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A Dynamic Method for Liquid-Particle Mass Transfer in Trickle Beds

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Previous investigations of liquid-to-particle mass transfer in trickle beds have used non-porous particles. However, the applications of trickle beds have been for reactions or adsorption employing porous catalysts as adsorbents. In this study a method of dynamic adsorption with porous particles is developed. The procedure is to extrapolate the response to a step-function input of non-volatile adsorbate to a short time where intraparticle diffusion and adsorption do not affect the overall process.

The method is applied to the adsorption of benzaldehyde from aqueous solutions in a bed of granular, activated carbon particles. Measurements were made for three particle sizes, in adsorbers of two diameters, at 298K and 1 atm. The gas flow rate had no discernible effect on the mass transfer coefficients, $k_{Ls}a_{Ls}$, over a superficial velocity range of 1.47 to 8.0×10^{-2} m/s, but $k_{Ls}a_{Ls}$ increased with liquid rate. The results, correlated as Sherwood vs. Reynolds number, agree well with the non-porous particle data of Van Krevelen and Krekels (1948), but suggest larger Sherwood numbers than similar data from later investigations.

The wide range of mass transfer results for either liquid-to-particle or liquid-to-gas in trickle beds seems to be due in large measure to the difficulty in reproducing the rivulent flow pattern of liquid from bed to bed.

SCOPE

Van Krevelen and Krekels (1948) used the dissolution technique (benzoic acid particles dissolving in water and glycol solutions) in one of the first studies of mass transfer in trickle beds. They measured liquid-particle transfer coefficients in downflow of liquid over the particles; a gas phase was present but there was no gas flow. Since then all published data have been obtained by the same dissolution method using various kinds of non-porous particles. The dissolution technique has several advantages, but the time required to reach steady state (Satterfield et al., 1978) and particle shrinkage may affect the results. However, the most important reason for seeking another method is that most applications of trickle beds employ porous catalyst particles.

A critical requirement of a method for porous, catalyst-type particles is that a minimum of rate parameters be involved. To eliminate rate constants for catalytic reaction (and associated problems, such as variable catalyst activity) an adsorption process is desirable. However, the process now becomes a dynamic one. If the adsorption is of the rapid, physical type, equilibrium may be assumed at the adsorption site within the particle. A non-volatile adsorbate is necessary if mass transfer from liquid to gas is to be avoided. A system that meets these and other requirements reasonably well is the adsorption of benzaldehyde (vapor pressure at 298K = 127 Pa) from aqueous solutions in a bed of activated carbon particles. In this system intraparticle diffusion in the liquid-filled pores, as well as the desired mass transfer process from liquid to particle, may affect the concentration of benzaldehyde in the effluent liquid. However, intraparticle diffusion can be eliminated from dynamic adsorption results by extrapolating the data to essentially zero time.

This paper reports mass transfer coefficients measured in this way at 298K and atmospheric pressure. Results were obtained for gas and liquid flow rates in the gas-continuous flow regime.

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CONCLUSIONS AND SIGNIFICANCE

Liquid-particle mass transfer rates, $k_{Ls}a_{Ls}$, were measured for three particle sizes ($d_p = 1.34 \times 10^{-3}$, 2.4×10^{-3}, and 4.06×10^{-3} m) and for two tube diameters (1.57 and 5.64×10^{-2} m), at 298K and atmospheric pressure for activated carbon particles. These results, obtained by a simple, dynamic method, were reproducible to about 7%. While $k_{Ls}a_{Ls}$ increased with liquid flow rate, it did not change with the co-current, downflow gas velocity which was varied from 1.47 to 8.0×10^{-2} m/s (superficial values).

To our knowledge these data are the only published, trickle-bed, liquid-to-particle rates for porous, catalyst-type particles. Other results are based upon dissolution data for non-porous particles such as benzoic acid and β-naphthol. For the same particle sizes, and corrected approximately for the effect of porosity on the mass transfer area, our results for k_{Ls} agree with those of Van Krevelen and Krekels (1948), but are greater than those of Goto and Smith (1975). Speccia et al. (1976) and Satterfield et al. (1978).

