TWO-PHASE THEORY OF FLUIDIZATION

V. A. Borodulya, Yu. S. Teplitskii, and Yu. E. Livshits

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The effect of the circulational motion of particles on the form of the gas distribution between phases is taken into account.

The separation of a gas flow between phases in a fluidized bed is a problem which is constantly under scrutiny by researchers. Despite the sufficiently large number of works devoted to the two-phase theory of fluidization [1-5], this problem remains far from satisfactory (mainly experimental) solution. The basic difficulties, as noted in [2], are associated with the formulation of a balance equation reflecting the laws that govern the division of gas between the phases and with the experimental measurement of the quantities appearing in this equation (gas velocity in the emulsion phase, porosity of the emulsion phase, velocity of the through gas flow in the bubbles, etc.).

The resolving step to the formulation of a satisfactory equation of two-phase theory was made in [4]. This theory is based on hydrodynamic models of the bubble proposed in [6, 7], which take into account almost all of the basic structural features of the bubble and continuous phases of the fluidized bed that are presently known, but, as analysis shows, do not consider the effect of circulational motion of the particles on the division of the gas flow between phases. This factor is particularly significant in the case of small particles, and leads, as is known [8], to inverse longitudinal mixing of the gas.

In connection with this, the present work poses the problem of taking the existence of circulational (directed) motion of the particles into account in deriving the equations of two-phase theory [4], and comparing some of its consequences with existing experimental data.

To simplify the analysis, two limiting cases will be considered: a) when all the bubbles are fast: $\alpha = (u_b + u_s)\varepsilon_r/u_r > 1$; b) when all the bubbles are slow: $\alpha < 1$. The real fluidization process may be represented simply as the superposition of these cases [4].

a. Fast Bubbles

Taking account of gas transfer in the wake of ascending bubbles,* the equation of two-phase Rowe theory takes the form (Fig. 1a)

$$u = (1 - \varepsilon_b - \varepsilon_c - \varepsilon_w) u_e + \varepsilon_b (nu_r + u_b) + \varepsilon_c (n_1 u_r + \varepsilon_r u_b) + \varepsilon_w (u_r + \varepsilon_r u_b). \quad (1)$$

In writing Eq. (1), the porosity of the bed in the bubble wake and throughout the continuous phase is assumed to be the same. The structure of the gas flow behind the bubble adopted here (Fig. 1a), but not in [4], corresponds to the experimental data of [9].

The equations

$$\varepsilon_b n + \varepsilon_c n_1 = 0, \quad (2)$$

$$(1 - \varepsilon_b - \varepsilon_w) u_s = \varepsilon_w u_b, \quad (3)$$

*The fact that the gas in the bubble wake is constantly replaced by fresh gas from the dense phase does not provide an adequate basis for the assumption, made in [4], that the gas in the wake does not move at all with the bubble. This would only be the case if there were an infinite volume of gas between the bubble wake and the emulsion phase.

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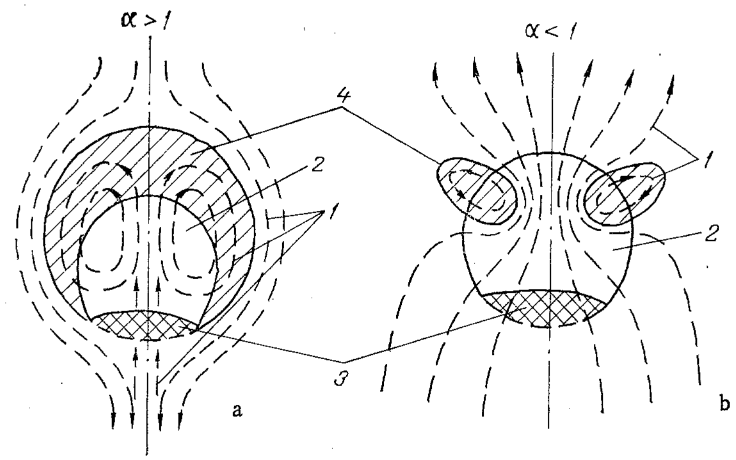

Fig. 1. Diagram of gas flows with respect to bubble: a) fast bubble; b) slow bubble; 1) gas flow line; 2) bubble; 3) wake; 4) cloud (a) and rings (b).

