FORMATION OF THE REGION OF PHASE-PARAMETER STABILIZATION IN A DEVELOPED FLUIDIZED BED

Yu. P. Gupalo, A. A. Oigenblik,
Yu. S. Ryazantsev, and Yu. A. Sergeev

UDC 532.529:66.096.5

A model is proposed for the formation of the zone of heightwise variation of phase parameters and the zone of stabilization of the bubble phase in a fluidized bed. A qualitative study is made of the effect of the fluidization parameters on the position and dimensions of the zone.

To design fluidized-bed reactors, designers usually use mean parameters of the bubble phase and the dense phase — the dimensions of the bubbles, the volume of the bubble phase, the magnitude of the gas flow passing through the dense phase, etc. These parameters change considerably through the height of the bed [1, 2]. Experimental determination of the parameters of the bubble phase involves measurement of the local densities of the bed, frequencies of density fluctuation, and rate of surfacing and dimensions of the bubbles. Such studies can be conducted on so-called "cold" models of fluidized-bed units.

The fluidization conditions in a given fluidized-bed reactor may differ from the conditions present in the "cold" model. Thus, we can expect a change in the parameters of the bubble phase and the associated values of the phase exchange coefficients.

The goal of this article is to qualitatively evaluate the effect of the temperature, pressure, viscosity, and density of the fluidizing flow and the dimensions and density of the solid particles on the parameters of the bubble phase.

Three characteristic zones can be distinguished in units with free fluidized beds having distributors in the form of porous or perforated plates [1, 2]: 1) a zone of variable parameters adjacent to the gas-distributing zone in which there is substantial bubble growth and a reduction in the gas flow in the dense phase over the height of the bed; this zone contains a substantial quantity of fine bubbles; 2) a stabilization zone in which the parameters of the phases (bubble dimensions, mean gas velocity in the dense phase, etc.) are nearly constant; 3) a zone of intensive bubble disintegration adjacent to the upper boundary of the bed, the height of this zone being small compared to the heights of the first two zones.

We should note the substantial difference in the rates of interphase (bubbles—dense phase) mass transfer in the different zones. The zone of variable parameters contains fine bubbles, which have a high rate of mass exchange with the dense phase. The zone characterized by stable parameters contains coarse bubbles; the gas velocity in the dense phase is close to the minimum fluidization velocity, and nearly all of the reactants entering into reaction on the surfaces of the particles in the dense phase are delivered by bubbles as a result of volumetric flow.

The location and dimensions of the variable-parameters zone and stabilization zone and the values of the parameters of the phases in these zones have a significant effect on the occurrence of chemical processes in the fluidized-bed reactor. In particular, when catalytic processes are taking place, an increase in the dimensions of the variable-parameters region leads to an increase in conversion and selectivity [3].

Of practical interest are the dependence of the dimensions of the basic zones of the bed, their location, and values of parameters of the phases in the zones on the pressure, temperature, viscosity, and velocity of the gas and the dimensions and density of the solid particles, etc.

Institute of Problems of Mechanics, Academy of Sciences of the USSR, Moscow. Translated from Inzhenerno-Fizicheskii Zhurnal, Vol. 47, No. 5, pp. 783-789, November, 1984. Original article submitted March 24, 1983.

0022-0841/84/4705-1317$08.50

A qualitative explanation of the reasons for the formation of the stabilization zone follows. The velocity of the gas in the dense phase exceeds the fluidization velocity in direct proximity to the gas distributor. According to the results of the studies [4-7], in this case the bubbles grow as they rise in the bed. The bubble growth is due to the infiltration of gas from the dense phase into the bubbles and their coalescence and it leads to a reduction in gas velocity in the dense phase with increasing height. At a certain distance x_* from the gas distributor the gas velocity in the dense phase becomes equal to the minimum fluidization velocity. According to empirical results [1, 4, 5], this violates the conditions for bubble growth and at heights exceeding x_* the parameters of the phases remain constant up to the small bubble-disintegration zone adjacent to the upper boundary of the bed. The height x_* corresponds to the lower boundary of the stabilization zone.

