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EFFECT OF FLUID PROPERTIES ON MASS TRANSFER IN THE GAS PHASE

By

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EFFECT OF FLUID PROPERTIES ON MASS TRANSFER

IN THE GAS PHASE

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ABSTRACT

To determine the effect of fluid properties on the rate of mass transfer in the gas phase, water was vaporized into air, helium, and Freon-12. Data obtained from a series of wet- and dry-bulb experiments on these systems, in conjunction with similar data obtained by Arnold (3), Mark (36), Bedingfield and Drew (5), and Dropkin (16), indicate that the psychrometric ratio is best expressed as:

$$1.10 \frac{k_G M_m p_f \gamma}{h'} C_p = \left(\frac{\frac{C_p \mu}{k}}{\frac{\mu}{\rho D}} \right)^{1/2}$$

Since the Prandtl number was very nearly constant in the above experiments, it can be concluded that the mass transfer coefficient varies with the Schmidt number to the $-1/2$ power for flow perpendicular to single cylinders.

In the major phase of the investigation, data were obtained for water vaporization in countercurrent flow in a one-foot diameter column packed with one-inch carbon Raschig rings. At a liquid rate of 1575 lbs.(m)/hr. ft.^{2}, the gas rates were varied from fifteen percent of flooding velocity to the loading velocity for each of the three gases.

For air the liquid rate was varied from 575 to 3850 lbs.(m)/hr. $ft.^2$ at several gas rates. Packed heights of 7 $^1/_2$, 13 $^1/_2$, and 2 inches were used in determining end effects which were found to be equivalent to 3.51 inches of packing. When the data for the three gases were compared at a constant value of the gas Reynolds number and constant liquid rate, it was found that the height of a transfer unit was proportional to the 0.9 power of the Schmidt number. When compared at a constant value of the inertia group, ρu_m^2, and constant liquid rate, the height of a transfer unit was found to be proportional to the 0.47 power of the Schmidt number. With the supporting evidence of the psychrometric study, it is concluded that the inertia group is a more satisfactory correlating modulus than the Reynolds number for mass transfer rates in packed towers, and the height of a transfer unit can be expressed in terms of the inertia and Schmidt groups as:

$$HTU_G = \text{function } (\rho u_m^2) \times (Sc)^{.47}$$

The dependence of the HTU_G on the liquid rate cannot be expressed as a simple function, but it may be approximated as being proportional to the liquid rate to the $-^1/_3$ power.

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INTRODUCTION AND THEORY

Review of Fundamental Mass Transfer Theory

Mass transfer in packed towers, and related contacting equipment, is of extensive and obvious industrial importance, and at the present time represents one of the most important facets of chemical engineering. As a consequence, the past twenty-five years have witnessed a continuous and unrelenting effort on the part of chemical engineering research organizations to identify and properly evaluate the variables which govern the rate of mass transfer in this type of equipment. The period of time alone bears mute testimony to the complexity and difficulty of the problem.

The study of mass transfer between flowing gas and liquid phases was originally approached in a purely empirical manner, that of observing a similar or identical system with the one under consideration in pilot-sized equipment. Little or no effort was made to identify the controlling variables. The economic inadequacies of this type of approach are all too obvious, and yet it is still used by industrial organizations under the pressure of expediency. It was not until the advent of dimensional analysis, and the proposal of the "two-film" theory of mass transfer by Whitman (63), that the problem of mass transfer in flow systems received in some measure the direction,

clarification, and organization that its successful solution warrants.

The fundamental rate equation for the interdiffusion of stagnant gases has been developed by Maxwell (37). In this work, we are particularly interested in the special case of the diffusion of one gas through a second inert gas, and Maxwell's general equation reduces to:

$$N_A = \frac{-D_m \partial p_A}{p_B \partial x} \tag{1}$$

N_A = rate of transport of gases A - $\frac{\text{lb. moles}}{\text{hr. ft}^2}$

D_m = molal diffusivity - lb. moles/hr. ft.

p = partial pressure - atmospheres

x = length of the diffusion path - ft.

A refers to the diffusing gas, B to the inert gas.

The practical application of this equation to systems involving flowing fluids is extremely difficult because of the uncertainty in the value of the diffusion coefficient resulting from the influence of turbulence, and the absence of information on the length of the diffusion path.

For a fluid flowing under such conditions that turbulence exists, the flow pattern may be divided into two regions, a turbulent zone and a boundary layer (herein defined to include those regions where viscous forces are important, i.e., the laminar sub-layer and the buffer region). The turbulent zone embraces the main portion of flow wherein the transfer of momentum is by turbulent interchange. This same swirling action which causes the presence of eddies is responsible for the transfer of material by mass motion and gives rise to the term "eddy diffusion" which is used to describe it. As equation (1)

was used to describe the transfer of material by molecular diffusion, a similar equation can be written for eddy diffusion:

$$N_A = \frac{-E_m \partial p_A}{p_B \partial x} \qquad (2)$$

where E_m = eddy diffusivity - lb. moles/hr. ft.

As might be suspected, eddy diffusion is a much more rapid process than molecular diffusion, so that in the turbulent core it completely dominates molecular diffusion and determines the rate of mass transfer.

In the boundary layer, that region of flow which is adjacent to the confining surface, the flow picture is much more complicated in that both the viscous and turbulent forces are important and their relative importance varies throughout. For a short distance from the surface the flow may be considered laminar with turbulent eddies entering gradually thereafter through a transition or buffer region until the turbulence level of the main stream is reached at the inner edge of the boundary layer. An exact treatment of diffusion in this region requires the use of an equation which is a combination of equations (1) and (2) with the values of D_m and E_m properly weighted according to the flow conditions. Such a treatment of mass transfer obviously requires a knowledge of the velocity distribution which is generally unknown and results in equations which are far too complex for general use. To obviate these difficulties, an empirical approach has been taken with the introduction of a mass transfer coefficient, k_G.

In order to include the resistances to mass transfer imparted

by both the boundary layer and the turbulent core while retaining the simplicity of equation (1), it is necessary to visualize a hypothetical model composed of a turbulent core which offers no resistance to transfer and a completely laminar boundary layer of such thickness that its resistance is identically equal to the sum of the resitances of the actual model. With this model, equation (1) can be applied in integrated form:

$$N_A = \frac{D_m(p - p_i)}{B_g \quad p_f} \tag{3}$$

where B_g = the thickness of the hypothetical boundary layer or equivalent film thickness - ft.

p_f = film pressure factor, a term which accounts for the effect of simultaneous diffusion and the presence of other components on the diffusion of component A. For this particular case it is equal to the log mean partial pressure of inert gas - atmospheres.

p = partial pressure of component A in the main stream - atmospheres.

p_i = partial pressure of component A at the surface or interface - atmospheres.

If the mass transfer coefficient is then defined as

$$k_G = \frac{D_m}{B_g \; p_f}, \tag{4}$$

equation (3) becomes:

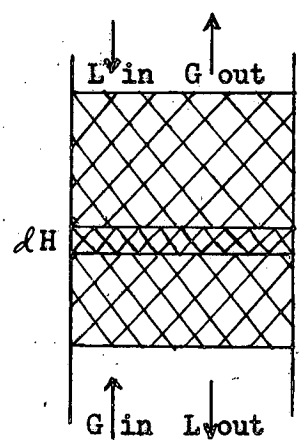
$$N_A = k_G(p - p_i) \tag{5}$$

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

k_G is an experimentally determined coefficient which depends upon the geometry of the system and the properties of the gas involved.

At the present time there is no completely suitable correlation available for its prediction in a packed column.

A general problem in equipment design is to utilize the "point value" mass transfer coefficient discussed above in the calculation of the apparatus size required to effect a desired separation. The following development will be made for the absorption or desorption of gases in a packed tower, but the resulting equations are not so restricted in their application. Consider a differential element of height, dH, over which transfer is occurring from a gas stream into a liquid. The liquid and gas are in countercurrent flow through the packing.



For the gas film transfer of component A

$$N_A = k_G\, P(y - y_i)$$

where y = mole fraction of A in the main stream

y_i = mole fraction of A at the gas-liquid interface

P = total pressure - atmospheres.

For steady state operation,

$$k_G a(y - y_i)P\, dH = -d(G_m y),$$

where G_m = molal gas mass velocity based on the total tower cross section perpendicular to flow - lb. moles/hr. ft.^{2}

a = mass transfer surface per unit packed volume ft^2/ft^3.

H = tower height - ft.

This equation has been integrated for the tower height by Wilke (64):

$$H_2 - H_1 = \int_1^2 \left(\frac{G_m}{k_G \; a \; p_f}\right) \left[\frac{-y_f dy}{(1 - \frac{y}{\alpha})(y - y_i)}\right] \tag{6}$$

where $y_f = p_f/P$

α = moles of gas A diffusing divided by the total moles of gas diffusing. In this work $\alpha = 1$.

The term in parenthesis was given the name "height of a transfer unit" by Chilton and Colburn (10) so that:

$$HTU_G = \frac{G_m}{k'_G \; a \; p_f} \tag{7}$$

and the term in brackets was called the "number of transfer units" or

$$NTU = \int_1^2 \frac{-y_f dy}{(1 - \frac{y}{\alpha})(y - y_i)} \tag{8}$$

Mass Transfer in Packed Columns

Since the geometry of a packed tower is so complex, it is not possible to use a simple boundary layer model as the basis of an analytical solution for k_G (or HTU_G) which would indicate the variables or groups that are significant in determining the mass transfer rates. However, some indication of the significant groups can be obtained by an analogy to the problem of heat transfer in smooth pipes which has been solved analytically by Prandtl (46), von Karman (29), Boelter, Martinelli, and Jonassen (6), and

Murphree (40). An extension of these solutions to mass transfer (see Appendix page 69) shows that the mass transfer rate in pipes is dependent on the Reynolds and Schmidt numbers. If it can be assumed that the same forces act in a packed column and in a pipe, then it can be expected that the mass transfer coefficient for a packed column is a function of these same groups. The transfer rate in a column should also be a function of the shape of the packing and the roughness of the packing surface since these factors will obviously influence the flow pattern. However, it is entirely probable that this dependence will never be evaluated in a satisfactory manner and in the final analysis it will be necessary to use a separate correlation for each packing, or at least for each packing type. By limiting the experimental investigation to a single packing, as has been done in this work, it is possible to eliminate the effects of the shape factor and surface roughness, and to relegate them to a later study after the effect of the Reynolds and Schmidt numbers has been determined.

The pressure drop in the flow of a fluid past a complex shape may be evaluated in terms of a drag coefficient, C_D, which includes both the skin friction and the pressure difference between the front and the rear of the object (form drag). It has been found to be a function of the Reynolds number and is defined:

$$C_D = \frac{\Delta P}{\frac{\rho u_m^2}{2g_c}} \tag{9}$$

where ΔP = total pressure drop across the body - lb.(f)/ft.^{2}

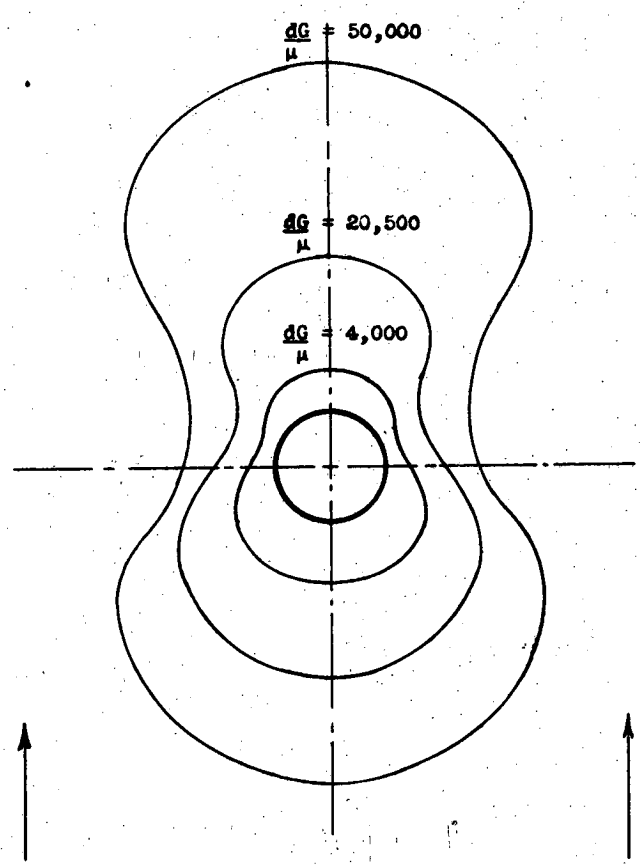


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Fig. 1

Relative rates of mass transfer in flow perpendicular to a single cylinder.

ρ = density lb.(m)/ft.^{3}

g_c = gravity constant - lb.(m)ft./lb.(f) sec.^{2}

u_m = mean velocity - ft./sec.

Studies of mass transfer from cylinders (35) show that most of the transfer takes place in the region of the stagnation point and the turbulent wake, with relatively little transfer on the sides of the cylinder. Since it is the pressure difference between these same two points which is primarily responsible for the drag, it might be suspected that the two phenomena enjoy a somewhat similar relationship to the flow field. Pressure drop measurements made on cylinders (19) show that the drag coefficient becomes a constant at high Reynolds numbers. This means that the drag/unit area is a constant times ρu_m^2 and is independent of the Reynolds number. Since mass transfer takes place primarily in those areas which greatly influence the drag, it seems reasonable to assume that the mass transfer coefficient, k_G, is also some function of ρu_m^2 rather than of the Reynolds number. It might be expected then that for a single cylinder

$$k_G = \text{function } (\rho u_m^2), (\mu/\rho D), \tag{10}$$

where μ = viscosity - lb.(m)/hr.ft.

D = diffusivity - ft.^{2}/hr.

Since pressure drop measurements made on packed towers also show C_D to be essentially constant over the usual range of flow rates below the loading point, this equation should be equally valid for a packed tower, i.e., a group of cylinders. It can be transformed to the HTU form by equation (7):

$$HTU_G = \psi \left[(\rho u_m^2), (\mu/\rho D)\right] \tag{11}$$

where ψ = a function to be determined by experiment.

Simultaneous Mass and Heat Transfer

Wet- and dry-bulb thermometer readings are generally used to determine the vapor content of a gas stream. When the vapor content is known, the readings can be used to determine the relative rates of heat and mass transfer. If an exact analogy can be drawn between the two processes, 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 (12) and Chilton and Colburn (9) set up equations for heat and mass transfer in terms of the j number, a quantity which is analogous to the friction factor in momentum transfer. According to this concept,

$$j_h = \frac{h}{C_p\, G}\left(\frac{C_p \mu}{k}\right)^{2/3} \tag{12}$$

and

$$j_d = \frac{k_G\, M_m\, p_f}{G}\left(\frac{\mu}{\rho D}\right)^{2/3} . \tag{13}$$

By assuming j_h is equal to j_d, the psychrometric ratio is obtained:

$$\frac{k_G\, M_m\, p_f}{h}\, C_p = \left(\frac{\frac{C_p \mu}{k}}{\frac{\mu}{\rho D}}\right)^{2/3} \tag{14}$$

where h = coefficient of heat transfer - Btu./hr. ft.^{2} $^\circ$F.

C_p = heat capacity of the gas - Btu./lb.(m) $^\circ$F.

k = thermal conductivity - Btu./hr. ft.^{2} $^\circ$F./ft.

M_m = mean molecular weight of the gas

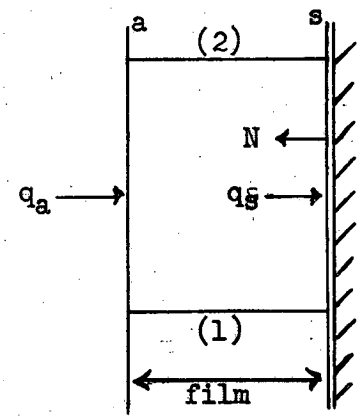
Bedingfield and Drew (5) accepted the form of the above equation but decided that it would fit the available data better if the exponent on the Schmidt and Prandtl numbers was 0.56. Heat and mass transfer data obtained from studies of packed beds[22] indicate that $j_h = 1.076\ j_d$.

The form of the Schmidt and Prandtl number functions given in equation (14) is that which was developed empirically by Colburn. Arnold (3) has developed an equation for the psychrometric ratio based on the equation of Prandtl (46) for heat transfer in pipes. This theory postulates the existence of a laminar flow region at the surface of the thermometer with the remainder of the flow being completely turbulent. The resulting expression includes the velocity at the edge of the laminar layer, u_B:

$$\frac{k_G\ M_m\ p_f}{h}\ C_p = \frac{1 + \frac{u_B}{u_m}\left(\frac{C_p \mu}{k} - 1\right)}{1 + \frac{u_B}{u_m}\left(\frac{\mu}{\rho D} - 1\right)} \cdot \qquad (15)$$

In the above analogies the value of k_G is the one determined from wet-dry bulb data since it is multiplied by the term p_f which corrects for the pressure of the inert gas. The values of h are not the values which would be obtained from the data, but are rather the true heat transfer coefficients that 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.

