

PARTICLE-TO-GAS MASS-TRANSFER MEASUREMENTS AND COEFFICIENTS IN FIXED BEDS AT LOW REYNOLDS NUMBERS

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Abstract—A model developed for calculating mass-transfer coefficients in packed beds at low Reynolds numbers showed that coefficients derived from measurements in beds of inert and active particles should be dependent on the geometric arrangements between active and inert particles. Experimental results with particles of size range 0.0171–0.9 cm supported the model prediction. Conflicting results obtained by a number of investigators concerning particle diameter effect on j_d factor are explained by these findings which also indicate that dispersed bed experimental measurements and calculations involving cup-mixing concentrations cannot result in valid mass-transfer coefficients.

NOMENCLATURE

a ,	specific surface [$\text{cm}^2 \text{cm}^{-3}$];
A ,	surface area for mass transfer [cm^2];
c ,	concentration of transferable component [g cm^{-3}];
c_0 ,	Concentration of transferable component at entrance [g cm^{-3}];
c_s ,	saturation concentration of transferable component [g cm^{-3}];
Δc_{lm} ,	log-mean driving figure [g cm^{-3}];
d_p ,	particle size [cm];
D_v ,	gas diffusivity [$\text{cm}^2 \text{s}^{-1}$];
F ,	D_v/u_e ;
j_d ,	mass-transfer factor;
\bar{k} ,	average mass-transfer coefficient;
l ,	height of cell [cm];
N_A ,	instantaneous mass-transfer rate [g cm^{-1}];
N_{Re} ,	modified Reynolds number, $d_p u \rho / \mu (1 - \epsilon)$;
N_{Sc} ,	Schmidt number, $\mu / \rho D_v$;
S_a ,	particle surface area in unit cross-section of height l [cm^2];
u ,	superficial gas velocity [cm s^{-1}];
u_e ,	gas velocity in cell [cm s^{-1}];
V_a ,	fluid volume in unit cross section of height l [cm^3].

Greek symbols

ϵ ,	void fraction;
θ ,	angle at z -coordinate from bed axis;
μ ,	absolute viscosity [$\text{g cm}^{-1} \text{s}^{-1}$];
ρ ,	gas density [g cm^{-3}];
ϕ ,	dimensionless concentration;
ψ_s ,	shape factor.

INTRODUCTION

SUBSEQUENT to the work of Gamson, Thodos and Hougen [1] and Hurt [2] a number of experimental studies have been made to extend their investigations of mass-transfer behaviour between solid particles and

gas in packed beds into the low Reynolds number region. To maintain a significant driving force for mass transfer throughout the entire depth of a packed bed at low Reynolds numbers requires either a shallow bed or else that a non-shallow 'dilute' bed be used. The dilute bed is obtained by dispersing 'active' particles throughout the matrix of inert or inactive material. In this latter approach the flow patterns within the interstices of the dilute or dispersed bed will be comparable to those of a deep packed bed, a situation that would not prevail in a one- or two-layer shallow bed in which entrance and exit effects could be significant.

In their work Thodos and co-workers [3,4] vapourized water or hydrocarbons from saturated porous spheres dispersed in a matrix of nonactive plastic or glass spheres. A number of investigations have involved the naphthalene-air system in which either a layer of active naphthalene particles was sandwiched between layers of inert particles [5] or else active naphthalene particles were dispersed in a matrix of inert particles [6,7]. In most of these works the investigators assumed that the mass-transfer coefficient as calculated from the measured mass-transfer rate, the area of active material in the dilute bed and the log-mean driving force based on inlet and outlet driving potentials would be identical to that expected from measurements made in a bed composed only of active particles. Hsiung and Thodos [7] measured concentration profiles within the packed bed to establish the mass transfer coefficients. The departure of the measured driving force from the log-mean value should produce information that reflects, in some lumped manner, the contribution of axial and radial dispersion effects associated with the transport of mass across a stagnant film.

No study has been made of the possible effect of different geometric arrangements among the active and inert particles so as to determine if measurements made in a dilute bed would, indeed, be applicable to a

packed bed of active particles. In this work a simple mathematical model is developed for mass transfer from active particles in an inert matrix at low Reynolds number. The major purpose of this model was to explore the possibility that measured mass-transfer rates in dilute beds could be dependent on the mutual geometric or spatial arrangements and relationships between the active and the inert particles. The model could also be used to cast some light on the conflicting experimental results in the literature in which some investigators observed a particle diameter effect additional to that accounted for in the Reynolds number [2, 5, 6] whereas other investigators [1, 3, 4, 7] did not observe a diameter effect. Experimental measurements were also made to confirm the model predictions.