The distribution of phase parameters over the height of a fluidized bed due to bubble growth was examined earlier in [8]. It should be noted, however, that in this work the authors assumed the concentration of bubbles to be constant over the bed height in the absence of coalescence. Since the bubbles accelerate with height as a result of the growth, their concentration should change over the height. The basic equation describing the dependence of bubble size on the distance to the gas distributor was derived using the assumption of the constancy of the gas velocity in the dense phase over the bed height. The dependence of bubble size on distance to the gas distributor obtained from dimensional theory leads to vanishing of bubble size when the gas velocity in the dense phase equals the minimum fluidization velocity. This contradicts the empirical findings — in the stabilization region, where gas velocity is roughly equal to the minimum fluidization velocity, bubble size is maximal. These considerations limit the range of application of the results from [8]. The model proposed below is free of such problems. In particular, the dependence of the bubble concentration and gas velocity in the dense phase on the distance to the inlet to the fluidized bed is determined from equations of the proposed model.

Let us construct the simplest possible qualitative theory describing the phenomenon in question. We will assume that the concentration of surfacing bubbles is not great, so that coalescence processes are absent and the rate of surfacing coincides with the rate of rise of an isolated bubble. In this case, the rate of bubble growth is determined only by the infiltration of gas from the dense phase. We will assume that the dimensions of the rising bubbles are the same.

The quantities characterizing the state of both phases will be averaged over a period of time longer than the characteristic time of ascent of a bubble in the bed. Here, the mean parameters of the phases will be assumed to be the steady-state values. The state of both phases at a distance x from the gas distributor will be characterized by the mean values of the parameters V, u, n, ε, and v over the cross section.

Given these assumptions, the rate of increase in bubble volume and the rate of rise of a bubble in a bed of fine particles are determined by the following expressions in accordance with [6, 7]:

$$\frac{dV}{dt} = \lambda \varepsilon v V^{2/3},\ \lambda = \frac{28\pi}{9}\left(\frac{3}{4\pi}\right)^{2/3} \approx 3.76;$$

$$u = \frac{dx}{dt} = \gamma V^{1/6},\ \gamma = 1.01\, g^{1/2}\left(\frac{3}{4\pi}\right)^{1/6} \approx 2.49\ \mathrm{m^{1/2}/sec}. \tag{1}$$

The first expression is written on the assumption that the bubble size is always greater than the so-called "equilibrium" size [6, 7]. The value of the coefficient γ in the second expression was obtained in accordance with the test data in [9].

Given the above assumptions, we write the simplest one-dimensional model describing the change in the parameters of the phases in a fluidized bed over its height:

$$\frac{dV}{dx} = \lambda\gamma^{-1}\varepsilon v V^{1/2},\quad \frac{d(nVu)}{dx} = \lambda\varepsilon v n V^{2/3},$$
$$u = \gamma V^{1/6},\ (1 - nV)\varepsilon v + nVu = U, \tag{2}$$
$$\varepsilon = \omega v^{4/15};\ \omega = \left[\frac{18\nu d_1}{g d_p^2 (d_2 - d_1)}\right]^{4/15}.$$

The first equation describes the growth of bubbles as they rise in the bed; the second equation is the balance equation for the volume of the bubble phase, with allowance for bubble growth and the change in rate of rise; the fourth equation is the condition for conservation of the total gas flow in the bed; in this equation, it is considered that in beds of fine particles the rate of bubble rise generally exceeds the fluidization velocity. Thus, each bubble is surrounded by a closed region of gas circulation in which the total gas flow in the coordinate system associated with the bubble is equal to zero (in the beds of fine particles being examined, the bubble volume is practically the same as the volume of the region of closed circulation); the fifth relation, in the case of fine particles, determines the connection between the porosity and gas velocity in the dense phase [10]. We assume that the bubble concentration and volume at the lower boundary of the bed are known, i.e.,

$$x=0,\ n=n_0,\ V=V_0. \tag{3}$$

Solution of problem (2)-(3) shows that the bubble volume and concentration are connected by the relation