If a model can be assumed as shown below, such that the transfer of heat and mass through regions (1) and (2) can be neglected (i.e., considering transfer only in the normal direction), a simple equation may be written for the heat transfer to the thermometer surface.



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
$$q_s = q_a - NC_m(t_a - t_s) \ , \tag{16}$$

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 $^\circ$F.

t_a, t_s = gas and surface temperatures respectively - $^\circ$F.

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

$$q_s = q_a = h(t_a - t_s) \tag{17}$$

where h = heat transfer coefficient - Btu/hr. ft.^{2} $^\circ$F., (this is assumed independent of the mass transfer movement).

Substituting from equations (17) and (5) for q_a and N into equation (16),

$$\begin{aligned} q_s &= h(t_a - t_s) - C_m\, k_G(p_s - p_a)(t_a - t_s) \\ &= h(t_a - t_s)(1 - C_m\, k_G\, \Delta p/h) \qquad (18) \\ &= h(t_a - t_s)\, \gamma \end{aligned}$$

where $$\Upsilon = 1 - C_m k_G \Delta p/h \tag{18a}$$

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

then $$h' = h \Upsilon, \tag{19}$$

and the correct psychrometric ratio is:

$$\frac{k_G M_m p_f \Upsilon}{h'} C_p = \text{function of (Sc, Pr)} \tag{20}$$

A treatment similar to that shown above was developed by Colburn and Drew (13) for correcting the heat transfer coefficient in the case of condensation of mixed vapors. However, they did not apply their equation to the correction of wet- and dry-bulb data, although it could be so used.

EXPERIMENTAL

Plan of Investigation

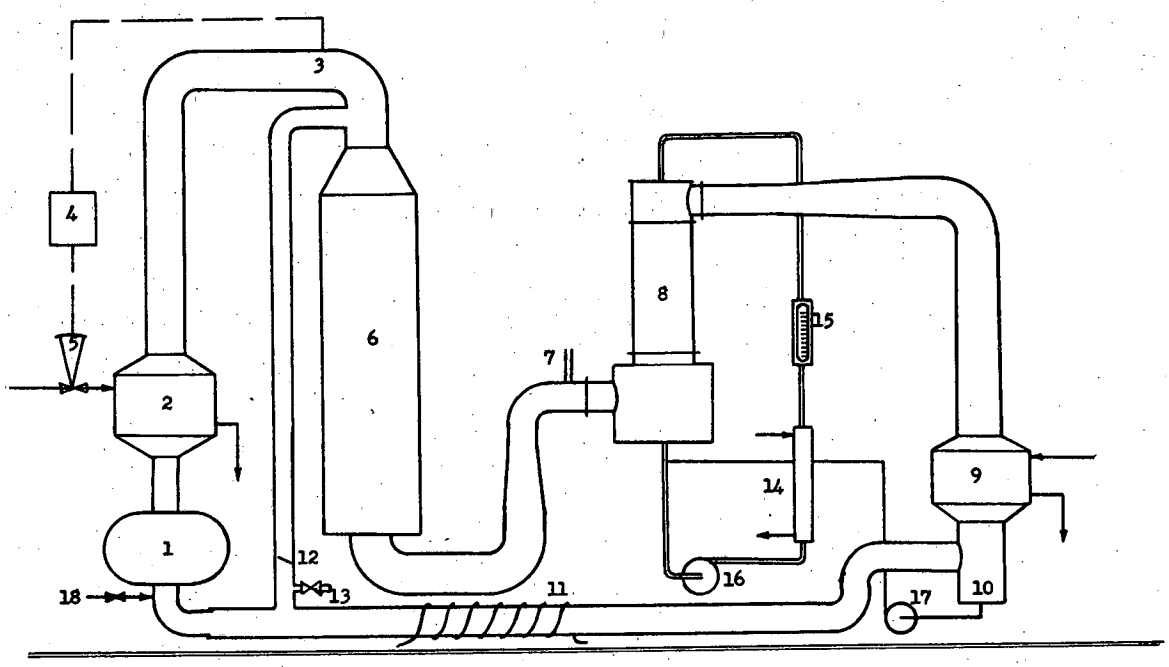
The purpose of this investigation was the determination of the effect of the fluid properties on the rate of mass transfer in packed towers. (The role of the Schmidt number was of particular interest.) In order to do this, a twofold plan was followed. The most direct method was the actual measurement of gas film mass transfer coefficients in a packed tower. In order to avoid the uncertainties experienced by other investigators, it was decided to evaporate a single pure liquid into inert gases. The gases chosen were air, helium, and Freon-12. This gave a thirtyfold variation in density, a twofold variation in viscosity, and an eightfold variation in diffusivity, with water as the liquid. These properties combined to give a four-and-one-half fold variation in Schmidt number.

The second method of approach was the use of wet- and dry-bulb measurements to determine the psychrometric ratio. If the Prandtl number is very nearly constant, this method will give the variation of the mass transfer coefficient with Schmidt number in evaporation from single cylinders. It has the advantage of not requiring a knowledge of the effect of the mass velocity term provided the analogy between heat and mass transfer can be assumed valid. While it is not necessary that the same Schmidt number effect will hold for single cylinders and packed towers, the expected effect should be of similar magnitude and the results should serve as an indication to confirm the results obtained by the more direct method.

Equipment

To carry out this experimental program, equipment was set up as

shown in the accompanying diagrams. Gas circulation was provided by a positive displacement Roots-Connersville blower which was capable of a maximum output of 700 cfm. The gas temperature was maintained by steam heated coils, the supply for which was controlled by a Brown Electronik temperature recorder-controller. In general, it was possible to keep the temperature fluctuations within $\pm$ 1/2 °F. at the control point (no.3, Fig.2). (The metal of the flow cabinet and ducting provided a damping action which limited fluctuations to $\pm$ 0.1 °F. at the tower inlet.) The heated gas entered a 20 inch diameter section (Fig.5) containing straightening vanes and a flow nozzle made to the specifications of Bean, Buckingham, and Murphy (4). For low flow rates, a calibrated orifice was placed over the end of the nozzle. The calibration for this orifice, as well as a calibration of the nozzle at very low flow rates, is included in the appendix. The gas next passed to the wet- and dry-bulb set-up. An orifice was placed in the ducting, (Fig.9), preceding the thermometers to provide a uniform high velocity flow across the thermometer bulbs. A thermocouple was also inserted in the duct wall adjacent to the thermometers to determine the wall temperature for the radiation corrections. From here the gas passed into the gas distributor (Figs. 6 and 7) where it was sampled and then entered the tower through the risers which discharged about one inch from the bottom of the packing. The temperature at the riser discharge was measured by a thermocouple inserted in the gas stream. The temperature was measured again about six inches above the packing. The gas leaving the tower passed through a series of four mixing baffles in the exit duct before being sampled again. The



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 lay-out.

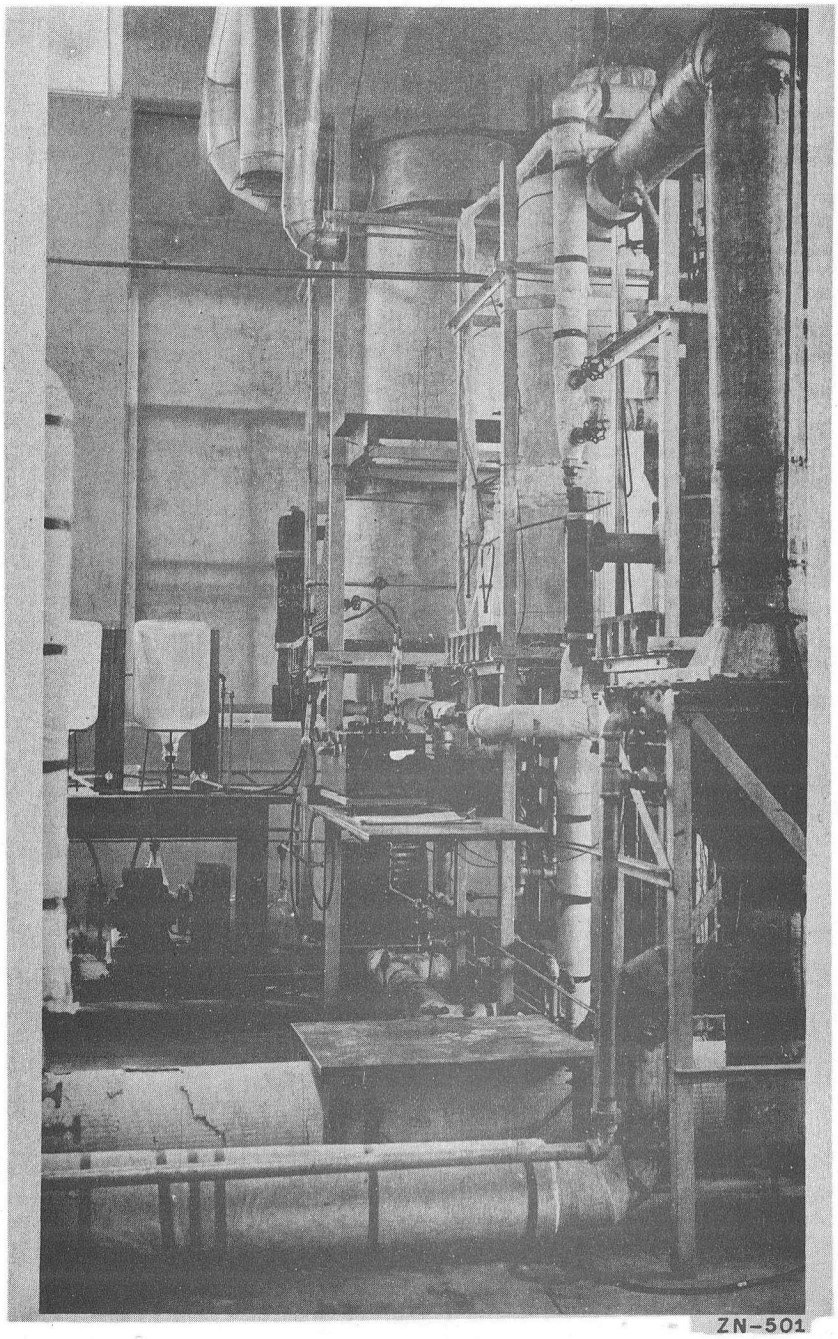

Fig. 3

Photograph of the absorption tower.

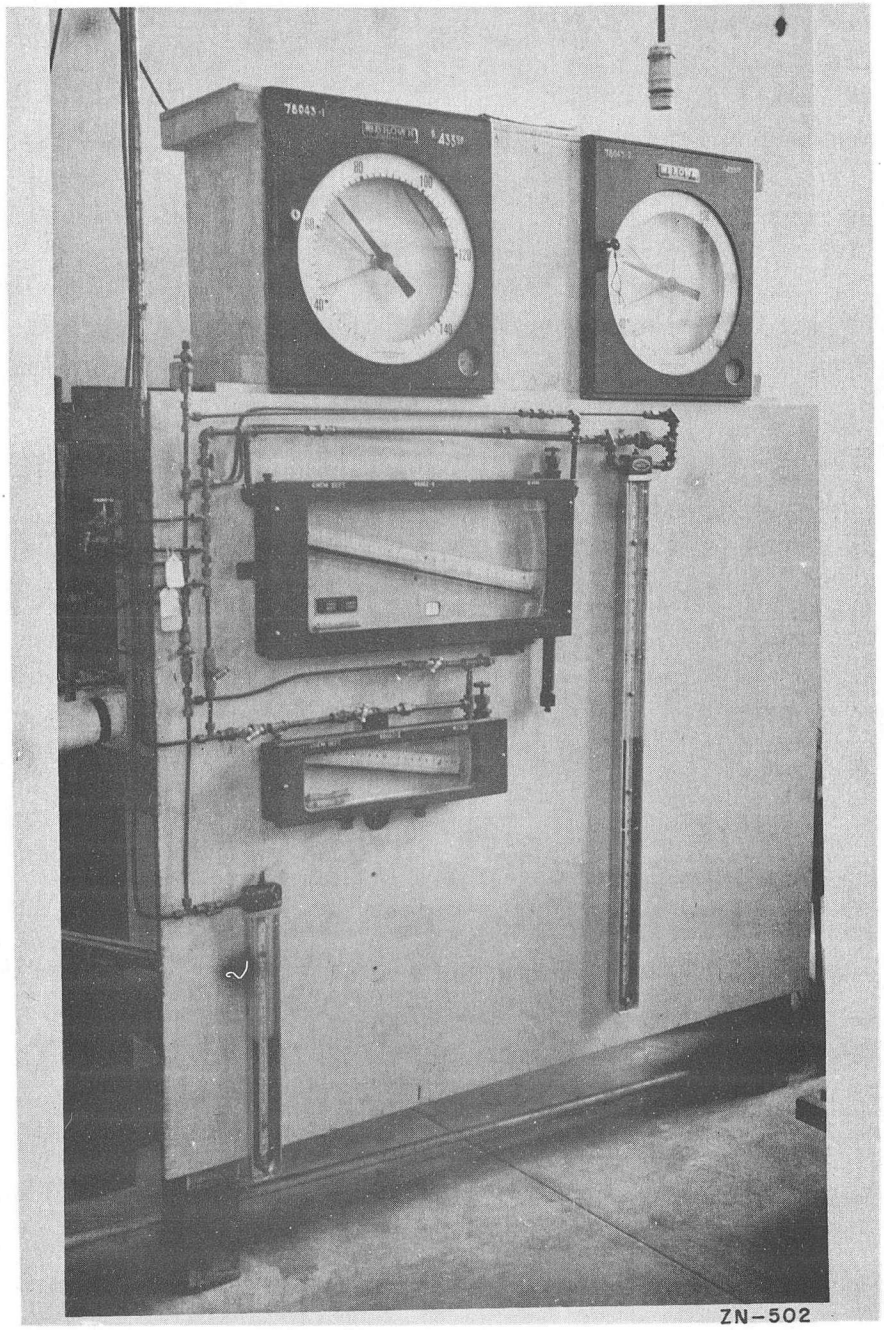

Fig. 4

Photograph of the instrument board.

water was removed from the exit gas by passing it through a water-cooled coil, the condensate being returned periodically to the liquid system. A heating coil was provided on the return line to the blower to slightly superheat the gas and to eliminate the possibility of condensation in the blower. Variable gas rates were obtained by the use of a variable speed drive on the blower and a by-pass on the circulation system.

The large section at the bottom of the tower (Fig. 6) served as a liquid reservoir with circulation being provided by a centrifugal pump. After leaving the pump, the liquid passed through a double pipe heat exchanger which controlled the inlet temperature to the tower. The flow rate was measured by a calibrated rotameter. The liquid discharged into the tower through a 44-point distributor which discharged within one-half inch of the top of the packing. Each of the tubes in the distributor delivered liquid at a rate which was within ten percent of the average rate for a single tube. The temperature of the inlet liquid stream was measured by a four-couple thermopile. The thermocouples were inserted into rubber sleeves which replaced the last three-and-one-half inches of four of the copper distributing tubes. The liquid temperature at the bottom of the packing was measured by a single couple located in a small cup. The cup had a small drain hole in the bottom to insure liquid circulation. In addition, the liquid temperature was measured by thermometers placed in the line prior to its entry into the tower and in the reservoir. These served solely as a check on the thermocouples.

The tower was one foot in diameter and was packed by wet

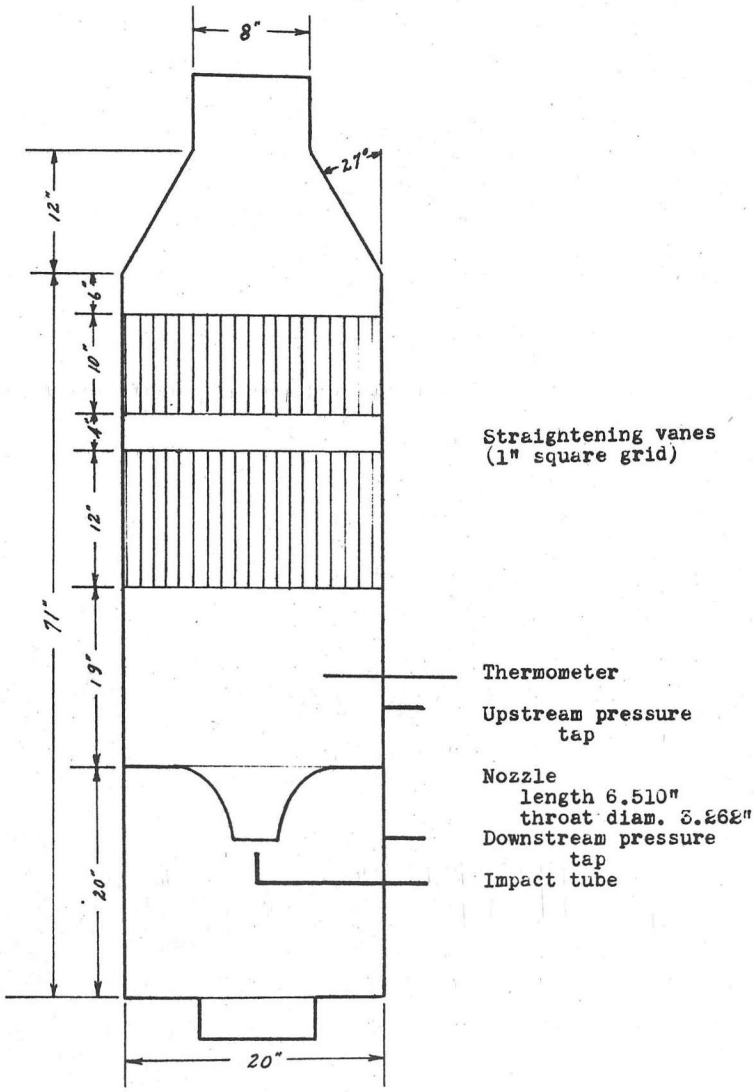

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Fig. 5

Flow measuring chamber.