FORMULATION OF THE MODEL

The model that is developed is as simple as possible; consistent with the purposes of the model. This is because the model was to be used primarily to study if measured mass-transfer coefficients were dependent on the spatial relationships between active and inert particles. The model was not intended to be used for numerical prediction purposes but, as will be seen, it was surprisingly accurate in this respect in spite of its simplicity.

The model is based on subdividing the free volume in the bed into a number of equal-volume elements that are formed between the particles (Fig. 1). Each hollow element is assumed to be cubical in shape and surrounded by inert or active particles. Perfectly mixed fluid enters at the bottom of the cube and the stream exits from the top where it is immediately mixed. The remaining four sides of the cube are formed by the surfaces of neighbouring particles. The size of the cube is a function of particle diameter and bed porosity. In the event that the packed bed is composed only of active particles all four surfaces of the cubical element would be active surfaces from which mass transfer to the fluid can take place. In the case of a dilute bed in which each active particle is surrounded only by inert particles, only one of the surfaces would be an active surface. Three additional possibilities exist for the active surfaces: two adjacent active surfaces, two

opposite active surfaces, and three active surfaces.

The assumptions used in deriving the model equations are: (1) Plug flow in the hollow channel. (2) Physical properties of the fluid are constant and independent of concentration of the transferring medium. (3) Concentration of transferring medium at an active surface corresponds to saturation conditions. (4) Material is transferred in a direction normal to the flow direction only by diffusion.

Axial diffusion was neglected. Its inclusion would have complicated the model unnecessarily. It would have had no effect on the qualitative conclusions reached with the model and only a second-order effect on the numerical predictions.

Development of model equations

The volume of fluid and the surface area of particles in a unit cross-sectional area of bed of height l will be, respectively,

$$V_a = l\varepsilon \quad (1)$$

and

$$S_a = la \quad (2)$$

and a , the surface area of particles per unit volume of bed, will be

$$a = \frac{6(1-\varepsilon)}{\psi_s d_p} \quad (3)$$

If all of the cubic elements in the bed are assumed to be identical, the width of the fluid in the element can be calculated as

$$l = \frac{V_a}{\frac{1}{4}S_a} \quad (4)$$

Substituting from equations (1), (2) and (3) gives the characteristic dimension of the cubical element

$$l = \frac{2}{3} \frac{\varepsilon d_p \psi_s}{(1-\varepsilon)} \quad (5)$$

In the subsequent development it will be assumed that the particles are spherical, thus, ψ_s is unity.

The velocity, u_e , of the fluid in the element can be calculated from the superficial gas velocity, u , which is flowing at the angle, θ , from the z -coordinate by

$$u_e = \frac{u}{\varepsilon \cos \theta} \quad (6)$$

An average value for $\cos \theta$ of 0.5 will be assumed.

Formulation for one active surface and three inert surfaces

Cubical cell elements with one active surface and three inert surfaces would correspond to a dilute bed in which active particles would be surrounded only by inert particles. The cell material balance for this case with molecular diffusion in the x -direction and forced convection in the z -direction will be

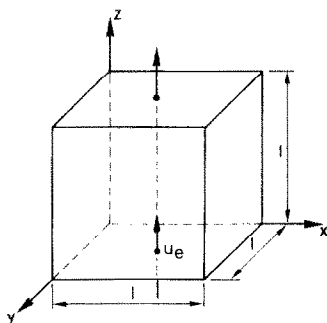


FIG. 1. Representation of model.

$$u_e \frac{\partial c}{\partial z} = D_v \frac{\partial^2 c}{\partial x^2}. \quad (7)$$

