Gas Phase Mass Transfer in Packed Columns

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There has been considerable discussion on the effect of diffusivity on the gas phase mass transfer coefficient in packed columns. Vidwans and Sharma (1967) reviewed the literature on this subject and pointed out that the exponent of the diffusivity of the solute in the gas phase is reported to vary from 0.15 to 1.0. These authors also report experimental data with an eight-fold variation in gas phase diffusivity and conclude that the coefficient varies as the square root of the diffusivity and that it is unlikely that the Schmidt number is the proper correlating factor. Correlation of the mass transfer coefficient with diffusivity is dimensionally inconsistent. This note proposes a dimensionally consistent correlation.

Hughmark (1971) showed that gas phase mass transfer on a distillation tray can be modeled as that for a turbulent gas core. This analysis can also be applied to the gas phase in packed columns with countercurrent liquid flow. A prior paper (Hughmark, 1971) presents an equation for transfer in the core for turbulent flow in a circular pipe.

$$k = \frac{3.58\mathfrak{D}}{R} + uf \tag{1}$$

The first term represents molecular diffusion and the second term corresponds to the turbulent transport contribution. Equation (1) can be transposed to obtain

$$\frac{k}{u} = \frac{7.16\mathfrak{D}}{Du} + f = \frac{7.16}{N_{Re}N_{Sc}} + f \tag{2}$$

Equation (2) applies to a circular core. A general equation for different core configurations is then

$$\frac{k}{u} = \frac{A}{N_{Re}N_{Sc}} + f \tag{3}$$

Flow in a packed column can be modeled as flow in parallel tortuous paths in which the equivalent diameter D_e is $4\epsilon/a_v$, the actual velocity u_e is $L_e/L(u_s/\epsilon)$, and the Reynolds number is $D_e u_e/\nu$. This model for packed beds is discussed by Hughmark (1972).

Figure 1 shows the data of Vidwans and Sharma for 0.95-cm Raschig rings with $u_s = 29.7$ cm/s and a liquid rate of 0.267 cm/s. The data are correlated by the equation

$$\frac{k}{u_e} = \frac{1.2}{N_{Re}N_{Sc}} + 0.00525 \tag{4}$$

Figure 1 also shows the data of Zabban and Dodge (1954) for 1.27-cm Raschig rings with $u_s = 12.8$ cm/s and a liquid rate of 0.41 to 0.44 cm/s. The data are for the methanol-air-water system with a pressure range of 1.95 to 10.7 atmospheres and are observed to be consistent with Equation (4). Experimental data are also reported for 2.54-cm Raschig rings with different gas phase diffusivities. Figure 1 also shows these data for $u_s = 22.9$ cm/s and a liquid rate of 0.195 cm/s from Fellinger as reported by Sherwood and Pigford (1952) for air-water and Houston and Walker (1950) for ammonia, methanol, and acetone with the air-water system. These data are observed to fit the equation

$$\frac{k}{u_e} = \frac{7.16}{N_{Re}N_{Sc}} + 0.0225 \tag{5}$$

which indicates that these data correspond to a circular turbulent core. Mass transfer coefficients are obtained from the reported experimental data for 1.27-cm and 2.54-cm Raschig ring packing with the interfacial areas reported by Danckwerts and Sharma (1966).

NOTATION

A = constant in Equation (3)
a_v = specific surface area of dry packing
D = core diameter
$\mathfrak{D}$ = molecular diffusivity
f = friction factor
k = mass transfer coefficient
L = bed depth
L_e = actual length of path taken by fluid in traversing depth L of bed
N_{Re} = Reynolds number
N_{Sc} = Schmidt number
R = core radius
u = average core velocity
u_e = actual velocity in packing channels
u_s = superficial velocity

