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AVERAGE VELOCITY OF BUBBLES IN A FLUIDIZED BED CONTAINING PACKING MATERIAL

D. M. Galershtein, A. I. Tamarin, S. S. Zabrodskii, and V. P. Borisenko

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The expansion of a fluidized bed with various packings in columns of several dimensions was measured. The average velocity of the bubbles and the influence of the packing parameters on this velocity were estimated.

When a bed of dispersed material is fluidized by a gas, gas cavities (bubbles) rise continuously through it; the existence of these is due to the fundamental instability of the system [1, 2]. All the gas passing through the bed is divided into two fluxes, one of these constituting the bubbles, while the other incorporates the gas filtering between the suspended particles. The two fluxes or flows differ chiefly as regards their time of existence in the bed and their conditions of contact with the dispersed material. With increasing filtration velocity the flow of the bubble phase increases, while the second flow varies very little [3]. This type of flow has a deleterious effect on the intensity of the gas–particle exchange processes and reduces the efficiency of a number of technological processes (catalytic reactions, sorption, etc.).

In order to increase the homogeneity of the system, a collection of immobile elements (packing) may be placed in the fluidized bed; these partly break up the bubbles and greatly increase the efficiency of technological processes [4-6, 10, 17]. The hydrodynamics of a layer containing such packing material have been studied by a number of research workers in recent years, and a considerable proportion of the results have been presented in review articles [4, 5, 7]. Even so, information on this subject is still somewhat sketchy and largely of a qualitative nature.

In this paper we shall set out the results of an experimental investigation into the effects of various forms of packing on the mean velocity of the bubbles in a fluidized bed. This investigation extends earlier-published data [8, 11].

The experimental method was based on a two-phase model of the bed, according to which [9]

$$u_{\mathrm{ba}} = u - u_0 + u_{\mathrm{b}}. \tag{1}$$

This model allows us to relate the bubble velocity to the expansion of the bed [9, 10] by means of the equation

$$u_{\mathrm{b}} = (u - u_0) H_0 (H - H_0)^{-1}. \tag{2}$$

Since the position of the upper boundary of the bed is hard to measure accurately, especially for high gas velocities, we used [11] the well-known relationship between the height of the bed and its mean porosity:

A. M. Lykov Institute of Heat and Mass Transfer, Academy of Sciences of the Belorussian SSR, Minsk.
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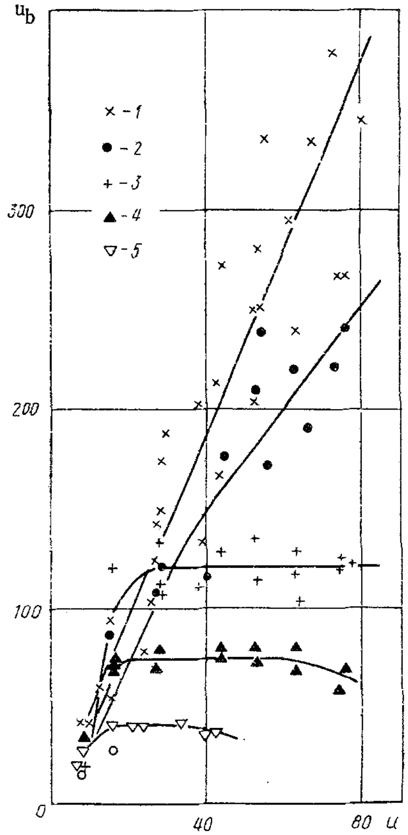

Fig. 1. Relative bubble velocity as a function of the gas filtration velocity: 1) free bed; 2) group of rods with longitudinal ribs (Table 2, item 8); 3) cylindrical spirals, l_p = 3.06 cm; 4), 5) cylindrical spirals, l_p = 0.57 cm. Dispersed material: 1, 2, 3, 4) quartz sand 1; 5) silica gel (see Table 1); u in cm/sec.

$$H \cdot H_0^{-1} = (1 - \varepsilon_0)(1 - \varepsilon)^{-1},$$

whence from (2) we obtain

$$u_b = (u - u_0)(1 - \varepsilon)(\varepsilon - \varepsilon_0)^{-1}. \qquad (3)$$

This equation enables us to determine the mean relative velocity of the bubbles in terms of easily measurable quantities.

A number of modifications of the two-phase model have also been considered in the literature [12, 13]. We note that the model described in [12] contains quantities for which no values have yet been obtained, while that proposed in [13] leads to the same results as Eq. (3) for high velocities, which are of chief practical interest.