Expressing the concentration in dimensionless form as

$$\phi = \frac{c_s - c}{c_s - c_0}$$

and substituting into equation (7) yields

$$u_e \frac{\partial \phi}{\partial z} = D_v \frac{\partial^2 \phi}{\partial x^2} \quad (8)$$

the boundary conditions will be

$$\phi = 0 \quad \text{when } x = 0, \quad z \geq 0$$

$$\frac{\partial \phi}{\partial z} = 0 \quad \text{when } x = l, \quad z \geq 0$$

$$\phi = 1 \quad \text{when } z = 0, \quad x \geq 0$$

$$\phi = 0 \quad \text{when } z = \infty, \quad x \geq 0.$$

Equation (8) can be solved by separation of variables yielding

$$\phi = \frac{4}{\pi} \sum_{n=0}^{\infty} \frac{1}{2n+1} \exp \left[-\frac{(2n+1)^2 F \pi^2}{4l^2} z \right] \cdot \sin \left[\frac{(2n+1)}{2l} x \right] \quad (9)$$

where

$$F = \frac{D_v}{u_e}.$$

Upon substituting $l = z$ and integration of equation (9) between the limits of $x = 0$ and $x = l$, the concentration of transferring species in the mixed stream leaving the cell can be calculated. This cup-mixing concentration, which would be the concentration measured by an analytical device, will be

$$c_1 = c_s - 0.81(c_s - c_0) \sum_{n=0}^{\infty} \frac{1}{(2n+1)^2} \times \exp \left[-\frac{1.85(2n+1)^2}{N_{Re} \cdot N_{Sc}} \right] \quad (10)$$

where N_{Re} is a modified Reynolds number defined as $d_p u \rho / [u(1 - \varepsilon)]$, as proposed by Gamson, Thodos and Hougen [1].

The instantaneous rate of mass transfer per unit area will be

$$N_A = D_v \left(\frac{\partial c}{\partial x} \right)_{x=0} = \frac{2D_v}{l} (c_s - c_0) \sum_{n=0}^{\infty} \times \exp \left[-\frac{F \pi^2 (2n+1)^2}{4l^2} z \right] \quad (11)$$

and the rate of mass transfer from the active surface upon integration of equation (11) between the limits of $z = 0$ to $z = l$ becomes

$$\bar{N}_A = \frac{8D_v l (c_s - c_0)}{F \pi^2} \sum_{n=0}^{\infty} \frac{1}{(2n+1)^2}$$

$$\times \left\{ 1 - \exp \left[\frac{F \pi^2 (2n+1)^2}{4l} \right] \right\}. \quad (12)$$

The average mass transfer coefficient defined as

$$\bar{k} = \frac{\bar{N}_A}{A \Delta C_{lm}} \quad (13)$$

will, according to this model, be

$$\bar{k} = \frac{8D_v}{\pi^2 F} \frac{c_s - c_0}{\Delta C_{lm}} \sum_{n=0}^{\infty} \frac{1}{(2n+1)^2} \left\{ 1 - \exp \left[-\frac{F^2 (2n+1)^2}{4l} \right] \right\}. \quad (14)$$

The j_d factor for mass transfer, $\bar{k} N_{Sc}^{2/3} / u_e$, becomes

$$j_d = 0.81 N_{Sc}^{2/3} \frac{c_s - c_0}{\Delta C_{lm}} \Sigma_{k1} \quad (15)$$

where Σ_{k1} is the summation term in equation (14). The term involving concentrations in equation (15) can be evaluated with the aid of equation (10) yielding

$$\frac{c_s - c_0}{\Delta C_{lm}} = \frac{\ln 0.81 \Sigma_{c1}}{0.81 \Sigma_{c1} - 1} \quad (16)$$

where Σ_{c1} is the summation term in equation (10). Thus, for the case of one active surface and three inert surfaces and introducing N_{Re} and N_{Sc}

$$j_d = 0.81 N_{Sc}^{2/3} \frac{\ln 0.81 \Sigma_{c1}}{0.81 \Sigma_{c1} - 1} \sum_{n=0}^{\infty} \frac{1}{(2n+1)^2} \left\{ 1 - \exp \left[-\frac{1.85(2n+1)^2}{N_{Re} N_{Sc}} \right] \right\}. \quad (17)$$

Formulation for other configurations

Expressions for the j_d factor for the four remaining configurations for the cell model were derived in a similar manner. The complete derivations are available elsewhere [8]. The final equations are presented below:

Four active surfaces. This model would correspond to the case of a packed bed of active particles only. The expression for the j_d factor is

$$j_d = 0.0823 \frac{\ln 0.657(\Sigma\Sigma)_4}{0.657(\Sigma\Sigma)_4 - 1} N_{Sc}^{2/3} \sum_{n=0}^{\infty} \sum_{m=0}^{\infty} \frac{1 - \exp \left\{ -\frac{7.4[(2n+1)^2 + (2m+1)^2]}{N_{Re} N_{Sc}} \right\}}{(2m+1)^2 [(2n+1)^2 + (2m+1)^2]} \quad (18)$$

where

$$(\Sigma\Sigma)_4 = \sum_{n=0}^{\infty} \sum_{m=0}^{\infty} \frac{1}{(2n+1)^2 (2m+1)^2} \times \exp \left\{ -\frac{7.4[(2n+1)^2 + (2m+1)^2]}{N_{Re} N_{Sc}} \right\}. \quad (18a)$$

Three active surfaces, one inert surface. For this case the j_d factor is

$$j_d = 0.436 \frac{\ln 0.657(\Sigma\Sigma)_3}{0.657(\Sigma\Sigma)_3 - 1} N_{Sc}^{2/3} \sum_{n=0}^{\infty} \sum_{m=0}^{\infty} \frac{1 - \exp\left\{-\frac{1.85[4(2n+1)^2 + (2m+1)^2]}{N_{Re} N_{Sc}}\right\}}{(2m+1)^2[4(2n+1)^2 + (2m+1)^2]} \quad (19)$$

where

$$(\Sigma\Sigma)_3 = \sum_{n=0}^{\infty} \sum_{m=0}^{\infty} \frac{1}{(2n+1)^2(2m+1)^2} \times \exp\left\{-\frac{1.85[4(2n+1)^2 + (2m+1)^2]}{N_{Re} N_{Sc}}\right\} \quad (19a)$$

Two adjacent active surfaces, two inert surfaces. The j_d factor is

$$j_d = 0.329 \frac{\ln 0.657(\Sigma\Sigma)_{2a}}{0.657(\Sigma\Sigma)_{2a} - 1} N_{Sc}^{2/3} \sum_{n=0}^{\infty} \sum_{m=0}^{\infty} \frac{1 - \exp\left\{-\frac{1.85[(2n+1)^2 + (2m+1)^2]}{N_{Re} N_{Sc}}\right\}}{(2m+1)^2[(2n+1)^2 + (2m+1)^2]} \quad (20)$$

where

$$(\Sigma\Sigma)_{2a} = \sum_{n=0}^{\infty} \sum_{m=0}^{\infty} \frac{1}{(2n+1)^2(2m+1)^2} \times \exp\left\{-\frac{1.85[(2n+1)^2 + (2m+1)^2]}{N_{Re} N_{Sc}}\right\} \quad (20a)$$

Two opposite active surfaces, two inert surfaces. The j_d factor for this configuration is

$$j_d = 0.204 \frac{\ln 0.81 \Sigma_{2b}}{0.81 \Sigma_{2b} - 1} N_{Sc}^{2/3} \sum_{n=0}^{\infty} \frac{1}{(2n+1)^2} \left\{ 1 - \exp\left[-\frac{7.4(2n+1)^2}{N_{Re} N_{Sc}}\right] \right\} \quad (21)$$

where

$$\Sigma_{2b} = 0.81 \sum_{n=0}^{\infty} \frac{1}{(2n+1)^2} \exp\left[-\frac{7.4(2n+1)^2}{N_{Re} N_{Sc}}\right] \quad (21a)$$

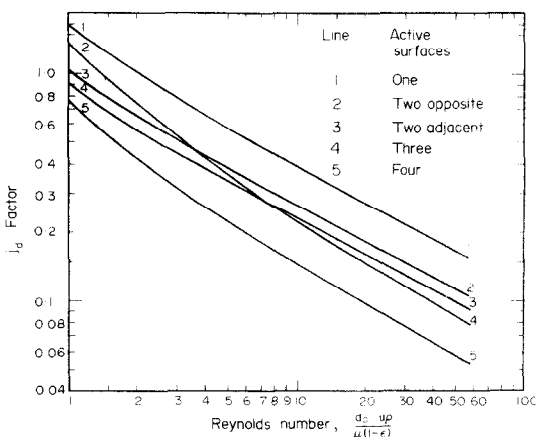


FIG. 2. j_d factor as a function of Reynolds number-model behaviour for naphthalene-air ($N_{Sc} = 2.53$).

MODEL BEHAVIOUR

The model equations developed above were evaluated for the naphthalene-air ($N_{Sc} = 2.53$), naphthalene-helium ($N_{Sc} = 5.45$) and naphthalene- CO_2 ($N_{Sc} = 1.52$) systems over the Reynolds number range of 1-60. A value of 0.40 was assumed for the bed porosity. The results are presented graphically in Figs. 2 and 3.

