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FEATURES OF THE MOTION OF A CIRCULATING BED IN A LARGE-VOLUME SPHERICAL PACKING

G. I. Kovenskii

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An experimental study has been made of the motion of a circulating fluidized bed in a column filled with a large-volume spherical packing. It has been established that, once the feed of the loose material ceases, the motion of the bed continues up to the total removal of the disperse medium from the column with a velocity acquired in constant circulation. A procedure based on the processing of experimental results has been proposed for calculation of the operating conditions of apparatuses in which the indicated motion can be employed.

The homogeneity of a free gas-fluidized bed makes it difficult to employ this bed as a working medium for processes with interphase heat and mass exchange [1]. One method of changing the structure of the bed is the arrangement of bulk packings in the volume of a disperse system and creation, in such a manner, of a retarded bed [1]. One is able to attain the most homogeneous structure of such a bed by using an appropriately selected large-volume spherical packing [2, 3]. The formation of bubbles, the breakthrough of a gas, and the attrition of particles are totally excluded [3].

A great advantage of fluidized-bed apparatuses is the simplicity of organization of a change of dispersed material in the reaction column with the use of flow devices [1], which is necessary for the continuity of the technological process.

The circulation of a dispersed material through internal flows and a free fluidized bed has repeatedly been investigated. Designs of flow devices have been developed and methods of their calculation have been proposed [4]. At the same time, the influence of packings of all kinds on the flow of a circulating loose material has not been studied, in practice. Therefore, the present work seeks to experimentally study the influence of a large-scale spherical packing on the motion of a dispersed material in its circulation in the column and external flow devices.

A diagram of the setup for investigation of the motion of a circulating fluidized bed in a large-volume spherical packing is presented in Fig. 1. It consists of column 1, gas-distributing box 2, two bunkers 3, system of feeding of a dispersed material into the column 4, and two trays 5 for discharge of a loose material from the column into the bunkers. The column of the setup was filled with packing spheres 6 through its entire height.

The influence of a large-volume packing on the expansion of a fluidized bed has a pronounced extremum character [3]. According to the procedure of [3, 5], we have selected the optimum packing-sphere size ensuring the largest expansion of the stationary bed for the loose material employed in the experiments.

The values of the parameters of the setup are given in Table 1.

The fluidizing agent (carrier gas) was fed into the column via the tube 7 welded into the bottom of the gas-distributing box and covered with a perforated cone-shaped cap at the top.

Air (injection gas) which was fed via the metal tubes 4 welded into the lower part of the bunkers 3 was used for feeding the dispersed material. Sand arrived at the column via cross-shaped holes cut in the walls of the gas-distributing box 2. Compressed air at a pressure of 6 atm was fed into the setup.

The flow rate of the carrier gas and the injection gas was monitored by rotameters.

The experiment was as follows:

1. A stationary retarded fluidized bed was created in the column for a prescribed velocity of the carried gas u; the bed filled the column through its entire height.

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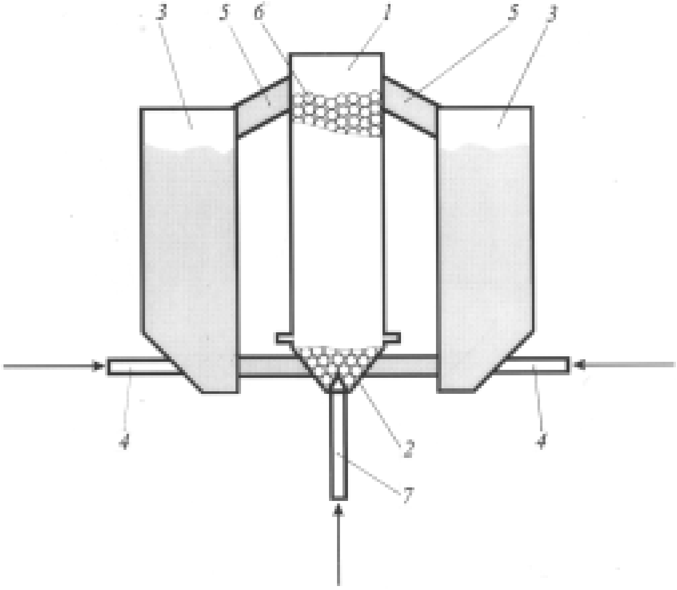

Fig. 1. Diagram of the experimental setup.

TABLE 1. Parameters of the Experimental Setup

Parameter	Value	Parameter	Value
Column — plastic (organic glass):		Particles — sand	
H_p	0.8 m	ρ_s	2600 kg/m^3
D_1	0.108 m	d	$2.3\cdot10^{-4}$ m
S	0.009 m^2	Ar	1160
Packing — steel spheres:			
D	0.015 m	Re_{mf}	0.73
ε_p	0.36	u_{mf}	0.05 m/sec
Gas — air under normal conditions:			
ρ_g	1.19 kg/m^3	Re_t	30
μ	$1.79\cdot10^{-5}$ kg/(m·sec)	u_t	1.95 m/sec
ν	$1.5\cdot10^{-5}$ m^2/sec	ε_0	0.41

2. Circulation of a loose material through the packing was organized by feeding the material with the use of the injection gas. The velocity of the fluidizing agent in the column was equal to $u + u_i$, and the flow rates of the solid phase fed via the tubes 4 and removed via the trays 4 were identical. The flow rate of the dispersed material in the trays 5 was monitored by the sampling method. The height of the charge in the bunkers 3 and the height of the bed in the column 1 remained constant.