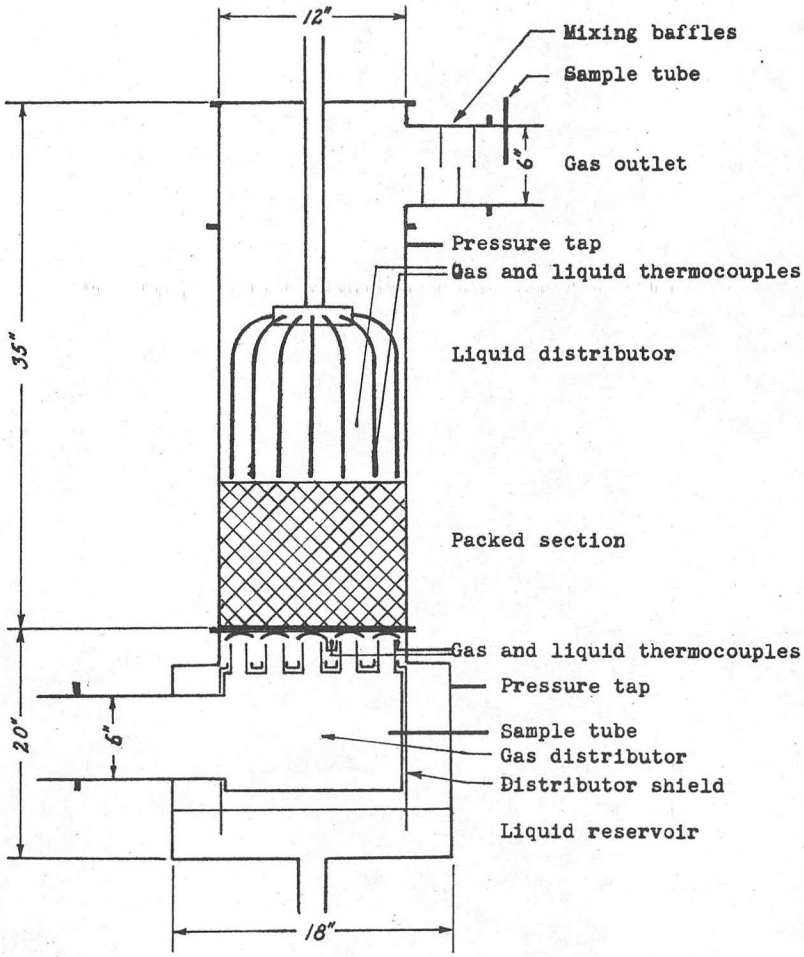

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Fig. 6

Absorption tower

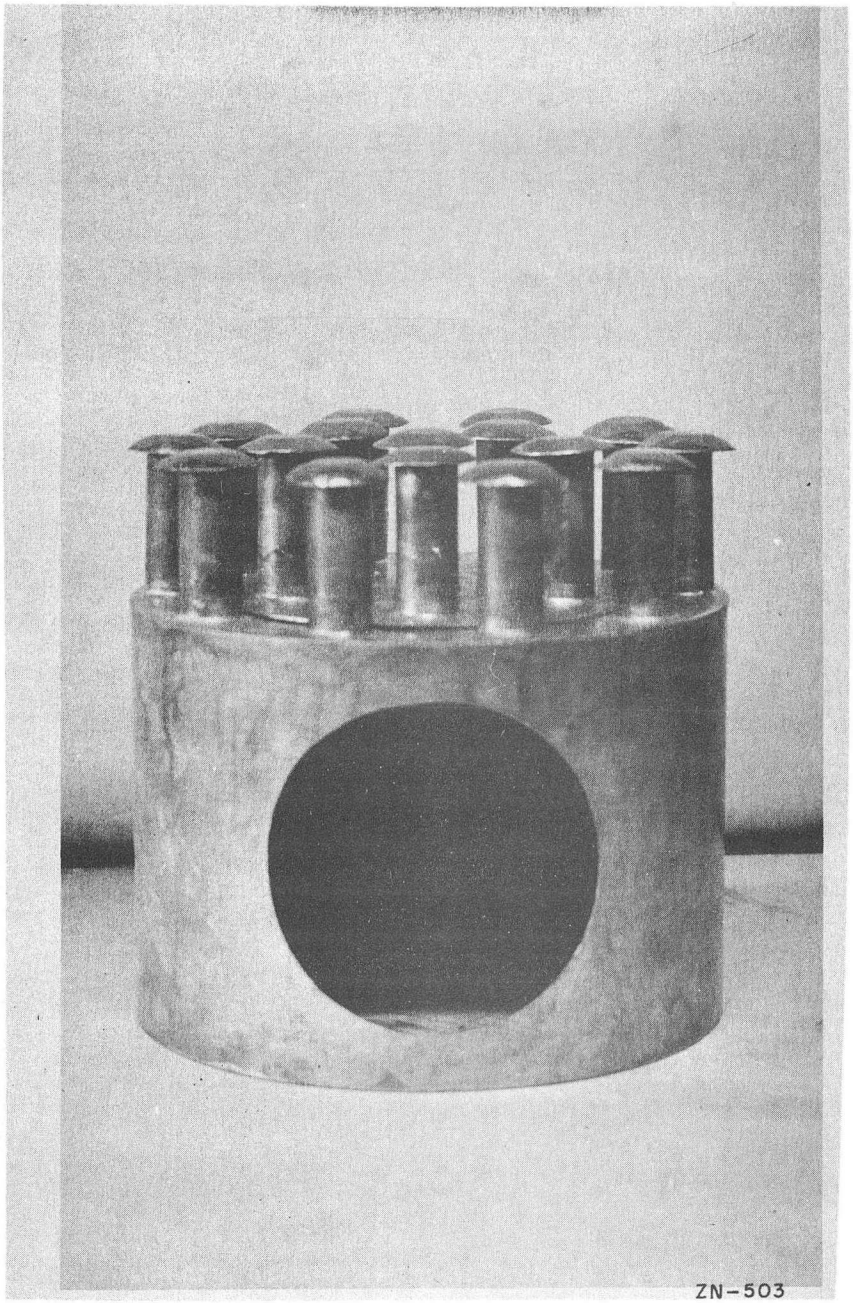

Fig. 7

Photograph of the gas distributor.

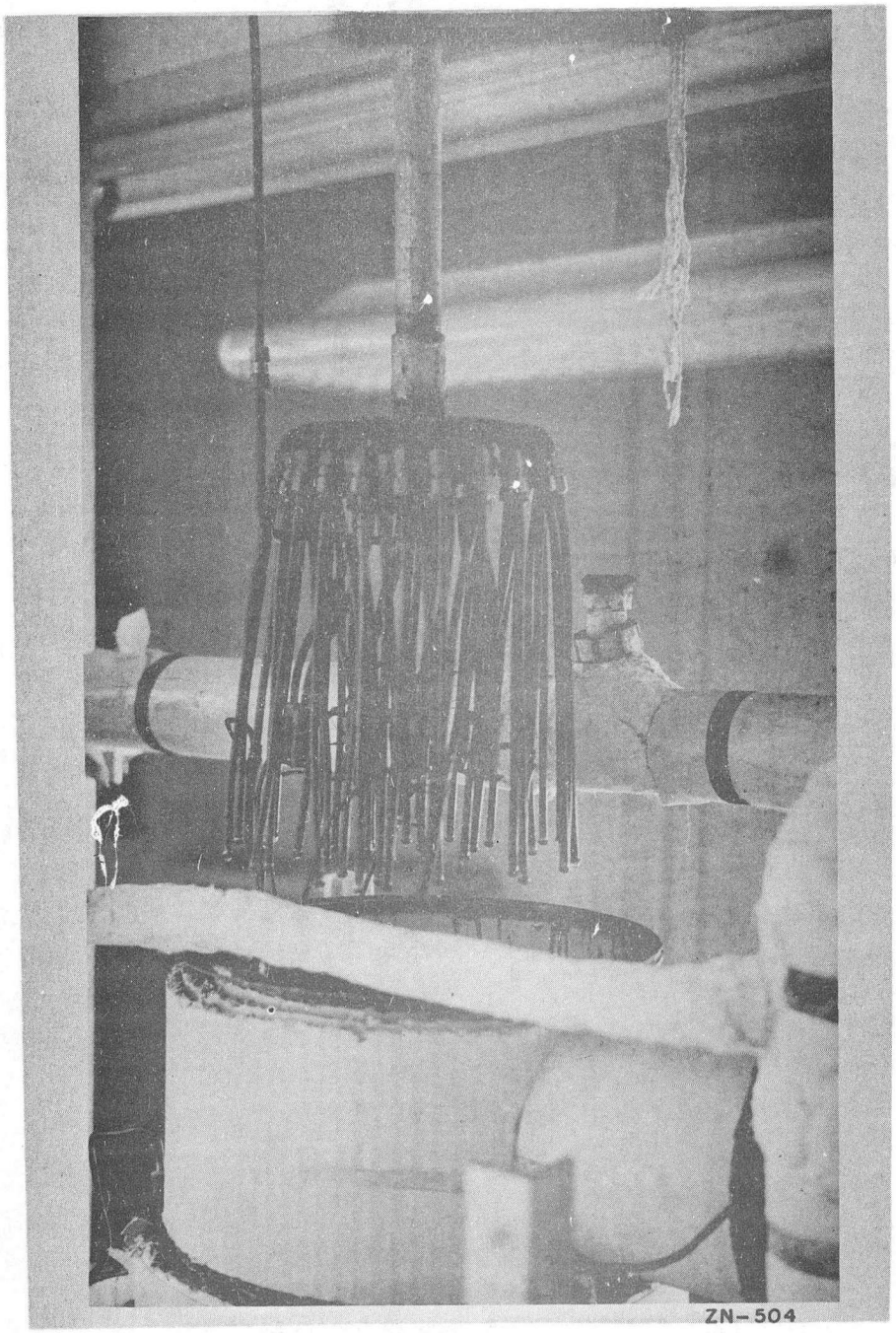

Fig. 8

Photograph of the liquid distributor.

dumping with one-inch carbon Raschig rings. The packing was supported by a one-inch mesh steel screen which was in turn supported by a metal grid 1/8 inch thick. The tower was wrapped with a one-inch thick glass wool insulation which was covered by two layers of corrugated asbestos paper. The exit section of the tower was wrapped with a heating coil to eliminate the possibility of condensation of water in the exit gas prior to sampling.

The sample lines were covered with approximately 1/2 inch of magnesia insulation and double traced with heating coils. The moisture content was determined by passing the sample through two drying tubes in series filled with Drierite. The junction of the sample line and the drying tube took place in a heated box. A tee was also placed in the sample line at this point to permit a small gas pump to continuously draw a sample through the lines at the approximate sampling rate while sampling was not actually in progress. A sample was drawn by permitting water to flow from a bottle of known volume. To insure a constant tower pressure when operating with gases other than air, the water was run into a second bottle filled with gas, forcing the gas into the system at the same rate at which the sample was being drawn.

The pressure at the top of the tower and the pressure drop across the packed section were read on a U-tube manometer and a two-inch draft gage respectively. The draft gage reading could be estimated to 0.005 inch of water. The static pressure preceding the nozzle and the difference in head between the impact tube and the static pressure at the nozzle discharge were read on a five inch draft gage. The reading could be estimated to 0.01 inch of

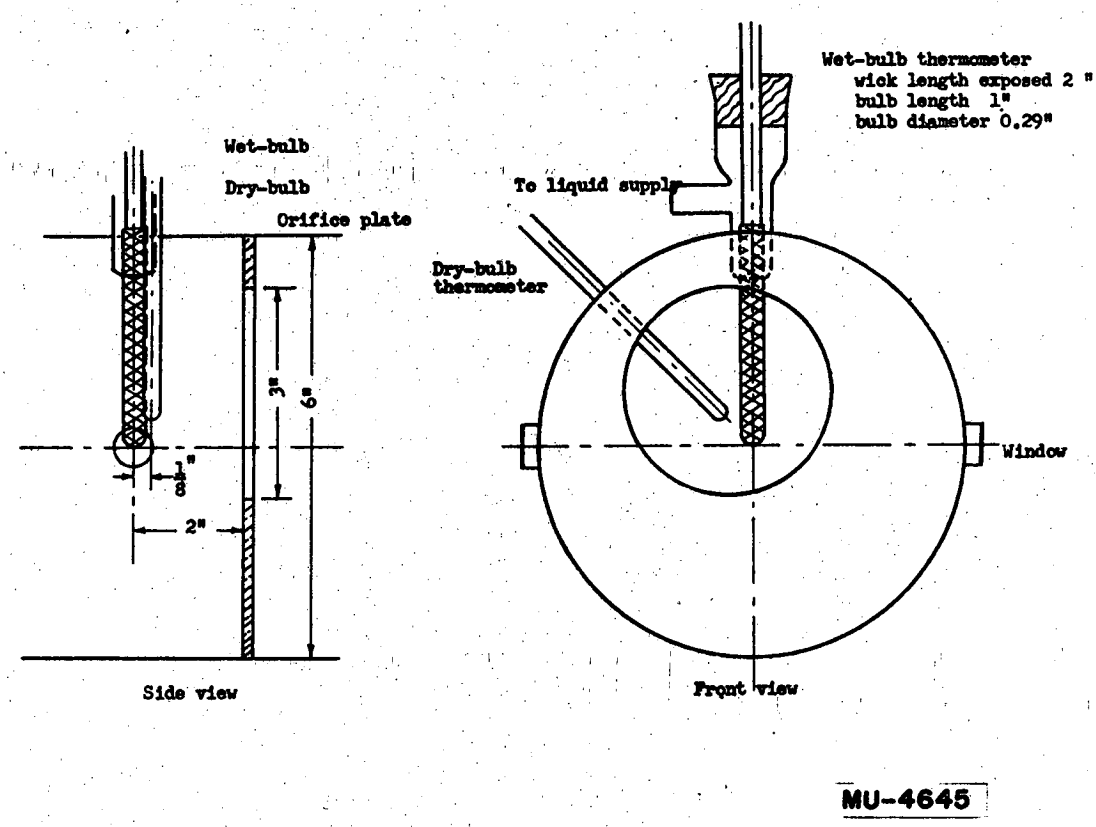

Fig. 9

Wet-and dry-bulb thermometer arrangement.

water.

The wet-dry bulb assembly is shown in Fig. 9. 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. Windows were provided in the ducting at the point where the thermometers were inserted so that the positioning and wetting of the wick could be observed during a run. The windows were made sufficiently small so that they would not affect the radiation correction. The ducting at this point was covered with a one-inch thickness of magnesia insulation.

The thermocouples used in determining the gas and liquid temperatures in the tower were calibrated against a Bureau of Standards' thermometer. The emf.s 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 outlet gas temperatures were found to be in error at the low flow rates due to radiation and conduction losses in the thermocouple and consequently were not used in the final calculations.

Procedure

For all runs made with gases other than air, it was necessary to first purge the system before each series. The size of the equipment and frailty of the ductwork made it unwise to either purge with the gas to be studied or to attempt to remove the air through evacuation. Consequently, a steam purge system was chosen since the presence of water in the system was necessary anyway.

To accomplish this, the blower was operated and live steam injected at the inlet, forcing the air ahead of it as it moved through the system. The butterfly valve on the by-pass was closed and the air exhausted through the vent valve (no. 13, Fig. 2). Since some mixing of the steam and air was inevitable, the steam injection was continued until such time as a thermometer in the return line to the blower indicated a temperature of approximately 100° C. At this time, the vent and steam valves were closed and the gas to be studied was introduced, filling the system as the steam condensed. During the steam purging, all auxiliary lines were also purged with the gas. The steam purging took an average of about ten minutes and resulted in essentially complete removal of the air. Air concentration was checked by means of an Orsat analysis for oxygen which was sensitive to ± 0.05 per cent. After purging was complete, condensed steam was removed from the lines and the necessary adjustments were made to bring the equipment on stream. To prevent introduction of impurities, the equipment was operated with a slight positive pressure at the blower inlet.

At least three hours were required between start-up and the first run to obtain steady-state conditions. Subsequent changes required from one-half to one-and-one-half hours for the attainment of steady-state as indicated by constant gas and liquid temperatures. Then samples were drawn simultaneously from the top and bottom of the tower through Drierite tubes for about twenty minutes. The amount of water collected by the Drierite was determined by weighing and was recorded to 0.0001 grams. The temperature and pressure of the gas in the aspirator bottles were also recorded. The temperatures

of both gas and liquid streams were measured and the inlet water temperature was closely watched and controlled by varying the flow of cooling water through the liquid system heat exchanger. In general, it was possible to operate with the inlet and exit liquid temperatures less than 0.5° C. apart, with the variation of the inlet temperature during a run of less than ± 0.05° C. All flow rates and pressures were recorded.