$$nV^{1/6}=n_0V_0{}^{1/6}=C=\text{const}.$$

We have the following for the dependence of the bubble volume on the distance to the gas distributor, this dependence being obtained from the solution of problem (2)-(3):

$$x=\frac{1}{\lambda C\varphi^{1/2}}\left\{\ln\frac{(1+\rho^{1/2})(1-\rho_0^{1/2})}{(1-\rho^{1/2})(1+\rho_0^{1/2})}+T\left[E(\rho)-E(\rho_0)\right]\right\};$$

$$E(\rho)=3\rho^{1/3}-3^{1/2}\,\mathrm{arctg}\,\frac{3^{1/2}}{2\rho^{1/3}+1}+\ln(1-\rho)+\frac{3}{2}\ln(\rho^{2/3}+\rho^{1/3}+1), \tag{4}$$

$$\rho=\frac{V}{\varphi},\ \varphi=\frac{U}{\gamma C},\ T=\frac{U^{5/6}C^{1/6}}{\gamma^{5/6}},\ \rho_0=\frac{V_0}{\varphi}.$$

Bubble volume is connected to gas velocity in the dense phase by the following equation from the fourth equation of (2):

$$\rho+(1-T\rho^{5/6})\zeta=1;\ \zeta=\omega v^{19/15}U^{-1}=\varepsilon vU^{-1}. \tag{5}$$

The lower boundary of the stabilization region x_* is determined from Eqs. (4) and (5) at $v = u_{mf}$. At $\zeta \ll 1$ and $\rho_0 \ll 1$, which is typical of large fluidized beds of fine particles for which $v \ll U$, we have

$$\rho_*\approx 1-(1-T)\zeta_*,\ \zeta_*=\omega u_{mf}^{19/15}U^{-1}=\varepsilon_{mf}u_{mf}U^{-1},\ x_*\approx\frac{1}{\lambda C\varphi^{1/2}}\{1{,}39+5{,}55\,T-(1-T)\ln[(1-T)\zeta_*]\}. \tag{6}$$

Equations (4)-(5) are valid at $x \leqslant x_*$. At $x > x_*$, we should set $\rho = \rho_*$ and $\zeta = \zeta_*$, i.e., V, n, v, and ε = const.

Let us present an example. We will examine the fluidization by air ($d_1 = 1$ kg·m^{-3}, $\nu = 1.8\cdot10^{-5}$ m^2·sec^{-1}) of highly porous particles of silica gel of diameter $d_p = 5\cdot10^{-4}$–$2.5\cdot10^{-3}$ m. The initial fluidization velocity $u_{mf} = 7\cdot10^{-3}$ m·sec^{-1}. The gas velocity, calculated for the empty cross section of the unit, is taken equal to $U = 0.03$ m·sec^{-1}. For n_0 and V_0 we take the values $n_0 = 5\cdot10^2$ m^{-3}, $V_0 = 4\cdot10^{-6}$ m^3. We find the values of the parameters in Eqs. (4) and (5) in the form $C = 63.0$, $\varphi = 1.91\cdot10^{-4}$, $T = 5.02\cdot10^{-2}$, $\omega = 1.54$, $\rho_0 = 2.09\cdot10^{-2}$. For $\zeta = \zeta_*$, corresponding to the lower boundary of the stabilization region, $x_* = 1.07$ m. The use of approximate formulas (6) gives $\rho_* = 0.909$, $x_* = 1.2$ m. The estimates obtained are comparable to the test data presented in [1, 2].

Figures 1-3 show the dependences of the parameters of the fluidized bed on the distance to the gas-distributing grate. It should be noted that at $x = x_*$, the derivatives of the phase parameters along the vertical coordinate are very small. This is consistent with the assumption of the constancy of the parameters at $x \geqslant x_*$.

In constructing the curves in Fig. 3, the value of the dimensionless initial volume of the bubbles was assumed to be so small that $\rho_0^{1/3} \ll 1$, and in Eq. (4) we could set $E(\rho_0) = -3^{1/2}\pi$. It should be noted that the last assumption is consistent with many experimental findings; in particular, bubbles close to the gas distributor are so small that they cannot be identified by conventional experimental methods.