reflect the balance of the circulational flows: the gas in the bubble and its cloud (2) and the solid particles in the given horizontal cross section of the bed (3). Taking Eqs. (2) and (3) into account, together with the relation $u_e = u_r - \varepsilon_r u_s$, Eq. (1) yields

$$u = (1 - \varepsilon_b - \varepsilon_c)\, u_r + \varepsilon_b u_b + \varepsilon_c \alpha u_r. \tag{4}$$

b. Slow Bubbles

In view of the role of the wake in longitudinal gas transfer, the balance equation of the gas flows in any horizontal cross section is written in the form (Fig. 1b)

$$u = (1 - \varepsilon_b - \varepsilon_{rg_1} - \varepsilon_w)\, u_e + \varepsilon_{rg_1} (n_2 u_r + \varepsilon_r u_b) + \\ + \varepsilon_{rg_2} (n_3 u_r + u_b) + \varepsilon_{b_1} (n u_r + u_b) + \varepsilon_w (u_r + \varepsilon_r u_b). \tag{5}$$

Taking account of a relation analogous to Eq. (2)

$$\varepsilon_{rg_1} n_2 + \varepsilon_{rg_2} n_3 = 0, \tag{6}$$

which is a consequence of the continuity of the circulational flows in the rings of a slow bubble, and also of Eq. (3), Eq. (5) takes the form

$$u = (1 - \varepsilon_b - \varepsilon_{rg_1})\, u_r + \varepsilon_b u_b + \varepsilon_{rg_1} \alpha u_r + \varepsilon_{b_1} n u_r. \tag{7}$$

The expressions for the gas filtration rate in Eqs. (4) and (7) are the equations of two-phase theory of fluidization for fast (small particles) and slow (large particles) bubbles. They take into account the influence of the circulational particle motion in Eq. (3) on the division of the gas flow between the phases. The corresponding equations of Rowe theory, taking no account of gas transfer in the bubble wake, are analogous in form, but the expressions $(1 - \varepsilon_b - \varepsilon_c)$ and $(1 - \varepsilon_b - \varepsilon_{rg_1})$ are multiplied by u_e rather than by the factor u_r in Eqs. (4) and (7), and the quantity $u_b \varepsilon_r / u_r$ appears in place of α. It is clear that this modernization may significantly change the partition of gas between phases in small-particle beds, where $\varepsilon_r u_s / u_r \sim 1$. In this case, the quantity $u_e = u_r - \varepsilon_r u_s$ may even be negative when the velocity u_s of downward motion of the particles is sufficiently large (inverse longitudinal mixing of the gas). In the case of large particles, on the other hand, $\varepsilon_r u_s / u_r \sim 0.03$–$0.1$ and $u_e \approx u_r$. Therefore, Eq. (7) is not actually a lot different from the corresponding equation of Rowe theory.

A more detailed analysis of Eq. (7) will now be undertaken. The literature includes evidence, based both on model considerations [10, 11] and on experimental observations [12, 13], that the factor n determining the velocity of through gas flow in a small bubble may differ significantly from the value predicted by the theory of [3], which, as is well known, suggests values of 2 and 3 for n in two-dimensional and three-dimensional bubbles, respectively. The existing experimental results, admittedly scant, give values of the order of

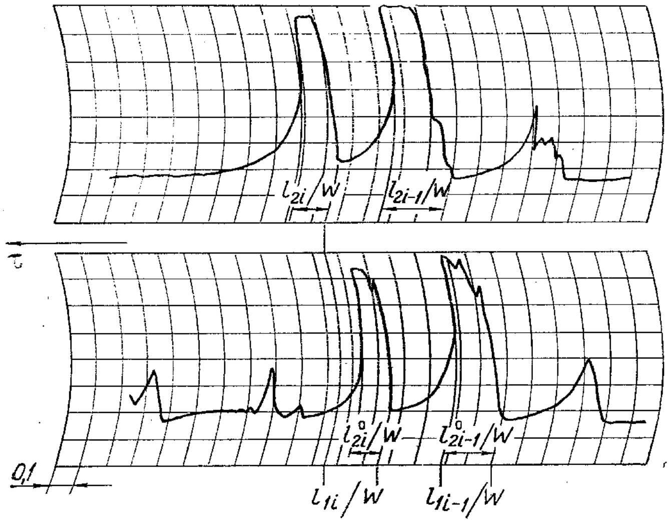

Fig. 2. Characteristic form of sensor signals; $u = 92$ cm/sec. τ, sec.

