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A NEW CORRELATION FOR MASS TRANSFER IN THE FLOW OF GASES THROUGH PACKED BEDS AND FOR THE PSYCHROMETRIC RATIO

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Edward J. Lynch and Charles R. Wilke

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ABSTRACT

A new method is proposed for the correlation of mass transfer in the flow of gases through packed beds. This method is based on an empirical extension of the equations of Lin, Moulton, and Putnam and von Karman. The proposed equation agrees well with data on the vaporization from solids over a range of Schmidt numbers from 0.24 to 3.15.

Modifications of the proposed equation appear to be applicable to mass transfer in packed columns with liquid flow and to psychrometric work. More extensive data will be necessary to confirm the suitability of these applications.

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INTRODUCTION AND THEORY

At the present time there is not an adequate theoretical treatment of mass transfer in packed beds either with or without countercurrent liquid flow. Attempts at a semi-theoretical approach through analogies and dimensional analysis have led to equations relating the mass transfer coefficient to the modified Reynolds number and the Schmidt number in a generalized form. The specific mass transfer equations, in the final analysis, have been determined by application of experimental data. Probably the best known and most widely used of these is the j-number form of Chilton and Colburn:[4]

$$j_D = \frac{k_G M_m p_f}{G} (Sc)^{2/3} = \psi (Re) \qquad (1)$$

where ψ (Re) is usually represented by a simple exponential function of the Reynolds number or by an empirically determined curve.

Chilton and Colburn's use of the Schmidt number to the 2/3 power has received wide acceptance. The exponential function was actually based on an empirical study of heat transfer data in pipes, but it appears to be reasonably valid also for mass transfer in packed beds with no liquid flow. Many of the experimental studies are of such poor precision as to be inconclusive with respect to the Schmidt number function. For example, in the case of mass transfer with both gas and liquid flow the dependence of the mass transfer coefficient on the Schmidt number has been reported from $Sc^{.17}$ to $Sc^{1.0}$.

The analytical treatment of the transfer of heat and/or mass from a stationary surface to a fluid in turbulent flow has been made by several authors[2,6,18,24] for fully developed flow in straight circular pipes. A similar direct approach for the packed bed has been prevented by a lack of knowledge about the flow pattern, the continual disturbance of the boundary layer so that uniform conditions are never established at the surface, and the presence of a very large form drag. From the aforementioned analytical treatments, however, it is possible to make a qualitative extension to mass transfer from a fixed surface contained within a packed bed.

Lin, Moulton, and Putnam[18] have recently derived an equation for mass transfer from the solid wall of a pipe which allows for turbulent penetration of the laminar sublayer in proportion to the third power of the distance from the wall:

$$k_c = \frac{u^{f/2}}{1 + \sqrt{f/2}} \quad \phi(Sc) \tag{2}$$

where:

k_c = mass transfer coefficient based on concentration gradient

u = fluid velocity

f = friction factor

$$\phi(Sc) = \left[\frac{14.5}{3} (Sc)^{2/3} F(Sc) + 5 \ln \frac{1 + 5.64\, Sc}{6.64(1 - 0.041\, Sc)} - 4.77\right] \tag{3}$$

$$F(Sc) = 1/2 \ln \frac{[1 + 0.345\, Sc^{1/3}]^2}{1 - 0.345\, Sc^{1/3} + 0.119\, Sc^{2/3}} + 1.73 \tan^{-1}(0.398\, Sc^{1/3} - 0.578) + 0.905 \tag{4}$$

For gases the function $F(Sc)$ becomes approximately equal to $Sc^{1/3}$ and the value of $\phi(Sc)$ becomes approximately equal to that derived by Von Karman.[24]

It is not possible to make a direct application of this equation to mass transfer in packed beds. In the first place the derivation

assumed the existence of a uniform velocity distribution throughout the region of mass transfer whereas a continually interrupted flow pattern exists in a packed bed. In the second place the friction factor in a packed bed accounts for only a small fraction of the energy transferred. Both of these effects would tend to give much higher values of the mass transfer coefficient in a packed bed than that indicated by equation (2). It is conceivable, however, that while the equation would not give the correct magnitude of k_c, it may be quite correct in determining the form of the Schmidt number dependence. It was therefore assumed that the equation which will correlate mass transfer data in packed beds should be of the form:

$$\frac{1}{a_v HTU} = \frac{k_G M_m p_f}{G} = B \frac{C_m}{1 + \sqrt{C_m}\,\phi(Sc)} \qquad (5)$$

where:

B = a constant which is greater than 1.

C_m = a factor which is dependent only on the Reynolds number.

From the study of a complete set of experimental data covering a range of Schmidt numbers the value of the constant B can be determined. The factor C_m is used here instead of the friction factor of equation (2) because it is virtually impossible to evaluate a true friction factor in a packed bed. The dependence of C_m on the Reynolds number can be determined from a single set of reliable mass transfer data.

According to equation (5) the mass transfer coefficient is dependent on the Reynolds number through the factor C_m. Since $\sqrt{C_m}$ appears in the denominator as a multiplier of the Schmidt function, however, the dependence of the mass transfer rate on the Reynolds number will not be common for all systems, but will depend on the magnitude of $\phi(Sc)$. Also, since C_m is not a constant, the variation of the mass transfer rate with Schmidt number will not be the same for all Reynolds numbers.

EXPERIMENTAL

Physical and Thermodynamic Properties of Materials

Viscosity: The viscosities of air, water vapor, and hydrogen were taken from a review article by F. G. Keyes[16] in which the available literature data for several gases are collected and formulated by means of an empirical equation. The viscosity of Freon-12 was taken from a paper by Buddenberg and Wilke[3] 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.[12]

Thermal Conductivity: The thermal conductivities of air, water vapor, and hydrogen were taken from the review article by Keyes quoted above. The thermal conductivity of Freon-12 is that reported by Sherratt and Griffiths.[23]

Heat Capacity: The heat capacities of air and water vapor were taken from the book by Keenan[15] on the thermodynamic properties of air. The heat capacity of hydrogen is from the API Tables.[1] The heat capacity of Freon-12 was calculated from the work of Justi and Langer[14] who reported the heat capacity at constant volume.

Diffusivity: The diffusivities for the systems air-water, air-nitrobenzene, and Freon-12 - water are those reported by Lee and Wilke.[17] The diffusivities for the systems hydrogen-nitrobenzene, and Freon-12 - naphthalene were calculated by the equation of Hirschfelder, Bird, and Spotz.[12] The diffusivity for the system air-naphthalene was taken from the I.C.T.[13]

Density: Gas densities were taken from the Physics and Chemistry Handbook[11] with ideal gas pressure and temperature corrections.

