F	.04993	.02359	2.739	.228	.335	.646	25,780	1,504
6F	.04885	.02225	2.810	.222	.326	.628	25,460	1,480
7F	.04783	.02146	3.409	.183	.269	.518	16,152	938.7
8F	.04773	.02143	4.088	.153	.225	.434	9,827	568.5
9F	.04776	.02148	4.309	.145	.213	.410	6,538	377.9
10F	.04912	.02409	3.506	.178	.262	.505	15,140	876.2
11F	.05045	.02399	3.726	.168	.247	.476	10,574	610.0
12F	.05067	.02574	2.780	.225	.331	.638	21,990	1,279
13F	.05138	.02584	2.715	.230	.338	.651	27,790	1,629
14F	.05395	.02836	2.726	.229	.337	.649	27,725	1,635
15F	.05049	.02738	2.358	.265	.390	.752	38,730	2,269
16F	.05036	.02841	2.362	.265	.390	.752	49,190	2,880
17F	.04993	.02511	2.540	.246	.362	.698	25,800	1,506

Note: All runs made at a packed length of 7-1/2" except Nos. 45-49 which were made at 13-1/2", and Nos. 50-54 which were made at 2".

*Nomenclature

L = liquid flow rate - lb.(m)/hr-ft^2

G = gas flow rate - lb.(m)/hr-ft^2

T_{AB} = gas temperature at the bottom of the packed section, ^oC.

T_{WT} = water temperature at the top of the packed section, °C.

T_{WB} = water temperature at the bottom of the packed section, °C.

y_{iT} = interfacial mol fraction of water at the top of the packed section.

y_{iB} = interfacial mol fraction of water at the bottom of the packed section.

y_T = bulk mol fraction at the top of the packed section.

y_B = bulk mol fraction at the bottom of the packed section.

NTU = number of transfer units.

HTU' = apparent height of a transfer unit - ft.

HTU = height of a transfer unit corrected for end effects - ft.

Sc = Schmidt number - $(\frac{\mu}{\rho D})$

G/μ = value proportional to the Reynolds No. - 1/ft.

$\sqrt{\rho u_m^2}$ = inertia group $\sqrt{\text{lb.(m)/ft.-hr}^2}$

TABLE IV

Packing Properties

Type -	carbon Raschig rings
Nominal size -	1 inch
*Outside diameter -	1.005 inches
*Inside diameter -	0.735 inch
*Length -	0.996 inch
Number per cubic foot -	1,316
Area per cubic foot -	56.5 sq. ft./cu. ft.
Dry voids -	0.716
Drain voids -	0.710 to 0.707

*The dimensions above were obtained from the measurement of 200 rings.

NOMENCLATURE

a - area available for mass transfer per unit volume of packing - $ft.^2/ft.^3$

C_m - molal heat capacity at constant pressure - Btu./lb.mole $^oF.$

C_p - heat capacity at constant pressure - Btu./lb.(m) $^oF.$

D - diffusivity - $ft.^2/sec.$

d - diameter - ft.

G - mass velocity of the gas stream based on the superficial area - $lb.(m)/hr.ft.^2$

H - height - ft.

h - heat transfer coefficient - $Btu./hr.ft.^2\ ^oF.$

h' - pseudo-heat transfer coefficient defined by equation (10).

HTU_G - height of a transfer unit for the gas film - ft.

HTU_G' - height of a transfer unit for the gas film - not corrected for end effects - ft.

j_d - mass transfer factor defined by equation (4).

j_h - heat transfer factor defined by equation (3).

j_d' - mass transfer factor defined by equation (13).

j_h' - heat transfer factor defined by equation (12).

k - thermal conductivity - $Btu./hr.ft.^2\ ^oF./ft.$

k_G - gas film mass transfer coefficient - $lb.moles/hr.ft.^2$ atm.

L - liquid flow rate based on the superficial tower area - $lb.(m)/hr.ft.^2$

M_m - mean molecular weight of the gas stream - lb.(m)/lb.mole.

N - rate of transport of vapor - $lb.moles/hr.ft.^2$

NTU - number of transfer units

P - total pressure - atm.

p - partial pressure - atm.

p_f - film pressure factor - atm.

q - heat transfer rate - Btu./hr.ft.^{2}

t - temperature - $^oF.$, $^oC.$

u - velocity - ft./sec.

u_m - average velocity based on the total cross-section area - ft./sec.

y - mole fraction

y_f - p_f/P

Z - packing length - ft.

Greek letters

β - $\frac{k_G M_m p_f \gamma}{h^r} C_p$

γ - a heat transfer factor defined by equation (9a)

μ - viscosity - lb.(m)/ft.hr.

ρ - density - lb.(m)/ft.^{3}

Dimensionless groups

Re - Reynolds number - $du\rho/\mu$

Sc - Schmidt number - $\mu/\rho D$

Pr - Prandtl number - $C_p\mu/k$

Subscripts

a - main stream

av - average

d - refers to mass transfer

G - refers to gas film

h - refers to heat transfer

i - interface

m - mean

m - molal

s - surface

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