The deviations between the several sets of data are believed to be caused by variations in liquid flow around and in between the particles. In our work the liquid distributor and tube-to-particle diameter ratio were such that the liquid flow rate should have been uniform across the tube diameter, according to the criterion of Herskowitz and Smith (1978). Perhaps of more importance than completely uniform liquid distribution are the uncontrollable variations in liquid flow in trickle beds. Visual observation shows rivulets of liquid flowing over parts of the outer surface of the particles and stationary regions of liquid at points of contact between particles. The rivulets and stagnant regious will influence $k_{Ls}a_{Ls}$ and yet they vary from bed to bed in an uncontrollable way. These effects, could explain the large variation in mass transfer rates that seems to be characteristic of trickle-bed operation. The deviations mentioned here and observed in various studies in our laboratory (Morita and Smith, 1978; Herskowitz et al., 1979; Mata, 1980) lead to this conclusion and illustrate the difficult problem in reproducing liquid flow patterns in co-current gas and liquid flow in packed beds.

At liquid flow rates used in trickle-bed reactors liquid-particle mass transfer can have a significant effect on the global reaction rate. Following the initial study by Van Krevelen and Krekels (1948), there have been several investigations of liquid-particle mass transfer (Sato et al., 1972; Hirose et al., 1974; Lemay et al., 1975; Goto and Smith, 1975; Sylvester and Pitayagulsarn, 1975; Specchia et al., 1976; Hirose, 1976; Satterfield et al., 1978; Reuther et al., 1980). All these results are based upon dissolution data for non-porous particles and are usually, but not always, presented as correlations of Sherwood number as a function of Reynolds number, or Chilton-Colburn j factors, and Schmidt number. The data and correlations refer primarily to the gas-continuous or trickling-flow regime. Also some results are for the higher flow rates corresponding to pulsing flow.

Our objective was to obtain data for porous particles. In order to eliminate additional rate parameters and problems with catalyst activity, an adsorption process with rapid (equilibrium) physical adsorption was indicated. Further, if a non-volatile adsorbate is used, the complications of gas-liquid mass transfer and transport between gas and "gas-covered" surface of the particles could be eliminated. The system chosen to meet these requirements was the adsorption of benzaldehyde from aqueous solutions on activated carbon particles.

With adsorption, the process became a dynamic one so that the concentration of benzaldehyde in the liquid effluent from the bed varied with time. In addition to liquid-particle mass transfer, the overall process could be affected by intraparticle diffusion in the liquid-filled pores. The effective diffusivity of benzaldehyde in liquid-filled pores is expected to be two or three orders of magnitude less than its diffusivity in a gas phase. Hence, it was necessary for accurate results to use a method of analyzing the dynamic data such that the effect of intraparticle diffusion was eliminated. Then only one rate parameter $k_{Ls}a_{Ls}$ would be involved. An additional complication was the mixing in the unavoidable volume of liquid, downstream from the bed, necessary to separate the gas and liquid effluent streams. The procedure developed for treating the dynamic data (breakthrough curves) with these complications is described in later sections.

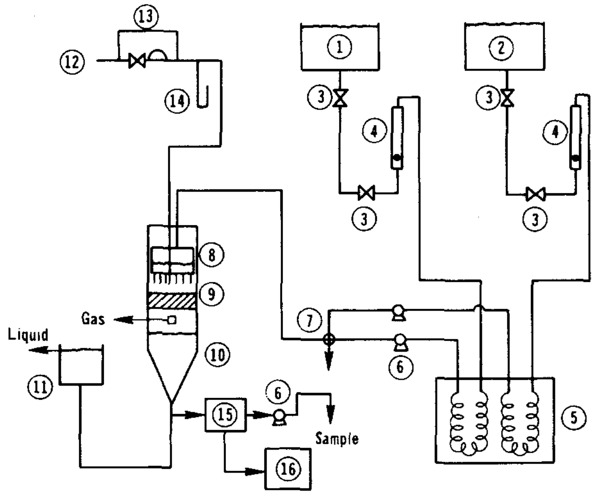
EXPERIMENTAL

The apparatus is shown schematically in Figure 1. Two Pyrex adsorbers were used, 1.57 and 5.64×10^{-2} m in inside diameter. Granular type BPL activated carbon particles of three diameters 1.34×10^{-3} m (10 to 16 mesh), 2.4×10^{-3} m (7 to 9 mesh) and 4.06×10^{-3} m (4 to 6 mesh) were used to prepare the beds. Most of the data were obtained at bed depths from 2 to 3.3×10^{-2} m. The manufacturer quoted a surface area of 1.1×10^6 m^2/kg (by N_2 absorption), particle density of 0.89×10^3 kg/m^3 and porosity of 0.62 for these particles. After sieving, the carbon particles were washed five times in deionized water to remove dust and fine particles. Then they were boiled in water for 40 minutes and dried to constant weight at 393K.