Vapor Pressure: The vapor pressure of nitrobenzene was measured by the authors and reported in detail elsewhere.[20] The vapor pressure of naphthalene is from the I.C.T.[13]

Gas Specifications: The hydrogen was supplied by the Stuart Oxygen Co. and was made electrolytically. The Freon-12 was supplied by Kinetic Chemicals with a purity of 99.95%. The major impurities were Freon-11 and Freon-13.

Liquid and Solid Specifications: The nitrobenzene was Eastman Kodak's White Label grade. This was distilled in an Oldershaw column and the center cut retained for the vaporization experiments. Naphthalene was used directly as supplied by Baker Chemical Co., C.P. grade. The melting point was measured and found to be in agreement with accepted values.[11]

Equipment

The equipment is shown schematically in Figure 1. The flushing box was a large glove box connected to the flow system. It could be filled with the same gas being used in the experiments. This made it possible to remove the packed bed from the flow system and weigh it without removing it from the gas atmosphere. The water cooled and refrigerated sections removed condensables from the gas stream so that the gas could be recirculated. The packed tee served as an equilibrium contact surface for the gas and liquid (or solid). The temperature measured at this point determined the concentration of the transfer material in the gas stream entering the bed. (In the case of air-water runs this stream was also sampled and the water vapor determined gravimetrically. There was very good agreement between the two methods.) Temperature measurements were also made at the locations indicated in the figure, i.e. before the packed bed (1), following the packed bed (2), and on the surface of the tube (3) adjacent to the location of thermocouple (2). No. 30 copper-constantin wire was used with an L and N Portable Precision Potentiometer for all measurements.

A detail of the packed section is shown in Figure 2. The bed, which was approximately 1" deep and $2\frac{1}{2}$" in diameter was held in place by a spring clamp mechanism. Tapered, smoothly machined surfaces served to align the bed and provide a seal; the bellows provided sufficient play for insertion and removal. The packing was composed of $\frac{1}{4}$ x $\frac{1}{4}$" naphthalene cylinders or $\frac{1}{4}$" alundum spheres soaked with either nitrobenzene or water.

Procedure

The system was first brought to an equilibrium condition through

a period of prolonged recirculation with a blank section in place of the packed bed. The packed bed was then weighed and inserted. Pressure drop readings were taken across the orifice and the bed, and absolute pressure readings were taken before the orifice and the bed. Temperatures were taken at the four points indicated in Figure 1. After a predetermined period the bed was removed and reweighed. When the alundum pellets were used, the run was always restricted to the constant-rate drying period.

When gases other than air were used, it was necessary to first purge the air from the system and the flushing box. Purging was continued until the concentration of air was reduced to less than 2%. A continuous flow of gas into the system was then maintained throughout all experiments and the concentration was checked at regular intervals.

RESULTS

The experimental results of this study are shown in Figures 3 and 4. Six systems were studied, having a range of Schmidt numbers from 0.24 for water-Freon-12 to 3.15 for nitrobenzene-hydrogen. The results are plotted as the modified Reynolds number, $D_pG/\mu(1-\epsilon)$, versus a_vHTU. This method of graphical representation was selected on the basis of the conclusions drawn in reference 7 and the work of other investigators.[5,9,22]

Since the data for the nitrobenzene-air system are the most consistent and extensive, they were selected to serve as a basis for the semi-theoretical equations for correlation of mass transfer data in packed beds. In Figure 3 the line for Schmidt number 1.79 was drawn through the points for this system. The lines for the other Schmidt numbers were located in accordance with the j-number theory, i.e. all lines were drawn parallel to the base line but spaced vertically according to the Schmidt number to the 2/3 power. This type of correlation gives a good fit of the data for the higher Schmidt numbers. For Schmidt numbers less than 1.79, however, the agreement becomes progressively worse, reaching a condition of essentially no agreement between the calculated line and the data for the water-Freon-12 system

which has a Schmidt number of 0.24.

Figure 4 was drawn with equation (5) of the theory as a basis. By means of a trial procedure it was determined that the value of the constant B should be approximately 20, so that equation (5) then becomes:

$$\frac{1}{a_v HTU} = 20 \frac{C_m}{1 + \sqrt{C_m}\,\phi(Sc)} \qquad (6)$$

The nitrobenzene-air data were again selected for the location of a base line and the factor, C_m, was evaluated as a function of the modified Reynolds number from this line. This relationship is plotted in Figure 5. Equation (6) was then used in conjunction with Figure 5 to locate the lines drawn on Figure 4 for the other Schmidt numbers. For the higher Schmidt numbers this correlation is comparable with the conventional j-number correlation using the Schmidt number to the 2/3 power. For Schmidt numbers lower than 1.79, this correlation appears to be superior to the former methods.

A cross-plot has been made from Figure 4 at modified Reynolds numbers of 400 and 4000 and is presented in Figure 6. The trends with Schmidt number which are shown here open the way for a discussion of some of the anomalous results that have been noted in other mass transfer studies.

DISCUSSION

Equation (6) is plotted as $a_v HTU$ versus modified Reynolds number in Figure 4 with Schmidt number as the parameter and as $a_v HTU$ versus Schmidt number in Figure 6 with the modified Reynolds number as the parameter. Figure 5 defines the relation between C_m and the modified Reynolds number.

The lines of constant modified Reynolds number on Figure 6 show that when the Schmidt number is greater than 2, the effect of Schmidt number on mass transfer approximates $Sc^{2/3}$ regardless of the value of the Reynolds number. For high values of the Schmidt number, therefore, equation (6) becomes approximately:

$$\frac{1}{a_v HTU} = \frac{20\sqrt{C_m}}{Sc^{2/3}} \tag{7}$$

This relation is of the form predicted by the j-number analogy and probably explains why that analogy has received such wide support. For values of Sc less than two, the effect of Schmidt number is greatly influenced by magnitude of C_m and consequently by the magnitude of the modified Reynolds number. For a Re' of 400, the effect is greater than $Sc^{2/3}$ and for a Re' of 4000 the effect is less than $Sc^{2/3}$. At a modified Reynolds number of 600, the effect of the Schmidt group is approximately $Sc^{2/3}$. Since this latter value of the Reynolds number corresponds to the region in which most of the mass transfer work has been done, it has been very easy to assume that the j-number analogy is valid for all values of Schmidt and that discrepancies are the result of experimental error.

