MASS TRANSFER BETWEEN PHASES IN A FLUIDIZED BED AND IN A LAYER WITH A CYLINDRICAL SPIRAL FILLING

V. D. Chernov and B. R. Serebryakov

UDC 66.096.5:66.015.23

It has been shown by experiment that the mass transfer between phases in a fluidized bed with cylindrical spiral packing is improved on account of improvement in the rate constant for the process with fixed values for the proportion of gas passing through the layer in bubbles and the linear velocity for the phases.

It is often found that the observed reaction rate in a heterogeneous catalytic process is less in a fluidized bed catalyst than in a stationary one; one of the possible reasons is longitudinal mixing of the gas-contact mass in the reaction zone. If the particles are porous, the gas mixing occurs by an adsorption–desorption mechanism [1-3], and the fall in the reaction rate by mixing is governed by many factors [1, 4]. However, a fluidized bed of nonporous particles is a system close in properties to an ideal displacement apparatus for gas [3].

A major reason for rate reduction in a fluidized bed is the macroscopic nonuniformity: part of the fluidizing flow passes through the layer in the form of bubbles, and in that case the decisive factors are the proportion of bubbles in the total flow, the linear size of these, the rise rate, etc.

If the fluidized bed is nonuniform, the overall rate of the chemical process is dependent on the reaction rate in the dense phase and the diffusion rate from the bubbles into the continuous phase. Under certain conditions, the interphase mass transfer becomes the rate-limiting step [2].

Many studies have been made [1-8] on nonuniformity in such systems and the interphase mass-transfer rate, which have shown that these parameters are dependent on the size of the apparatus and the detailed hydrodynamic conditions.

We have examined the nonuniformity and interphase mass transfer in an apparatus of diameter about 0.5 m; we used the transient-state method of analysis, in which a step pulse of tracer gas is admitted at the bottom of a layer of nonporous particles (the F-function method).

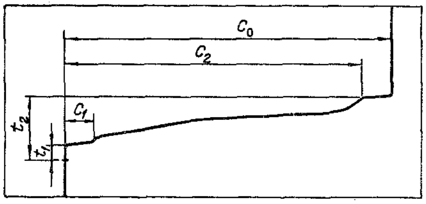

Fig. 1. Typical response to a step input at the base (r = 2.12 mm, H = 850 mm, propylene tracer, grid of diameter 45 mm inserted in layer).

This F-function method is simpler to operate than the δ-function method [2], because in the first case one needs to provide only a steep leading edge.

If a positive-going step is supplied to the base of the layer, the mass transfer is described by the following system of equations:

$$\frac{\partial Z}{\partial l} + \frac{1}{W_z}\frac{\partial Z}{\partial t} = \alpha (Y - Z), \qquad Y(l,\ 0) = Z(l,\ 0) = 0,$$

$$\frac{\partial Y}{\partial l} + \frac{1}{W_y}\frac{\partial Y}{\partial t} = \beta (Z - Y), \qquad Z(l,\ t)_{l=W_y t} = Z(W_y t,\ t) = C_0, \tag{1}$$

where

$$\alpha = \frac{kf}{bW_z}; \quad \beta = \frac{kf}{(1-b)W_y}; \quad b = \frac{\omega W_y}{\omega W_y + (1-\omega) W_z}.$$

Translated from Inzhenerno-Fizicheskii Zhurnal, Vol. 21, No. 6, pp. 992-997, December, 1971. Original article submitted November 26, 1970.

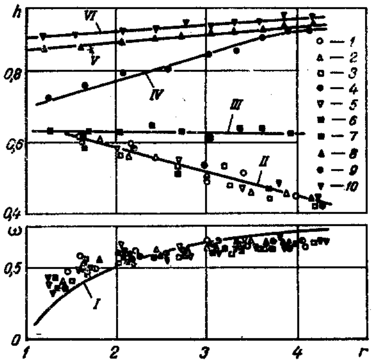

Fig.2

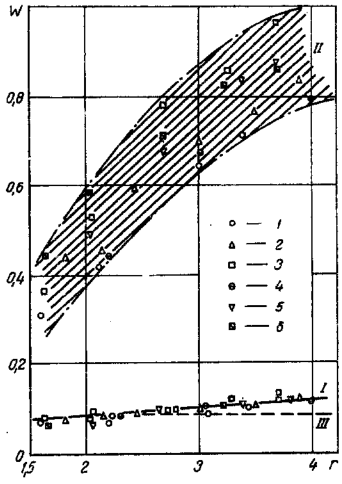

Fig.3

Fig.2. Effects of fluidization number r on w (proportion of gas passing through layer in bubbles) and h (constant for mass transfer between phases, sec^{-1}) for layers in various states: I) w = f(r) from the two-phase theory; II-VI) h = f(r); 1-3) layer without grid, tracer propylene, layer heights 0.5, 0.8, and 1.3 m, respectively; 4-6) the same with hydrogen; 7) layer of height 0.8 m with grid 75 mm in diameter set vertical; 8) the same set horizontal; 9) the same within layer; 10) grid 45 mm in diameter in layer of height 0.8 m.

