INVESTIGATION OF THE PARAMETERS OF MOTION OF PARTICLES IN A FLUIDIZED BED USING RADIOACTIVE ISOTOPES

III. Mean Particle Velocity

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The paper presents the relations between the mean particle velocity in the vertical (longitudinal) direction of a monodisperse fluidized bed and the parameters of the bed and the gas. A dimensionless empirical formula is given for calculating the mean absolute particle velocity.

There is a direct connection between motion of the solid particles in a fluidized bed and the transfer of energy and mass, mixing, viscous properties, and other phenomena. Investigators have therefore given considerable attention to the study of the processes in the bed itself [1–6]. As yet, however, no quantitative relation has been established between the particle velocity and the bed parameters, when the latter are uniquely assigned. The present paper attempts to derive this relationship.

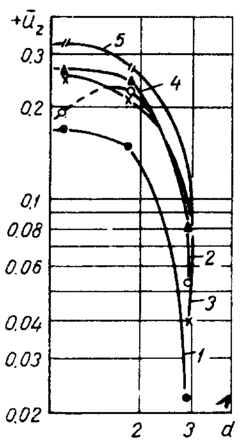

Fig. 1. Dependence of particle velocity $+\bar{u}_z$ (m/sec) on diameter d (mm): 1) and 2) with H_0 = 43 mm; 3) and 4) 89; 5) 133; 1) and 3) with w_p = 1.15 m/sec^{-1}; 2), 4), and 5) 1.35.

The equipment on which the experiments were carried out and the experimental technique have been described previously [7, 8].

The monodisperse beds investigated contained an industrial aluminum-silicate catalyst in the form of spheres screened into three classes on 3.0–2.5; 2.0–1.6; 1.2–1.0 mm screens, the equivalent values of particle diameter d_e in each class being: 2.81; 1.82; 1.09 mm, and the weights being: $14.6 \cdot 10^{-5}$; $4.16 \times \times 10^{-5}$; $0.091 \cdot 10^{-5}$ N (newton). One of the particle beds was labeled with cobalt 60 isotope and had diameter and weight values of 2.88; 1.82; 1.10 mm and $13.7 \cdot 10^{-5}$; $4.26 \cdot 10^{-5}$; $0.101 \cdot 10^{-5}$ N, respectively.

The critical velocities of each of the beds examined were 0.84, 0.56, and 0.31 m · sec^{-1}.

1. ANALYSIS OF THE PARTICULAR RELATIONS

In a large series of 120 tests the initial parameters were varied successively—air velocity (up to fluidization numbers of five), particle diameter, static height of bed H_0/D = 0.25–1.5, and type of gas distributor (porous and perforated). The tests were conducted in equipment 172 mm in diameter.

An oscilloscope record of the signals—coordinates of a labeled particle—was made following preliminary stabilization of the fluidization regime. A test would go on for 10–30 min, with recording every 3–5 min of the height, state of the bed, and readings of the instruments (mass flow rate meter and differential manometer).

The oscillograms were read under ten-fold magnification. The time interval between points to be processed varied between 0.1 and 1.0 sec, and was determined by the frequency of the curves, this depending on the gas stream velocity. Reduction of the oscillograms according to the time intervals adopted gave linear coordinates with respect to the axes x, y, z, and also gave the displacement of the labeled particle in the fluidized bed as a function of time. With successive reduction for each test, the three components of "local" velocity were obtained: longitudinal (vertical) $\pm u_z$, transverse u_φ, and radial $\pm u_R$, during the chosen time intervals. Then for each regime with all the parameters fixed we calculated the mean values $\pm u_z$; $\pm\bar{u}_R$, $\pm\bar{u}_\varphi$, and hence the absolute particle velocity

$$\bar{u} = \sqrt{\bar{u}_R^2 + \bar{u}_\varphi^2 + \bar{u}_z^2}$$

and the transverse (horizontal) component

$$\bar{u}_t = \sqrt{\bar{u}_R^2 + \bar{u}_\varphi^2}\ .$$

The partial relations between the particle velocity components and the initial system parameters were obtained in the form of a table; we present here only the graphical relations for the mean longitudinal (vertical) upward particle velocity $+\bar{u}_z$.