The anomalous change of the HTU with Schmidt number at the lower value of Schmidt number results in lines of gradually increasing slope when HTU is plotted against Reynolds number as in Figure 4. This anomaly also explains the disagreement among investigators concerning the dependence of j_D on the Reynolds number. One of the first investigations in the field was that of Gamson, Thodos, and Hougen[10] who vaporized water from various porous packings. They concluded that j_D varied as $Re^{-0.41}$. Although the line for water-air in Figure 4 is curved, the average slope of the line over the region in which Gamson, Thodos, and Hougen worked is 0.45. On the other hand, Shulman, Ullrich, Proulx, and Zimmerman[22] have recently reported that j_D varied as $Re^{-0.36}$. Their work was with the naphthalene-air system, however. From Figure 4 it can be seen that the dependence on Reynolds number for this system is approximately $Re^{-0.34}$. The difference between the two investigations, therefore, would seem to be consistent with the present method of correlation. A comparison of the data of this work with these two earlier correlations is shown in Figure 7.

In prior work by the authors[19] on mass transfer in a packed column with liquid flow it was found that a correlation of HTU with Reynolds number required that the Schmidt number enter to the 0.9 power. (A later investigation by Yoshida[25] appeared to confirm these results

provided that the same diffusivity is used for the helium-water system.) In the light of the known turbulence in a packed column, such a correlation did not seem reasonable at the time. A correlation with the inertia group, on the other hand, required only a 0.47 exponent on the Schmidt number and also served to align the flooding points. This was also in agreement with a psychrometric study which was conducted concurrently and which indicated that the mass transfer rate should vary with $Sc^{0.5}$. At the time it was noticed that the slope of the water-Freon-12 curve was somewhat greater than the other two, but straight parallel lines were drawn through all three sets of data and the results correlated as:

$$\frac{HTU}{Sc^{0.47}} = \psi(\rho u^2) \text{ for } 0.24 < Sc < 1.15 \tag{8}$$

On the basis suggested by equation (5), however, it is possible to make a different interpretation of these data. Figure 6 shows that at low Reynolds numbers and low Schmidt numbers the Schmidt number dependence approaches the first power. On this basis an exponent of 0.9 for Schmidt numbers between 0.24 and 1.15 would be quite reasonable. An attempt was made, therefore, to use equation (5) to correlate these data. The constant B was found to be 3.5 for this case:

$$\frac{1}{a_v HTU} = 3.5 \frac{C_m}{1 + \sqrt{C_m}\, \phi(Sc)} \tag{9}$$

C_m was evaluated as a function of the modified Reynolds number by using the data for the air-water system. This equation is plotted as the solid lines on Figure 8. It gives an excellent correlation of all data up to the loading points where the HTU decreases sharply (as indicated by the dotted lines) because of changes in the liquid flow pattern within the column. Equation (9) allows not only for the large Schmidt number dependence which was observed for these systems, but also for the fact that the Freon-12 system shows a different dependence on Reynolds number.

Figure 8 should not be construed as supporting the use of the equivalent packing diameter or the porosity in the Reynolds number for the case where there is liquid flow. In fact, it does not definitely establish whether this correlation method or that indicated by equation (8) is to be preferred. Final determination of the proper method will require data similar to that given in this paper and reference 7. An unfortunate aspect of any similar investigation will be the limiting condition imposed on the Reynolds number by the loading and flooding of the column.

In the psychrometric study mentioned above the relation obtained was:[1]

$$\beta = \frac{k_G M_m p_f C_p \gamma}{h'} = 0.91 \left(\frac{Pr}{Sc}\right)^{1/2} \qquad 0.24 < Sc < 2.65 \qquad (10)$$

---

[1]The factor γ which corrects the heat transfer coefficient for the effect of mass transfer was incorrectly defined in the previous paper by the authors.[19] However, the error was minor for the systems studied. A preferable form for γ is the Ackermann correction[a] which was verified experimentally by Mickley and Associates[b] for heat transfer at a flat plate with blowing or suction:

$$\gamma = \frac{\phi}{e^{\phi} - 1}$$

where:

$$\frac{N_A C_{pA}}{h} = \phi$$

N_A = rate of evaporation from the wet bulb.

C_{pA} = heat capacity of the evaporating liquid.

h = heat transfer coefficient in absence of mass transfer.

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[a]Ackerman, G., Forschungsheft, No. 382, 1-16 (1937).

[b]Mickley, H.S., Ross, R.C., Squyers, A.L. and Stewart, W.E., N.A.C.A. Technical Note 3208 (July, 1954).

Since the Prandtl number is approximately constant for all gases, this equation indicates that the mass transfer rate will be proportional to the Schmidt number to the ½ power.

It is customary in psychrometric studies to operate with very high velocities over the thermometers in order to reduce the radiation correction. This condition should correspond to the packed bed at very high Reynolds numbers. Reference to Figure 6 show that for Re' = 4000 and $0.24 < Sc < 2.65$ the dependence of the mass transfer rate on the Schmidt number is approximately proportional to $Sc^{1/2}$. This agreement with the above conclusions from the psychrometric study suggests a correlation for the psychrometric ratio based on equation (5):

$$\beta = \frac{k_G M_m p_f C_p \gamma}{h'} = \frac{B'[1 - \sqrt{C_h}\phi(Pr)]}{B[1 - \sqrt{C_m}\phi(Sc)]} \tag{11}$$

The best fit to the available data is obtained with $B'/B = 0.91$ and $C_m = C_h = 0.056$ so that:

$$\beta = \frac{k_G M_m p_f C_p \gamma}{h'} = 0.91 \frac{1 + 0.056\ \phi(Pr)}{1 + 0.056\ \phi(Sc)} \tag{12}$$

Equation (12) is plotted in Figure 9. The data are those reported in Table I of the earlier work.[19]

The applicability of the correlation method to systems of high Schmidt number was checked by using the data of Gaffney and Drew[8] for the systems acetone-succinic acid (Sc = 170), benzene-salicylic acid (Sc - 370), and butyl alcohol-succinic acid (Sc = 10960), and of McCune and Wilhelm[21] for the naphthol-water system (Sc = 1385). These data and equations (1) and (6) are plotted in Figure 10. Neither equation satisfies the data although equation (1) appears to be superior. Equation (6) is not recommended, therefore, for liquid-solid systems.

The design chart presented in Figure 11 is a plot of equation (6) for selected values of the Schmidt number. The height of a transfer unit can be obtained directly from the chart once the Schmidt number and modified Reynolds number are known. The range of Schmidt numbers given should cover gaseous systems normally encountered in practice.

CONCLUSION

It is believed that Equation (6) represents a useful method of interpretation of mass transfer in packed beds where there is no liquid flow. It also appears that other equations of this same form may be useful for the correlation of data obtained on packed columns with liquid flow and the correlation of psychrometric data.