Fig.3. Speeds of gas phases (m/sec) in fluidized bed: I) speed of dense phase; II) speed of bubbles; III) speed for fluidization onset; 1-3) propylene tracer, layer heights 0.5, 0.8, and 1.3 m respectively; 4-6) the same with hydrogen tracer.

This system is solved by means of integral transformation [9].

The tracer concentration at the exit from the layer after phase mixing is

$$C = \omega Z + (1 - \omega) Y, \tag{2}$$

and when the bubble front reaches the surface of the layer we have

$$C_1 = \omega C_0 \exp\left[-\frac{kf}{bW_z} H\right], \tag{3}$$

and when the front of the dense phase does the same we have

$$C_2 = C_0\left\{1 - (1 - \omega) \exp\left[-\frac{kf}{(1-b) W_y} H\right]\right\}. \tag{4}$$

Figure 1 shows a typical observed curve for the output in response to a step input; this curve enables one to find

$$W_y = H/t_2, \tag{5}$$

$$W_z = H/t_1, \tag{6}$$

$$\omega = \frac{1}{C_0 (t_2 - t_1)} \int_{t_1}^{t_2} C dt. \tag{7}$$

The rate constant for mass transfer between phases h is equal to kf/b and can be determined from C_1, C_2, and C_0 in accordance with (3) and (4).

We used beds of quartz sand (d_{av} = 200 μ, W_{cr} = 0.04 m/sec) of depth 0.5-1.3 m, which were fluidized by air in a glass apparatus 0.45 m in diameter. The distributor was a double perforated grid with an effective cross section of 0.175%. The tracer gas was injected into the space between the grids, which was filled

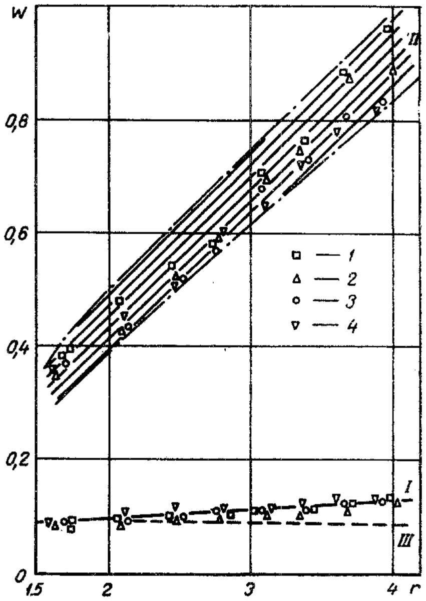

Fig. 4. Speeds of gas phases (m /sec) in layer with grid: I) speed of dense phase; II) speed of bubbles; III) speed for fluidization onset; 1) grid 75 mm in diameter set vertically; 2) the same but horizontal; 3) in layer; 4) grid 45 mm in diameter in layer.

with a porous mass, at 42 points uniformly distributed over the cross section. Tests showed that the steepness of the leading edge of the pulse was represented by a rise time of less than 0.1 sec. The nonuniformity in the tracer gas distribution over the cross section was less than 3% (at over 10 mm from the wall). The concentration at the exit from the layer was measured by a thermochemical detector. The exit flow was a sample at 18 points in the space above the layer to obtain an average sample. The time for the signal to pass through the measuring system ($\sim$0.5 sec) was recorded in each run; the time t_2 was 8-50 sec in accordance with the height of the layer. The coefficients of variation in ω, W_y, W_z, and h in a single experiment were respectively $\sigma\omega = 12.2\%$, $\sigma_{W_y} = 11.8\%$, $\sigma_{W_z} = 11.3\%$, $\sigma_h = 10.2\%$. To reduce the effective σ to less than 5% for each type of run, we made 6 to 10 parallel runs in each case.

Figures 2 and 3 show ω, h, W_y, and W_z as a function of fluidization number in the fluidized layer and for a layer filled with cylindrical spiral packing, these values being determined by the use of propylene and hydrogen as tracer gases.

The results show that the values are not dependent on the type of tracer, which indicates that convective transport dominates the interphase mass exchange.

Figure 3 shows that there is a clear tendency for W_y to increase with the fluidization number for any state of the bed; for $r > 2.3$, none of the observed values was less than W_{cr}, which conflicts with the distribution of the fluidizing agent between the phases assumed in the two-phase theory. A similar deviation from the two-phase theory is clearly seen on comparing the observed ω with theory [10] (Fig. 2).