Influence of particle diameter. It may be seen from Fig. 1 that rapid increase of the function $+\bar{u}_z = f(d_e)$ with w_p = const, H_0 = const commenced on going from

$d_e = 2.81$ mm to $d_e = 1.82$ mm, while, in the range 1.82 to 1.09 mm, the rate of increase in velocity is reduced. For small bed height (curve 2) and large operating gas velocity (fluidization number 4.36) there was a maximum of the function investigated.

An explanation of the relations obtained should be sought in two physical phenomena—variation of particle mass and the nature of mixing. Reduction of the diameter (particle mass) at unchanged gas velocity produced an increase in particle velocity. However, at small particle diameters or large gas velocities, particle circulation sets in, causing equalization of the longitudinal velocity values.

Influence of bed height. Increase of the mean longitudinal (vertical) particle velocity occurred with increase of bed static height, according to the law $+\bar{u}_z \sim$ $\sim H_0^n$, in the range $0.25 < H_0/D < 1.0$ (as is seen from Fig. 2), while for $H_0/D > 1.0$ a tendency was observed for the particle velocity to fall as a function of bed static height. Naturally, one would expect various factors to be influential here, both kinematic and hydrodynamic. The first relates to acceleration and deceleration phenomena, varying with bed height, and the second to the role of gas jet energy and the nonuniformity of fluidization always connected with large bed height.

The motion of a particle in the bed, and therefore, of the whole mass of particles, is nonuniform. The particle velocity varies in magnitude and direction during the entire period of motion from the free surface of the bed to the distributor and back [7, 8]. In each half period there is an accelerating section and a decelerating one. In beds in which the height is less than the diameter, the extent of the accelerating section increases with increase of the ratio H_0/D, at the same time as the maximum local and mean velocities increase. When a particle approaches the distributor and the free surface of the bed, the deceleration stems from variation of direction of the half-period trajectory, at the end of which the local velocity is zero. In thicker beds, when there is averaging, the particle velocity ($+\bar{u}_z$ or $-\bar{u}_z$) increases during a half-period, in comparison with the thin bed, since the fraction of decelerating sections over the whole path decreases.

The energy of a gas jet issuing from the apertures of the distributor is expended in the bed itself when $H_0/D > 1$. This indicates that in thick beds ($H_0/D > 1$), the influence of jet energy on increase of particle velocity is considerably less than in thin beds.

The uniformity of fluidization becomes worse with increase of bed height, which progressively promotes convergence of the particles, with formation of groups. The decrease of particle velocity with $H_0/D > 1.0$ is probably due to the fact that large agglomerates are formed; to move and break these down requires additional expenditure of the kinetic energy of the gas stream.

The lack of uniqueness of the general relation $+\bar{u}_z =$ $= f(H_0)$ allows us to propose the existence of an optimum bed height for which the mean particle velocity is greatest.

It should be stressed that the local particle velocities should vary according to a law with a maximum, i.e., should first increase from zero, and later diminish with height z.

Influence of gas velocity. We may evidently conclude from Fig. 3 that the particle velocity increases with increase of gas stream velocity.

The particle motion in the bed begins with the gas stream velocity somewhat exceeding the critical [7, 8]. The most pronounced increase in velocity is observed with increase of gas velocity up to a certain value, after which the rate of increase of particle velocity becomes less dependent on the gas velocity. These two rates of growth of $+\bar{u}_z = f(w_p)$ are separated by the "optimum" gas velocity, whose value depends in a quite definite way on the particle diameter and the static height of the bed. The ratio of the "optimum" fluidization velocity to the critical lies in the range 2–3. It should be noted that a fluidization number of 2–3, for variation of the rate relationship of the two velocities, agrees with the optimum recommendations for fluidization number in heat and mass transfer [1, 9, 10]. Also the particle velocities are identical at the different fluidization numbers, and different particle velocities are observed at identical fluidization numbers. These observations confirm the well-known conclusion regarding the limitation of the fluidization number as a hydrodynamic characteristic of the fluidized bed.