2-8 for n [12, 13]. In the light of this, an attempt is now made to calculate n (or an associated quantity, the absolute gas velocity in the bubble, $w_b = nu_r + u_b$) from Eq. (7) and compare the result with the available experimental data.

Simple manipulations bring Eq. (7) to the form

$$w_b = \left\{u - (1 - \varepsilon_b) u_r - \varepsilon_{rg_1}\left[\varepsilon_r (u_b + u_s) + \frac{\varepsilon_{rg_2}}{\varepsilon_{rg_1}} u_b - u_r\right]\right\} \Big/ (\varepsilon_b - \varepsilon_{rg_2}). \quad (8)$$

To determine the gas velocity in the bubble from Eq. (8), it is necessary to know ε_b, u_r, ε_{rg_1}, ε_{rg_2}, ε_r, u_b, u_s. The relation between u_r and ε_r is given by the well-known Todes formula for calculating the uniform expansion of fluidized beds [14]. The quantities ε_{rg_1} and ε_{rg_2} may obviously be calculated only theoretically from the value of α (e.g., according to the model of [6]). The quantities u_b, ε_b, and u_s are susceptible to direct experimental measurement.

Since experimental values of u_b and ε_b for large particles are almost completely in the literature, they were determined experimentally in the present work in conditions close to those of [13], where direct experimental data on the value of w_b were obtained.

The experimental apparatus consisted of a metallic column of rectangular cross section 40 × 25 cm and of height 150 cm. The front wall was made of organic glass. Two perforated plates, between which a layer of dense tissue was inserted, formed the gas distributor. The hole diameter of the perforations was 10 mm and the spacing 14 mm. The hydraulic drag of the lattice for air filtration rates in the range 0.50-1.13 m/sec is described by the relation $\Delta p = 4 \cdot 10^3 u^{1.35}$ Pa. As in [13], polystyrene particles ($\rho_s = 1050$ kg/m^3; $u_0 = 60.0$ cm/sec; $\varepsilon_0 = 0.428$; $d = 2.5$ mm) in the form of elliptical cylinders were used as the disperse material. The characteristics of the gas bubbles (vertical dimension and velocity) and also their concentration in the bed were measured by a specially constructed photoelectric sensor, consisting of a system of two pairs (a photoresistor and a small electric lamp) with their optical axes aligned in mutually perpendicular planes and separated vertically by a distance $S = 68.75$ mm (the sensor base). The output signals of the sensor were recorded on tape by a fast-acting recording instrument. The characteristic form of these signals is shown in Fig. 2.

The bubble parameters were determined from the relations

$$u_{bi} = SW/l_{1i}, \quad (9)$$

$$D_{hi} = S(l_{2i} + l^0_{2i})/2l_{1i}, \quad (10)$$

where l_{1i}, l_{2i}, and l^0_{2i} are lengths (mm) measured along the diagram tape (Fig. 2). The bubble concentration was found from the formula

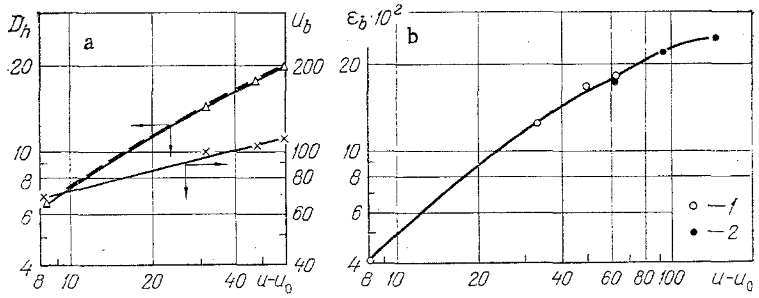

Fig. 3. Gas-bubble characteristics (a) and bubble concentration in bed (b): 1) experimental data of the present work; 2) data of [13]. D_h, cm; u_b, cm/sec; $(u - u_0)$, cm/sec.

$$\varepsilon_b = \frac{1}{2TW} \sum_{i=1}^{N} (l_{2i} + l_{2i}^0), \tag{11}$$

where T is the total signal-recording time of the photosensor. The height of the motionless bed was 38 cm. The sensor was positioned in the central region of the horizontal bed cross section at a distance h = 30 cm from the gas distributor (measured from the axis of the lower photoresistor—electric-lamp pair).