The rate of gas flow through the continuous phase has been discussed in a number of papers, mostly reviewed in [12]. Most of the authors correctly assume that the velocity of the gas through the continuous phase is usually a little greater than u_0, being equal to ku_0, where $k > 1$. However, there are at present no reliable values of k available. It was pointed out in [3, 14] that k increases with diminishing density and size of the particles. It was assumed in [3] that after reaching filtration velocities at which intense bubble formation began, the value of k fell rapidly.

Let us estimate the possible systematic distortion associated with the fact that Eq. (3) implies $k = 1$. Clearly, the systematic distortion will increase with increasing k. In order to treat the problem specifically we make use of the data of [14] relating to particles with d_{av} = 0.186 mm and u_0 = 1.39 cm/sec (the silica gel used in the experiments described below has similar parameters); in this case $k = 1.7$. For such a k value, Eq. (3) should overestimate u_b systematically by no more than 10% for $N \geq 10$.

TABLE 1. Characteristics of the Materials

Dispersed materials	d_{av}, mm	u_0, cm/sec	ε_0	Notation in Table 2 and Fig. 1
Quartz sand	0,23	6,0	0,45	q.1
Quartz sand	0,15	4,0	0,43	q.2
Corundum	0,12	4,3	0,53	c.
Silica gel	0,19	2,0	0,5	s.

TABLE 2. Characteristics of Packings and Bed

Serial No.	Packing	Characteristic dim.		ε_p	l_p, cm	Dispersed material (Table 1)	Diameter of bed in cm
		nomenclature	dim. cm				
1	Vertical rods with ribs along a helical line [15]	Width of rib	2,0	0,91	2,45	q.1	30
2	Packs of plates, neighbors inclined in opposite directions	Height of pack	5,5	0,96	1,59	s.	30
			3,5	0,94	1,14	q.1	30
			5,5	0,97	2,07	s.	30
			3,5	0,96	1,8	q.1; s.	30
3	Packing as item 2 but perforated plates	Height of pack	5,5	0,97	2,38	q.1; s.	30
			5,5	0,98	2,65	s.	30
4	Cylindrical spirals	Diameter of spiral	5,5	0,97	3,06	q.1, s.	30
			2,0	0,95	0,57	q.1, s.	30; 15
			2,0	0 95	0,57	q.2, c.	15
			1,0	0,92	0,35	c., s.	15
5	Horizontal grids	Distance between grids	5,0	0,989	2,9	q.1	70
			4,0	0,991	3,7	q.1	70
6	Rushing rings with slots	External diameter	2,1	0,760	0,385	q.1; s.	15
			2,4	0,68	0,29	q.1; s.	15
			2,4	0,68	0,29	q.2; c.	15
7	Group of vertical rods	Diameter of rods	2,0	—	5,8	q.1; s.	30
8	Packing as item 7, rods with three longitudinal ribs	Width of rib	2,0	—	2,7	q.1; s.	30
9	Packs of vertical plates (chord packing)	Height of pack	5,5	—	2,04	q.1; s.	30
10	Packing as item 9, plates inclined to one side	Height of pack	5,5	—	1,92	q.1; s.	30

The quantities entering into Eq. (3) were measured in the following way in our experiments.

Values of u_0 and ε_0 were measured in the ordinary manner with a smoothly falling filtration velocity.

The porosity was measured from the pressure drop between two points in the middle of the bed:

$$\Delta p = (1 - \varepsilon)\, l\rho. \tag{4}$$

In order to increase the accuracy of measurement of the mean pressure drop, choke elements (capillaries) were installed in the pulse lines leading to the manometer; these gave the measuring system a time constant of 1 min.

The rate of air filtration was measured in a high proportion of the experiments by means of a standard diaphragm, but in some of the experiments a lemniscate collector was employed.

Under the conditions of our experiments, the maximum error in determining u_b calculated from the errors committed in measuring the filtration velocity and pressure drop in the interior of the bed was 30% for low filtration velocities (N < 5) and no greater than 20% for N > 10.