Immediately following sampling, water from an overhead supply was introduced to the wet-bulb wick and allowed to thoroughly soak the cloth. 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 then recorded. The water supply was stopped and the wet bulb temperature checked for any further changes. (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 seven-and-one-half inch depth of packing with the exception of the runs made at the end with heights of thirteen-and-one-half inches and two inches which were used to determine the amount of transfer taking place outside the packed section. Caution was always exercised to avoid flooding, which had been observed in early experiments to cause a rearrangement of the packing. All runs were made at a liquid rate of 1575 lbs.(m)/hr.ft.^{2} with the exception of those made with air to check the effect of varying the liquid rate. Since the type of

distributor used gave erratic liquid distribution below rates of 1000 lbs.(m)/hr.ft.^{2}, only three runs were made below this rate.

RESULTS

Wet- and dry-bulb Temperature Measurements

The three values of the psychrometric ratio, $\frac{k_G M_m p_f \gamma}{h'} C_p$, obtained in this work are shown plotted against the ratio Sc/Pr in Fig. 10. These results alone would indicate a function having a slight curvature, but within the precision of the data, it can be approximated by a straight line having a slope of $-1/2$. This graph has been extended to include the work of Arnold (3), Mark (36), Bedingfield (5), and Dropkin (16). Making allowance for the fact that the results of Mark have not been corrected for radiation and are consequently low, a best line has been drawn through the points. This line also has a slope of $-1/2$. Since the value of the Prandtl group is nearly constant, the heat transfer coefficient may also be considered constant. The curve then represents the variation of the mass transfer coefficient with Schmidt number for the flow of gas perpendicular to single cylinders, i.e., k_G is proportional to $Sc^{-1/2}$.

The curve drawn is best suited by the j number analogy if $j'_h = 1.10\ j'_d$ where:

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

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

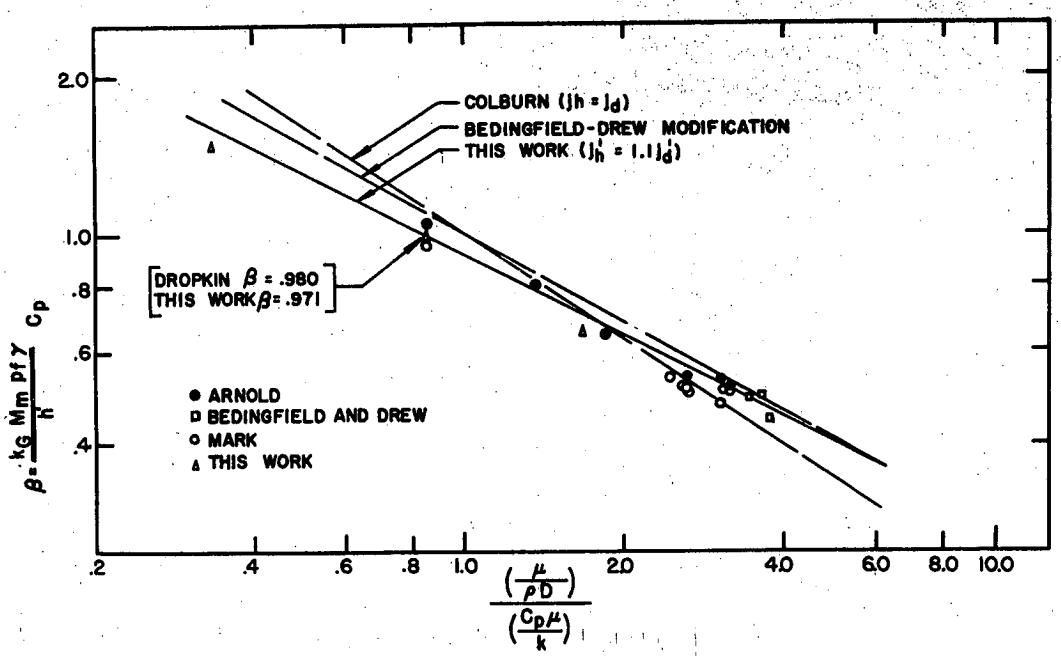

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Fig. 10

The psychrometric ratio as a function of Sc/Pr.

This is in good agreement with the results obtained from packed beds where it has been found empirically (22) that the ratio of j_h to j_d is 1.08. Neither the curve based on the equality of the Chilton and Colburn j numbers (equation 14) nor the modification of Bedingfield-Drew fit the data very well, particularly at the lower values of the Schmidt number.

An alternate theoretical treatment has been presented by Arnold (equation 15). An attempt was made to apply this method by evaluation of the ratio u_B/u_m from the data of Bedingfield and Drew and this work:

System	Velocity (ft./sec.)	Average u_B/u_m
air-naphthalene	31 to 87	.44
-p-dichlorobenzene	28 to 84	.52
-camphor	46 to 86	.55
-p-dibromobenzene	45 to 90	.55
-water	56 to 115	-.22
helium-water	74 to 145	.84
Freon 12-water	35 to 74	.57

A value of u_B/u_m = 0.52 seems to give the closest fit of the data. This agrees well with the average value of 0.56 suggested by Arnold. However, it is not possible to notice a trend in the value of the ratio with Reynolds number within the precision of the data as has been suggested by Arnold. Figure 11 shows the curves obtained, using the value of 0.52 (solid line) and the value of 0.56 (dashed line). Neither curve satisfies the data very well. The theoretical requirement that the curve pass through the point (1,1) results in a

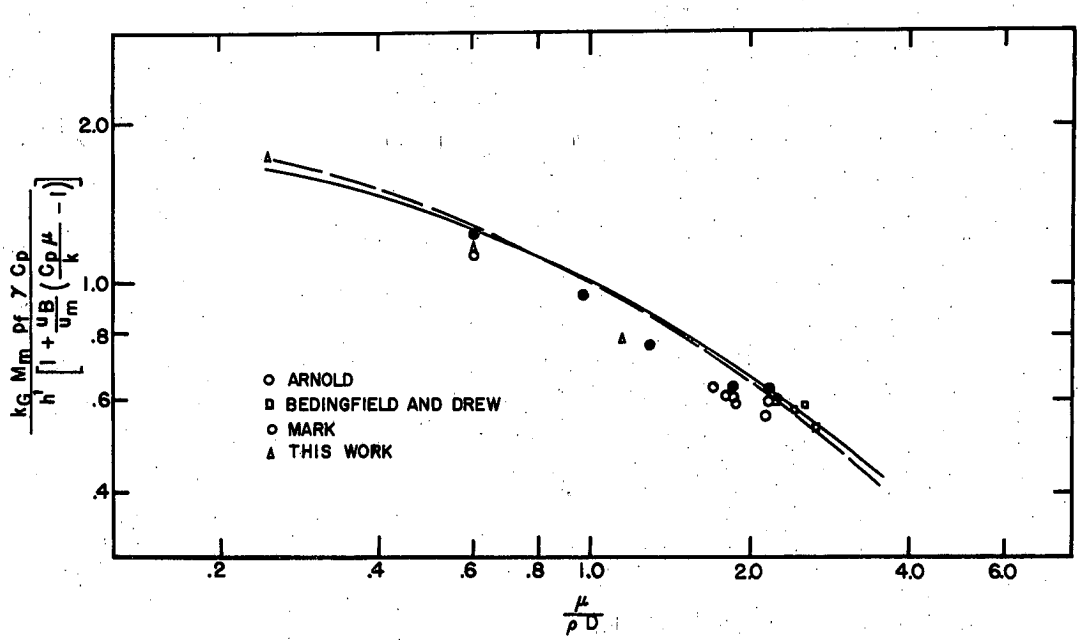

FIGURE 11

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Fig. 11

The spychrometric ratio according to the Arnold theory.

serious disagreement with the available data for the system air-water. Only the experimental result of Arnold agrees with the Arnold theory. The results of Mark, Dropkin, and this work indicate that the curve is approximately ten percent too high. Dropkin's work, in particular, must be given special consideration since it was a very carefully performed investigation with every precaution taken to eliminate thermometric and radiation errors.

It is concluded that the curve of Fig. 10, where $j_h' = 1.10\ j_d'$, represents a superior correlation of the available data to those suggested by either Arnold, the extension of Colburn's j number analogy, or the modification of Bedingfield and Drew.

Packed Column Measurements

The packed tower results are presented here in terms of the HTU instead of k_G. To calculate the HTU, the number of transfer units was first obtained from equation (8) by taking y_f and $(1 - y)$ at the arithmetic mean values in the tower, and using as the driving force, $(y - y_i)$, the logarithmic mean of the terminal values. The apparent value of the height of a transfer unit, HTU_G', was then obtained by dividing the length of the packed section by the number of transfer units.

Figure 12 shows the apparent HTU as a function of the group G/μ which is directly proportional to the Reynolds number. 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

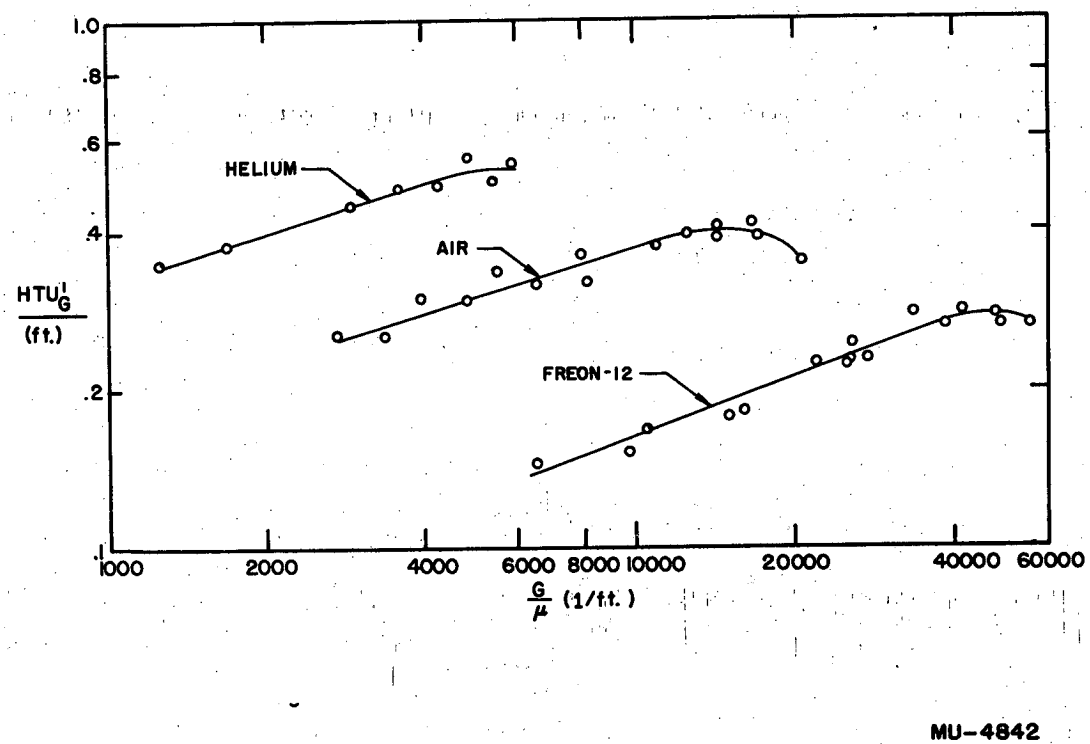

Fig. 12

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

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. Whether this decrease is the result of the agitation of the liquid surface, increased wetting of the surface of the packing, or some other cause is not known. It is believed that the straight line portion of these curves could be represented more truly by gentle curves (convex upward) but the data are not sufficiently precise to justify such a procedure. 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 by the Schmidt number differences, but the lateral separation would lead one to suspect that the Reynolds number may not be the proper correlating modulus in this case. If instead, the group suggested by equation (11) is used, i.e., ρu_m^2, the curves of Figure 13 are obtained. (The group ρu_m^2/length has the dimensional significance of inertia force per unit volume of fluid (61), 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 Figure 12. Similar usage in the correlation of flooding velocity (53) and pressure drop (48) data in packed columns would seem to justify the procedure. This group has also been used in the correlation of plate efficiences (23)

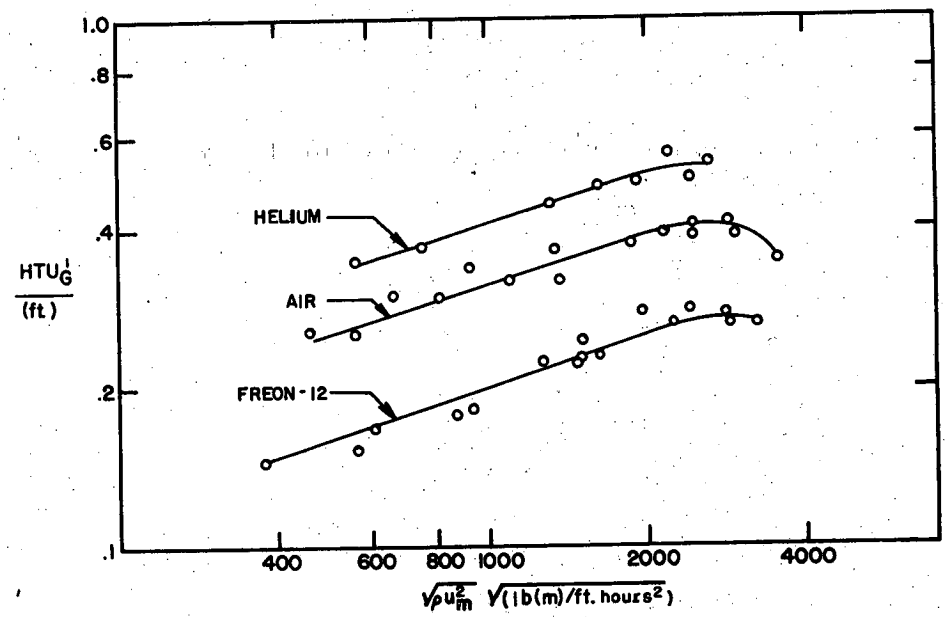

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Fig. 13

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

where it has been given the name F-factor. The curves of Figure 13 are identical in shape with those of Figure 12, 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 Figure 12 and at a constant value of $\sqrt{\rho u_m^2}$ of 1,000 in Figure 13 are presented in Figure 14. 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 Figures 12 and 13. The plot based on the Reynolds number would require this function to be $Sc^{.9}$. 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 only 0.47 on the Schmidt group to align the three curves. This exponent is in good agreement with that obtained in the psychrometric study (section a) and with some results obtained in the vaporization of liquids and organic solids in packed beds (11), (54). 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

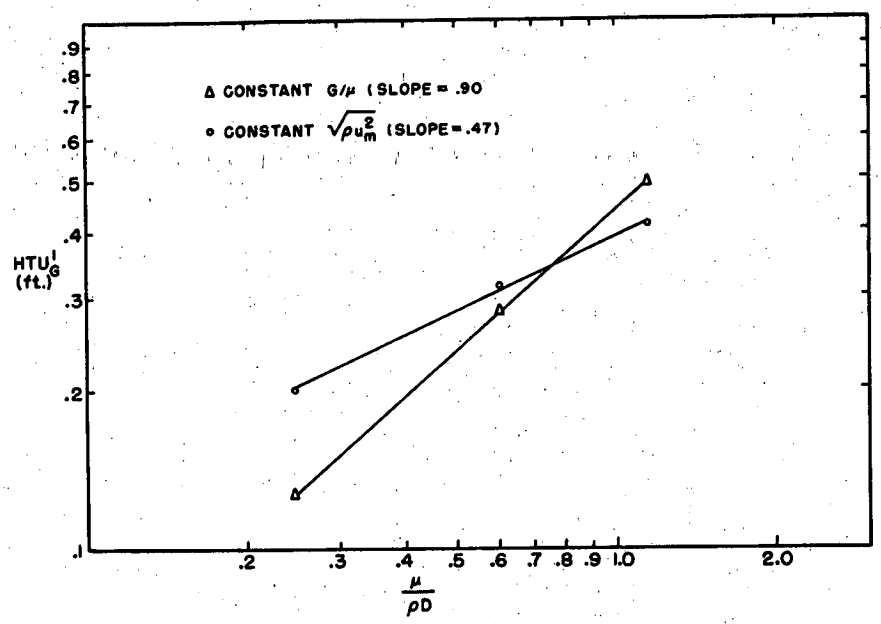

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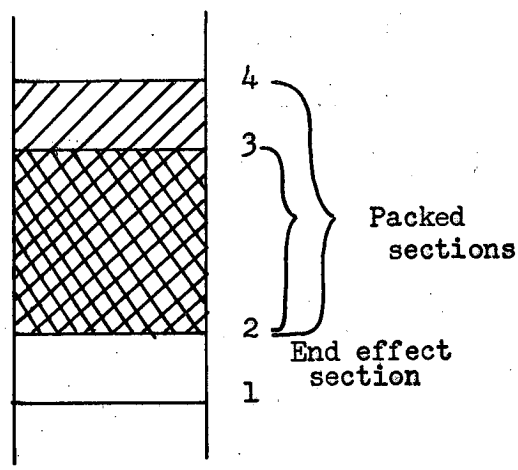
Fig. 14

HTU'_G as a function of Schmidt number at constant values of G/μ and inertia group. L = 1575.