In Fig. 2 the j_d factor is presented as a function of the Reynolds number for the naphthalene-air system for the different cell configurations. It is apparent that j_d is a function not only of N_{Re} but also of the mutual geometric or spatial arrangement between the active and inert particles. The highest values for the j_d factor are obtained for the case of a cell with one active and three inert surfaces, i.e. a very dilute bed, and the lowest values are obtained for the cell with four active surfaces, i.e. a packed bed consisting only of active particles. Thus, if a dilute bed is to be used in experimental work at low Reynolds numbers so as to avoid bed exit concentrations approaching saturation, the model shows that the experimentally-measured transfer coefficients should be a function not only of the flow regime but also of the geometric arrangement of the active particles in the bed. This effect was, indeed, noted by Petrovic and Thodos [3] who made mass-transfer measurements in packed beds of active particles as well as runs in which the active particles were dispersed in a matrix of inert particles. They reported that the dilute beds yielded j_d values that were always 15-25% higher than the values obtained in packed beds of active material. They noted, correctly, that the differences can vary for each bed arrangement and attributed the effect to a contribution other than mass transfer. Their reported results are consistent with the model predictions cited above.

In Fig. 3, in which the j_d factor is presented as a function of the Reynolds number for three values of the Schmidt number, it can be seen that the model predicts

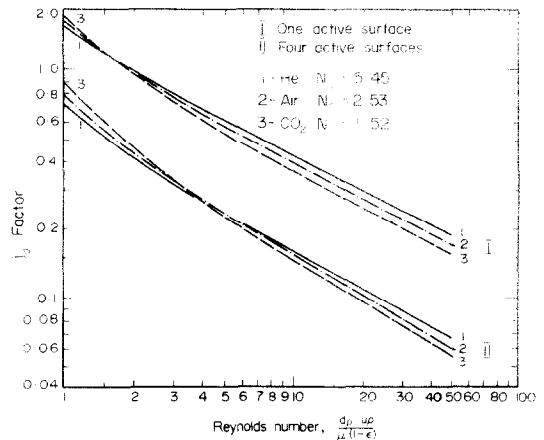


FIG. 3. j_d factor as a function of Reynolds number-model behaviour for several values of Schmidt number.

a small effect of the Schmidt number additional to that accounted for in the definition of the j_d factor. Again, the effect of the geometric configuration of the active particles is apparent.

An experimental programme was designed to test if the results predicted by the model would be borne out by experimental measurements. To test this, mass transfer measurements were made in beds in which different known configurations of active particles in a matrix of inert particles could be effected.

EXPERIMENTAL EQUIPMENT AND PROCEDURE

The mass-transfer measurements were mainly for the naphthalene-air system although some measurements were made for the naphthalene-helium and naphthalene-carbon dioxide systems. Particle sizes used covered the range of 0.0171–0.9 cm. Two 'reactors' were used with a 92 mm dia. glass reactor being used with the larger particles and a 42 mm dia. glass reactor for the smaller particles. The reactor and its associated equipment were installed in a constant-temperature air-bath whose temperature was maintained at $33 \pm 0.1^\circ\text{C}$.

Naphthalene concentration in the gas exiting from the bed was measured by passing part of the stream through the flame ionization detector of a gas chromatograph. The electrometer output of the FID is proportional to the concentration of organic material in the gas and was calibrated with air saturated with naphthalene. Detector linearity was checked by measuring electrometer output for gas streams containing known ratios of clean air to naphthalene-saturated air. This analytical technique permitted accurate and continuous measurement of the naphthalene concentration in the gas stream leaving the reactor.

The large particles, diameter greater than 0.3 cm, were steel balls. Active particles were produced by coating the balls with a thin layer of naphthalene by immersion in molten naphthalene and uncoated balls served as the inert particles. The area of the active surface was calculated from the number of active particles inserted in the bed and the ball diameter as measured with a micrometer. The total volume of inert plus active balls used in the reactor was 500 cm^3 .

The small naphthalene particles were made by the same technique used by Bar-Ilan and Resnick [5]. They were air-dried and screened into several size groups. Particle size was determined from measurements on photomicrographs of the various fractions. Glass beads of appropriate size were used as the inert material with the active naphthalene granules. The total volume of inert plus active particles used in the reactor was 100 cm^3 and the active surface area was calculated from the weight of active particles (less than 1 g).