At $t = 0$, the liquid flow was changed with a four-way valve from deionized water to an aqueous solution (containing 7 to 10 ppm of benzaldehyde). The liquid pump was located upstream from this valve in order to minimize mixing and reduce retention time between the valve and the top of the bed. Helium gas and the liquid were introduced to the bed through movable Plexi-glass distributors (Figure 2A). For the smaller adsorber, the distributor contained twelve stainless steel capillary tubes (10^{-3} m ID and 5×10^{-3} m long) located across the area in the bottom plate of the distributor. The larger distributor contained 42 capillary tubes (1.6×10^{-3} m ID, 7×10^{-3} m long). The distance between the top of the bed and the bottom of the capillary tubes was about 7×10^{-3} m. This arrangement ensured (Herskowitz and Smith, 1978) rather uniform distribution of liquid over the cross-sectional area of the bed. The gas was introduced through a 6.3×10^{-3} m ID tube located in the center of the distributor plate.

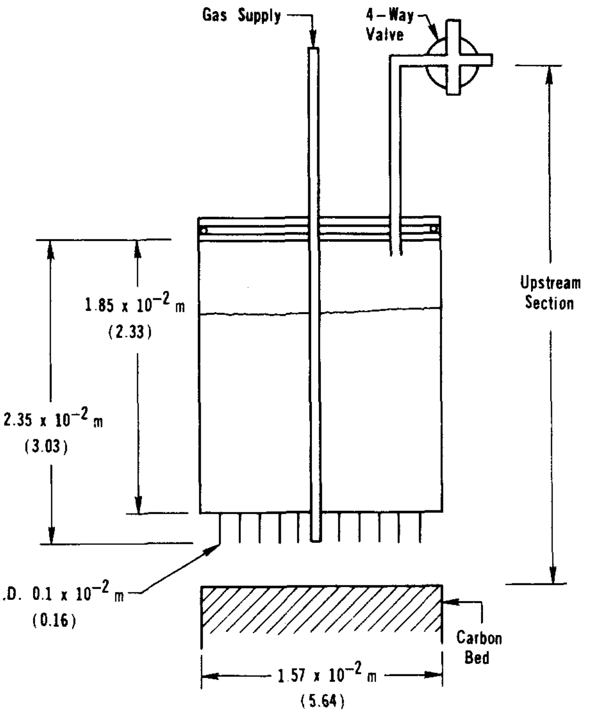
The space below the screen holding the carbon particles in the bed served as a separator for the effluent gas and liquid, as shown in Figure 2B. The length of the separators was about 6×10^{-2} m for the small adsorber and 16.9×10^{-2} m for the large adsorber. The gas was separated from the liquid above the liquid level through a 3.15×10^{-3} m tube shielded from liquid with a cap.

The liquid level in the separator was maintained as constant as possible during a run by using an auxiliary vessel. This level varied from 0.5×10^{-2}



1. Benzaldehyde Solution Feed 2. Water Feed 3. Valve 4. Rotameter 5. Constant Temperature Bath 6. Pump 7. 4-Way Valve 8. Liquid Distributor 9. Carbon Bed 10. Separator 11. Control Vessel 12. Helium Cylinder 13. Flow Regulator 14. Manometer 15. UV Photometer 16. Recorder

Figure 1. Schematic diagram of apparatus.



* Dimension in Bracket is For the Large Reactor

Figure 2A. Diagram of uniform-liquid distributor for small and large diameter reactors.

to 3×10^{-2} m in the small adsorber and from 0.7×10^{-2} to 6.5×10^{-2} m in the large adsorber, depending upon the liquid rate. The levels were low in order to minimize the corrections necessary for retention time and mixing.

The liquid effluent was analyzed continuously for benzaldehyde by withdrawing a sample stream at a rate of 3.33×10^{-8} m^3/s from the effluent line from the separator. The sample stream flowed through an ultraviolet photometer with a radiation source of 2570A°. Benzaldehyde has a high absorbance at this wavelength so that accurate measurements could be made at concentrations as low as 1.0 ppm. The retention time in the photometer was negligible (~0.1s) since the cell volume was but 2×10^{-9} m^3.

To prepare the bed, a portion of the dried particles was weighed and then boiled in deionized water for 30 minutes to fill the pores with liquid. The wet particles were added in portions to the adsorber tube, vibrating the tube after each addition.

Data were obtained in the small adsorber for liquid superficial velocities from 0.24×10^{-2} to 2.41×10^{-2} m/s and from 0.062×10^{-2} to 0.79×10^{-2} m/s in the large adsorber. The range of superficial gas velocities was 1.47×10^{-2} to 8.0×10^{-2} m/s for both adsorbers. In order to maintain a tube-to-particle diameter ratio of 12 or more, only the smallest particles ($d_p = 0.134 \times 10^{-2}$ m) were used in the small adsorber.