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 (21), (57). 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 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}.



Then it is known that:

a) $NTU_{14} = NTU_{12} + NTU_{24}$

b) $NTU_{13} = NTU_{12} + NTU_{23}$

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

substituting c) in b),

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

subtracting d) from a) and rearranging,

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

NTU_{24} is not a value that can be directly measured, but the corresponding value HTU_{24} is a true HTU since the end effect section is not included in the length Z_{24}. Substituting the values of the HTU for the NTU in equation e),

$$NTU_{14} = Z_{24}/HTU'_{24};\ NTU_{13} = Z_{23}/HTU'_{23};\ NTU_{24} = Z_{24}/HTU_{24}$$

f) $$\frac{Z_{24}}{HTU_{24}} = \frac{\frac{Z_{24}}{HTU'_{24}} - \frac{Z_{23}}{HTU'_{23}}}{1 - Z_{23}/Z_{24}}$$

rearranging,

$$HTU_{24} = \frac{HTU'_{24}HTU'_{23}(Z_{24}-Z_{23})}{Z_{24}\ HTU'_{23} - Z_{23}HTU'_{24}} \qquad (27)$$

To evaluate the end effect in this work, bed lengths of 13 $^1/_2$, 7 $^1/_2$, and 2 inches were used. The 13 $^1/_2$ inch 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

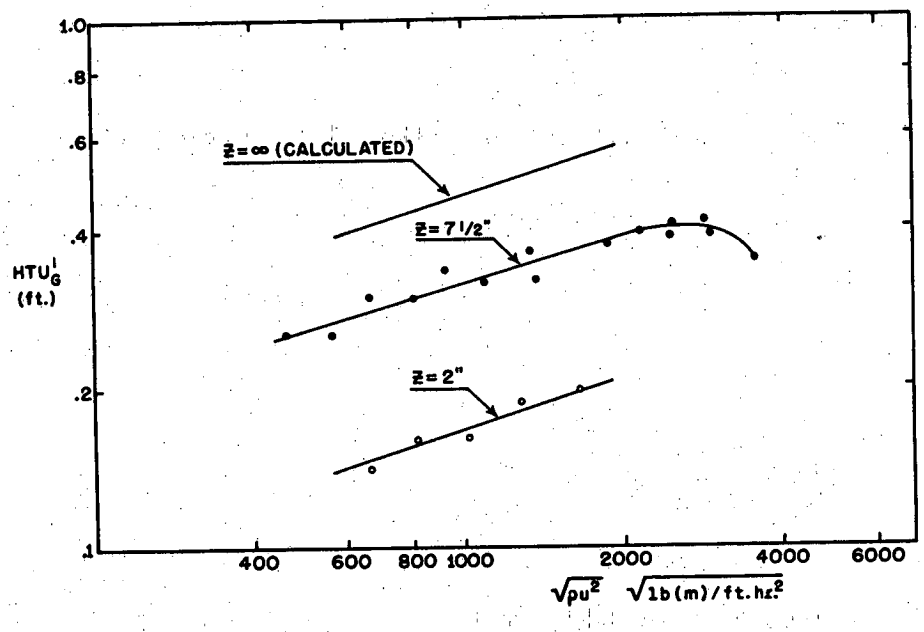


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Fig. 15

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

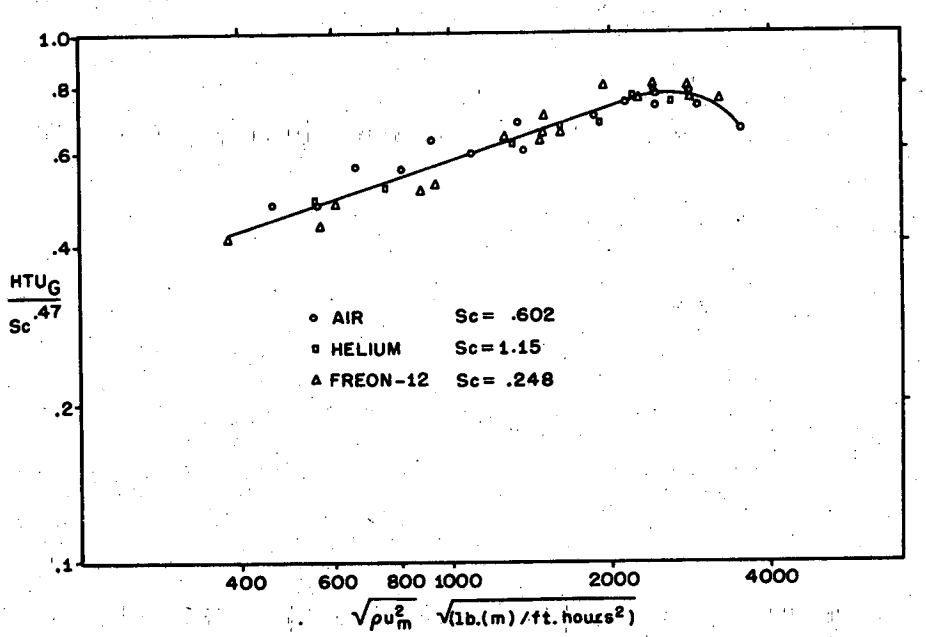

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Fig. 16

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

(see Figure 15) with an average value of 3.51 inches. 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 1575 lbs.(m)/hr. ft.^{2} may be correlated by the single curve of Figure 16 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 percent with a maximum deviation of 12.5 percent.

The curves of Figure 17 show the effect of the liquid rate on the HTU at several gas 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 and above 1500 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 of Figure 17 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 $(\sqrt{\rho u_m^2})_G/(\sqrt{\rho u_m^2})_L$ might serve as a single parameter for the general

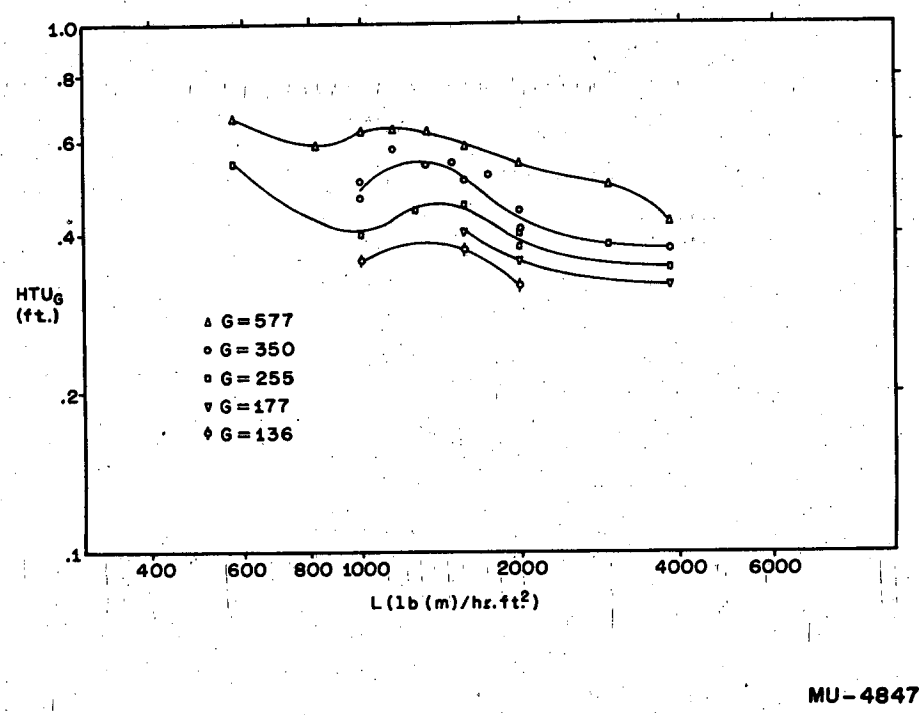

Fig. 17

HTU_G as a function of the liquid rate.

correlation as illustrated in Figure 18. 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 percent and the maximum deviation is 24 percent. In view of the limited scope of the liquid variables covered in this work, it is not possible to attach a final significance to this correlation. In fact, a qualitative consideration of the problem would lead to the conclusion that the forces of gravity and viscosity are more important in the case of the liquid than the inertia force. If this is the case, then the correlating modulus would involve the ratio of the gas inertia force to the liquid gravity and viscosity forces and would give

$$\frac{(\rho u^2)_G}{(\rho_L - \rho_G) g_c} \times \text{length}$$

and

$$\frac{(u\mu)_L}{(\rho_L - \rho_G) g_c} \times \text{length}^2$$

as the correlating groups. A combination of these groups gives

$$\frac{u_G}{u_L} \frac{\sqrt{\rho_G(\rho_L - \rho_G) g_c}}{\mu} \times \text{length}^{3/2} .$$

This shows the same dependence on liquid and gas velocities as the flooding modulus, but the density of the liquid appears in the numerator instead of the denominator. On the other hand, the work of Dukler and Bergelin (17) indicates that wave formation in falling liquid films is directly dependent on the liquid density. Since the HTU would decrease with increasing wave motion, this would suggest that the HTU would be inversely proportional to the liquid density.

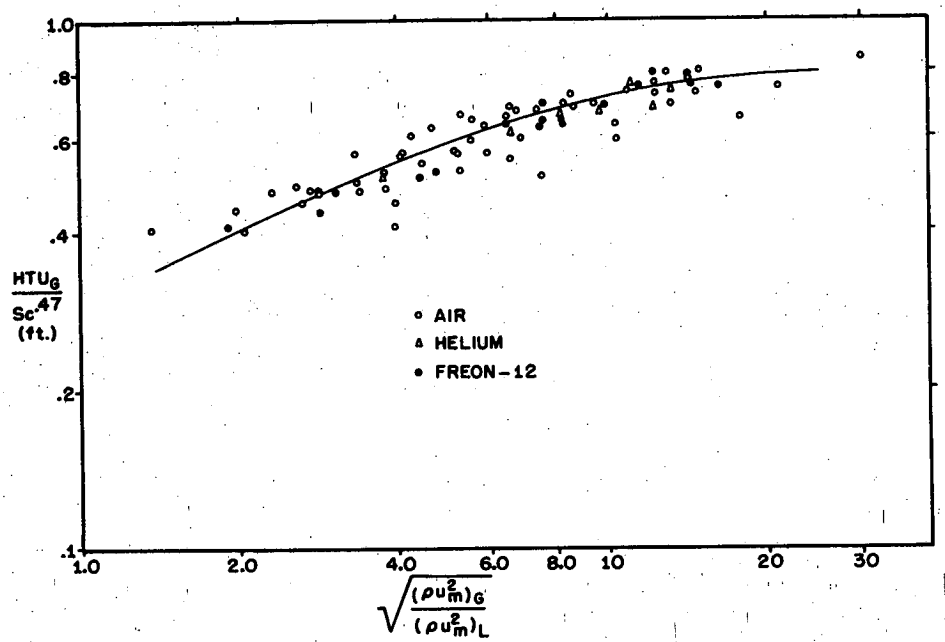

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Fig. 18

HTU_G as a function of the flooding modulus.

Since neither the density nor the viscosity of the liquid were varied in this investigation, it is not possible to test either of these theories.

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 all used the same packing (one inch 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, Figure 19, for gas film HTU have been corrected for Schmidt number 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. A more complete description will be found in the appendix.

Dwyer and Dodge (18) - absorption of NH_3 in water.

Fellinger (21) - absorption of NH_3 in water and H_2SO_4.

Houston and Walker (25) - absorption of NH_3, acetone, methanol and ethanol in water.

Johnstone and Singh (27) - absorption of SO_2 in NaOH.

McAdams, Pohlenz, and St. John (38) - vaporization of water.

Othmer and Scheibel (42) - absorption of acetone in water.

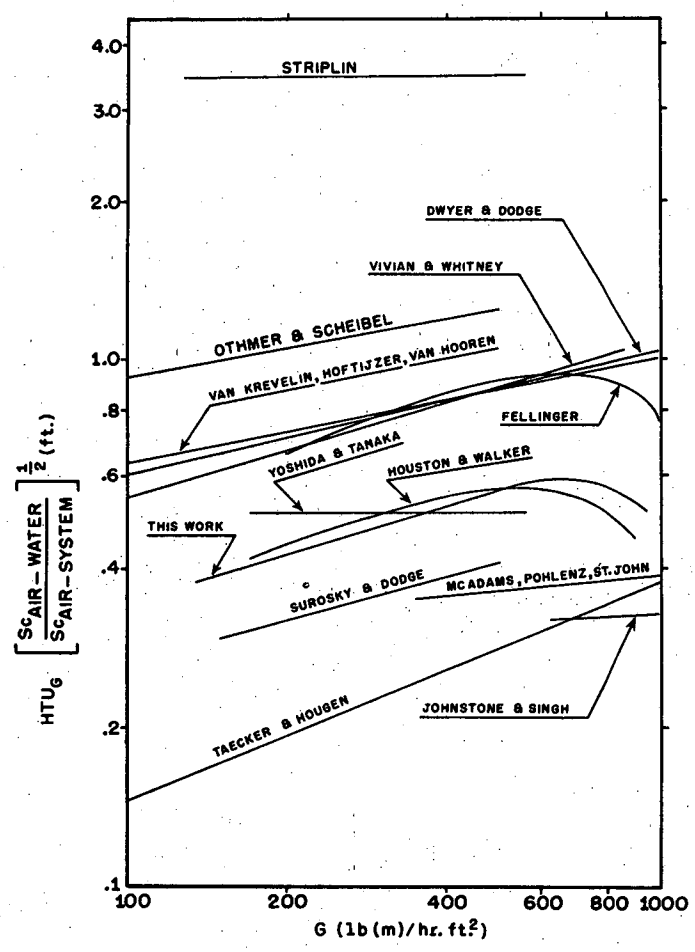

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Fig. 19

HTU_G as a function of gas rate for various investigations.

Striplin (56) - absorption of water in phosphoric acid.

Surosky and Dodge (57) - vaporization of water, methanol, benzene, and n-ethyl butyrate.

Taecker and Hougen (58) - vaporization of water from porous solids.

Vivian and Whitney (62) - absorption of SO_2 in water

Yoshida and Tanaka (65) - vaporization of water.

Van Krevelin, Hoftijzer, and Van Hooren (60) reviewed a number of gas film mass transfer studies to obtain 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.

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 HTUs 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 this 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 (14), (15). Mass transfer resistance 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 (63). 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 with the consequent lowering of the transfer rate from that which would be expected. This would give values of the HTU which were too high. Only fundamental experiments on the rate on 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 (52) show that the liquid film HTU is independent of the gas rate until the loading point is reached, after which the rate of transfer rapidly improves. This would indicate that at the loading 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. Figure 19 gives some indication of how important the area effect can be. 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 Figure 20. All of the curves show a decrease in the HTU with increase in liquid rate but with widely varying dependence. Dwyer and Dodge, Yoshida and Tanaka, and Vivian and Whitney indicate that the HTU varies as approximately $L^{-.2}$ over the entire range of liquid rates measured. The curve of McAdams, Pohlenz, and St. John is taken

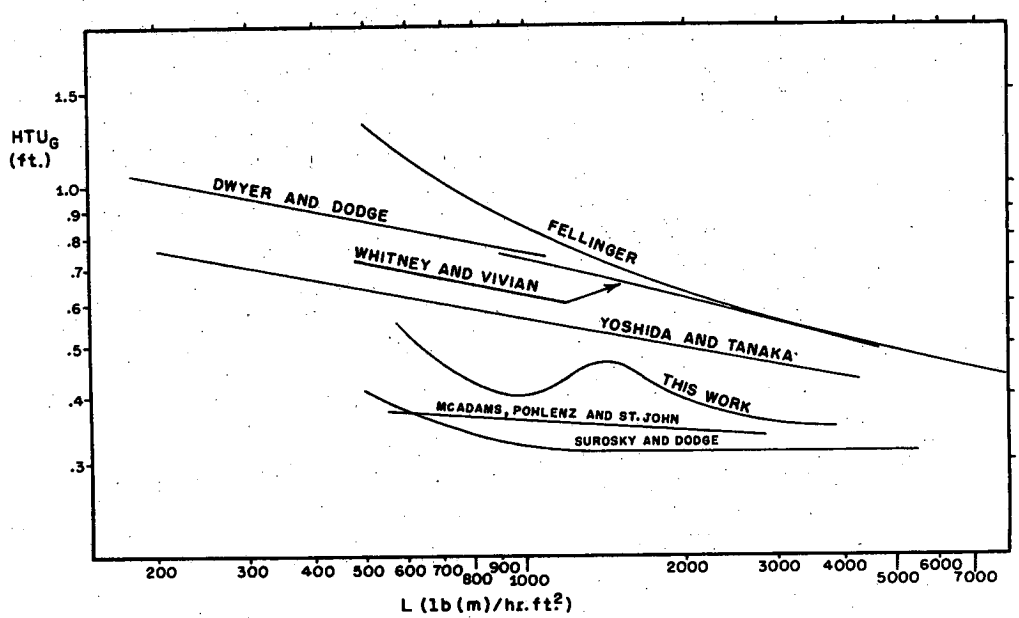

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Fig. 20

HTU$_G$ as a function of liquid rate for various investigations. G = 200.