3. Once a stable circulation was formed, we turned off the flows of the loose material and the injection gas into the column. The velocity of the fluidizing agent in the column was equal to u, just as in the first stage of the experiment. However the initial state of the fluidized bed was not restored — the disperse-medium column continued its motion through the column with a packing. The lower boundary of the disperse-medium column was well defined. The velocity of its motion was noted visually. The entire granular material was poured into the bunkers 3 via the trays 5. Its flow rate in the trays 5 remained constant and equal to the flow rate in stationary circulation.

TABLE 2. Results of the Experiments

Measured parameters				Calculated parameters				
u, m/sec	u_i, m/sec	Q, kg/sec	w^*, m/sec	u^*, m/sec	w, m/sec	ε_b	ε	N, W
0.55	0.12	0.024	–	1.53	–	–	0.92	0.49
0.55	0.23	0.028	0.027	1.53	0.01	0.88	0.92	0.77
0.55	0.32	0.034	0.04	1.53	0.0144	0.9	0.92	1.07

It should be noted that such behavior of the bed retarded by the large-volume packing differs from the behavior of the free circulating bed, which, because of the intense mixing, immediately becomes stationary after the cessation of feeding the loose material.

The results of the experiments are presented in Table 2.

The average porosity of the bed moving in the column was calculated from the formula

$$\varepsilon_b = 1 - \frac{Q}{Sw\rho_s}, \tag{1}$$

whereas the porosity of the homogeneous fluidized bed [6] was calculated from

$$\varepsilon = \left(\frac{\mathrm{Re} + 0.02\,\mathrm{Re}^2}{\mathrm{Re}_{mf} + 0.02\,\mathrm{Re}_{mf}^2} \right)^{0.21}. \tag{2}$$

As is clear from Table 2, the porosity of the fluidized bed ε_b moving in the large-volume packing approaches the porosity of the homogeneous disperse medium in value.

The high homogeneity (observed in the experiments) of the granular bed and its ordered motion after the interruption of circulation are determined by the influence of the large-volume packing on the flow of the fluidizing agent and hence the behavior of the disperse medium. Numerous experiments [3] have shown that the measure of such an action can be the power of viscous dissipation in filtration of air through the packing:

$$N(u) = 0.7\,[\nu + 0.1uD] \left(\frac{8u}{D} \right)^2 \rho_g \varepsilon_p H_p, \tag{3}$$

where $u \in [2u_{mf}, 20u_{mf}]$.

The flow rate of the circulating dispersed material has proved to correlate with the dissipation power indeed (Table 2). The corresponding dependence is approximated (with an average error of 2.7%) by the relation

$$Q = 0.029\,(N(u + u_i)/H_p)^{0.5}. \tag{4}$$

The ordered motion of the disperse-medium column after the cessation of circulation is, apparently, due to the conservation of the particle velocity acquired during the stable circulation of the bed. It is apparent that the mixing of the granular material is negligible in motion in the packing and the particles do not collide with each other, with the packing spheres, or with the column walls.

The phenomenon established is of practical importance. It enables one to facilitate and accelerate the discharge of granular material from the apparatus. We can propose the following procedure based on the processing of experimental data for calculation of the parameters of the setup with the possibility of discharging the solid reagent:

(a) the packing-sphere diameter ensuring the largest expansion of the stationary fluidized bed is determined according to the procedure of [5] for the dispersed material used;

(b) the dissipation power in the packing N is calculated from formula (3) for the selected diameter of packing spheres;

(c) the mass flow rate of the circulating granular material is calculated from formula (4) on the basis of the value of the dissipation power in the packing N.

NOTATION

$\mathrm{Ar} = \frac{gd^3}{\nu^2}\frac{\rho_s - \rho_g}{\rho_g}$, Archimedes number; d, diameter of particles, m; D, diameter of a packing sphere, m; D_1, diameter of the column of the experimental setup, m; g, free-fall acceleration, m/sec^2; H_p, height of the bed and the packing, m; N, power of viscous dissipation in filtration of air through the packing, W; Q, flow rate of the circulating dispersed material, kg/sec; $\mathrm{Re} = ud/\nu$, Reynolds number; S, cross-sectional area of the bed, m^3; u, velocity of the carrier gas per flow area of the column, m/sec; u_1, velocity of the injection gas per flow area of the column, m/sec; $u^* = u/\varepsilon_p$, velocity of the carrier gas in the packing, m/sec; u_{mf}, velocity of the onset of fluidization, m/sec; u_t, free-fall velocity, m/sec; w, velocity of motion of the dispersed flow per flow area of the column, m/sec; $w^* = w/\varepsilon_p$, velocity of the dispersed flow in the packing, m/sec; ε, porosity of the homogeneous bed, calculated from the Todes formula; ε_0, porosity of the charge; ε_b, porosity of the bed in the experiment; ε_p, porosity of the packing; μ, dynamic viscosity of the gas, kg/(m·sec); ν, kinematic viscosity of the gas, m^2/sec; ρ_s, density of particles, kg/m^3; ρ_g, density of the gas, kg/m^3. Subscripts: 0, charge; b, bed; g, gas; i, injection; mf, onset of fluidization; p, packing; s, particle material; t, free fall; *, velocity of the gas or the disperse medium in the packing.

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