Experimental runs were of short duration and were terminated before any bare surface appeared in the case of the large coated spheres or before there was any appreciable reduction in active area in the case of the

small particles.

For each run the bed was made up as a mixture of inert and active particles, appropriate connections made to the reactor in the constant-temperature bath and the run begun after temperature equilibrium was reached. Dried laboratory air was used for the naphthalene-air runs and helium and carbon dioxide from cylinders for the helium and carbon dioxide runs. Gas flow was started and the flow rate set after the FID flame was ignited, a sidestream of reactor exit gas was fed to the detector and the electrometer output was recorded after the output had stabilized. The gas-flow rate to the reactor was then changed and the cycle repeated. A total of 369 mass-transfer rate measurements were made in 49 runs with each run comprising from 4 to 9 different gas flow rates. Complete details of the experimental equipment and procedure are presented by Golt [8].

EXPERIMENTAL RESULTS

The experimental results obtained with the large particles, $0.3 \leq d_p \leq 0.9\text{ cm}$, are presented graphically in Figs. 4–6 as j_d factor vs Reynolds number. The results presented in Fig. 4 are for a dilute bed with the active particles arranged in a random manner. From 40–100 active particles were present in a total volume of 500 cm^3 of inert plus active particles in the bed. The geometric arrangement between active and inert particles was different for each of the 12 runs presented in Fig. 4 but in all runs each active particle would 'see' only inert particles. The correlating line obtained by Bar-Ilan and Resnick [5] in their work with pellets of naphthalene dispersed in pellets of inert material is presented for comparison. Also shown in Fig. 4 is the model prediction for the case of a cell element with one active and three inert surfaces, equation (17). In spite of the simplicity of the model, the predicted results are surprisingly close to the experimentally-obtained transfer coefficients.

The results obtained for a number of different arrangements between active and inert large particles are presented in Fig. 5. The upper line presents the

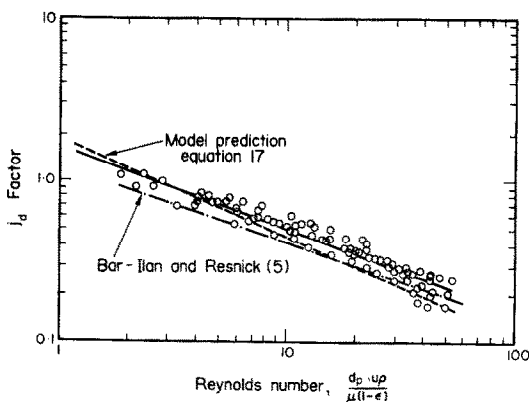


FIG. 4. Experimental results for j_d factor for random arrangement of large particles.

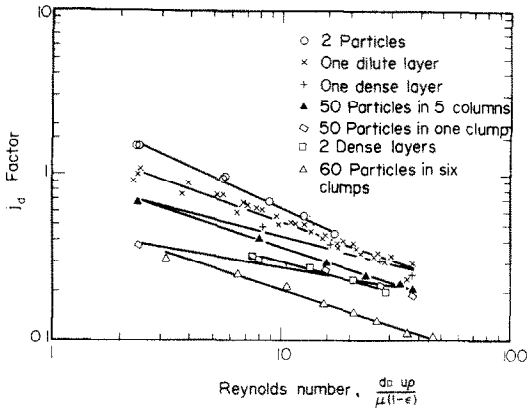


FIG. 5. Experimental results for j_d factor for single dilute layer of large particles.

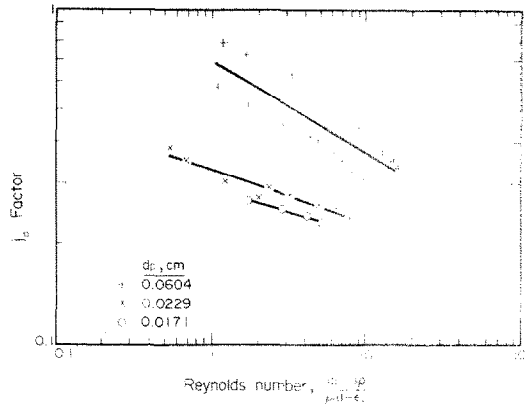


FIG. 7. Experimental results for j_d factor for small particles in single layer.