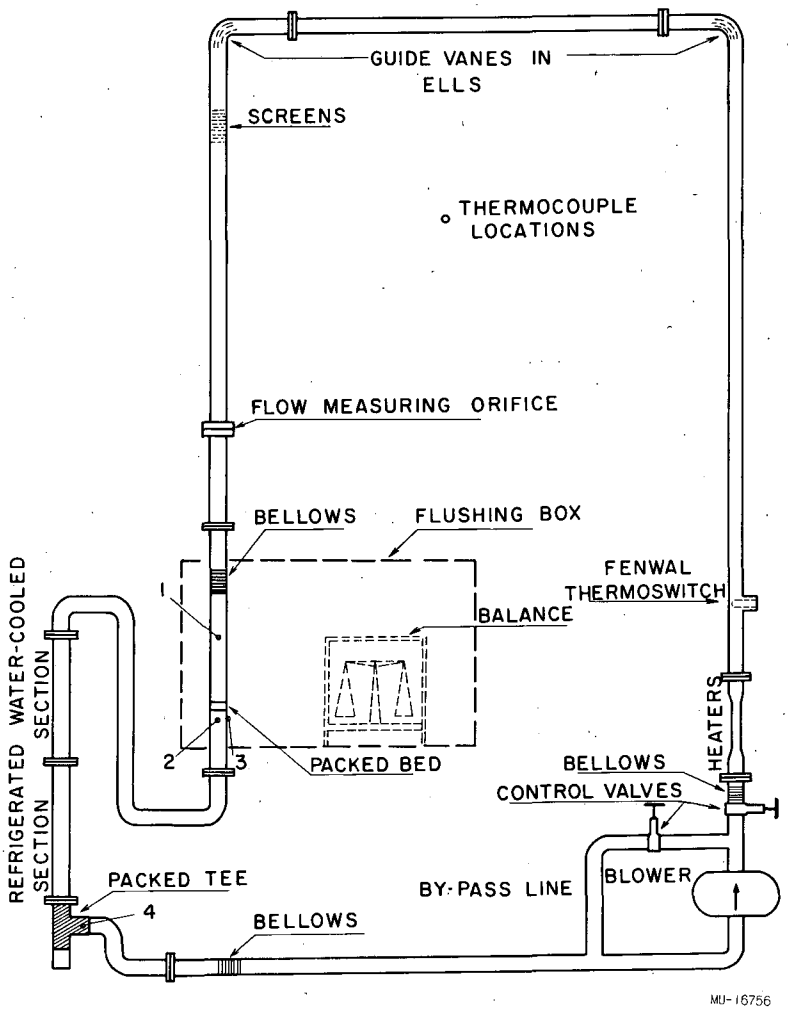

Fig. 1. Schematic diagram of the apparatus.

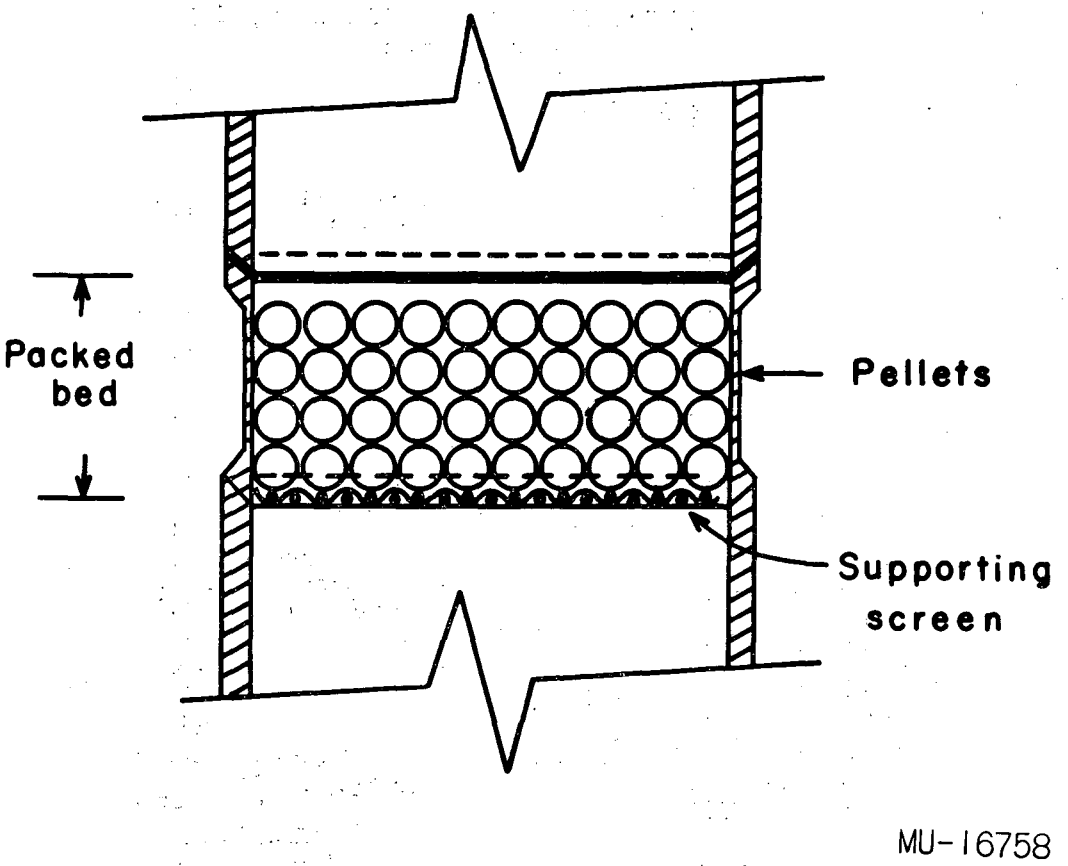

Fig. 2. Packed bed detail.

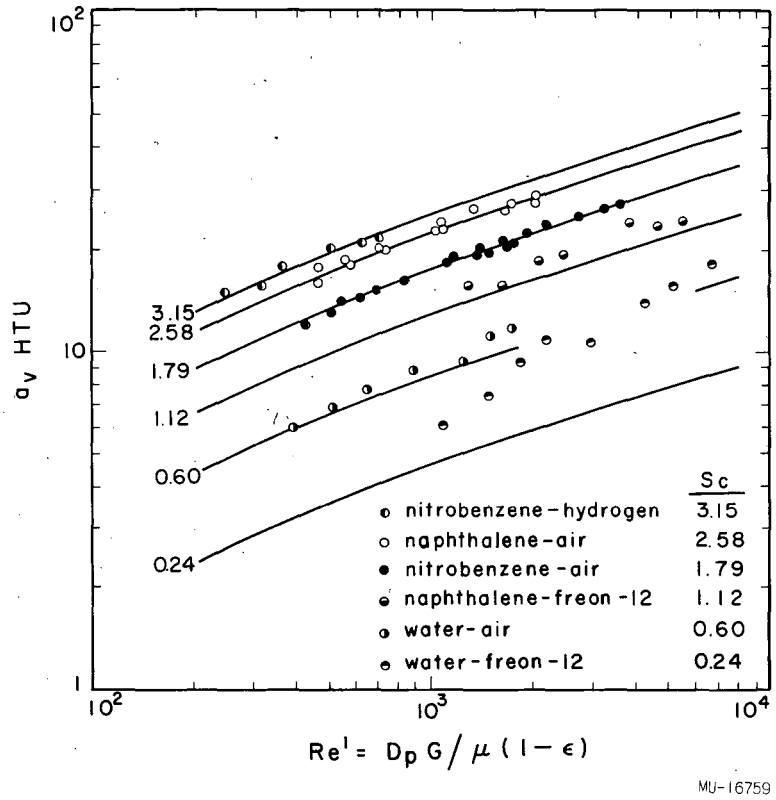

Fig. 3. Experimental data and calculated curves based on the j_D number analogy.