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 1200. 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, 1500, and 4500, shows the greatest dependence on liquid rate, the HTU varying with $L^{-.7}$ at low flow rates and changing gradually to $L^{-.3}$ 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. Nor is it possible to say whether they offer either supporting or contradictory evidence for the unique liquid rate dependence found by the author.

The number of works devoted to the determination of the Schmidt number effect is considerably more limited. Houston and Walker (25) 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 (49) 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 (18) 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

of uncertainty in that it is necessary to correct for the liquid film resistance which is present, and 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 (39) vaporized water, methanol, benzene, and toluene, and found that the HTU was proportional to $D^{-.17}$. Surosky and Dodge (57) used water, methanol, benzene, and ethyl butyrate as the liquids and found that their data could be correlated using HTU proportional to $D^{-.15}$. The data of Simkin (54) and Chrisney (11) on the vaporization of solids and liquids from packed beds in the absence of liquid flow indicate that the HTU is proportional to $D^{-.36}$. 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 the packing. The possible error from this source has been discussed previously.

It is possible to avoid the above difficulties by 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 different gases do not influence the liquid flow pattern in different ways. If the gas does cause a change in the liquid flow, it should do so primarily by the pressure it exerts. Elgin and Weiss (20) and Jesser and Elgin (26) have found that the liquid holdup 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 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 wetting.

It is concluded that the inertia group is the correct correlating modulus for mass transfer in packed towers with countercurrent flow of gas and liquid, and that the Schmidt number dependence found in this work, i.e., HTU proportional to $Sc^{.47}$, is valid within the limits of experimental error.

NOMENCLATURE

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

b - distance from the center of the pipe to the Murphree boundary layer - ft.

B_g - hypothetical film thickness - ft.

C, C_o - coefficient of discharge

C_D - coefficient of drag defined by equation (9)

C_m - molal heat capacity at constant pressure - Btu./lb.mole °F.

C_p - heat capacity at constant pressure - Btu./lb.(m) °F.

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

d - diameter - ft.

D_m - molal diffusivity - lb.moles/hr. ft.

D_2 - nozzle throat diameter - inches

E - eddy viscosity at the centerline of a pipe - lb.(m)/ft.hr.

E_m - molal eddy diffusivity - lb. moles/hr. ft.

F - frictional force per unit area - $lb.(f)/ft.^2$

$f/2$ - coefficient of friction (friction factor)

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

G_m - molal mass velocity of the gas stream based on the superficial area - $lb.moles/hr.\ ft.^2$

g_c - universal gravity constant - $lb.(m)ft./lb.(f)\ sec.^2$

H - height - ft.

h - heat transfer coefficient - $Btu./hr.ft.^2$ °F.

h' - pseudo-heat transfer coefficient defined by equation (19)

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.

ΔH - pressure drop across the flow measuring orifice - inches of water

j_d - mass transfer factor defined by equation (13)

j_h - heat transfer factor defined by equation (12)

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

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

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

P - total pressure - atm.

p - partial pressure - atm.

p_f - film pressure factor - atm.

ΔP - pressure drop - lb.(f)/ft.^{2}

Δp_m - ratio of the maximum partial pressure difference to the mean partial pressure difference

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

R - universal gas constant - atm. ft.^{3}/lb. mole °F.

r - pipe radius - ft.

T - absolute temperature - °R.

t - temperature - °F., °C.

u - velocity - ft./sec.

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

u_{max} - velocity at the centerline of a pipe - ft./sec.

w - weight rate of flow - lb.(m)/hr.

x - length of diffusion path - ft.

y - mole fraction

y_f - p_f/P

Y - compressibility factor

Z - packing length - ft.

Greek letters

α - the number of moles of gas A diffusing divided by the total number diffusing

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

γ - a heat transfer factor defined by equation (18a)

ϵ - surface roughness - ft.

μ - viscosity - lb.(m)/ft. hr.

ρ - density - lb.(m)/ft.^{3}

ϕ - $1 + (Sc - 1)u_B/u_m$

ψ - a function to be determined experimentally

Dimensionless groups

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

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

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

Subscripts

A - refers to gas A

a - main stream

av - average

B - refers to gas B

B	-	refers to boundary condition
c	-	convection
D	-	dry bulb
d	-	refers to mass transfer
G	-	refers to gas film
g	-	gas
h	-	refers to heat transfer
i	-	interface
m	-	mean
m	-	molal
r	-	radiation
r	-	room
s	-	surface
T, t	-	total
w	-	wet bulb

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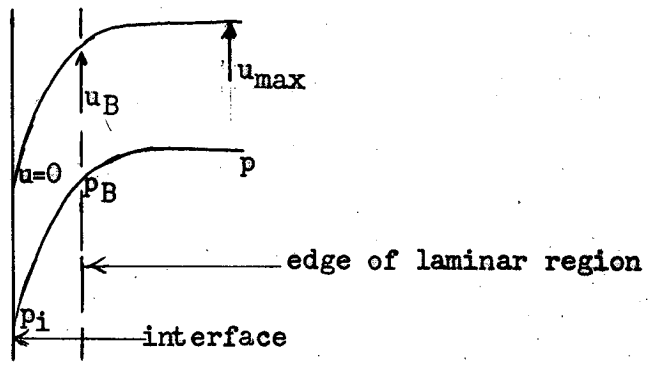
APPENDIX

Mass Transfer Theories for Forced Convection in Pipes

A direct theoretical treatment of the problem of mass transfer in forced convection is possible if the velocity distribution is known for the boundary layer. However, in order to integrate the equations it is necessary to make certain simplifications in setting up the model that represents the flow conditions in the boundary layer. Since the smooth straight pipe represents the only case where extensive experimental work has been done, and since it is a fairly simple case, i.e., essentially a flat plate, all of the theoretical developments have taken the smooth pipe as a basis.

The simplest development was that of Osbourne Reynolds (47) who made the assumption that there was no boundary layer and that the turbulent region extended to the pipe wall. Reynolds' interest was in heat transfer, and for this he postulated that heat transfer and momentum transfer were analogous processes and that the ratio of the momentum transferred to the wall divided by the total momentum of the stream was equal to the heat transferred to the wall divided by the total heat of the stream above the wall temperature. A similar statement may be made regarding the analogy between momentum transfer and mass transfer. While this analogy completely ignores the role of the boundary layer, it will give the same result as the more elaborate analogies for the singular case of the Schmidt number equal to one. It has also served succeeding analysts in setting up equations to represent the turbulent core.

Colburn (12) has approached the problem of mass transfer by means of an analogy to momentum transfer similar to that used by Prandtl (46) and Taylor (59) for comparing heat transfer with momentum transfer. In this derivation it was assumed that the bulk of the fluid is in turbulent flow while a thin laminar flow region exists adjacent to the phase boundary.



The transport of mass through the laminar region is considered to occur by pure molecular diffusion. In the turbulent core, Reynolds' analogy is assumed to be valid. By equating the two transport rates at the boundary layer, the following general equation was obtained:

$$k_G = \frac{F}{u_m \, p_f \, M_m \, \phi} \tag{21}$$

and for a smooth pipe,

$$k_G = \frac{(f/2)\, G}{p_f \, M_m \, \phi} \quad . \tag{21a}$$

$f/2$= coefficient of friction

where $\phi = 1 + (u_B/u_m)(Sc - 1)$

F = frictional force per unit area at the wall - lb.(f)/ft.^{2}

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

u_m = mean gas velocity - ft./hr.

u_B = gas velocity at the boundary between the turbulent and laminar regions - ft./hr.

Sc = Schmidt number = $\left(\frac{\mu}{\rho D}\right)$.

Because of the oversimplification of the boundary layer model, equation (21) cannot be expected to give a correct Schmidt number dependence for values of Sc very far from unity.

To avoid the limitation on the validity of the Prandtl number function imposed by assuming a completely laminar boundary layer, von Karman (29) has divided the boundary layer into a laminar sub-layer and a buffer layer in which both viscous and turbulent forces are important. Reynolds' analogy was assumed to hold as before in the turbulent region. The data of Nikuradse (41) on the velocity gradient in flow through a pipe served as a basis for determining the variation of the eddy conductivity in the film. While von Karman's derivation was for the heat transfer case, it can be readily extended to mass transfer (51) to give:

$$k_G = \frac{(f/2)\, u_m}{RT\left[1 + 5\sqrt{f/2}\left\{Sc - 1 + \ln\frac{1 + 5Sc}{6}\right\}\right]} \tag{22}$$

where

R = gas constant

T = absolute temperature.

Boelter, Martinelli, and Jonassen[6] extended von Karman's derivation to account for the effects of the variation of fluid properties due to heating and cooling of the fluid. In doing so, they also rederived the isothermal case, using the data of Nikuradse for the turbulent core. Because their equation for the

heat transfer coefficient was based on the centerline temperature whereas the mean temperature of the fluid is usually used in heat transfer work, they multiplied their integrated expression by Δt_m, which is the ratio of the mean temperature difference to the maximum temperature difference. Their isothermal case in the equivalent mass transfer units is:

$$k_G = \frac{\sqrt{f/2} \quad (\Delta p_m)\ u_m}{5\left[Sc + \ln(1 + 5Sc) + \ln \frac{Re}{60}\sqrt{f/2}\right] RT} \tag{23}$$

where

Δp_m = maximum concentration difference divided by the mean concentration difference.

Re = Reynolds number.

Murphree (40) also attempted to avoid the oversimplification of a laminar boundary layer by assuming that the eddy conductivity entered the boundary layer to a degree proportional to the cube of the distance from the wall while remaining constant over the main turbulent core. By using Lees' (33) equation for f in round tubes and Stanton's (55) data for u_{max}/u_m, a general relationship is obtained between the eddy viscosity, the film thickness, and the Reynolds number. The resulting equation for heat transfer can again be extended to mass transfer to give:

$$k_G = \frac{(1 + b/r)\ g(\Theta)\ D}{(1 - b/r)\ 3\ \Theta^{1/3}\ RTd} \tag{24}$$

where $$g(\Theta) = {}^1/_2 \ln \frac{(1 + \Theta^{1/3})^2}{1 - \Theta^{1/3} + \Theta^{2/3}} + \sqrt{3}\tan^{-1}\frac{2\,\Theta^{1/3} - 1}{\sqrt{3}} + \frac{\pi\sqrt{3}}{6} \tag{24a}$$

$\Theta = (E/\mu)Sc$

E = eddy viscosity at the pipe center

r = pipe radius

b = distance from the pipe center to the edge of the turbulent region

d = pipe diameter.

Because of the complexity of the mathematical expression, Murphree considered only the resistance of the film in deriving the equation for heat transfer. It was also found that this gave better agreement with the literature data.

Since the factor f/2 is known to be a function of the Reynolds number, it may be said that from the preceding analogies the mass transfer coefficient is a function of the Reynolds number and the Schmidt number. From equation (7) the HTU can then be expressed in a general form to satisfy all of the analogies as:

$$aHTU_G = \text{function (Re)} \times \text{function (Sc)}.$$

In its simplest form, i.e., Chilton and Colburn's (41) empirical correction of the Prandtl-Taylor analogy, this equation becomes:

$$aHTU_G = \text{function (Re)} \times (Sc)^{2/3}.$$

A similar result can be obtained from dimensional analysis based on our partial knowledge of the system. From measurements made on turbulent flow in pipes, it is known that the factors influencing the flow pattern are the density ρ, viscosity μ, velocity, diameter, and relative roughness ϵ/d. In addition to the terms which influence the flow pattern, it is necessary to include the molecular diffusivity which influences mass transfer. If it is assumed that the forced convection completely dominates free convection, and there is no complication introduced by the presence of a flowing liquid, the variables listed should describe the system, and dimensional analysis can be applied to

give the equation:

$$aHTU_G = \text{function } \left(\frac{dG}{\mu}\right), \left(\frac{\mu}{\rho D}\right), \left(\frac{\epsilon}{d}\right)$$

or in the form of equation (12) for a smooth pipe,

$$aHTU_G = \text{function } \left(\frac{dG}{\mu}\right) \times \text{function } \left(\frac{\mu}{\rho D}\right)$$

where

G = gas mass velocity - lb.(m)/hr. ft.^{2}

D = molecular diffusivity - ft./hr.

A Review of Mass Transfer Studies of Interest to the Present Research

A) Vaporization of Pure Liquids

Chrisney, J. B. (11)

This is an extension of the work of Simkin using water, naphthalene, aniline, and nitrobenzene in the same apparatus. Gas rates were varied from 400 to 2000 lbs./hr.-ft.^{2}. Combined with the work of Simkin, a Schmidt number dependence of $Sc^{.36}$ was obtained.

McAdams, W. H., Pohlenz, J. B., and St. John, R. C. (38)

Gas film coefficients for one inch carbon raschig rings were determined in adiabatic isothermal water vaporization runs employing a four inch diameter tower with packed heights of six, nine, and twelve inches. The end effect was found to be equivalent to 7.2 inches from heat transfer data. G was varied from 350 to 1000 and L from 500 to 2600. The mass transfer data scattered badly, but a line drawn through the mean of their data can be assigned the equation:

$$k_G a = 0.083\ G^{.9}\ L^{.07}$$

The author suggests an exponent of 0.7 on G based on the heat transfer results which were far superior.

Simkin, D. J. (54)

Pellets were used to evaporate water, naphthalene, and p-dichlorobenzene into a flowing air stream in order to study the effect of the diffusivity on the mass transfer coefficient. The apparatus used was a packed bed having the dimensions 1.97 x 2.26 inches by 1 inch deep. A Schmidt number dependence of $Sc^{.38}$ was established.

Surosky, A. E. and Dodge, B. F. (57)

This work was intended primarily to determine the effect of Schmidt number by the vaporization of three organic liquids and water in an eight inch tower packed with one inch carbon rings. Packed heights of 4, 6, and 12 inches were used. End effects were found equivalent to 2.2 inches. $k_G a$ was found to vary as $G^{.72}$ and $D^{.15}$. G was varied from 140 to 500 at a value of L of 1600. The diffusivity was varied 3.7 fold. $k_G a$ was found to increase with liquid rate until a value of L = 1200 was reached where it became independent of liquid rate.

Taecker, R. G. and Hougen, O. A. (58)

The authors studied the drying rate of water from a number of porous solids of common shapes in a packed tower. In this manner they attempted to present a completely wetted surface and remove the effect of liquid rate on the gas film coefficient. Actual measurements were made on the heat transfer coefficient, the mass transfer rate being determined by the j number analogy, using the empirical relationship of Gamson, Thodos, and Hougen (22) that

$$j_h = 1.076\ j_d.$$

The resulting mass transfer equation presented for one inch rings is

$$j_d = 1.070 \left(\frac{G\sqrt{A_p}}{\mu} \right)^{-.41}$$

where A_p = area of a particle - $ft.^2$

G was varied from 18 to 1520 lbs/hr. $ft.^2$

Yoshida, F. and Tanaka, T. (65)

The authors ran air-water contact operations of three kinds, constant water temperature humidification, water cooling, and dehumidification in a tower 10 inches in diameter packed to a depth of 12.5 inches with 15, 25, and 35 mm. ceramic rings. They found the gas film coefficient proportional to the gas rate and the 0.2 power of the liquid rate. No correction was made for end effect. In the constant temperature runs they found

$$k_G a = 0.0155\ GL^{.2}$$

with no effect of packing size. G was varied from 137 to 586 and L from 201 to 4160 lbs/hr. $ft.^2$

B) Absorption of Soluble Vapors from an Inert Gas Stream

Dwyer, O. E. and Dodge, B. F. (18)

The rate of absorption of ammonia from air by water was studied in a one foot diameter tower packed with carbon raschig rings. The effect of humidity and redumping of the rings was found to be minor. Temperature increase decreased both the overall and the gas film coefficients. The effect of flow rate for one inch rings is correlated by the equation:

$$\frac{1}{K_G a} = \frac{1}{\gamma G^n L^m} + \frac{1}{H\beta L^r}$$

where

$$\gamma = 0.036,\ n = 0.77,\ r = 0.78,\ m = 0.20,\ \beta = 0.103$$

at 85° F.

The equation indicates that both gas and liquid rates affect the gas film coefficient, but only the liquid rate is important in the case of the liquid film. The equation for the gas film coefficient is given as:

$$k_G a = 0.127\ G^{.77} \quad (L = 500)$$

$$= 4.12\ L^{.20} \quad (G = 500)$$

combined, $$k_G a = 0.036\ G^{.77}\ L^{.20}\ .$$

Variation in G was from 100 to 1000, and L from 160 to 1050. The packed height used was four feet. No corrections were made for end effects.

