draulic radius R_h

N_{Re} = $2Rv_0\rho/\mu$, Reynolds number based on sphere radius

N'_{Sh} = $2kR_h/D_{AB}$, Sherwood number based on hydraulic radius R_h

q = internal tortuosity factor

R = average particle radius, cm

R_h = hydraulic radius, cm

v = linear interstitial velocity, cm/s

v_0 = αv, linear superficial velocity, cm/s

α = bed void fraction

β = internal porosity of the particle

δ_0 = $\beta(1-\alpha)/\alpha$

δ_1 = $(\beta^2R^2/15)\ (1/D_i + 5/kR)\ (1-\alpha)/\alpha$

$\Delta\mu_1'$, $\Delta\mu_2$ = differences between moments at inlet and outlet; s, s^2

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The Grid Region in a Fluidized Bed Reactor

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The current models for gas fluidized bed reactors as described by Kunii and Levenspiel (1969) do not distinguish between the region near the inlet grid and the rest of the reactor. It is well known, however, that the grid region plays a critical role in determining reactor performance, particularly in large beds. Industrial pilot plant experiments by Cooke et al. (1968) have shown that with fast reactions most of the conversion can take place in the first half-meter of bed height. Changing the grid can have a significant effect. Cooke et al. found that increasing the grid hole size caused a drop in conversion and Hovmand et al. (1971) showed that increasing the number of holes in the grid caused a marked improvement. These observations imply that the contacting efficiency near the grid is greater than in the rest of the bed.

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In this note, we present a simple model which accounts explicitly for the grid region and allows easy analysis of the effect of grid modifications.

Recent experiments by Behie (1971, 1972) have characterized the heat and mass transfer in the region above a perforated plate grid in a pilot scale fluidized bed. Grid hole diameters from 6.35 to 25.4 mm were used with jet velocities between 15.3 and 91.5 m/s, giving data which are typical of industrial conditions. The data could be represented adequately by the simple transfer resistance model described below.

THE MODEL

The gas enters the reactor as high speed jets which penetrate a distance h before breaking up into bubbles. That part of the bed below h is called the grid region and is assumed to contain no bubbles. It is assumed that the jets are well mixed radially with plug flow in the axial direction. The fluidized emulsion in the neighborhood of the distributor is assumed to be perfectly mixed by the highly turbulent jets. Mass transfer takes place between the jets and the emulsion and the transfer is described here in terms of a mass fraction driving force and a resistance.

Many models have been proposed for the bubbling region of the bed and all of them can be used in conjunction with the model of the grid region. We use here the simple model of Orcutt (1960) which assumes that the bubble gas is in plug flow and the emulsion is perfectly mixed. Transfer between bubbles and surrounding emulsion is described by a resistance and a driving force.

In writing the model equations we ignore reaction in the jets. In practice, catalyst particles are entrained in the jets, but the solids concentration is sufficiently low for reaction to be negligible. The equations describing the system are

Jets

$$Q\frac{dy_j}{dx} + k_j a_j A(y_j - y_\infty) = 0 \quad 0 \leq x \leq h \tag{1}$$

$$\text{B.C.} \quad y_j = y_{j0} \quad \text{at} \quad x = 0$$

Bubbles

$$Q_b\frac{dy_b}{dx} + k_b a_b A(y_b - y_\infty) = 0 \quad h \leq x \leq H \tag{2}$$

$$\text{B.C.} \quad y_b = y_j(h) \quad \text{at} \quad x = h$$

Emulsion

$$Q_e(y_{e0} - y_\infty) + k_j a_j A \int_0^h (y_j - y_\infty)dx + k_b a_b A \int_h^H (y_b - y_\infty)dx = L_e V_e \rho_s R(y_\infty) \tag{3}$$

$$\text{B.C.} \quad y_{e0} = y_j(h)$$

Also

$$Q = Q_e + Q_b \tag{4}$$

where Q_e is the mass flow rate of gas through the emulsion and is usually taken as the flow rate required to fluidise the bed incipiently.

Equations (1) to (3) must be solved simultaneously. To simplify the analysis, we assume that the reaction is first order and that there are no diffusional limitations in the emulsion so that $R = \nu y$. Hence the fractional conversion η is given by

$$\eta = \frac{\mu[1 - (1 - \delta)e^{-\gamma}]}{\mu + 1 - (1 - \delta)e^{-\gamma}} \tag{5}$$

In most industrial reactors the flow rate used is many times that required for incipient fluidisation and $\delta << 1$. In this case

$$\eta = \frac{\mu(1 - e^{-\gamma})}{\mu + 1 - e^{-\gamma}} \tag{6}$$

In what follows conversions computed from Equations (6) are called *Grid Model* predictions. Bubble Model predictions are obtained by taking the jet penetration to be zero so that bubbles are formed at the grid.