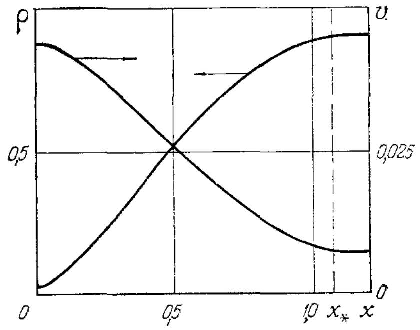

Fig. 1

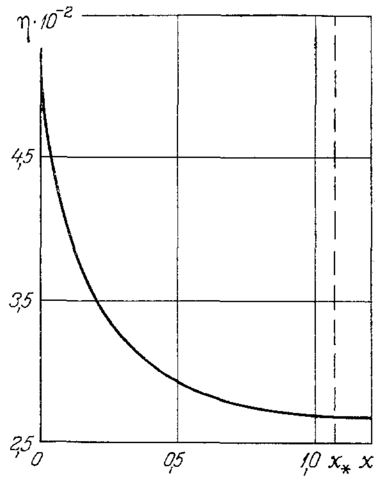

Fig. 2

Fig. 1. Dependence of the dimensionless volume of the bubbles ρ and the gas velocity in the dense phase v on the distance to the gas-distributing grate x with air fluidization of high-porosity particles of silica gel: $u_{mf} = 7 \cdot 10^{-3}$ $m \cdot sec^{-1}$; $U = 0.03$ $m \cdot sec^{-1}$; $n_0 = 5 \cdot 10^2$ m^{-3}; initial volume of bubbles $V_0 = 4 \cdot 10^{-6}$ m^3. v, $m \cdot sec^{-1}$; x, m.

Fig. 2. Dependence of bubble concentration on distance to bed inlet. The values of the parameters correspond to the data in Fig. 1.

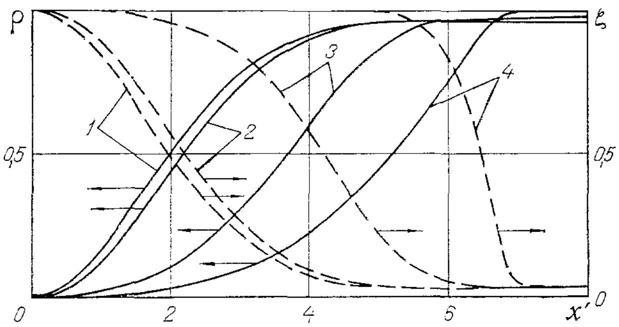

Fig. 3. Dependence of the dimensionless volume of the bubbles ρ and the dimensionless gas flow in the dense phase $\zeta = \varepsilon v U^{-1}$ on the dimensionless distance to the bed inlet $x' = \lambda C \varphi^{1/2} x$ with different values of the parameter T: 1) T = 0.05; 2) 0.1; 3) 0.5; 4) 0.9.

The values of n_0 and V_0 for some types of gas-distributing devices can be calculated using the correlation in [11] for the frequency of bubble formation and the initial volume of the bubbles when discharged from a single opening. In the general case, the value of the constant C can be established by measurements of n and V a sufficient distance from the bed inlet, i.e., in the stabilized region of the bed. This is quite a bit simpler than directly measuring n_0 and V_0 near the gas-distributing grate. This approach is sufficient to establish the distribution of the phase parameters over the height and the lower boundary of the stabilization region at $\rho_0^{1/3} \ll 1$. At greater values of ρ_0, data are also needed on the values of one of the parameters n_0 or V_0.

Now let us analyze the effect of the density and viscosity of the fluidizing gas, the pressure and temperature in the bed, and the density and dimensions of the solid particles on the location and dimensions of the zones and the parameters of the phases in the stabilization zone.