We found that h fell, not increased, as the fluidization number was increased, because there was a tendency for the bubbles to occur less as the fluidization number increased in the range we employed, and turbulence in the flow of bubbles and dense phase do not compensate for the reduction in the relative surface available for mass transfer.

We have already remarked that sometimes the rate-limiting step is reagent transfer from the bubbles to the dense phase, and the process operates in the interphase-diffusion region [2]; of course, any improvement in the output parameters for such a process is dependent primarily on improving the mass-transfer characteristics. The interphase mass transfer in a fluidized bed can be intensified by filling the volume with a cylindrical spiral packing [11], but there is no quantitative evidence on the performance of such packing as regards layer uniformity and mass transfer, in spite of the considerable interest in the topic. We have made measurements to determine the characteristics of the interphase mass transfer in fluidized beds filled with packing made from wire 5 mm in diameter coiled into cylindrical spirals 100-120 mm high, 45 or 75 mm in diameter, and pitch 12-15 mm. These tests were done with beds of depth 0.8 m. In these tests we varied the disposition of the packing (horizontal or vertical ordered arrangement, placed at random) as well as the fluidization number.

Figure 2 shows the measured proportion of the gas that passed through the layer in bubbles, together with ω for freely fluidized beds. The packing had virtually no effect on this characteristic of the nonuniformity.

Figure 4 compares W_y and W_z for layers with the packing together with the corresponding values for free fluidized layers, which shows only that there is a reduction in the dispersion of the local bubble velocities. The effect of the packing is greatest as regards h, which occurs because the packing hinders bubble coalescence and facilitates breakup of consolidation zones around them. Superficially, this appears as a breaking up of the bubbles by the packing. If the bed is freely fluidized, the finite size of the bubbles is bounded only by the diameter of the layer, whereas a layer containing packing has a bubble diameter dependent on the size and disposition of the packing. Figure 2 gives values for the rate constant for interphase mass transfer for beds with different dispositions of the packing. As would be expected, the largest h was obtained for a bed with packing of the least diameter (45 mm), as placed at random in the layer, and the important point is that h increases with r.

NOTATION

k	is the mass-transfer coefficient, m/sec;
f	is the relative surface of the bubbles, i.e., the surface of the bubbles in unit gas volume of the layer, m^2/m^3;
b	is the fraction of the gas volume occupied by bubbles, m^3/m^3;
ω	is the fraction of the gas passing through the layer in bubbles, m^3/m^3;
W_z, W_y	are the linear velocities of the bubbles and the dense phase, m/sec;
Z, Y	are the concentrations of tracer in the bubbles and the dense phase, m^3/m^3;
C_0	is the concentration of tracer in the completely mixed gas, m^3/m^3;
l	is the current height, m;
t_1, t_2	are the times for the bubbles and the dense phase to pass through a layer of height H, sec;
C_1, C_2	are the steps in the output function at the escape of the bubble front and the dense phase front, m^3/m^3;
H	is the height of the layer under working conditions, m.

LITERATURE CITED

1. I. I. Ioffe and A. F. Grigorov, Khim. Prom., No. 3 (1959).
2. I. I. Ioffe and L. M. Pis'men, Khim. Prom., No. 4 (1960).
3. G. N. Abaev and T. Ya. Gusman, Khim. Tekh. Topliv i Masel, No. 10 (1964).
4. N. I. Gel'perin, V. G. Ainshtein, and V. B. Kvasha, Principles of Fluidization Technology [in Russian], Khimiya, Moscow (1967).
5. I. A. Vakhrushev and V. A. Basov, Khim. Prom., No. 6 (1968).
6. M. É. Aerov and O. M. Todes, Hydraulic and Thermal Principles of Operation for Equipment with Stationary and Fluidized Granular Beds [in Russian], Khimiya, Leningrad (1968).
7. V. Sh. Kernerman, I. P. Mukhlenov, M. G. Slin'ko, and Yu. V. Filatov, Khim. Prom., No. 6 (1966).
8. A. A. Oigenblik, Author's Abstract of Candidate's Dissertation [in Russian], All-Union Scientific-Research Institute of Petroleum Chemistry, Moscow (1968).
9. V. D. Chernov, I. P. Dobrovol'skii, B. R. Serebryakov and M. A. Dalin, Dokl. Akad. Nauk AzSSR, 26, No. 2 (1970).
10. J. F. Davidson and D. Harrison, Fluidization of Solid Particles [Russian translation], Khimiya, Moscow (1965).
11. A. I. Tamarin and V. D. Dunskii, Authors' Certificate No. 242146; Byul. Izobr., No. 15 (1969).