The volume of the liquid in the apparatus may be divided into four sections: 1) the upstream section from the 4-way valve to the top of the bed, 2) the bed of carbon particles, 3) from the bottom of the bed to the sample line connection in the effluent line (the separation volume), and 4) the sample line from the effluent line to the photometer. Table 1 shows the estimated volumes and range of retention or residence times of the liquid in each of the four sections. The τ for the bed was estimated by subtracting from the total retention time the values for the other three sections. For comparison with these residence times, and to evaluate dispersion, blank runs were made without carbon particles. For such runs the liquid distributor was lowered to a location 7×10^{-3} m above the screen, to correspond to the distance from the bottom of the distributor to the top of the bed in the adsorption runs.

TABLE 1. LIQUID VOLUMES AND MEAN RESIDENCE TIMES

Volume Section (Figure 1)	Adsorber Diameter m × 10^2	Estimated Volume m^3 × 10^6	Mean Residence Time* s
Upstream	1.57	1.5	0.3 to 3.2
Carbon Bed	1.57	1.0 to 6.5	0.5 to 3.0 (= τ)
Separator	1.57	1.0 to 5.0	1.0 to 3.0
Sample Line	1.57	0.01	0.3
Upstream	5.64	2.5	0.13 to 1.1
Carbon bed	5.64	10. to 80.	0.7 to 8.0
Separator	5.64	2.0 to 35.	0.8 to 3.0
Sample Line	5.64	0.01	0.3

* Varies with liquid flow rate (except in the sample line), but a specific value is known for each run.

A blank run was made immediately prior to and at the same flow rates as the subsequent adsorption run. The steady-state concentration, attained quickly, established the step concentration C_f of the feed stream for the adsorption run. A typical set of blank and adsorption runs are shown in Figure 3. The blank run (as well as the adsorption run) shows considerable deviation from a plug-flow response, indicating that dispersion occurs in the liquid volumes. To help identify the location of this dispersion, a second type of blank run was made. The upstream section was filled with benzaldehyde solution and the response curve measured. This second blank run includes all the liquid volumes except that in the upstream section. With all other conditions the same, there was no noticeable difference between the two blank runs. Hence, dispersion in the upstream section was negligible. Separate measurements in connection with a benzaldehyde-calibration curve, with the same diameter and length of lines as in the sample-line section, also did not exhibit measureable dispersion. This meant that the observed dispersion in the first type of blank run was occurring in the two remaining sections, the bed itself and the separator. Wakao and Funazkri (1978) have shown that the experimental liquid-particle mass transfer data in packed beds are not significantly affected by axial dispersion at Reynolds numbers greater than 3. In our experiments Re ranged from 0.88 to 34.5, with only three runs at (Re) < 3. To test the effect of axial dispersion in our work, a few runs were made for a bed depth of about 0.4×10^{-2} m. The mass transfer rate, $k_{Ls}a_{Ls}$, for this short bed was 8.7% less than the value for a bed depth of 3.1 cm. This difference is within the accuracy of the data. As a further test, runs were made with carbon particles replaced with a bed of glass beads. With glass beads, no mass transfer to the particles would be expected. The response curves for such runs were essentially the same as the response curves for the blank runs with no bed, again suggesting that axial dispersion in the bed was not significant. In view of the bed depth and glass bead data the observed dispersion in the blank run in Figure 3 is assumed to occur entirely in the separator.

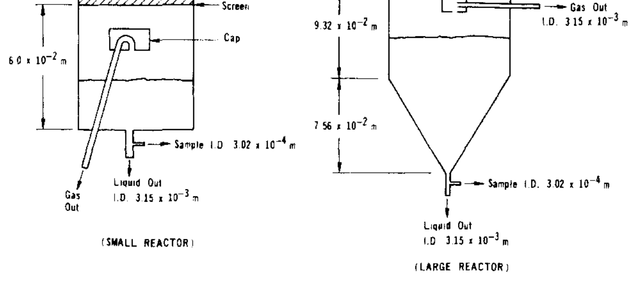


Figure 2B. Diagram of separator for small and large diameter reactors.