experimentally-obtained value for the j_d factor when only two active large spheres were present in the entire bed. The next line is the result obtained when the active particles were arranged as one horizontal layer, one particle deep, in dilute packing so that active particles were not adjacent to active particles. The next line down is for the same number of active particles but arranged in one densely packed layer, i.e. active particle adjacent to active particle, which did not cover the entire bed cross-section. The additional arrangements studied experimentally were: two layers of densely packed active particles; 50 active particles arranged in five columns of 10 particles each; 50 active particles in one clump; 60 particles arranged in six clumps of 10 particles each. It is apparent from Fig. 5 that the experimentally-determined j_d factor decreases as the geometric arrangement proceeds from less dense to more dense packing of the active spheres, as is predicted by the model.

The effect of the Schmidt number on the j_d factor is presented in Fig. 6. All runs were for a single, dilute layer of active spheres in the bed. The progression is as predicted by the model with naphthalene-helium

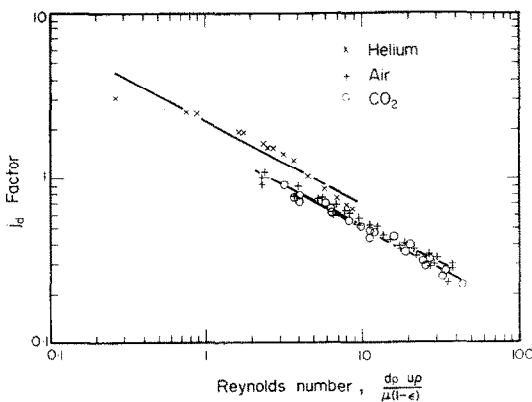


FIG. 6. Experimental results for j_d factor for single dilute layer of large particles with different values of Schmidt number.

yielding the highest values for the j_d factor and naphthalene-CO₂ the lowest. The model also predicts a cross-over and a small extrapolation of the CO₂ correlating line would lead to a cross-over with the air line at a Reynolds number of ~ 2 .

Some of the small particles results are presented in Figs. 7 and 8. In Fig. 7, j_d factor for the case of a very shallow single layer of active particles are presented and a strong particle diameter effect could be inferred. A closer examination of the experimental results, however, would lead one to question this inference. The data shown for the two smaller particles (0.0171 and 0.0229 cm) were obtained from one run for each of the particles, i.e. the bed was not disturbed or remade as the air velocity was changed to change the Reynolds number. The data points for the largest particle shown (0.0604 cm) were obtained from two different runs with the data for one run being well above the least-squares correlating line and the data for the other run being well below the correlating line. Each set of data could have been well-correlated by a separate line. It is apparent that the j_d factor as measured was dependent upon the actual intra- and interparticle arrangements that are obtained when single layer of active particles is obtained by pouring fine naphthalene particles onto a bed of inert glass beads. Replicate arrangements would be almost impossible to obtain when fine particles are used.

The same effect is shown in Fig. 8 which is for small particles arranged in three shallow layers. Three pairs of lines are shown with each pair representing two separate runs made with the same particle diameter. As can be seen the data for each run can be nicely correlated but different runs using the same particle diameter require different correlating lines. This, again, is the result of the impossibility of producing replicate geometric arrangements between active and inert particles of small size. It can also be pointed out that the four runs shown in Figs. 7 and 8 for active particle diameter of 0.0604 cm produce four different correlating lines. These results again confirm the model

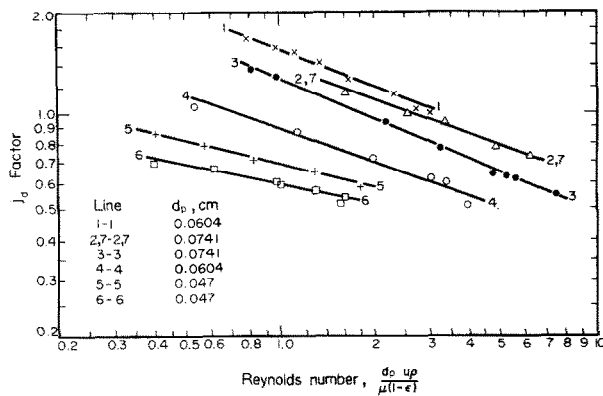


FIG. 8. Experimental results for j_d factor for small particles in three layers.

predictions that the measured values for the j_d factors will be a function, among others, of the mutual spatial arrangements that exist between active and inert particles.