In presenting examples, we have sought to use values of the various parameters which are realistic for industrial conditions. Behie's (1972) experiments have shown that the mass transfer from high velocity, noninteracting grid jets can be described reasonably well by a plug flow model with constant mass transfer coefficient. The experiments were carried out in a 0.61 m diameter bed of 60 micron cracking catalyst. Typical experimental values for the mass transfer coefficient $k_j a_j$ are given in Table 1, and these are used here as representative of the mass transfer from the turbulent jets in commercial equipment. Values of the jet penetration h and the diameters d_b of bubbles breaking away from the jets have been calculated from the empirical equations of Basov (1969) which are in good agreement with the measurements reported by Behie. As a first approximation, values of $k_b a_b$ have been calculated from the equations of Kunii and Levenspiel (1969). Bubble transfer areas a_b were computed from bed expansion taken from Behie (1972), which gave $\epsilon_b = 0.25$ when the bed superficial velocity U_s was about 0.61 m/sec. The incipient fluidizing velocity U_{mf} was taken as 1.53 cm/s and the void fraction under incipient conditions ϵ_{mf} was 0.5.

Figure 1 shows the effect of the grid on conversion for both a fast and slow reaction in a fluidized bed. The effective rate constants $\nu' = \nu L_e \rho_s$ were taken from reported data. For the fast reaction the value of $\nu' = 9.63$ (kg/sec · m³ emulsion) was taken from the data of Cooke (1968) for a fluidised carbonizer. For the slow reaction $\nu' = 9.63 \times 10^{-2}$ (kg/sec · m³ emulsion) is within the range reported by Hovmand et al. (1971) for the decomposition of ozone. The curves are computed for a grid with 12.7 mm holes and a jet velocity of 45.7 m/sec. The predictions of the Grid Model and Bubble Model are shown for each reaction. For the fast reaction the predictions of the two models are significantly different. With a bed 0.6 m deep the Grid Model predicts a conversion of 84%, while the Bubble Model predicts a conversion of only 11%. Cooke's data for this reaction showed that the conversion was very high in a 0.6 m bed and that increasing the bed height above this increased the conversion only slightly.

For the slow reaction the difference between the two models is much less significant since mass transfer resistance does not limit the rate severely.

TABLE 1. MASS TRANSFER COEFFICIENTS FOR GRID JETS AND GAS BUBBLES IN A FLUID BED OF CRACKING CATALYST
($U_s = 0.61$ m/s, $\epsilon_{mf} = 0.5$, $\epsilon_b = 0.25$, $U_{mf} = 1.53$ cm/s)

	$D_0 = 6.35$ mm			$D_0 = 12.7$ mm			$D_0 = 19.05$ mm
	$U_0 = 15.2$ m/s	$U_0 = 45.7$ m/s	$U_0 = 91.5$ m/s	$U_0 = 15.2$ m/s	$U_0 = 45.7$ m/s	$U_0 = 91.5$ m/s	$U_0 = 15.2$ m/s
$k_j \cdot a_j \left[\frac{kg}{m^3\,s}\right]$	17.7	12.7	7.77	12.6	7.05	3.54	9.41
$k_b \cdot a_b \left[\frac{kg}{m^3\,s}\right]$	0.446	0.258	0.194	0.234	0.143	0.0891	0.153
a_j (m)	25.1	8.40	4.04	12.5	4.20	2.10	8.43
a_b (m)	37.7	25.8	20.2	23.6	16.1	11.2	17.9
h (cm)	12.6	18.7	23.9	20.7	30.5	43.6	27.4
d_b (cm)	3.88	5.86	7.38	6.35	9.37	13.4	8.40