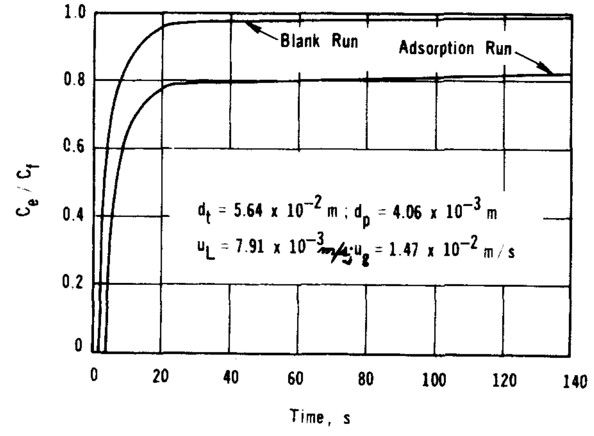


Figure 3. Typical response curves-run #81.

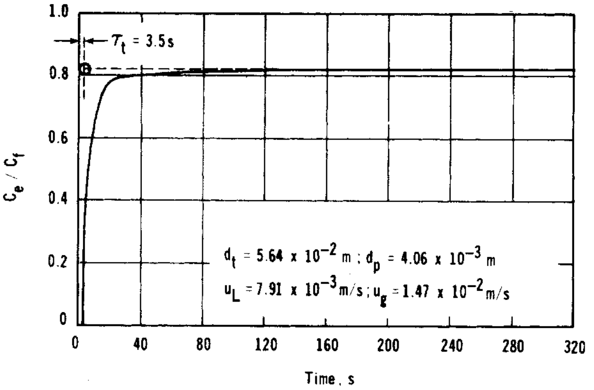


Figure 4. Extrapolation of response curve to obtain $(C_L)_{t=\tau}$ for run #81.

Mass Transfer Rates, $k_{Ls}a_{Ls}$

The differential mass balance for non-volatile benzaldehyde in the liquid phase may be written

$$\epsilon_L \frac{dC}{dt} + u_L \frac{dC}{dx} + k_{Ls}\, a_{Ls}\,(C - C_s) = 0 \tag{1}$$

where C_s is the concentration in the liquid at the outer surface of the particles. For the reasons mentioned earlier, axial dispersion is not included in Eq. 1. The initial and boundary conditions are:

$$C = 0 \text{ at } t = 0 \text{ for all } x \tag{2}$$

$$C = C_f \text{ at } x = 0 \text{ for all } t \tag{3}$$

In general, solution of Eqs. 1–3 for the concentration C_L in the effluent from the bed would require simultaneous solution of an intraparticle mass balance for C_s. Such a solution would introduce several problems: 1) a second unknown rate parameter, the effective intraparticle diffusivity; 2) the equilibrium adsorption isotherm for benzaldehyde on the activated carbon; and 3) the problems associated with unsymmetrical boundary conditions on the particle outer surface when part of the surface is not covered by liquid rivulets. To eliminate these problems, a method was developed to evaluate C_L when $C_s = 0$.

During the time interval 0 to τ, where τ is the residence time of the liquid in the bed, $C_s = 0$ for the particles which have yet to be in contact with the benzaldehyde solution. At the end of the bed, $C_L = 0$ for $t < \tau$, and for $t = \tau$ the value of C_L is given by the solution of Eqs. 1–3 with $C_s = 0$. This solution is the simple expression

$$(C_L)_{t\,=\,\tau} = C_f \exp\left(-\frac{k_{Ls}\, a_{Ls}}{u_L} L\right) \tag{4}$$

For a known bed depth L, superficial liquid velocity u_L and feed concentration C_f, $k_{Ls}a_{Ls}$ can be obtained from Eq. 4 provided C_L at $t = \tau$ can be determined from the measured response curve.

One method of obtaining $(C_L)_{t\,=\,\tau}$ is to assume that the dispersion model for a cylindrical tube will fit the mixing in the separator. Then the response curve for the blank run (as illustrated in Figure 3) may be employed to establish a Peclet number for the separator. This Peclet number can be used to evaluate the mixing in the separator for an adsorption run, and thus obtain $(C_L)_{t\,=\,\tau}$. The latter step is a two-point boundary value problem requiring a numerical solution. The results showed that the predicted curve of C_L vs. t oscillated (with time) at low times, and then agreed with the observed response curve, C_e vs. t at longer times. The oscillating period corresponded to the time interval where C_L changed sharply with time; for example, from τ_t to about 20 seconds for run 81 shown in Figure 4. The oscillation could be reduced but not eliminated by reducing the step size in the numerical solution, as explained by Tan (1980). Hence, this method could not be used to find $(C_L)_{t\,=\,\tau}$. However, the agreement of the predicted and observed curves after the initial period indicated that the dispersion model was able to represent the small amount of dispersion occurring after the C_e curve became nearly flat.