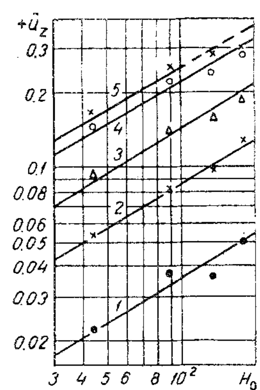

Fig. 2. Dependence of particle velocity $+\bar{u}_z$ (m/sec) on static height H_0 (mm) of bed: 1), 2), and 3) with $d_e = 2.81$ mm; 4) 1.82; 5) 1.09; 1), 4), and 5) $w_p =$ $= 1.15$ m/sec^{-1}; 2) 1.35; 3) 1.55.

A physical explanation of the two rates in the interrelationship of $\bar{u}$ and w_p may be given when account is taken of the already mentioned variation of the nature of circulation of solids with increase of gas velocity: there is a regime of smooth fluidization and a vortex regime (the "boiling" bed regime), the particles being accelerated during linear displacement in the first regime, and in the second during circulation. In the process of circulation of the mass of solid phase, some part of the gas energy is expended, so that it does not come into play in the linear acceleration of the particles.

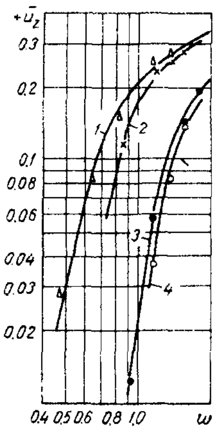

Fig. 3. Dependence of particle velocity $+\bar{u}_z$ (m/sec) on gas velocity w (m/sec), for H_0 = 89 mm: 1) at d_e = = 1.09 mm; 2) 1.82; 3) and 4) 2.81; 1), 2), and 4) with a porous distributor; 3) a perforated distributor.

In all the tests at identical conditions (H_0, w_p, etc.), the absolute values of particle velocity with ascending flow, $+\bar{u}_z$, were greater than with descending flow, $-\bar{u}_z$. The absolute values of mean radial velocities in particle motion from the wall to the center, and from the center to the wall, were close to each other in the majority of the tests. The velocity in the longitudinal direction, $\bar{u}_z$, was always about twice that in the transverse direction, $\bar{u}_t$. The maximum absolute particle velocities were close to the air velocity, calculated for the total cross section of the equipment, while the minimum was zero.

Role of the distributor. In the equipment with a perforated distributor (apertures of 2 mm, cross section fraction 2.1%), other conditions being equal, the particle velocity was greater than in the equipment with a porous (ceramic) distributor, as may be seen in Fig. 3, by comparing curves 3 and 4.

This may evidently be attributed to the greater kinetic energy of the jets issuing from the drilled apertures, $M_g w_p^2/2$: above the porous distributor the jets are less powerful because the mass of gas in each is small. This is also confirmed by visual observation—when operating with the porous distributor the splashing is considerably less than with the perforated distributor, and the fluidization is more uniform.

2. GENERALIZATION OF TEST DATA

An attempt to generalize the data in the form $\bar{u}/w_p = f(\mathrm{Re}_\varepsilon)$ led to an interesting result, in qualitative agreement with the relation Nu = f(Re) for heat exchange between a bed and a surface [10]. The curves $\bar{u}w_p = f(\mathrm{Re}_\varepsilon)$ are similar for all the beds investigated, and are characterized by the presence of a maximum (Fig. 4), which occurs for a fluidization number of 2.75. This experiment may also confirm the explanation of the causes of the heat-transfer maximum in the bed-surface system given in [11], but we did not pursue this point.

The optimum fluidization regime for various technical processes must be established in conformity with the specific requirements: maximum heat-transfer coefficient, necessary time of heat treatment, reaction time, etc. The quantitative evaluation of each of the technical parameters is usually connected with an experimental value of the fluidization number, i.e., with the gas velocity. Objectively, however, this kind of parameter is connected with the state of the bed, i.e., with the intensity of motion of the particles in it.

This is precisely the reason why, in generalizing our test data, we used a relation linking particle velocity with gas velocity and with the bed parameters, assigned under conditions of uniqueness.