According to the results in [10], the parameter ζ_* in Eqs. (6) is determined by the expression

$$\zeta_* = \frac{\varepsilon_{mf}^{19/4} g d_p^2 (d_2 - d_1)}{18 d_1 \nu U}.$$

It follows from Eqs. (6) that an increase in the density, viscosity, and velocity of the gas leads to an increase in the distance from the lower boundary of the stabilization zone to the gas distributor, i.e., to an increase in the size of the variable-parameters zone. It follows from this that an increase in the temperature and pressure in the bed will also lead to an increase in the size of the variable-parameters zone and, thus, to an intensification of phase exchange processes in the lowest zone of the bed. The size of the variable-parameters zone decreases with an increase in the size and density of the solid particles.

It should be noted, however, that in actual developed fluidized beds the dimensions of the bubbles stabilize at greater heights than the height at which the gas flow stabilizes in the dense phase [2]. This fact evidently has to do with the coalescence of bubbles in the region where the gas flow has already stabilized, with the presence of very fine bubbles in the dense phase, and with other factors.

It also follows from Eqs. (6) that bubble size in the stabilization zone decreases with an increase in the density, viscosity, temperature, and pressure of the gas and a reduction in the dimensions and density of the solid particles. The dependences of the bubble velocity and the percentage of the gas flow passing through the bed with the bubbles on these quantities are similar. The number of bubbles in the bed increases somewhat here. The results obtained are supported by measurements [12-14] of bubble size in the stabilized zone as a function of temperature, pressure, and the dimensions of the solid particles.

In conclusion, we should note that the change in the phase mass transfer coefficients over the height of the bed can be analyzed in accordance with the test results in [15], which gives data on the qualitative character of the dependence of the phase mass transfer coefficient on bubble size.

Allowing for these qualitative laws makes it possible to reduce the volume of studies necessary on pilot and experimental units. In particular, study of the parameters of the bubble phase on a "cold" reactor model can provide reliable low estimates of conversions and selectivities for an actual reactor [16].

NOTATION

$C = nV^{1/6}$; d_1, density of the gas; d_2, density of the material of the solid particles; d_p, particle diameter; n, number of bubbles per unit volume; $T = U^{5/6}C^{1/6}\gamma^{-5/6}$; t, time; U, gas velocity calculated for the empty cross section of the unit; u, rate of bubble rise; u_{mf}, minimum fluidization velocity; V, bubble volume; v, gas velocity in the dense phase; x, distance to gas distributor; $\gamma = 1.01g^{1/2}(3/4\pi)^{1/6}$; ε, porosity of dense phase; ε_{mf}, minimum fluidization porosity; $\zeta = \varepsilon v U^{-1}$, dimensionless gas flux in dense phase; $\zeta_* = \varepsilon_{mf} u_{mf} U^{-1}$; $\lambda = 3.76$; ν, kinematic viscosity of gas; $\rho = V\varphi^{-1}$, dimensionless volume of bubbles; $\varphi = U(\gamma C)^{-1}$; $\omega = \{18\nu d_1/[g d_p^2(d_2 - d_1)]\}^{4/15}$.

LITERATURE CITED

1. C. Chavarie and J. R. Grace, "Performance analysis of a fluidized bed reactor," Parts 1-3," Ind. Eng. Chem. Fundam., 14, No. 2, 75-91 (1975).
2. A. A. Oigenblik, A. S. Zheleznov, M. G. Slin'ko, and V. I. Barannikov, "Study of parameters of the bubble phase in a developed fluidized bed," Dokl. Akad. Nauk SSSR, 238, No. 3, 653-656 (1978).
3. V. E. Soroko, B. T. Vasil'ev, G. Ts. Slavin, and B. A. Konovalov, "Combination reactors with fluidized beds in circulation systems in sulfuric acid production," in: Sixth All-Union Conference on Modeling Chemical and Petrochemical Processes and "Khimreaktor-6" Reactors, Vol. 1, Dzerzhinsk (1977), pp. 24-35.
4. L. Davies and J. E. Richardson, "Gas interchange between bubbles and continuous phase in a fluidized bed," Trans. Inst. Chem. Eng. London, 44, No. 8, T293-T305 (1966).
5. P. N. Rowe and R. Matsuno, "Single bubble injected into a gas fluidized bed and observed by x-rays," Chem. Eng. Sci., 26, No. 6, 923-935 (1971).