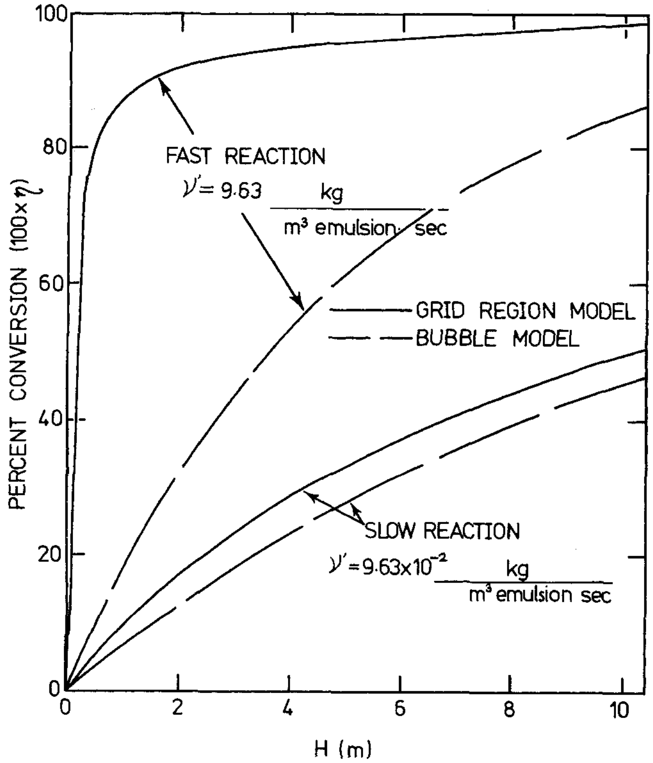

Fig. 1. The effect of enhanced contacting efficiency near the distributor for fast and slow reactions.

The model is consistent with many experimental observations. With the values in Table 1 calculations show that conversion is increased by decreasing grid hole size or increasing the number of holes.

CONCLUSIONS

It is apparent that for fast reactions the grid arrangement can have a significant effect on conversion. It is seen from Equation (6) that the critical factor in determining the conversion in the grid region is the value of α. Table 1 shows that α tends to increase as the grid hole size is decreased and as the number of holes is increased. If α is greater than 3, the conversion in the grid region is close to its maximum value and little improvement can be effected by changing either the size or the number of holes. On the other hand, if α is much less than 3, improvements in conversion may be obtained by modifying the grid. Because it accounts explicitly for the grid region, the model appears to be useful for predicting the behavior of fast reactions in large fluidized reactors. The bubbling region of the bed above h can be described by more sophisticated models if necessary. Experimental measurements of $k_j a_j$ for various grid designs and gas mixtures will be necessary if the model is to be used for design purposes.

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NOTATION

a_b = transfer area bubbles to emulsion per unit volume of reactor, m^2/m^3
a_j = transfer area jets to emulsion per unit volume of reactor, m^2/m^3
A = cross-sectional area of reactor, m^2
d_b = bubble diameter, cm
D_0 = grid hole size, mm
h = depth of jet penetration, cm
H = bed height, m
k_b = mass transfer coefficient, bubbles to emulsion, $kg/m^2 \cdot s$
k_j = mass transfer coefficient, jet to emulsion, $kg/m^2 \cdot s$
L_e = catalyst loading in emulsion, m^3/m^3 emulsion
Q = mass flow through reactor, kg/s
Q_b = mass flow through reactor as bubbles, kg/s
Q_e = mass flow through emulsion, kg/s
R = rate of reaction per unit weight of catalyst, $kg/s \cdot kg$
U_{mf} = minimum fluidizing velocity, cm/s
U_0 = jet velocity, m/s
U_s = superficial velocity through reactor, m/s
V_e = volume of emulsion, m^3
x = distance, m
y_b = mass fraction of reactant in bubbles
y_{e0} = mass fraction of reactant in gas entering emulsion
y_j = mass fraction of reactant in jet gas
y_{j0} = mass fraction of reactant in feed
y_x = mass fraction of reactant in emulsion gas

Greek Letters

α = dimensionless group $\frac{k_j a_j A h}{Q}$
β = dimensionless group $\frac{k_b a_b A h}{Q_b}$
γ = dimensionless group $\alpha + \beta\left(\frac{H}{h} - 1\right)$
δ = fraction of flow through emulsion $\frac{Q_e}{Q}$
ϵ_b = fraction of reactor volume above h occupied by bubbles, m^3/m^3
ϵ_{mf} = void fraction of emulsion at incipient fluidization, m^3/m^3 emulsion
η = fractional conversion
μ = dimensionless rate constant $\frac{\nu L_e \rho_s V_e}{Q}$
ν = rate constant $kg/s \cdot kg$ catalyst
ν' = effective rate constant $\nu L_e \rho_s$ $kg/s \cdot m^3$ emulsion
ρ_s = density of a catalyst particle, kg/m^3

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