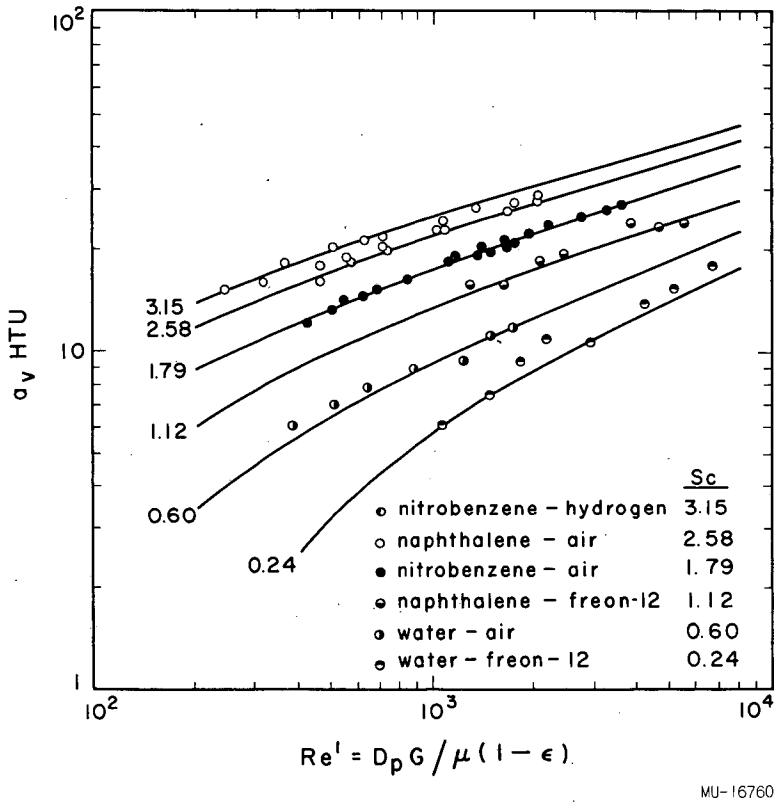

Fig. 4. Experimental data and calculated curves based on Equation (6).

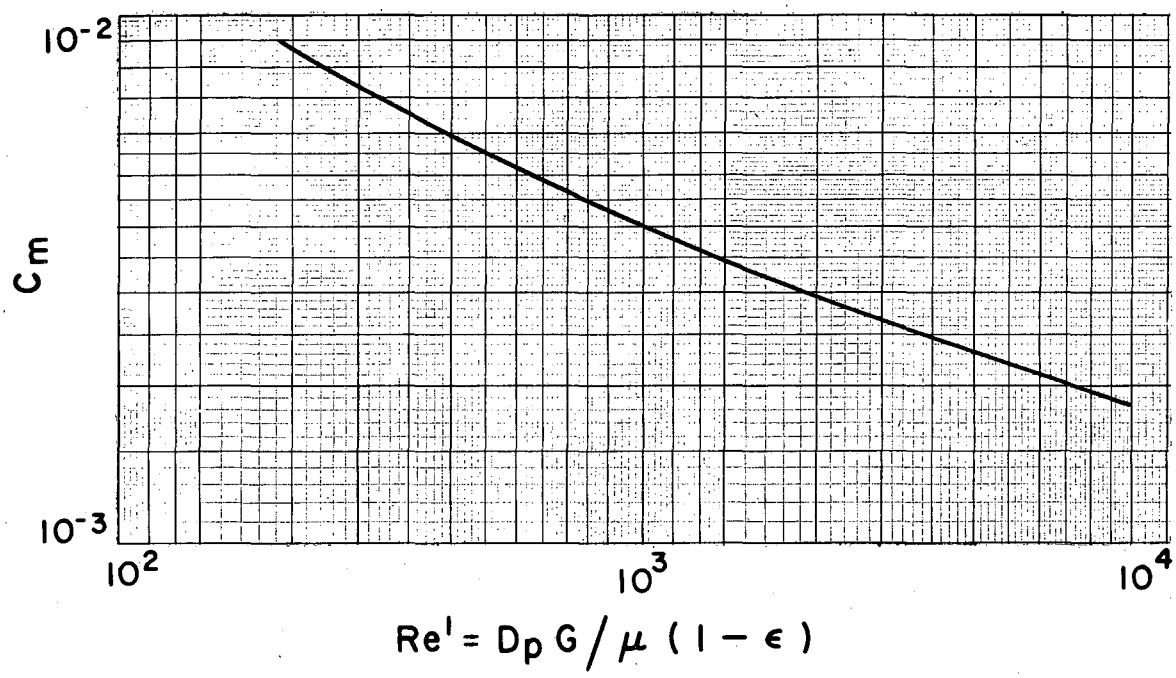

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Fig. 5. Relationship between C_m and Re' for packed beds.