In this nearly flat section the change in C_e was very small during the residence time in the separator (1.0 to 3.0 seconds as shown in Table 1). Hence, the concentration change between the bottom of the bed and the photometer was so little that mixing in the separator was unimportant. Note that the source of the dispersion has been shown to be in the separator. Then the effect of mixing in the separator could be eliminated, and $(C_L)_{t\,=\,\tau}$ could be evaluated by extrapolating the nearly flat portion of the observed curve of C_e vs. t back to τ_t; τ_t is the total response time of the four sections listed in Table 1. That is, the extrapolated concentration to $t = \tau_t$ is equal to the concentration at the same instant, at the bottom of the bed when there is no axial dispersion in the separator. Figure 4 illustrates the extrapolation, which for this run was to a time $\tau_t = 3.5$ s. Since the relatively flat portion of the curve is long and nearly linear, there is little uncertainty in the extrapolation. Every response curve exhibited a nearly linear section at intermediate times. Also, the good agreement between the results for columns of different diameter, indicated in Figure 5, is evidence for the validity of the extrapolation procedure.

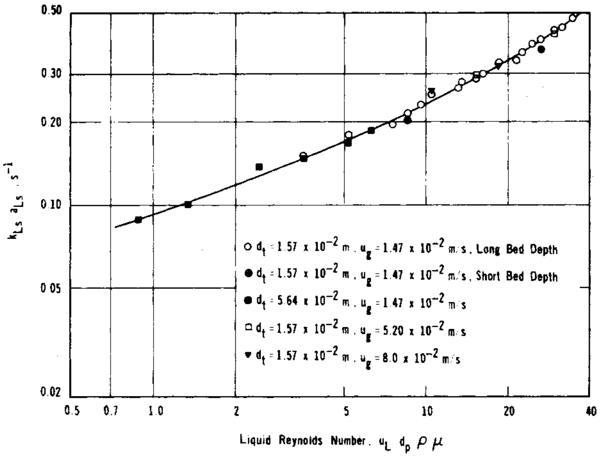


Figure 5. Experimental mass transfer rates for smaller particles (d_p = 1.34 × 10^{-3}m).

Values of $k_{Ls}a_{Ls}$ calculated from $(C_L)_{t\,=\,\tau}$ and Eq. 4 are shown in Figure 5 for the smallest particle size and the two sizes of adsorber. The agreement of the data points for the large and small adsorbers suggests that with a ratio d_t/d_p of 12 and with uniform distribution of liquid feed, wall effects were not important. Also, the points in Figure 5, which were for gas velocities ranging from 1.47 to 8.0 × 10^{-2} m/s, show that in this range gas flow rate has no discernible effect on mass transfer from liquid to particle. The marked points in Figure 5 show about a 9% difference between bed depths of 0.4 and 3.1 × 10^{-2} m. Runs made at the same conditions indicated a reproducibility of about 7%.

Figure 6 includes the data for both adsorbers and the two larger particle sizes. The results for d_p = 0.24 × 10^{-2} and 0.406 × 10^{-2} m were obtained at gas rates from 1.47 to 8.0 × 10^{-2} m/s. Reproducibility tests again indicated a deviation of about 7%. The marked points for the largest particles (d_p = 0.406 × 10^{-2} m) for a bed depth of 0.4 × 10^{-2} m was within 2% of the remainder of the data for a bed depth of about 2.4 × 10^{-2} m. As mentioned earlier, these results suggest that axial dispersion in the bed itself was negligible.

The physical properties of benzaldehyde-water solutions, used for evaluating the Reynolds number and for subsequent correlations, were obtained from Reid et al. (1977) and are given in Table 2. The mass transfer rates in Figures 5 and 6 are based upon a unit volume of empty adsorber and would normally depend upon the void fraction ϵ_B of the bed. This quantity, measured for each bed, ranged from 0.37 to 0.47. For consistency all the $k_{Ls}a_{Ls}$ values shown in Figures 5 and 6 were based upon $\epsilon_B = 0.39$ by multiplying the $k_{Ls}a_{Ls}$ from Eq. 4 by the ratio $(1 - 0.39)/(1 - \epsilon_B)$. This small correction supposes that the mass transfer area at the outer surface of the particles is porportional to the volume fraction of the particles in the bed.

