TRANSITIONAL PROCESSES IN FLUIDIZED SYSTEMS

A. P. Baskakov, A. A. Zharkov, and N. F. Filippovskii

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An analysis is made of the transition of a fluidized bed from one steady state to another with a change in fluidizing velocity. The results are compared with experimental data.

Analyzing the transition of a fluidized system from one steady state to another with a change in fluidizing velocity is important not only for description of the phenomenon of pulsative fluidization [1, 2], but also to allow a deeper understanding of the hydrodynamic behavior of conventional and, especially, nonuniform fluidized beds.

We will examine a uniform fluidized bed of identical particles being blown by a fluidizing agent with the velocity w_1 nominally referred to the entire cross section of the apparatus. In the steady state, the weight of the particles of the bed (minus buoyancy) is equal to the resistance to the filtration of the fluidizing agent between the particles:

$$\beta(v_1 - u) = (1 - \varepsilon_1)(\rho_s - \rho_f)\, g. \quad (1)$$

Here, the friction coefficient in the filtration of the fluid through the bed β is assumed to be independent of velocity.

We will gradually increase the velocity of the fluidizing agent to the value w_2. The agent is nearly incompressible at the velocities of interest to us, so we may assume that its velocity increases instantaneously throughout the bed. Since the particles initially are stationary, friction between the flow and particles is increased, which leads to accelerated motion. Each particle of the uniform bed is theoretically acted upon by exactly the same force, meaning that all of the particles will move synchronously, i.e., with a gradual increase in fluidizing velocity, a uniform fluidized bed should rise as a piston. The rate of ascent of the piston is not hard to obtain if we examine the balance of the forces acting on an elemental volume of particles. Since the porosity of the piston ε_1 is agreed to be constant, we write

$$w_2 = v\varepsilon_1 + u(1 - \varepsilon_1) \quad \text{or} \quad v = w_2/\varepsilon_1 - u(1 - \varepsilon_1)/\varepsilon_1.$$

The equation of motion of the particles for the unidimensional case has the form [3]

$$(1 - \varepsilon_1)\,\rho_s \frac{\partial u}{\partial \tau} - (1 - \varepsilon_1)\,\rho_f \frac{\partial v}{\partial \tau} + (1 - \varepsilon_1)(\rho_s - \rho_f)\, g - \beta(v - u) = 0. \quad (2)$$

Substituting the filtration velocity v into (2), we have

$$\frac{du}{d\tau}\left(1 - \frac{\rho_f}{\rho_s}\frac{1 - \varepsilon}{\varepsilon_1}\right) = \frac{\beta}{(1 - \varepsilon_1)\,\rho_s\varepsilon_1}(w_2 - u) - \left(1 - \frac{\rho_f}{\rho_s}\right) g. \quad (3)$$

Integrating this equation with the initial condition $u = 0$ at $\tau = 0$ and considering Eq. (1), we obtain an expression for the piston ascent rate

$$u = (w_2 - w_1)\left[1 - \exp\left(-\frac{g_1\tau}{w_1}\right)\right], \quad (4)$$

where

$$g_1 = g\left(1 - \frac{\rho_f}{\rho_s}\right)\Big/\left(1 + \frac{\rho_f(1 - \varepsilon_1)}{\rho_s\varepsilon_1}\right).$$

In fluidizing with a gas, $g_1 \simeq g$. Equation (4) is obviously also valid with a reduction in velocity.

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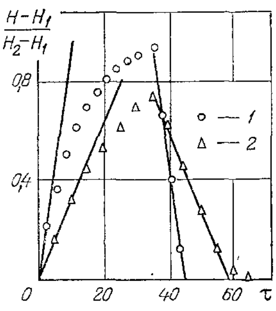

Fig. 1

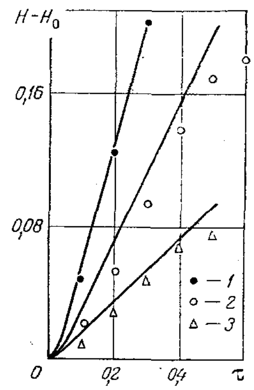

Fig. 2

Fig. 1. Change in height of bed fluidized by a liquid after a gradual increase (at $\tau = 0$) and decrease (at $\tau = 35$ sec) in filtration velocity: the curves represent results calculated from Eq. (4); 1) our data ($d_p = 0.74$ mm; $w - w_k = 0.035$ m/sec); 2) data from [4] ($d_p = 0.2$ mm; $w - w_k = 0.003$ m/sec).

Fig. 2. Change in height of fluidized bed after sudden delivery of fluidizing air: curves represent results calculated from Eq. (4); $d_p = 0.4$ mm; $w_k = 0.22$ m/sec; $H_0 = 0.235$ m; $w_1 = 0$; 1) $w_2 = 0.96$ m/sec; 2) 0.6; 3) 0.38; $H - H_0$, m.