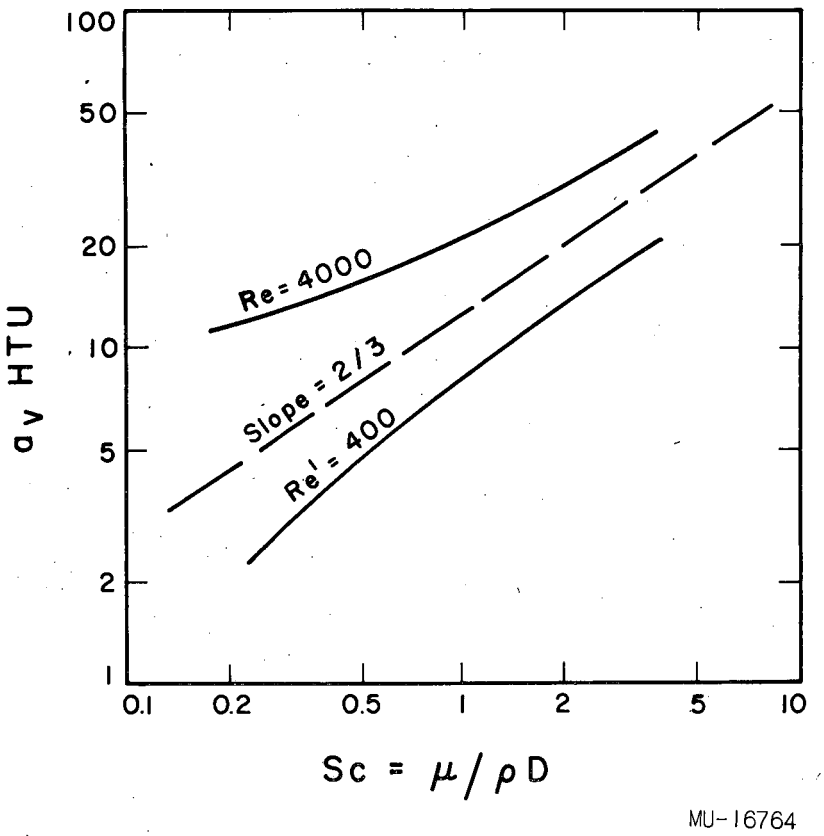

Fig. 6. Equation (6) as a **function of Schmidt number with Re'** as a parameter.

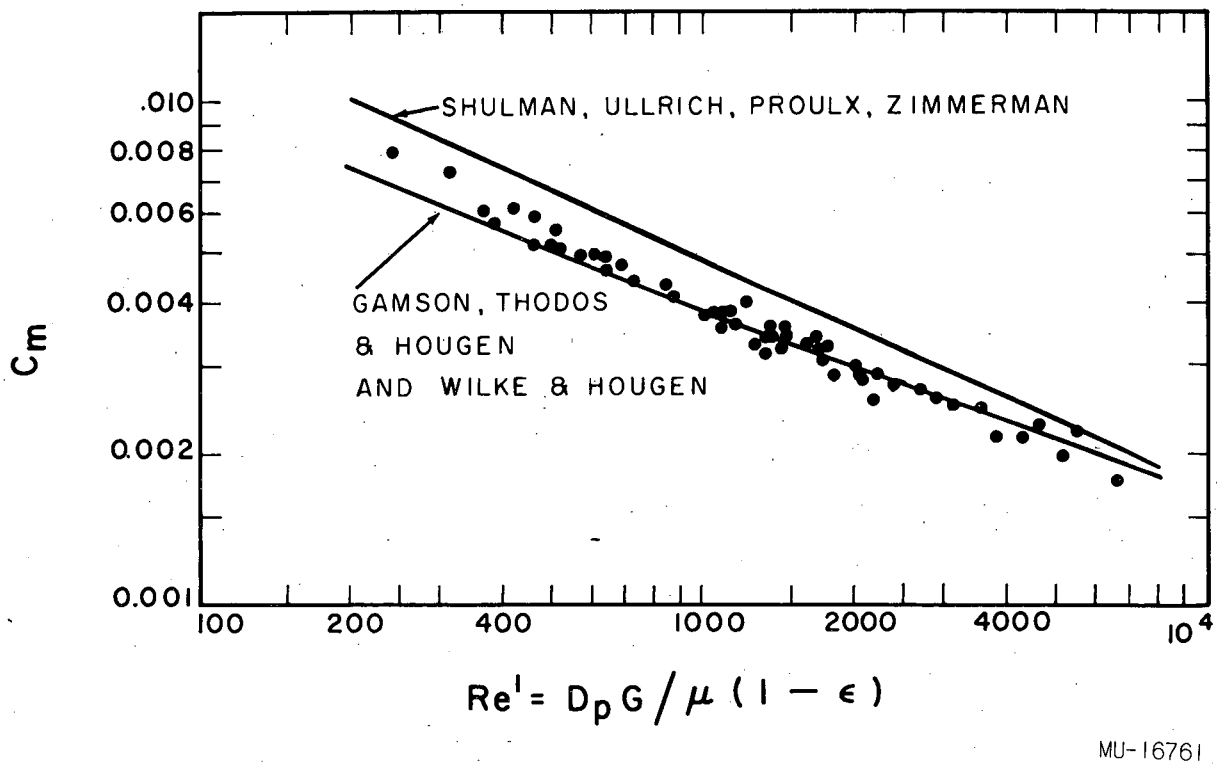

Fig. 7. Comparison of the data of this work with other correlations.

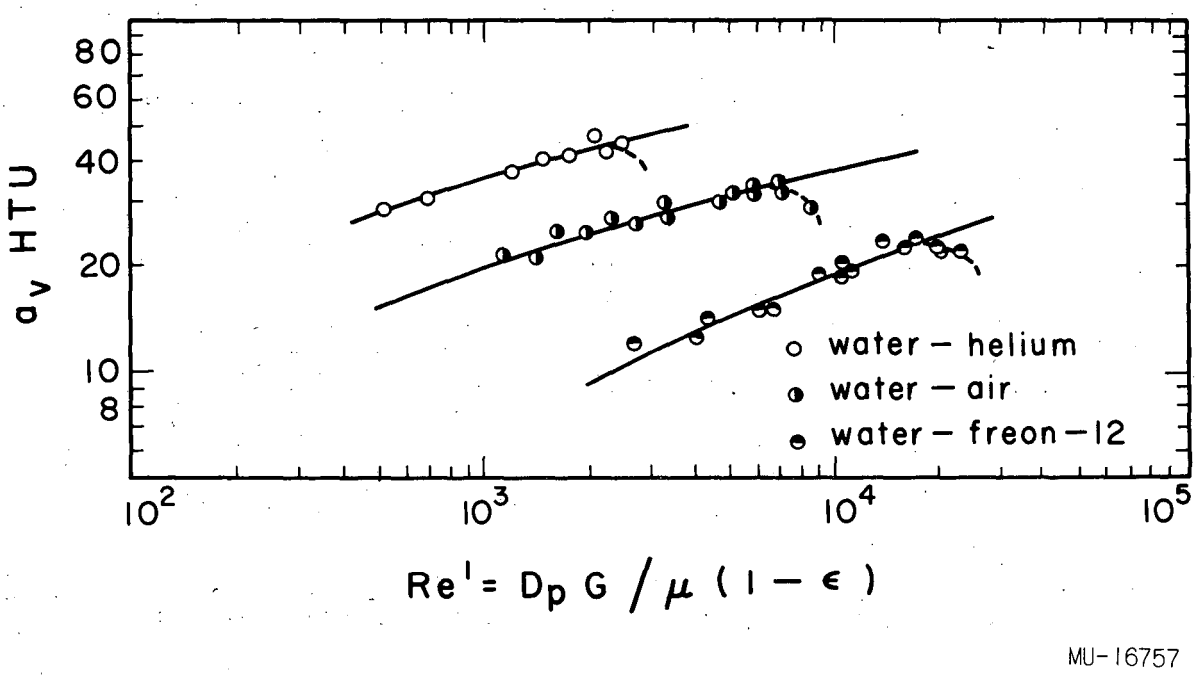

Fig. 8. Packed column data from Reference 19 with calculated curves based on Equation (9)