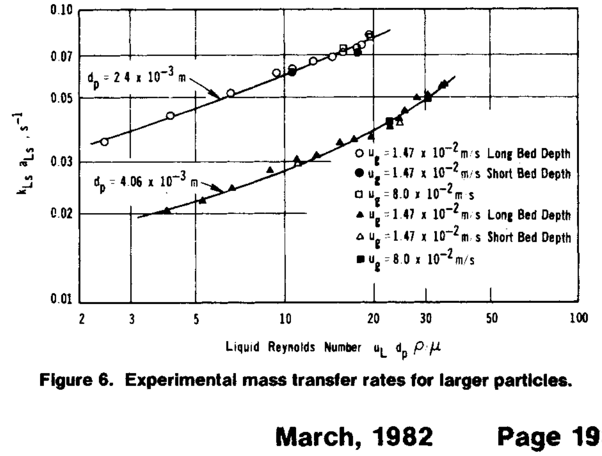


Figure 6. Experimental mass transfer rates for larger particles.

TABLE 2. PROPERTIES OF DILUTE BENZALDEHYDE-WATER SOLUTIONS AT 298 K AND 1 ATMOSPHERE

Diffusivity of benzaldehyde, D	9.55×10^{-10} m^2/s
Vescosity, μ	0.93×10^{-3} kg/(m)(s)
Liquid density, ρ	1.0×10^3 kg/m^3
Schmidt number, Sc	974

CORRELATIONS FOR k_{Ls}

The four correlations of $k_{Ls}a_{Ls}$ reported by Van Krevelen and Krekels (1948), Goto and Smith (1975), Speccia et al. (1976), and Satterfield, et al (1978) cover the range of results available in the literature. Our data are compared with these correlations in Figure 7. In order to make the comparison meaningful, differences in mass transfer area a_{Ls} need to be considered. Two factors affecting this area may be different for the various results and these factors are discussed in the following paragraphs.

First, the area for the porous particles of our study could be less than the area for the non-porous particles upon which the four correlations are based. We have assumed that a_{Ls} for porous particles is equal to $a_t f \epsilon_p$, where ϵ_p is the porosity of the particles and f is the friction of the outer surface of the particles covered by rivulets of liquid. If the wetted outer surface (void plus solid) of the particles is covered by liquid of thickness greater than the diameter of the pores, a_{Ls} would be $a_t f$. The question of whether ϵ_p should be included has not been completely answered. We chose to include ϵ_p but this assumption does not affect our results for $k_{Ls}a_{Ls}$. It is only the method of correlation that is affected. The total area a_t, assuming a spherical shape, is $6(1 - \epsilon_B)/d_p$. Then, if a_{Ls} is taken equal to $a_t f \epsilon_p$, the Sherwood member is

$$\mathrm{Sh} = \frac{k_{Ls}d_p}{D} = \frac{(k_{Ls}a_{Ls})_{\exp}}{a_t \epsilon_p f}\frac{d_p}{D} \tag{5}$$

or

$$(\mathrm{Sh})f = \frac{(k_{Ls}a_{Ls})_{\exp}}{a_t \epsilon_p}\frac{d_p}{D} \tag{6}$$

For non-porous particles $\epsilon_p = 1$ and for the activated carbon particles of our study $\epsilon_p = 0.62$. Second, the fraction f is unknown and must be included in the correlation as $(\mathrm{Sh})f$, according to Eq. 6. From the results presented by Satterfield (1975) and by Herskowitz et al. (1979), f is expected to range from about 0.65 to 1.0 for the liquid Reynolds numbers covered in this investigation and in that of Goto and Smith (1975).

Figure 7 shows the variations between the different investigations. Our data may be represented by the equation

$$(\mathrm{Sh})(f) = 4.25\,(\mathrm{Re})^{0.48}\,(\mathrm{Sc})^{1/3} \tag{7}$$

Equation 7 and the data points agree well with the correlation of Van Krevelen and Krekels (1948) but are higher than the results of Speccia et al. (1976). Plotting Sherwood number vs. Re accounts reasonably well for the effect of particle size for our experimental points. Also, the correlations proposed by Satterfield et al. (1978) and Speccia et al. (1976) account for particle diameter by plotting Sh vs. Re. In contrast, the methods used by Van Krevelen and Krekels (1948) and Goto and Smith (1975) indicate that a separate effect of d_p, beyond that of the Sherwood and Reynolds numbers, is needed. Hence, the lines for these two correlations in Figure 7 are for $d_p = 2.4 \times 10^{-3}$ m. This size corresponds to the middle of the range included in our experiments. The differences shown in Figure 7 are not due to particle size. Rather the deviation seems more likely to be caused by the difficulty in reproducing the flow pattern (rivulet pattern) of the liquid in trickle beds. Variations in particle shape, perhaps in properties of the liquid such as surface tension, and the variations in arrangement of the individual particles from bed to bed, can cause radical shifts in location, size and number of rivulets. This characteristic of a packed bed with two flowing phases appears to introduce much greater variability in mass transfer rate than is observed in packed beds through which only gas is flowing. Hirose (1976) employed metal particles which corroded in water to improve the wetability in some experiments. His results agree well with our data shown in Figure 7. Hirose (1980) has pointed out that our results with catalyst particles and his with corroded particles may represent a near upper limit for mass transfer rates, as far as wetting and liquid distribution are concerned.