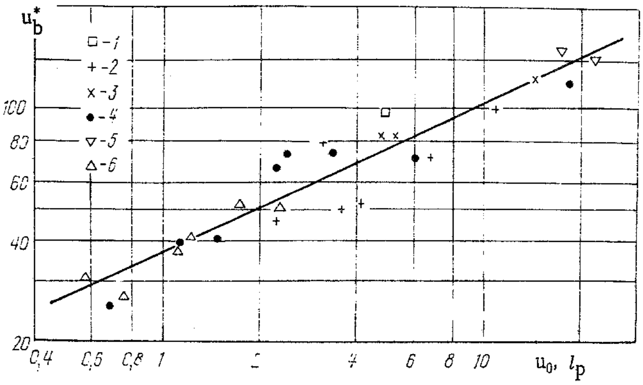

Fig. 2. Generalization of experimental data regarding stabilized bubble velocities. 1, 2, 3, 4, 5, 6) Serial Nos. in Table 2; $u_0 l_p$ in cm^2/sec.

The dispersed material in our experiments included two fractions of quartz sand, corundum, and silica gel. The characteristics of these materials are shown in Table 1.

The experiments were carried out with fluidized beds 15, 30, and 70 cm in diameter. The height of the stationary filling of dispersed material lay in the range 23-27 and 30-35 cm in the first two cases, while in the 70-cm-diameter bed it equalled 56 or 86 cm.

The investigations were carried out with packings of various constructions. As a generalized characteristic of the packing we used a parameter proportional to the mean hydraulic diameter of the channels formed by the packing:

$$l_p = (V_{bed} - V_p) F_p^{-1}. \tag{5}$$

The characteristics of the packings and the experimental conditions are indicated in Table 2.

Typical experimental data obtained in the 30-cm-diameter bed are presented in Fig. 1. These data demonstrate the following characteristics of the mean bubble velocity.

1. In the free bed u_b increases almost linearly with increasing gas filtration velocity.

2. A packing of vertical rods with longitudinal ribs (Table 2, item 8) slightly reduces u_b, although the relative velocity of the bubbles continues increasing monotonically with u. An analogous picture is presented on using packings such as items 7, 9, and 10 in Table 2. Evidently, these four packings only reduce the transverse motion of the bubbles, which slightly retards their merging (a similar effect was noted in [7] for vertical rods); however, these packings do not prevent the merging of the bubbles along the vertical.

3. In the case of packings in the form of a free filling of cylindrical spirals the mean relative bubble velocity does not increase above a certain value u_b^*. Stabilization of the relative bubble velocity also occurs in the other packings. The characteristics of these packings and the corresponding experimental conditions appear in items 1-6 of Table 2.

In generalizing the experimental data for packings stabilizing the bubble velocity, we started from the following considerations. According to the principles of the two-phase model, the motion of the gas in the bed is characterized by two velocity scales: the velocity of the bubbles and the velocity of the gas in the intergranular channels (u_0).

The geometry of the system may be characterized by the parameter l_p. We therefore sought the dependence of u_b^* on l_p and u_0.

Figure 2 generalizes the experimental data for the packings in items 1-6 of Table 2. The experimental points may be approximated by the following equation with a mean square deviation of 17%:

$$u_b^* = 37 (u_0 l_p)^{0.45}. \tag{6}$$

The foregoing investigations enable us to estimate the role of bed size in retarded systems. Thus, a packing of cylindrical spirals with $l_p = 0.57$ cm was used in beds of diameters 15 and 30 cm, and a packing of horizontal wire grids was placed in beds of diameter 70 cm and heights $H_0 = 56$ and 86 cm. In all cases, for the same values of (u_0/l_p), the corresponding values of u_b^* either coincided or differed by less than 15%. It is an extremely important point that Eq. (6) generalizes data obtained in beds of different sizes. All this indicates that effective packings may, in principle, stabilize u_b, and hence the dimensions of the bubbles, not only with respect to filtration velocity, but also with respect to space.

NOTATION

u is the air filtration velocity, referred to the empty cross section of the bed, cm/sec;
u_0 is the velocity at the onset of fluidization, cm/sec;
u_{ba}, u_b are the absolute and relative average (in space and time) bubble velocities, cm/sec;
H, H_0 are the heights of fluidized and nonfluidized beds, cm;
ε is the porosity of the bed;
ε_0 is the porosity of the bed for u_0;
u_b^* is the stabilized relative bubble velocity, cm/sec;
l_p is the mean diameter of channels in the packing, cm;
V_{bed}, V_p are the volumes of the bed and the packing, cm^3;
F_p is the surface area of the packing in V_{bed}, cm^2;
d_{av} is the average particle diameter of the dispersed material, mm;
ρ is the particle density, kg/m^3;
Δp is the pressure drop in a section of bed of height l;
$\varepsilon_p = V_p/V_{bed}$ is the porosity of packing.

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