Greek Letters

ϵ = packing void fraction
ν = kinematic viscosity

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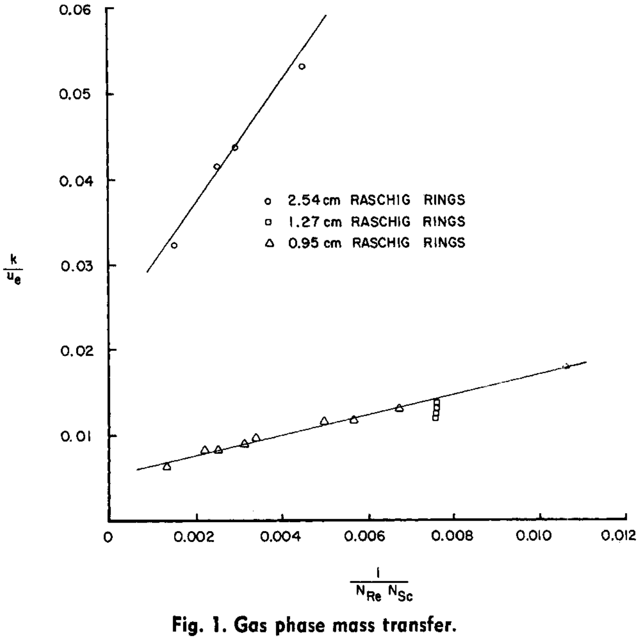

Fig. 1. Gas phase mass transfer.

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Direct Solid-Catalyzed Reaction of a Vapor in an Apparently Completely Wetted Trickle Bed Reactor

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Experimental evidence is presented to show that in a trickle bed reactor operating under apparently completely wetted conditions some vapor-phase reaction may be catalyzed directly by the solid catalyst at liquid flow rates typical of pilot-plant size reactors and of some industrial reactors.

EXPERIMENT

The hydrogenation of benzene at 76°C, four degrees below the boiling point, was studied in a bed of 2% Pt-on-alumina spherical pellets 0.635 cm (1/4-in.) in diameter held in a glass-walled reactor. This was maintained isothermal by automatically-controlled electrical heaters. The catalyst bed was 3.81 cm in internal diameter, 10 cm high and surmounted by a layer 3.5 cm high of inert alumina pellets, also 0.635 cm in diameter. The liquid distributor was a flat shower head with 47 pipes to provide uniform initial liquid distribution on the alumina pellets even at low liquid flow rates. Incoming benzene was preheated and saturated with hydrogen. It was assumed that the liquid and gas stream were in equilibrium with each other by the time they reached the catalyst because of the ample opportunity for contact provided by the shower head design and the inert packing. The exit gas stream from below the catalyst bed was condensed, measured, and analyzed by gas chromatography. The liquid phase was also analyzed. The mole fraction of hydrogen in the gas phase was calculated to be 0.125. Readings from five thermocouples located in the center of the inert pellet section and catalyst section showed that axial temperature gradients were less than 0.1 to 0.2°C. The maximum temperature difference radially between the center of the bed and the wall was less than 1°C.

Before each run reported, the reactor was heated to 120° to 140°C under H_2 flow for at least two days, then cooled to room temperature and heated up to reaction temperature under an automatic control system all under H_2 flow. Only then was benzene introduced into the system. The conversion was a fraction of 1% at all times so the reactor can be treated as a differential reactor. The superficial hydrogen flow rate was 0.146 cm/s (S.T.P., inlet conditions) although the total of H_2 plus benzene vapor flow rate was about 1.17 cm/s (again, S.T.P., inlet conditions). More details are given by Özel (1973).

RESULTS AND DISCUSSION

Table 1 summarizes the results of three runs of interest, each at a different liquid flow rate. In Run 1 the feed contained some cyclohexane; in the other two the feed was pure benzene. The figure shows, for Run 2, the typical change in concentration with time of the exit gas and liquid phases until steady state was reached. At all three flow rates studied the exit gas phase concentration exceeded that in equilibrium with the liquid phase and both concentrations decreased to steady-state values reached only after a period of 1 to 2 hours. The steady state rate of reaction also decreased with increased flow rate. At the highest liquid flow rate, Run 3, there was little change in liquid exit composition with time, but the gas phase concentration dropped from 4×10^{-3} mole % at 30 min. to the steady state value of 1.1×10^{-3} mole % after about 90 min. Analysis was by gas chromatography.

By running very dilute solutions of cyclohexane in benzene over inert spherical alumina pellets in the reacter at 76°C in the presence of H_2 the mole fraction ratio y/x was found to be constant at 1.6 for the range of concentrations obtained. In each case the amount of gas phase condensate agreed well with that corresponding to the vapor pressure of benzene at 76°C. Although the composition difference between the two phases is striking, the mass flow rate of liquid is so much greater than that of the vapor that most of the cyclohexane formed, for example, 95% for run 2, left in the liquid phase.

Calculations (Özel, 1973) showed that mass transfer limitations were substantial. For the three runs, γ was estimated by the stagnant film model to be about 0.6 to 0.7 and the internal effectiveness factor varied from about