It follows from Eq. (4) that the relaxation time of the piston velocity, i.e., its changes in e times, is equal to

$$\tau_r = w_1/g_1. \tag{5}$$

At $\tau \gg \tau_r$, the velocity of the uniform bed becomes constant and equal to $(w_2 - w_1)$.

The above results were checked experimentally with both liquid and gas fluidization. The liquid fluidized bed was created inside a glass column 70 mm in diameter by fluidizing a bed of glass beads about 0.74 mm in diameter with water. The gas bed experiments were conducted in a rectangular unit 20×250 mm with transparent windows. Air was blown into the unit to fluidize particles of electrocorundum with a mean size of 0.12 and 0.4 mm. The flow rate of the fluidizing agent was regulated by a valve with an electromagnetic drive. The change in bed height with time was determined from a film record. The moment of change in filtration velocity (i.e., opening or closing of the valve) was synchronized with the beginning of filming. The results are shown in Figs. 1 and 2. Also shown for comparison are the results calculated with Eq. (4) (points – experiments, lines – calculation).

It should be noted first of all that the experimental points obtained with a gradual decrease in filtration velocity in the liquid fluidized bed agree well with the data from Eq. (4). Filtration velocity is usually low in fluidizing a fine-grained material with liquid, and relaxation time is of the order of 10^{-2}–10^{-3} sec. Thus, the top boundary of the bed descends almost immediately at the constant velocity $u = (w_1 - w_2)$ after a reduction in the velocity of the liquid from w_1 to w_2, as was seen in the experiment (Fig. 1). Here, a layer of lower porosity ε_2, corresponding to the velocity w_2, is formed on the gas-distribution grid. The thickness of this layer increases with time until the descending upper boundary and ascending lower boundary meet.

The character of the transitional process with a gradual increase in liquid velocity proved to be more complicated. It should be noted that analysis of the transition from the viewpoint of a unidimensional bed (see above) does not explain the change from one porosity ε_1 to another, greater porosity. The experiments show that each fluidizing velocity corresponds to a certain steady state of the bed, i.e., to a certain porosity.

It can be seen from Fig. 1 that the bed of material, in the form of a piston, actually begins to ascend at the constant velocity $u = (w_2 - w_1)$ almost immediately after a sudden increase in filtration velocity, with a space filled with pure liquid theoretically remaining between this piston and the grid. This is confirmed by the results in [4, 5], where the dependence of the height of a liquid fluidized bed on time was found to be linear on the initial section and in agreement with the results found from Eq. (4). However, it is known that the interface between two liquids is unstable if the heavier liquid (in our case, the fluidized bed) is located above the lighter liquid (the pure liquid). Two mechanisms of development of this instability can be represented. With the first (which will be called the micromechanism), during ascent of the piston (bed) individual particles falling away from the bottom surface of the piston, acted upon with a lesser force by the liquid flow due to less physical constraint, remain in the layer of material trailing away from the piston and fall onto the grid. The particles leaving the piston "expose" the next layer (row) which, again because of the lesser constraint, are acted upon with less force by the liquid and allowed to fall to the grate. The fallen particles will create a uniform bed with a higher porosity ε_2, corresponding to the new filtration velocity, on the bed. Thus, the transitional process takes place through the successive falling of particles away from the rising piston. The level at which particle fallout begins rises at a faster rate than the velocity of the piston as a whole. Until this level and the top boundary of the piston meet, the height of the bed will increase at a constant rate equal to the rate of ascent of the top boundary and determined by Eq. (4). The rate of increase in the height of the bed will subsequently begin to decrease. This was seen in the experiment (Fig. 1). Depending on the specific conditions (size and properties of particles being fluidized, velocity and properties of fluidizing agent), the duration of the linear and nonlinear sections of increase in layer height varies [4, 5]. The micromechanism of instability development is typical of systems in which homogeneous fluidization occurs, since the difference in the densities of the fluidized bed and pure fluid are very great in this case. It follows from Fig. 1 that the transitional process develops slowly in such systems and lasts tens of seconds. It can almost be considered quasisteady. This is the reason for the good agreement between the empirical results and the analysis in [5], which was based on the assertion that the porosity of the uniform bed and the fluidizing velocity (the Richardson–Saki formula) are connected and the proposal that the porosity of the bed on the grate assumes a value corresponding to the new fluidizing velocity at the very beginning of the transitional process.

Another mechanism leading to instability (referred to as the macromechanism) is also possible, however. This mechanism is typical of systems tending toward inhomogeneous fluidization. In this case, the ascending bed, in the form of a piston, will "collapse" at isolated places into the gas that is pushing it. The gas will then penetrate the bed at these places, forming a "discrete phase" in the medium – which is assumed to be of constant porosity, in accordance with the two-phase model. It is interesting to note that in this case the top boundary of the bed will be lifted in accordance with Eq. (4) until it is no longer penetrated by the discrete phase, regardless of the exact form of the phase (bubbles, a gas piston occupying part of the bed cross section, etc.). This is confirmed by the agreement between the experimental and theoretical data in Fig. 2.