Fellinger, L. (21)

This work on the absorption of ammonia in water is reported in Perry's Handbook. Data were taken on various packings in a tower one-and-one-half feet in diameter at gas rates up to 1000 lbs/hr. ft.^{2} and at liquid rates of 500, 1500, and 4500. All data were corrected for end effects. Values of $k_G a$ were obtained by subtracting Sherwood and Holloway's (11) liquid film resistance from the overall value. An attempt to obtain gas film data by absorption in H_2SO_4 was not entirely successful, the mass transfer coefficient being 1.5 to 2.5 times as high as the water absorption data. The water absorption data probably represent the best and most complete work available at present.

Houston, R. W. and Walker, C. A. (25)

Overall coefficients for the absorption of NH_3, acetone, ethanol, and methanol in water were measured, and the Sherwood and Holloway (11) liquid film coefficients were subtracted out. Measurements were made in a twelve inch column packed to a height of two feet, with one inch raschig rings. At liquor rates of 500, 1000, and 2000 lbs./hr. ft.^{2} and gas rates up to 600 lbs./hr. ft.^{2}, the various systems were correlated by the 2/3 power of the diffusivity. At the higher liquor rate of 3000, there was much less dependence on the gas diffusivity. No correction was made for end effects.

Johnstone, H. F. and Singh, A. D. (27)

The authors absorbed SO_2 in dilute NaOH in a rectangular tower of 0.555 ft.^{2} cross-section packed to a depth of 0.5 feet with one inch rings and several other packings (flat and corrugated plates, grids). No correction was made for end effect. At a liquid rate of 1080 lbs./hr.ft.^{2}, they suggest the correlation

$$k_G a = 0.107\ G^{.95}\ .$$

G was varied from 625 to 2410. The data do not indicate any variation from a straight line, although the maximum gas rate reported would appear to be beyond the flooding velocity for this system and packing.

Othmer, D. F. and Scheibel, E. G. (42)

Acetone recovery from air by scrubbing with water in a ten inch diameter column nineteen feet high gives the following equation:

$$k_G a = \frac{1}{\frac{23}{L^{.95}} + \frac{30}{G^{.8}}}$$

The packing used was one inch raschig rings. The liquid rate varied up to 700 and the gas rate varied up to 500 lbs./hr. ft.^{2} The authors' method of interpreting the data is questionable.

Scheibel, E. G. and Othmer, D. F. (49)

A four inch i.d. column packed to a height of 37 inches with 10 mm. glass raschig rings was used to absorb acetone, methyl ethyl ketone, methyl isobutyl ketone, and methyl n-amyl ketone in water. The data on the systems acetone and methyl ethyl ketone were used to give the correlation for the gas film of:

$$k_G = 0.188\ D^{.5}\ G^{.8}\ ;$$

$$L = 780 \text{ and } 445$$

where the gas film resistance was determined by plotting $1/k_G$ vs. $H/G^{.8}$. In trying to fit data of other investigators (18) for the absorption of CO_2 and NH_3, the authors were led to use a value of 1 as the exponent on the diffusivities. The authors' methods of interpreting their data are open to question.

Striplin, M. M. (56)

The author dried air with 95 to 99 percent H_3PO_4 in a ten inch diameter tower packed to a height of seven feet with one inch raschig rings. A plot of $1/k_Ga$ vs. $1/G$ leads him to the conclusion that the liquid film resistance is negligible. The equation presented as best representing the data is:

$$k_Ga = 0.01\ G$$

The liquid rate was 460 and the gas rate was varied from 128 to 556 lbs./hr. ft.^{2}

Van Krevelin, D. W., Hoftijzer, P. J., and Van Hooren, C. J. (60)

Analysis of various literature data was made by the authors from overall gas film coefficients. They assumed that the equilibrium curve was straight and that the gas film coefficient was a function of the gas velocity to the 0.8 power. This permits plotting $1/k_G a$ vs. $(1/G)^{.8}$ to obtain $k_G a$. The authors assumed $k_G a$ independent of the liquid rate but dependent on the wetted area which was a function of liquid rate as

$$a/a_o = 1 - e^{-\delta L} \quad .$$

Empirical examination of the data available then gave

$$\frac{k_G RTd}{D} = 2\left(\frac{G}{a_o \mu}\right)^{.8}\left(\frac{\mu}{\rho D}\right)^{1/3} \quad .$$

It is interesting to note that the expression for a/a_o gives a liquid rate dependence almost identical with that reported by Fellinger for one inch rings.

Vivian, J. E. and Whitney, R. P. (62)

Overall coefficients were measured for the absorption of SO_2 in water in an eight inch diameter tower packed to a height of two feet with one inch raschig rings. The overall coefficients were broken down by assumptions based on a previous work and by plotting $1/k_L a$ vs. $1/G$. The equation representing the gas film is:

$$k_G a = 0.028\, G^{.7}\, L^{.25} \quad .$$

A unique feature of the work was the introduction of the gas through five upcomers directly into the packing. L was varied from 900 to 12000 and G from 65 to 850 lbs./hr. ft.^{2}

C) Psychrometric Studies

Arnold, J. H. (3)

The author's work is reported in two parts. The first is a study of wet bulb depressions of water, methanol, propanol, and toluene using a glass side-arm U-tube with a thermometer inserted in each branch of the U. Dried air from a laboratory supply was introduced and left through the side-arm. The wick was supplied with liquid by dipping it in an external supply which was held near the wet bulb temperature. The results were corrected for radiation. In this paper the author also develops a theory for the wet bulb depression which is similar to Prandtl's and shows a dependence on the velocity. The second study of the depressions of toluene, chlorobenzene, and m-xylene was designed to show the velocity dependence. For this the thermometers were inserted in a four inch square duct perpendicular to the flow direction. A variation with the 0.07 power of the velocity was shown, but because of the use of a very short wick, it could well be the result of conduction losses by the thermometer. These data were also corrected for radiation losses. In no case were the results corrected for either the film pressure factor, p_f, or the heat transfer factor, γ.

Bedingfield, C. H. and Drew, T. B. (5)

Psychrometric measurements were made by using solid naphthalene, Camphor, p-dichlorobenzene, and p-dibromobenzene. The solids were cast as a cylinder around a thermocouple. The cylinder was mounted in a 1 x 1 foot square duct as were the two

shielded thermocouples used for measuring the air temperatures. The wall temperature was also measured, presumably for a radiation correction. The results were also corrected for the film pressure factor, but not for the heat transfer factor. By assuming that the curve used to fit the data had to go through a semi-theoretical point of Sc = 0.70, $k_G M_m p_f/h$ = 4.16, the authors were misled into using an exponent of -.56 on the Schmidt number.

Dropkin, D. (16)

In this very carefully performed work, the author was primarily interested in measuring the difference between the adiabatic saturation temperature and the wet bulb temperature. Consequently, the humidities of the air stream passing the thermometers were not measured. But from the reported values of the saturation temperature it is possible to calculate the variation of the psychrometric ratio from the specific heat of the gas. The thermometers were covered with wicks 8 inches long to make conduction errors negligible. The results show no variation with either temperature or Reynolds number.

Mark, J. G. (36)

Psychrometric measurements were made employing water, benzene, chlorobenzene, toluene, carbon tetrachloride, ethylene tetrachloride, and ethyl acetate with air as the inert gas. The results are quite consistent, but were not corrected for film pressure factor, heat transfer factor, or radiation. Mark was aware of the radiation correction, but because of the physical set-up, he was unable to do more than estimate that it amounted to 1 to 4 percent.

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 (31) 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 Buddenburg and Wilke (8) 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 (24).

Thermal conductivity

The thermal conductivity of air, water vapor, and helium were taken from the review article by Keyes (31) quoted above. The thermal conductivity of Freon-12 is that reported by Sherratt and Griffiths (50) over the range from 33.3 to 216.2° C.

Heat capacity, C_p

The heat capacity of air and water vapor were taken from the book by Keenan (30) on the thermodynamic properties of air. The heat capacity of Freon-12 was calculated from the work of Justi and Langer (28) who reported the heat capacity at constant volume. The heat capacity of helium is reported in Perry's Handbook (44) as constant for all temperatures.

Diffusivity

The diffusivities used were those measured by C. Y. Lee (32). The measured values were corrected to the desired temperature by the equation of Hirschfelder, Bird, and Spotz (24).

Gas specifications

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

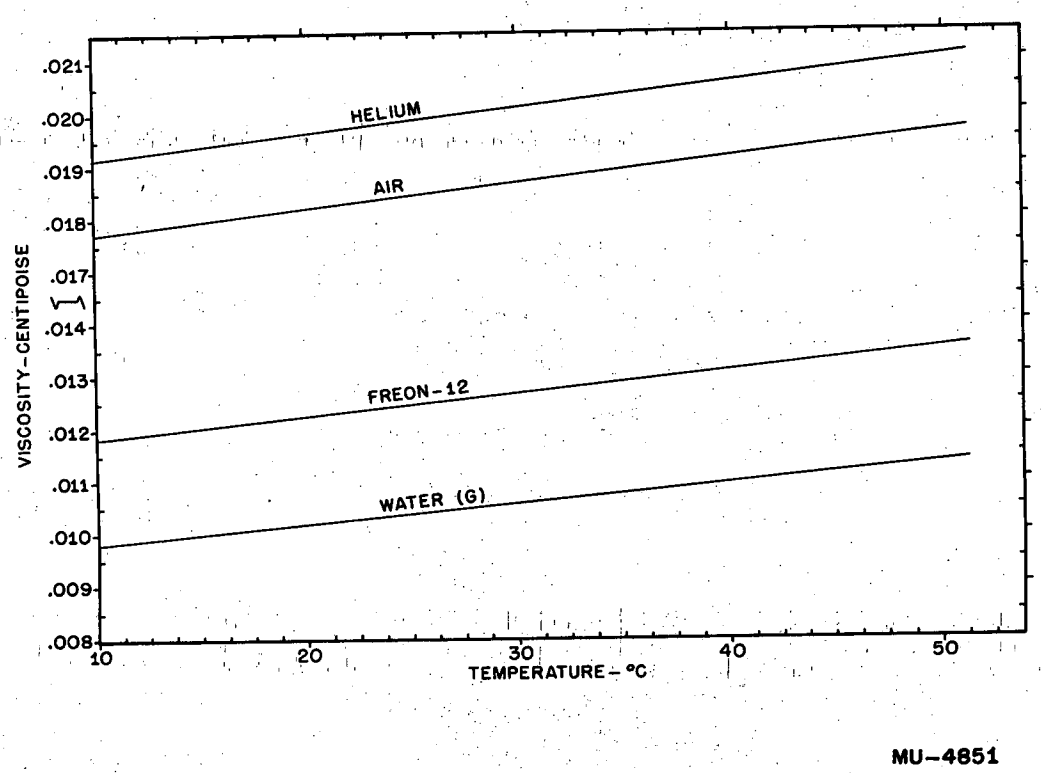
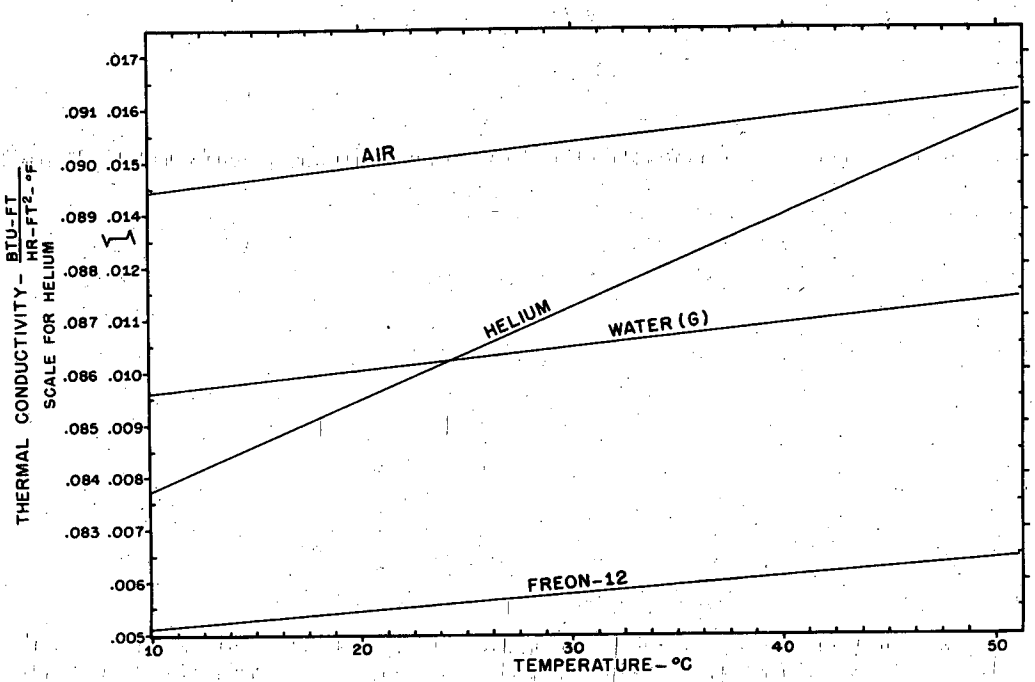


Fig. 21

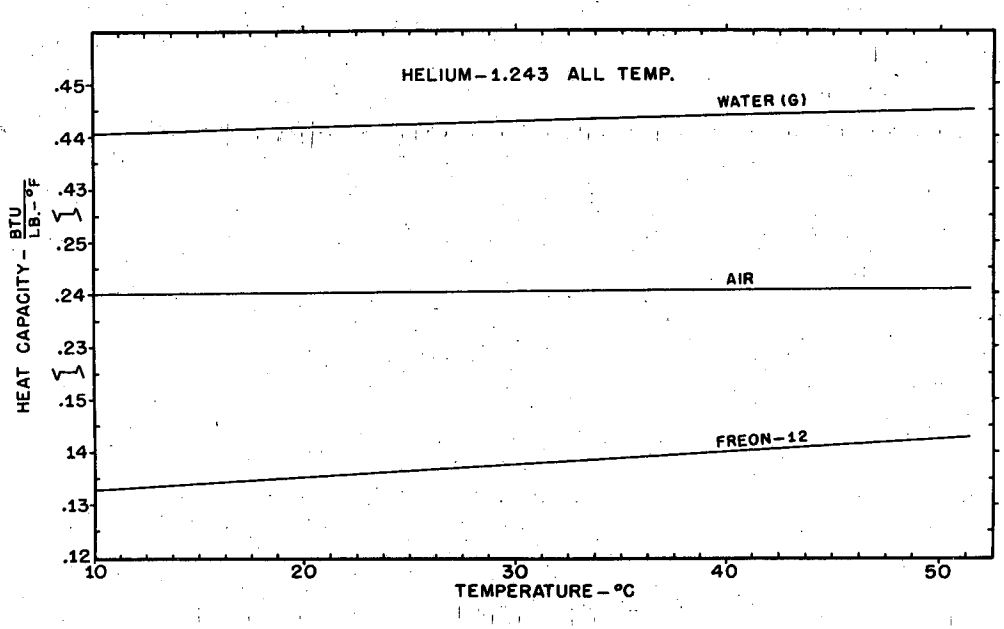
Viscosity of pure gases.



MU-4852

Fig. 22

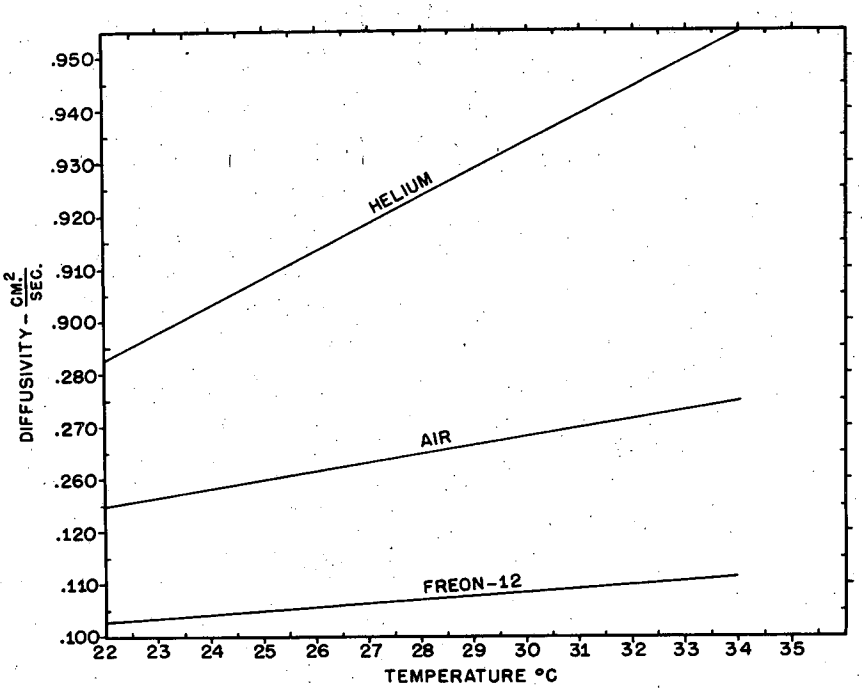
Thermal conductivity of pure gases.



MU-4853

Fig. 23

Heat capacity of pure gases
at constant pressure.



MU-4854

Fig. 24

Diffusivity of water vapor in pure gases.