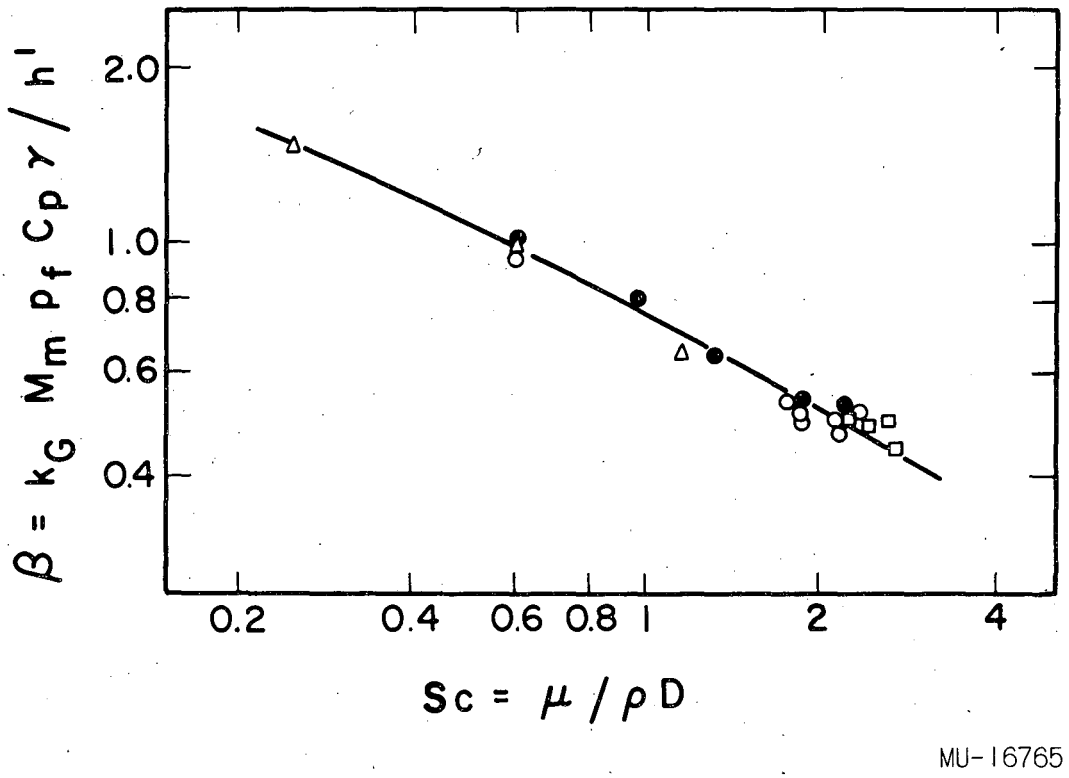

Fig. 9. Psychrometric data and the calculated curve based on Equation (12).

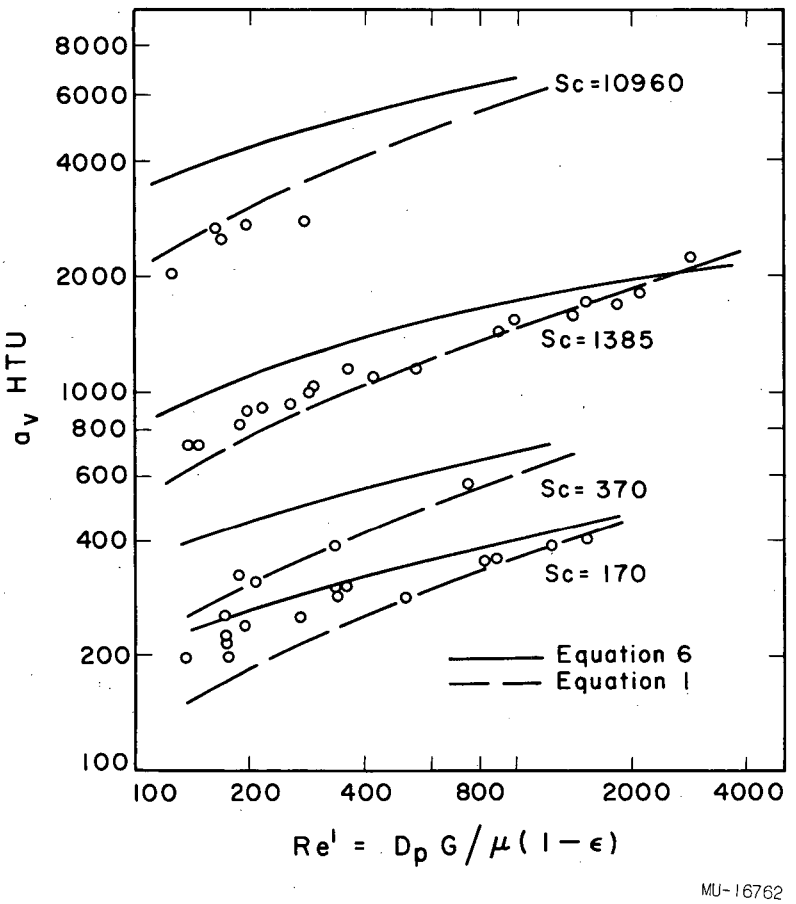

Fig. 10. Experimental data at high Schmidt numbers compared with Equations (1) and (6).

