PROBLEMS OF THE HYDRODYNAMICS OF A BED FLUIDIZED BY DUST-LADEN GAS

V. S. Nosov and V. K. Maskaev

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The results of an experimental investigation of the effect of the degree of dust contamination of the air on the hydraulic resistance and velocity of the start of fluidization of a bed of spherical particles with an average diameter of 2.00-9.35 mm are presented. Recommendations are given for calculating the frictional resistance coefficient and the velocity of the start of fluidization.

A bed fluidized by a gas-suspension flow can be of considerable interest for a number of technological processes using a solid heat exchanger. In this case the heat-transfer processes in the bed will be intensified additionally, since the heat can be conducted with the heated particles circulating through the bed. In addition, the use of a gas suspension as a fluidizing medium will permit a gradual change of its density and therefore a change of the homogeneity of the bed, the velocity of the start of fluidization, and other characteristics.

An experimental study of the hydrodynamics of a bed fluidized by air containing corundum particles with an average diameter of 60 μ in a quantity up to 12.5 kg/kg was carried out on the device shown in Fig. 1. The device is open with respect to both the gas and solid phases. The piston compressor 1 delivered air through receiver 2, drier 3, and flowmeter 4 into aeration chamber 7. The solid phase was also delivered there from hopper 5 by screw feeder 6. After passing through the hydrodynamic stabilization section 8, the mixture entered the working section 11. The air and dust were separated in the storage hopper 13. Thus the time of continuous operation of the device was determined by the travel time of the entire solid phase from hopper 5 to hopper 13 and amounted to 1-2 h. The cylindrical working section was made of organic glass. A steel grate with holes of diameter 1.5 mm and spacing 5 mm had a cross section for passage of the gas suspension of 7.07%.

The following quantities were measured during the experiments: the air flow rate; the temperature and pressure of the air before the flowmeter; the temperature and pressure of the dust-laden air in the working section under the grate, directly over the grate, and on the upper boundary of the bed. The flow rate of the solid phase was controlled by the number of revolutions of the feeder screw and the height of the stationary and fluidized beds was determined visually from a scale applied on the transparent wall of the working section.

The characteristics of the solid spherical particles used for creating the bed are given in Table 1, which also includes the experimental values of the velocity of the start of fluidization of the bed upon blowing it with pure air.

TABLE 1. Characteristics of Solid Particles

Material	d_s, mm	ρ_s, kg/m³	G, kg, H = 70 min	φ	w'_g, m/sec
Pea	5,65	1390	0,127	1,64	2,14
Steel	2,00	7870	0,900	2,30	2,40
			H=110 mm		
Steel	6,31	7870	0,65	1,85	5,44
Alundum	4,54	3630	0,231	4,35	2,72
Alundum	5,52	3630	0,250	2,77	3,52
Alundum	9,35	3630	0,222	4,52	4,92

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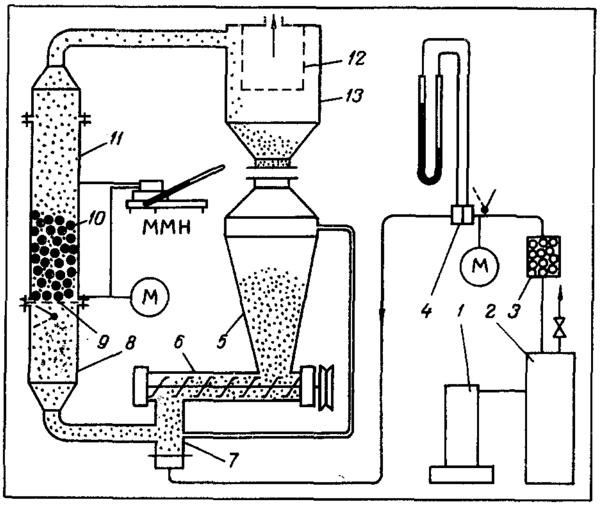

Fig. 1. Diagram of the experimental device: 1) compressor; 2) receiver; 3) drier; 4) flowmeter; 5) hopper; 6) screw feeder; 7) aeration chamber; 8) hydrodynamic stabilization section; 9) grate; 10) fluidized bed; 11) working channel; 12) filter; 13) storage hopper.

Visual observations showed that, on blowing the stationary layer, the gas suspension easily passes through the grate and bed, filling all voids of the bed. Circulation of dust within the bed was not observed, but is was noted that at all filtering velocities the flow rate of dust was higher close to the walls of the channel than on the axis. The latter is related with the nonuniform porosity of the bed over the channel cross section.

It was noted during filtering of the dust-laden flow through the fluidized bed that with an increase of the dust concentration the quality of fluidization improved, the gas bubbles burst, and the pulsations of the bed particles decreased. At a dust concentration in the air of 10-12 kg/kg the fluidized bed of particles with $d_s = 5.65$ mm and $\rho_s = 1390$ kg/m^3 became practicaly homogeneous with a rather clear-cut upper boundary.

A decrease of the air flow rate at a constant dust flow rate at first led to settling of the bed and then to the formation of a stationary dust film on the upper boundary of the bed, with the exception of the wall boundary zone. Near the wall the dust continued to pass through the bed. With a further decrease of the air flow rate the free section remaining near the wall for passage of the gas suspension narrowed and simultaneously the thickness of the stationary film increased toward the grate. At the actual air velocity in the bed, approximately equal to the velocity of suspension of the corundum dust particles, the dust particles covered the entire channel cross section. At this instant the bed was again fluidized. Intense agitation of the large spherical particles over the entire volume of the bed began, i.e., the formation of a unique "two-fraction fluidized bed" occurred at a very low air filtering velocity.