SUMMARY AND CONCLUSIONS

A simple mathematical model predicted that gas-solid mass-transfer rates obtained by measurements in a packed bed of active particles dispersed in a matrix of inert particles would be a function not only of the Reynolds and Schmidt numbers but also of the mutual spatial arrangement between inert and active particles. Experimental measurements with large particles which permitted controlled geometric arrangements between the active and inert particles supported the model predictions.

The experimental work on small particles indicated that discrepancies among the reported results of a number of investigators were probably the result of differences in the geometric arrangements between the inert and active particles and were not the result of particle diameter effects additional to those accounted for in the Reynolds number.

An additional conclusion is that it is an exercise in futility to attempt to obtain valid mass-transfer coefficients by measurements in dilute beds of active material dispersed in a matrix of inert material when the calculated coefficients are based on driving forces at the bed inlet and exit. Although in-bed driving force measurements as performed by Hsiung and Thodos [7] provide an improved experimental technique, their

calculated coefficients are based on cup-mixing concentrations and, hence, cannot result in true mass-transfer coefficients in the case of low Reynolds numbers.

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MESURE DU TRANSFERT MASSIQUE ENTRE UNE PARTICULE
ET UN GAZ ET COEFFICIENTS DANS DES LITS FIXES
AUX FAIBLES NOMBRES DE REYNOLDS

Résumé—Un modèle pour calculer les coefficients de transfert massique dans les lits fixes à des nombres de Reynolds faibles montre que les coefficients déduits des mesures pour des particules inertes et actives peuvent dépendre des arrangements géométriques entre les particules inertes et actives. Des résultats expérimentaux avec des particules de taille comprise entre 0,0171 et 0,9 cm soutiennent le modèle. Des résultats contradictoires obtenus par des chercheurs, concernant l'effet du diamètre de particule sur le facteur j_d , sont expliqués par ces résultats qui indiquent aussi que des mesures expérimentales dispersées et des calculs à partir des concentrations de mélange ne peuvent conduire à des coefficients corrects de transfert massique.

FESTSTOFF/GAS-STOFFTRANSPORTMESSUNGEN UND -KOEFFIZIENTEN
IN HAUFWERKEN BEI KLEINEN REYNOLDS-ZAHLEN

Zusammenfassung — Ein zur Berechnung der Stofftransportkoeffizienten in Haufwerken entwickeltes Modell bei kleinen Reynolds-Zahlen zeigte, daß die aus Messungen gewonnenen Koeffizienten in Festbetten von der geometrischen Anordnung der inerten und aktiven Teilchen abhängig sind.

Versuchsergebnisse mit Teilchen der Größenordnung 0,0171 bis 0,9 cm bestätigten die Aussagen des Modells.

Widersprüchliche Ergebnisse über den Einfluß des Teilchendurchmessers auf den j_d -Faktor, die von mehreren Forschern erhalten wurden, konnten durch diese Erkenntnisse erklärt werden, die auch darauf hinweisen, daß Versuchsergebnisse an dispersen Feststoffbetten und Berechnungen, die "cup-mixing"-Konzentrationen enthalten, nicht zu gültigen Stoffübergangskoeffizienten führen können.

ИССЛЕДОВАНИЕ МАССООБМЕНА МЕЖДУ ЧАСТИЦАМИ И ГАЗОМ И
ОПРЕДЕЛЕНИЕ КОЭФФИЦИЕНТОВ МАССОПЕРЕНОСА В НЕПОДВИЖНЫХ СЛОЯХ
ПРИ МАЛОМ ЧИСЛЕ РЕЙНОЛЬДСА

Аннотация — С помощью модели, разработанной для расчета коэффициентов массопереноса в плотных слоях при малом числе Рейнольдса, показано, что значения коэффициентов в слоях инертных и активных частиц должны зависеть от их взаимного геометрического расположения. Полученные экспериментальные данные для частиц размером 0,0171–0,9 см подтвердили результаты расчетов. Предложенная модель позволила объяснить противоречивость результатов, полученных различными авторами при исследовании влияния диаметра частиц на коэффициент j_d . Показано также, что при экспериментальном и теоретическом исследовании дисперсных слоев методом объемных концентраций нельзя получить достоверных значений коэффициентов массопереноса.