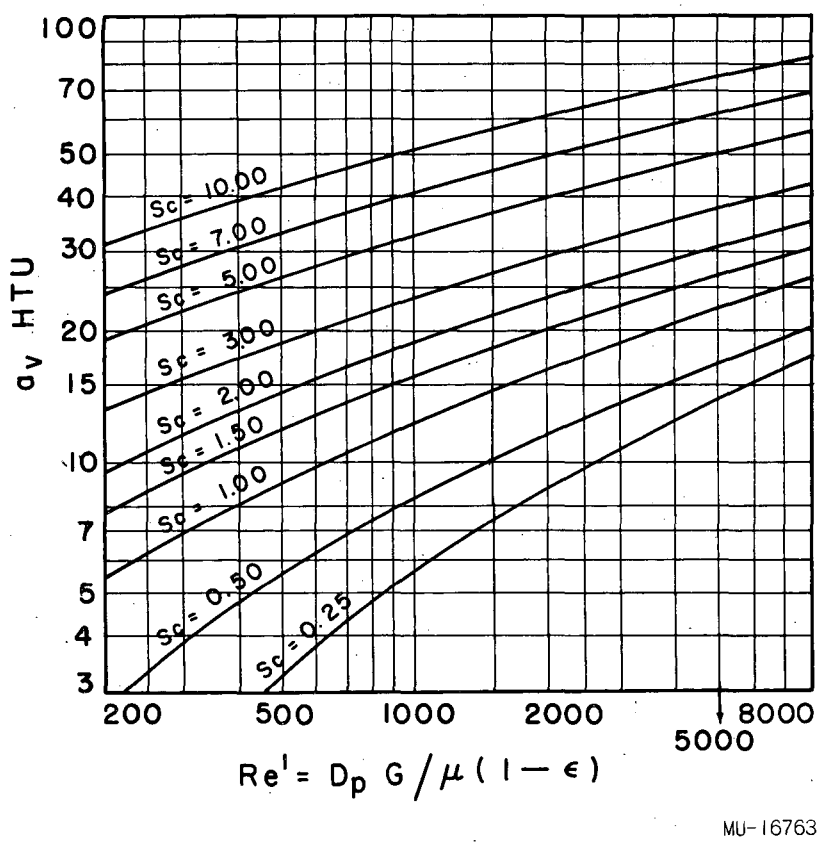

Fig. 11. Design chart developed from Equation (6) for mass transer in packed beds.

Table 1

Mass Transfer in Packed Beds
For Several Systems at Atmospheric Pressure

Run	NTU_G	$a_v HTU_G$	Sc	j_D	Re'	C_m
Air-Water						
1	1.763	7.837	.60	.0907	643.9	.00465
2	1.978	6.985	"	.1018	514.2	.00511
3	2.294	6.024	"	.1180	387.1	.00578
4	1.178	11.72	"	.0607	1714	.00328
5	1.232	11.21	"	.0634	1489	.00342
6	1.452	9.36	"	.0760	1237	.00401
7	1.547	8.93	"	.0796	887.4	.00416
Air-nitrobenzene						
1	1.134	12.18	1.79	.1211	424.0	.00621
2	1.045	13.21	"	.1116	509.1	.00564
3	.945	14.61	"	.1009	619.9	.00501
4	.906	15.24	"	.0968	687.4	.00476
5	.724	19.07	"	.0773	1162	.00366
6	.705	19.58	"	.0753	1477	.00355
7	.661	20.89	"	.0706	1752	.00329
8	.617	22.38	"	.0659	1910	.00305
9	.968	14.27	"	.1033	547.4	.00490
10	.841	16.42	"	.0898	842.2	.00436
11	.750	18.41	"	.0801	1102	.00382
12	.688	20.07	"	.0735	1357	.00345
13	.651	21.21	"	.0695	1643	.00324
14	.711	19.42	"	.0760	1359	.00359
15	.683	20.21	"	.0730	1677	.00342
16	.682	20.24	"	.0729	1397	.00341

Air-$PhNO_2$

Run	NTU_G	$a_v HTU_G$	Sc	j_D	Re'	C_m
17	.586	23.56	1.79	.0626	2212	.00287
18	.553	24.97	"	.0591	2746	.00268
19	.515	26.81	"	.0550	3591	.00247
20	.530	26.05	"	.0566	3168	.00255

Hydrogen-nitrobenzene

1	.910	15.17	3.143	.1415	243.1	.00796
2	.859	16.07	3.091	.1320	315.4	.00733
3	.754	18.31	3.250	.1198	364.3	.00612
4	.670	20.61	3.115	.1034	500.6	.00520
5	.636	21.71	3.198	.1000	700.5	.00484
6	.647	21.34	3.115	.0999	624.4	.00496

Air-naphthalene

1	.624	22.99	2.580	.0811	1091	.00379
2	.593	24.20	"	.0770	1095	.00355
3	.544	26.38	"	.0707	1344	.00319
4	.525	27.33	"	.0682	1743	.00306
5	.501	28.64	"	.0651	2030	.00287
6	.698	20.56	"	.0907	707.6	.00438
7	.756	18.98	"	.0982	561.3	.00486
8	.800	17.94	"	.1039	463.3	.00521
9	.631	22.74	"	.0820	1044	.00386
10	.552	26.00	"	.0717	1661	.00325
11	.513	27.97	"	.0666	2027	.00297
12	.886	16.20	"	.1149	463.7	.00598
13	.775	18.52	"	.1006	573.8	.00500
14	.703	20.41	"	.0913	733.1	.00442

F-12-naphthalene

Run	NTU_G	$a_v HTU_G$	Sc	j_D	Re'	C_m
1	.738	19.46	1.120	.0554	2415	.00271
2	.774	18.55	"	.0581	2070	.00285
3	.908	15.81	"	.0682	1624	.00336
4	.912	15.74	"	.0685	1282	.00338
5	.600	23.93	"	.0451	3809	.00220
6	.619	23.20	"	.0465	4675	.00227
7	.602	23.87	"	.0452	5555	.00220

F-12-water

1	.899	15.38	.2459		5136	.00198
2	1.005	13.74	.245		4284	.00215
3	1.301	10.62	.2445		2927	.00258
4	1.303	10.80	.2428		2166	.00257
5	.771	17.91	.2473		6622	.00176
6	1.498	9.22	.2436		1844	.00285
7	1.867	7.40	.2443		1464	.00333
8	2.299	6.01	.2422		1068	.00382

Nomenclature

a_v = area per unit volume, sq-ft/cu-ft

B = a constant defined by Equation (5)

C_h = heat transfer factor similar to C_m

C_m = mass transfer factor defined by Equation (5)

C_p = heat capacity at constant pressure Btu/(lb-mass)(^oF)

D_p = equivalent particle diameter for packing (ft)

f = friction factor

G = mass velocity of the gas stream based on the superficial area, lb-mass/(hr)(sq-ft)

h' = apparent heat transfer coefficient, see reference (19)

HTU = height of a transfer unit, ft

j_D = mass transfer factor defined by Equation (2)

k_c = mass transfer coefficient based on a concentration gradient

k_G = gas film mass transfer coefficient, lb-moles/(hr)(sq-ft)(atm)

M_m = mean molecular weight of the gas stream, lb-mass/lb-mole

p_f = film pressure factor, atm

Pr = Prandtl number, $\frac{C_p \mu}{k}$

Re' = modified Reynolds number $\frac{D_p G}{\mu(1-\epsilon)}$

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

u = mean velocity, ft/sec

$\phi(Sc)$ = a function of Schmidt number defined by Equation (3)

β = psychrometric ratio defined by Equation (10)

γ = factor which accounts for effect of mass transfer on heat transfer

ϵ = porosity

μ = viscosity, lb-mass/(ft)(hr)

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