Four modes of air filtration were investigated: u = 68, 92, 108, and 120 cm/sec. The data obtained were treated in accordance with Eqs. (9)-(11) and are shown in Fig. 3 (u_b and D_h were calculated as the arithmetic means of 50-80 individual measurements). Also in Fig. 3, values of ε_b obtained at elevated filtration rates in [13] are shown.

On the basis of the results obtained, values of w_b were calculated from Eq. (8), making some additional assumptions. It was assumed that ε_r and u_r are equal to ε_0 and u_0, respectively, since it follows from the results of [4, 5, 11] that when $h/H \sim 1$ the emulsion phase is in a state of minimal fluidization. In addition, when $u - u_0 \geq 60$ cm/sec, the emulsion phase of the whole bed enters this state [11]. The values of ε_{rg_1} and ε_{rg_2} were calculated from α and ε_b according to Fig. 6 of [4], plotted in accordance with the model of [6]. At filtration rates $u \geq 150$ cm/sec, as shown by visual observations through the transparent front wall, the bed enters a piston mode, the pistons being in the form of gas layers. Hence, in forced modes it is senseless to take account of closed-circulation rings (see Fig. 1b) moving together with the slow bubbles, and therefore when $u \geq 150$ cm/sec it may be assumed that $\varepsilon_{rg_1} = \varepsilon_{rg_2} = 0$.

The results obtained for w_b from Eq. (8) with the noted assumptions are shown in Fig. 4, together with the experimental data of [13], where the absolute gas velocity in the bubbles was measured with a thermoanemometric probe in very similar experimental conditions. The values of ε_b at $u \geq 150$ cm/sec are taken from [13] and are shown in Fig. 3b.

As is evident, the calculated and experimental data are in satisfactory agreement. This confirms (for the present number of experimental points) the validity of the equations of two-phase theory in the form in Eq. (7).

Further investigation of the two-phase theory of a fluidized bed should be conducted within the framework of a systematic experimental investigation of the dependence of ε_b, u_b, w_b, and u_s on the mode $(u - u_0)$, geometric (D_k, H_0, h), and physical (ρ_s, ρ_f, d) parameters of the system. This would offer the possibility of determining, on the basis of Eqs. (4) and (7) and the Todes formula for uniform expansion of fluidized beds, the relative gas and particle velocities and the porosity in the emulsion phase of the bed in relation to various

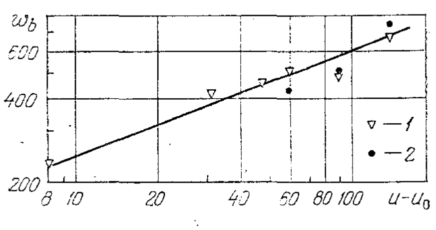

Fig. 4. Absolute gas velocity in bubble: 1) as calculated from Eq. (8); 2) experimental data [13]. w_b, cm/sec.

experimental conditions. The parameters u_r and ε_r have a considerable influence on the flow of reactions of the type gas—solid, and therefore are necessary for the calculation and design of reactors with fluidized beds.

NOTATION

d, equivalent particle diameter; D_k, diameter of the apparatus; D_h, vertical bubble dimension; h, height (along the vertical) above gas distributor; H_0, H, bed height at filtration rates u_0 and u; n, n_1, n_2, n_3, dimensionless (in fractions of u_r) relative gas velocity in bubble, fast-bubble cloud, and outer and inner parts of slow-bubble ring; u_0, u, rate of onset of fluidization and filtration rate; u_r, gas velocity with respect to particles in emulsion phase, referred to emulsion-phase cross section; u_e, absolute gas velocity in emulsion phase, referred to its cross section; u_b, absolute bubble lift velocity; w_b, absolute gas velocity in bubble; u_s, downward particle velocity in continous phase; W, velocity of recording-tape motion; $\alpha = (u_b + u_s)\varepsilon_r/u_r$; ε_0, porosity of bed with $u = u_0$; ε_r, porosity of emulsion phase of bed; ε_b, bubble concentration in bed; $\varepsilon_{b_1} = \varepsilon_b - \varepsilon_{rg_2}$; ε_c, concentration of fast-bubble clouds in bed; ε_{rg_1}, ε_{rg_2}, concentration of outer and inner (with respect to bubble) parts of slow-bubble rings; ε_w, concentration of bubble wakes; ρ_s, ρ_f, particle and gas density.

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