Calibration Procedures and Results

A) Nozzle at Low Flow Rates

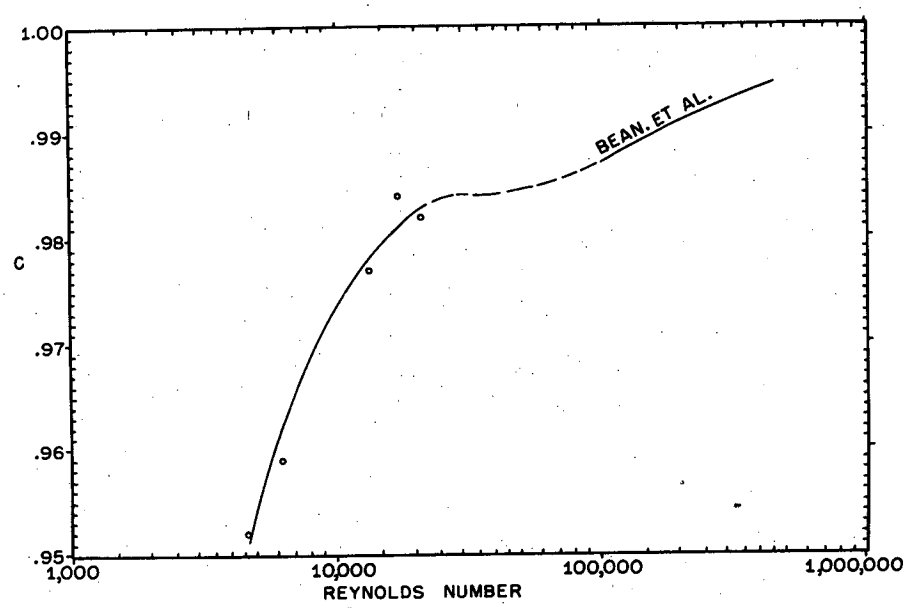
The Reynolds numbers for the lowest flow rates with helium were below those for which the nozzle had been calibrated by Bean, Buckingham, and Murphy (4). To get the discharge coefficient at these rates, pitot traverses were made with helium in the system. The impact tube used had a smoothly tapered tip with an opening of 1/16 inch diameter. The pressure tap was located opposite the end of the nozzle in the wall of the flow measuring chamber. Differential heads were read on a draft gage and could be estimated to 0.001 inch of water (see Figure 25).

B) Nozzle Orifice

For low flow rates with air and Freon-12, it was necessary to use an orifice plate with a one inch diameter opening to get a sufficiently high draft gage reading. This was calibrated by injecting CO_2 at a known rate and then analyzing the gas stream by means of an Orsat apparatus. The place where the gas was sampled was sufficiently far downstream to provide adequate mixing of the CO_2 and the air (see Figure 26).

C) Wet Bulb Velocity

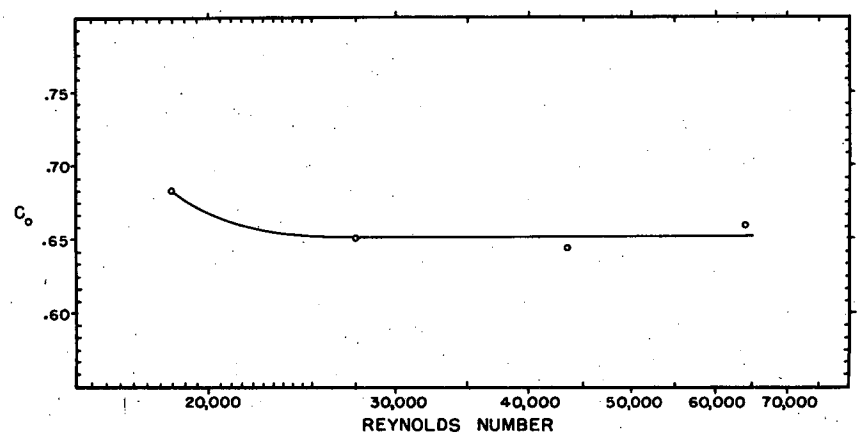
It was necessary to know the velocity across the wet bulb thermometer to be able to make the radiation correction. The thermometer was replaced by a glass impact tube and a pressure tap was located in the duct wall directly opposite to it. At known flow rates, a pitot traverse was made across the portion of the duct normally occupied by the thermometer. The variation



MU-4855

Fig. 25

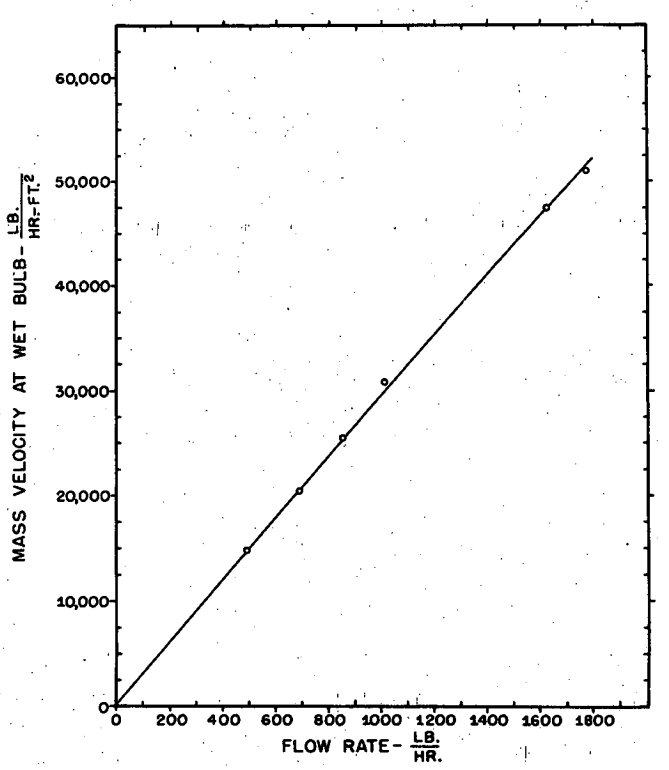
Discharge coefficient for flow nozzle A-1.



MU-4856

Fig. 26

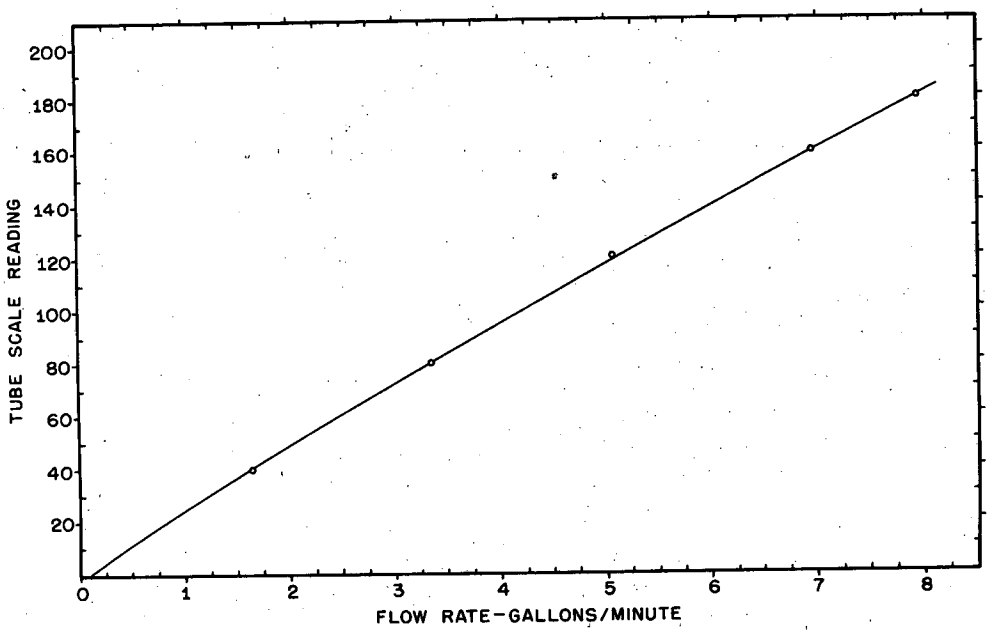
Discharge coefficient for flow measuring orifice.



MU-4857

Fig. 27

Average velocity at the thermometers vs. flow rate.



MU-4858

Fig. 28

Calibration curve for Fischer-Porter rotameter No. M9-1578-4.

over the last one-and-one-half inches of the thermometer was only 3 percent (see Figure 27).

D) Rotameter

The rotameter was calibrated by weighing the amount of water flow for a measured period of time (see Figure 28).

Sample Calculations

A) Flow Rates

Calculation of the gas flow rate was made with the use of equations taken from the Am. Soc. Mech. Eng., 1949 (1), Fluid Meters Report. The weight rate of flow through a nozzle is given as:

$$w = 0.5250\ D_2^2 YC \sqrt{\rho_1 \Delta P_t}\ \text{lbs./sec.}$$

where D_2 = nozzle diameter in inches

C = discharge coefficient - (see Figure 25)

ΔP_t = impact head minus pressure head at nozzle discharge in lbs./in.^{2}

Y = compressibility factor.

The values of C were obtained from the work of Bean, Buckingham, and Murphy (4). The weight rate of flow through the orifice is given by:

$$w = 0.06479\ Y \sqrt{\rho_1 \Delta H}\ \text{lbs./sec.}$$

where Y = compressibility factor $= 1 - 0.41(p_1 - p_2)/p_1 \frac{C_p}{C_v}$

ΔH = pressure drop across the orifice in inches of water

ρ_1 = density at upstream conditions.

For air run no. 13, which will be used throughout as the sample,

the nozzle equation can be simplified to

$$w = 24542\ C\sqrt{P_2 \Delta P_t / T_1}\ \text{lbs./hr., where } T_1 = {}^{o}K$$

$$w = 24542 \times 0.983\sqrt{14.625 \times 0.00964/326.7}$$

$$= 24125 \times 0.02077 = 501.08 \text{ lbs./hr.}$$

This value must be corrected to include the average amount of water evaporated in the tower.

$$w_T = 501.08\ (1 + 0.0089 \times 18/29) = 503.9$$

where 0.0089 is one-half the number of moles of water evaporated per mole of gas passing through the tower. The average gas properties are evaluated at the average temperature and pressure in the tower. These are:

$$\rho_{av.} = 0.0700 \text{ lbs./ft.}^3$$

$$\sqrt{\rho_{av.}} = 0.2646$$

$$\mu = 0.04496 \text{ lbs./ft. hr.}$$

μ is properly evaluated for the mixture of air and water by using the equation of Buddenburg and Wilke (7):

$$\mu_{av.} = \frac{\mu_1}{1 + 1.385\mu_1 x_2 / D_1 x_1} + \frac{\mu_2}{1 + 1.385\mu_2 x_2 / D_2 x_2}$$

where D = diffusivity and x = mol fraction.

The values of the correlating moduli used in representing the results are now obtained simply as:

$$G/\mu = 14,417$$

$$\sqrt{\rho}\, u_m = 2450.$$

Values for the liquid flow rate were taken directly from the calibration chart.

B) Height of Transfer Unit

Using the properties of air listed in Perry's Handbook (43), the weight of gas in the aspirator bottle was calculated at the temperature and pressure recorded:

$$\text{weight of air} = \frac{0.6868 \times 401.25}{13.740 \times 407.14} = 0.04926 \text{ top sample}$$

$$= \frac{0.6833 \times 400.75}{13.740 \times 407.14} = 0.04895 \text{ bottom sample,}$$

where the values 0.6868 and 0.6833 are the bottle volumes. From this the lb. moles of gas could be calculated by dividing by the molecular weight

$$\frac{0.04926}{29} = 0.0016986$$

$$\frac{0.04895}{29} = 0.0016879$$

The moles of water are obtained by taking the weight of water in grams removed from each of the above gas samples and dividing by 18 and 453.6.

$$\frac{0.5945}{18 \times 453.6} = 0.00007281$$

$$\frac{0.3977}{18 \times 453.6} = 0.00004871$$

The mole fraction in the gas stream entering and leaving the tower is obtained by division:

$$\frac{0.00007281}{0.0016986 + 0.00007281} = 0.04110$$

$$\frac{0.00004871}{0.0016879 + 0.00004871} = 0.02805$$

The measured inlet and exit water temperatures give vapor pressures of 33.446 mm. and 33.484 mm. Dividing these values by the tower pressure at the top and bottom respectively, the interfacial mole fractions of water are obtained:

$$\frac{33.446}{753.76} = 0.04437 \qquad \frac{33.484}{754.40} = 0.04438$$

The number of transfer units was taken as

$$NTU = \left[\ln \frac{y_{i_1} - y_1}{y_{i_2} - y_2}\right]\left(\frac{y_f}{1-y}\right),$$

where $_1$ refers to inlet and $_2$ refers to exit.

$$NTU = \left[\ln \frac{0.04438 - 0.02805}{0.04437 - 0.04110}\right]\left(\frac{0.956}{0.965}\right)$$

$$= (\ln 4.994)(0.992) = 1.609$$

The height of a transfer unit is the packed length divided by NTU:

$$HTU = \frac{0.625}{1.609} = 0.388$$

C) Calculation of the Psychrometric Ratio From the Wet Bulb Depression

The data required for this calculation are the wet bulb, dry bulb, wall, and ambient air temperatures, the gas velocity, water concentration in the gas, and the total pressure. The thermometer readings were first corrected for the calibration and for the emergent stem. The emergent stem correction was:

$$\text{correction} = Kn(t_g - t_r)$$

where K = 0.00016, n = degrees emergent, t_g = gas temperature, and t_r = ambient air temperature. For run no.5, this correction was 0.18° C. for the dry bulb and 0.03° C. for the wet bulb. The

corrected temperatures were then:

$$t_D = 50.70 + 0.18 + 0.05 = 50.93^\circ \text{ C.}$$

$$t_W = 29.50 + 0.03 - 0.04 = 29.49^\circ \text{ C.}$$

$t_S = 48.67$ (obtained from a thermocouple embedded in the duct wall)

$$t_r = 25.9$$

The gas mass velocity was obtained from a previously prepared calibration chart. The chart was made by making pitot traverses across the wet bulb and plotting the average against the mass velocity measured at the nozzle. For this run, the value of G at the wet bulb was 11,604. This same value was used in calculations involving the dry bulb. For gases other than air, it was necessary to calculate the Reynolds and Prandtl numbers to get the heat transfer coefficients for the two thermometers from the equation (45):

$$h_c = \frac{k}{d} 0.26 \, Re^{.6} \, Pr^{.3}$$

For air this simplifies to:

$$h_c = 0.026 \, G^{.6}/d^{.4}$$

where d is the diameter of the thermometer. The diameter of the wet bulb was 0.29 inch and the dry bulb was 0.23 inch. For this run h_c for the wet bulb is 33.2 and for the dry bulb is 43.2. The radiation coefficient h_r was calculated for both thermometers using the Stefan-Boltzman relationship with emissivities of 0.94 for the thermometer and 0.90 for the wetted wick. The radiation coefficients were 1.26 and 1.11 for the dry and wet bulb respectively. By applying the radiation correction, the true

temperature of the gas was found to be 51.00° C. At the wet bulb temperature, the vapor pressure of the water is 30.905 mm. The partial pressure in the gas is 19.910 mm. The difference is 0.01447 atm. The latent heat of vaporization at the wet bulb temperature is 1045 Btu. The film pressure factor, p_f, was taken as the log mean average partial pressure of dry air and is 0.9669. The heat transfer factor, γ, as described in the theory (p. 18) was 0.985. The psychrometric ratio is given by the equation:

$$\frac{k_G M_m p_f \gamma}{h'} = \frac{(t_g - t_W)\left\{1 + \frac{h_r}{h_c}\left[\frac{t_S - t_W}{t_g - t_W}\right]\right\}}{(p_W - p)\lambda} \frac{M_m p_f \gamma}{M_v}$$

$$= \frac{38.72\ (1 + 0.030)}{0.01447\ (1045)} \quad \frac{28.60 \times 0.9669 \times 0.985}{18}$$

$$= 3.99$$

In the calculations of the Schmidt and Prandtl numbers, the average values of the heat capacity and density were used. The viscosity of the mixture was calculated, using the equation of Buddenburg and Wilke (7). The thermal conductivity of the mixture was calculated, using the equation of Lindsay and Bromley (34). The diffusivities were those measured by Chen Ying Lee (32) corrected for temperature by the equations of Hirschfelder et. al. (24)

TABLE I

Summary of Wet-bulb Data for Various Investigations*

Investigator and System	Pr	Sc	Sc/Pr	$\frac{k_G M_m p_f \gamma}{h'} C_p$
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, °C.

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.

TABLE V

Miscellaneous

Volumes of aspirator bottles for withdrawing samples:	
top sample	- 0.6868 cu. ft.
bottom sample	- 0.6833 cu. ft.
Wet bulb diameter	- 0.29 inch
Length of thermometer bulb	- 1 inch
Length of wick exposed	- 2 inches
Dry bulb diameter	- 0.23 inch
Flow nozzle - Bean, Buckingham, and Murphy designation A-1	
Flow origice - square edge, 1.000 inch in diameter.	

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