6. Yu. A. Buevich, "Movement of bubbles in a fluidized bed," Izv. Akad. Nauk SSSR, Mekh. Zhidk. Gaza, No. 3, 43-51 (1975).
7. Yu. P. Gupalo, Yu. S. Ryazantsev, and Yu. A. Sergeev, "Bubble growth in a fluidized bed with nonlinear phase interaction," Izv. Akad. Nauk SSSR, Mekh. Zhidk. Gaza, No. 6, 66-73 (1977).
8. Yu. A. Buevich and I. M. Tsyrul'nikov, "Evolution of the inhomogeneous state in a fluidized bed," Inzh.-Fiz. Zh., 40, No. 4, 696-704 (1981).
9. J. F. Davidson and D. Harrison (eds.), Fluidization, Academic Press (1971).
10. V. D. Goroshko, R. B. Rozenbaum, and O. M. Todes, "Approximate laws of hydraulics of a fluidized bed and constrained fall," Izv. Vyssh. Uchebn. Zaved., Neft' Gaz, No. 1, 125-131 (1958).
11. D. Kunin and O. Levenshpil', Commercial Fluidization [Russian translation], Khimiya, Moscow (1976).
12. D. Geldart, "The expansion of a bubbling fluidized bed," Powder Technol., 1, No. 6, 355-368 (1968).
13. M. P. Subwari, R. Clift, and D. L. Pyle, "Bubbling behavior of fluidized beds at elevated pressures," Fluidization — Proceedings of the 2nd Eng. Found. Conf., Cambridge, England, Cambridge Univ. Press (1978), pp. 50-54.
14. T. Varadi and J. R. Grace, "High pressure fluidization in a two-dimensional bed," Fluidization — Proceedings of the 2nd Eng. Found. Conf., Cambridge, England, Cambridge Univ. Press (1978), pp. 55-58.
15. V. A. Borodulya, Yu. A. Buevich, and V. I. Dikalenko, "Study of the motion and mass transfer of bubbles in a fluidized bed," Inzh.-Fiz. Zh., 41, No. 4, 678-689 (1981).
16. A. A. Oigenblik, A. S. Zheleznov, and M. G. Slin'ko, "Study of longitudinal mixing and phase exchange in a free fluidized bed," Dokl. Akad. Nauk SSSR, 222, No. 4, 901-903 (1975).

PACKET MODEL OF EXTERNAL HEAT TRANSFER FOR A FLUIDIZED BED

V. A. Borodulya and V. I. Kovenskii

UDC 66.096.5

A modification is proposed for the packet model of external heat transfer of a fluidized bed. The modified model considers heat exchange between the particles and the gas flowing through the packet formed by the particles.

As is known, a fluidized bed is characterized by a discrete structure [1]. The effect of this structure on external heat transfer is evidently best accounted for by the packed model described in detail in [1-3]. According to this model, rising gas bubbles mix with dispersed material and continuously move packets of particles from the core of the bed to the wall of the heat exchanger. Approaching the wall, the packets, in the process of nonsteady heat conduction, give up the heat they accumulated in the core (henceforth, it is assumed that the temperature of the heat exchanger is lower than the temperature of the bed). The packet model most accurately describes heat exchange in a fluidized bed of fine ($d_p \leqslant 0.5$ mm) particles [3].

Another well-known mechanism — convective heat transfer by a filtering gas — determines heat transfer in a bed of coarse ($d_p > 5$ mm) particles [3].

In accordance with the assumption of the additivity of the components of external heat transfer [1-6], transport models corresponding to the limiting cases of fine and coarse particles are used jointly to describe the process in a fairly broad and practicable range of intermediate dispersed-material sizes ($0.5 < d_p < 5$ mm). Here, an increase in particle size changes the relative contribution of the main heat-transfer mechanisms.

A. V. Lykov Institute of Heat and Mass Transfer, Academy of Sciences of the Belorussian SSR, Minsk. Translated from Inzhnerno-Fizicheskii Zhurnal, Vol. 47, No. 5, pp. 789-796, November, 1984. Original article submitted November 4, 1983.

 0022-0841/84/4705-1322$08.50