This interesting fact of secondary fluidization of the bed is explained by redistribution of the true velocity of the air over the channel section when the velocity near the channel axis increases considerably when the wall boundary zone is clogged with dust. Consequently, in all other cases the true air velocity and probably the dust flow rate in the wall boundary zone is higher than the average velocity over the channel section. The latter is confirmed by the authors of other investigations [1, 2].

Experiments to determine the hydraulic resistance in the bed were carried out with a change of filtering velocity from 1.0 to 5.5 m/sec and dust concentration in the air from 0 to 12.5 kg/kg. The velocity of the start of fluidization was determined visually and from an experimental curve plotted in $\Delta P - w$ coordinates.

In Fig. 2a the relative velocity of the start of fluidization is shown as a function of the dust concentration. As we see, with an increase of concentration, fluidization of the bed blown by a gas-suspension flow begins at velocities considerably less than in the case of blowing the bed with pure air. In this case the relative value of the velocity of the start of fluidization for a given value of the dust concentration in the air flow was the same for all particles except large alundum particles with a diameter of 9.35 mm (lower curve in Fig. 2), which is probably explained by the effect of the channel walls.

On the basis of the experimental data obtained we can recommend the following empirical equation for calculating the velocity of the start of fluidization:

$$\frac{Re'_{mix}}{Re'_g} = 1 - 0.1\,\mu^{0.7}\,. \tag{1}$$

The equation is valid for a change of μ from 0 to 12.5 kg/kg, d_s from 2 to 6.31 mm, and ρ_s from 1390 to 7870 kg/m^3. The value of Re'_g is determined by the equation

$$Re'_g = 0.367\left[\frac{\varepsilon^3}{\varphi(1-\varepsilon)}\,Ar\right]^{0.57}, \tag{2}$$

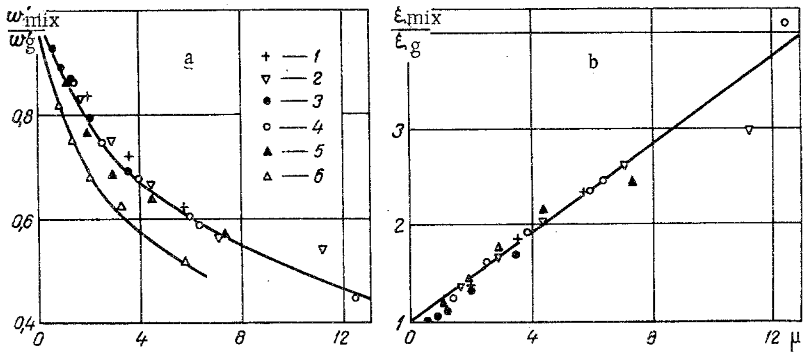

Fig. 2. Relative velocity of the start of fluidization (a) and relative frictional resistance coefficient (b) as a function of concentration: 1) pea, $d_s = 5.65$ mm; 2) steel, $d_s = 2.00$ mm; 3) steel, $d_s = 6.31$ mm; 4) alundum, $d_s = 4.54$ mm; 5) alundum, $d_s = 5.52$ mm; 6) alundum, $d_s = 9.35$ mm.

where φ is the form factor, which in our experiments varied from 1.64 to 4.54 (Table 1). The value of φ was determined, according to [3], for conditions of the start of fluidization on blowing the bed with pure air by the equation

$$\varphi = 2\,\frac{G d_s g}{\rho_g f \left(w_g'\right)^2 H \xi_g} \cdot \frac{\varepsilon^3}{(1-\varepsilon)^2}. \tag{3}$$

The value of the frictional resistance coefficient for the pure gas was assumed equal to

$$\xi_g = 11.6\,\mathrm{Re}_g^{-0.25}. \tag{4}$$

To determine ξ_m we also used Eq. (3) in which in place of w_g' we substituted the value of the velocity of the start of fluidization of the bed by the dust-laden air. The ratio of the frictional resistance coefficients ξ_{mix}/ξ_g as a function of the dust concentration is presented in Fig. 2b. As we see from Fig. 2b, the value of this ratio is directly proportional to the concentration and does not depend on the diameter, density of the material of the particles, and velocity of the fluidizing agent. The equation of a straight line approximating the experimental points with a 10% error has the form

$$\xi_{mix} = \xi_g(1 + 0.23\mu). \tag{5}$$

Equation (5) is valid for the same conditions as Eq. (1) and can be recommended for determining the hydraulic resistance of the bed being blown with dust-laden gas at the start of fluidization.

NOTATION

d	is the diameter;
H	is the height of the stationary bed;
f	is the cross sectional area of the channel;
w'	is the velocity of the start of fluidization;
G	is the weight of bed;
ε	is the porosity;
ξ	is the frictional resistance coefficient;
φ	is the form factor;
ρ	is the density;
μ	is the mass concentration of dust in the flow;
$\mathrm{Re}' = w'd/\nu_g$	is the Reynolds number;
$\mathrm{Ar} = (gd^3/\nu_g^2)(\rho_s - \rho_g)/\rho_g$	is the Archimedes number.

Subscripts

s denotes solid particles;
mix denotes dust-laden air (mixture);
g denotes pure air.

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