The duration of the transitional period τ_t also usually considerably exceeds the relaxation time $\tau_t \gg \tau_r$ in an inhomogeneous fluidized bed. In this case, the value of τ_t is easily evaluated from the following considerations. In accordance with the two-phase model of an inhomogeneous fluidized bed, the height of the bed H is related to the rate of ascent of bubbles w_b in the steady-state regime as follows [6]

$$\frac{H}{H_0} = \frac{w_b}{w_b - w_k(W-1)} . \tag{6}$$

It follows from (6) and (4) at $\tau_t \gg \tau_r$ that

$$\tau_t = \frac{H_0}{w_b - w_k(W-1)} . \tag{7}$$

The above analysis permits us to understand certain fluidization phenomena. Many authors [7, 8] assert that, with delivery of gas into a bed through a porous or perforated grid, bubbles are formed from the gas cavity which is formed on this grid. It is known that fluidization is usually realized in an oscillatory state [9]. Here, an increase in velocity causes the bed to rise as a piston and results in the formation of a gas interlayer, with bubbles subsequently being formed from the latter. The interlayer disappears as velocity is decreased. This is the pattern seen in visual observations of gas cavity development.

Strictly speaking, the above is valid only if a porous plate is used. With a perforated grate, the gas cavity is formed after streams from individual openings have merged, i.e., above the grate. In our experiments, with 10 mm between grate openings 2 mm in diameter, the height of the cavity was 10–15 mm.

NOTATION

d_p, mean particle size; g, acceleration due to gravity; H, bed height; H_0, initial bed height; u, particle velocity; v, true velocity of fluidizing agent; w, velocity of agent referred to the empty cross section of the apparatus; w_b, bubble ascent rate; w_k, initial fluidizing velocity; W, number of fluidizations; β, friction coefficient in filtration of fluidizing agent through bed; ε, porosity of bed; ρ_s, ρ_f, density of particles and gas; τ, time.

LITERATURE CITED

1. A. P. Vladislavlev et al., "Fluidized bed in a high-frequency pulsating flow," Teor. Osn. Khim. Tekh., 12, No. 5, 722-726 (1978).
2. P. G. Alfredson and I. D. Doig, "Behavior of pulsed fluidized beds," Trans. Inst. Chem. Eng., 51, 232-246 (1973).
3. N. F. Davidson (ed.), Fluidization [Russian translation], Khimiya, Moscow (1974).
4. L. T. Fan, J. A. Schmitz, and E. N. Miller, "Dynamics of liquid–solid fluidized bed expansion," AIChE J. 9, No. 2, 149-153 (1963).
5. P. L. Slis, Th. W. Willemse, and H. Kramers, "The response of the level of a liquid fluidized bed to sudden change in the fluidizing velocity," Appl. Sci. Res., Sect. A, 8, 209-218 (1959).
6. S. S. Zabrodskii, Hydrodynamics and Heat Transfer in a Fluidized Bed [in Russian], Gosénergoizdat, Moscow (1963).
7. A. P. Baskakov, Rapid Nonoxidative Heating and Heat Treatment in a Fluidized Bed [in Russian], Metallurgiya (1968).
8. D. R. McGraw, "The development of a mechanism for gas particle heat transfer in shallow fluidized beds of large particles," Chem. Eng. Sci., 32, No. 1, 11-18 (1977).
9. V. A. Borodulya, Yu. A. Buevich, and V. V. Zav'yalov, "Theory of relaxational oscillations in a granular bed fluidized by a gas," Inzh.-Fiz. Zh., 30, No. 3, 424-433 (1976).

EFFECT OF BOUNDING SURFACES ON POROSITY DISTRIBUTION IN A GRANULAR MEDIUM

V. N. Koleskin, P. G. Shtern, S. V. Turuntaev, G. N. Abaev, and E. K. Popov

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An experimental study is made of porosity distribution in a granular bed close to one of the surfaces bounding it. It is shown that wall deformation reduces the nonuniformity of porosity and the velocity profile of the gas flow.

The flow of a liquid or gas in reactors with a stationary granular bed of catalyst is significantly affected by the properties of the bed [1], particularly the distribution of local porosity over the cross section of the apparatus.

Experimental results have been obtained on porosity distribution inside stationary beds of uniform spherical and cylindrical particles [2-5], but such results are very limited in volume and do not permit a full evaluation of the effect on porosity distribution of such properties of the bounding surfaces as deformability, roughness, etc. The present work thus attempts a more detailed study of the porosity distribution of a granular material near a flat boundary and the effect of the properties of the wall on this distribution.

Experimental Unit and Method

The experimental unit consisted of a rectangular vessel 400 × 200 × 200 mm made of organic glass. The vessel had a double bottom, with holes joining the filling chamber with the working volume containing the granular material. The fluid was delivered from a buret through the chamber and into the bed, filling the cavities in the latter. The height of ascent of the fluid in the granular bed h was fixed with a reading microscope and we established a physically small volume ΔV_2 for averaging the porosity of the thin bed. Bed porosity was determined as

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