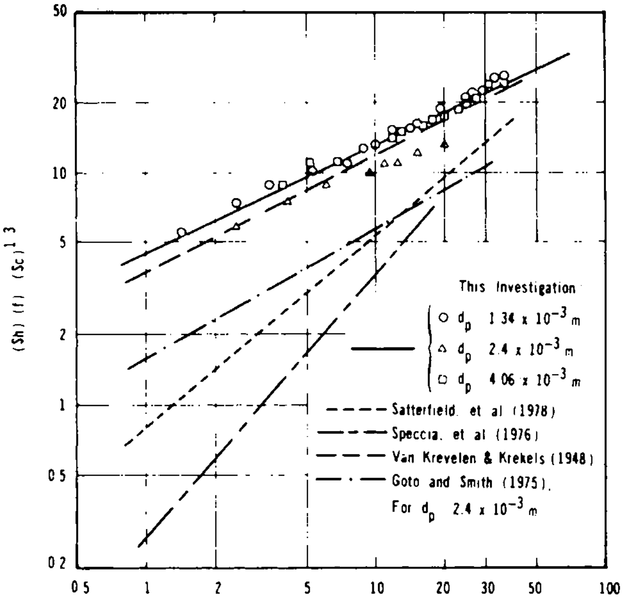


Figure 7. Comparison of correlations for liquid-particle mass transfer coefficients (based upon ϵ_B = 0.39).

ACKNOWLEDGMENT

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NOTATION

a_t = total outer area of particles, m^2/(m^3 of empty reactor)
a_{Ls} = effective mass transfer area from liquid to particle, m^2/(m^3 of empty reactor)
C = benzaldehyde concentration in the liquid phase, kg mol/m^3
C_e = benzaldehyde concentration at the photometer, kg mol/m^3
C_L = benzaldehyde concentration at exit of bed, kg mol/m^3
C_f = step-input benzaldehyde concentration, kg mol/m^3
C_s = benzaldehyde concentration in the liquid rivulet at the outer surface of the catalyst particle, kg mol/m^3
D = molecular diffusivity of benzaldehyde in water at 298K and 1 atm, m^2/s
d_p = particle diameter, m
d_t = diameter of adsorber, m
f = wetting efficiency, defined as the fraction of outer surface of catalyst particle covered by liquid rivulets
k_{Ls} = liquid-particle mass transfer coefficient, m/s
Re = liquid Reynolds number, $\rho u_L d_p/\mu$
Sc = Schmidt number, $\mu/\rho D$
Sh = Sherwood number, $k_{Ls}\,d_p/D$
t = time, s
u_g = superficial gas velocity, m/s
u_L = superficial liquid velocity, m/s

x = axial coordinate in adsorber in the direction of liquid flow, m

L = length of carbon bed, m

Greek Letters

ϵ_B = void fraction in the bed

ϵ_L = dynamic liquid holdup, (volume of liquid)/(volume of empty tube)

ϵ_p = particle porosity

μ = liquid viscosity, kg/(m)(s)

ρ = liquid density, kg/m^3

τ = retention time in the bed, $\epsilon_L l/u_L$, s

τ_t = total average retention time from injection point to detector, s

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Mixing of Single and Two Phase Systems: Power Consumption of Impellers

The mechanical power consumed in agitating aerated and nonaerated aqueous phases was measured and correlated using the data of previous investigators. The correlation is for a flat bladed turbine operating within the turbulent regime in fully baffled vessels. The accuracy of the predictions and the range of geometrical variations and equipment sizes support the use of the correlation for scaleup operations.

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SCOPE

The power consumption of impellers used for agitating gas-liquid systems is a widely used parameter for determining gas holdup, interfacial area and mass transfer rates in baffled vessels. The use of this parameter leads to complications since to date, there appears to be no general working equations for the prediction of power consumption. Calderbank's (1958) equations

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