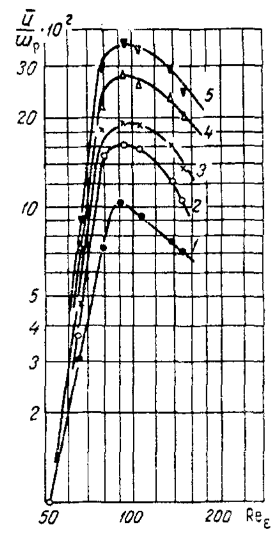

Fig. 4. Dependence of the ratio $\bar{u}/w_p$ on $\mathrm{Re}_\varepsilon \equiv w_p d_e/\varepsilon\nu$ with H_0 = 133 mm, d_e = = 1.09 mm: 1) for $\bar{u}_R/w_p$; 2) $\bar{u}_\varphi/w_p$; 3) $\bar{u}_t/w_p$; 4) $\bar{u}_z/w_p$; 5) $\bar{u}/w_p$.

For the physical process of motion of each particle we may write the dynamical equation

$$\vec{P} \equiv \vec{F}_{g_i} + \vec{P}_{f_i} = \vec{G}_i + \vec{F}_{c_i} + \vec{I}_i, \qquad (1)$$

in which the most significant forces are the gravity force G_i and the inertia force I_i, since $G \cdot . P$. during the motion. This is precisely why the dimensionless group should be formulated as the ratio of inertia to weight forces. The group is the Froude number

$$\mathrm{Fr}_\tau = u_\tau^2/gd. \qquad (2)$$

The physical process of motion of the gas through the bed is characterized by degeneracy of the inertia forces beyond the point at which the critical fluidization velocity is attained. In fact, in a gas stream in a static bed, there are forces of friction of the gas against the particles and the wall, and a particle form drag force P_f. Increase of velocity leads to an increase of these forces, but there is always equilibrium between them and the inertia force I, if we examine the vector sum of the forces upon interaction of the gas with one of the particles:

$$\vec{I}_i = \vec{P}_{f_i} + \vec{F}_{\mu_i}.$$

After a state of fluidization is reached, the number of forces acting increases (gravity and impact forces are added), but the d'Alembert principle applies for each of the particles.

For an entire bed, composed of N particles,

$$\sum_{i=1}^{N} \vec{I}_i = \sum_{i=1}^{N} \vec{P}_{f_i} + \sum_{i=1}^{N} \vec{F}_{\mu_i} + \sum_{i=1}^{N} \vec{F}_{c_i} + \sum_{i=1}^{N} \vec{G}_i. \quad (3)$$

At the beginning of fluidization and up to entrainment $\sum_{i=1}^{N} \vec{I} = 0$, since the bed as a whole is at rest and $\Sigma F_c = 0$ under an inelastic collision. Thus, the inertia forces are insignificant for the process of fluidization of a bed of particles; only the weight and viscosity forces are appreciable. A parameter which reflects the dynamic interaction in the bed is obviously the group

$$Ar/Re \equiv \frac{Re\,\Delta\rho}{Fr\,\rho} \equiv \frac{gd^2}{\nu w_p} \frac{\Delta\rho}{\rho}, \quad (4)$$

which is a measure of the ratio of the weight and viscosity forces [12].

In the logarithmic anamorphosis $Fr_T \sim Ar/Re$, all the test data are classified according to equidistant lines (each of which satisfies its own value of bed height) and particle diameter, and on each curve there are two laws (regimes) of dependence of the dimensionless groups: the first is $Fr_T = c_1(Ar/Re)^{-m_1}$, where $m_1 = const$; the second is $Fr_T = c_2(Ar/Re)^{-m_2}$, with $m_2 \neq const$ and $m_1 > m_2$.

We managed to bring all the curves together by introducing the additional geometrical parametric criteria H_0/d_e and H_0/D. Treatment using this curve in the first regime with $m_1 = const$ yielded a formula for determining the mean absolute particle velocity in a fluidized bed, $\bar{u}$

$$Fr_g = [8.0 \cdot 10^{-6} (Ar/Re)(H_0/d_e)^{0.87} : (H_0/D)^{1.05}]^{-6.25},$$
$$1.0 \cdot 10^5 < (Ar/Re)(H_0/d_e)^{0.87} : (H_0/D)^{1.05} < 3.2 \cdot 10^5, \quad (5)$$
$$0.25 < H_0/D < 1.5, \quad 16 < H_0/d_e < 250.$$

We may make the following general remarks regarding (5)*:

1. The quantity H_0/d_e varied from 16 to 90 for $d_e = 2.81$ mm, from 25 to 140 for $d_e = 1.82$ mm, and from 45 to 250 for $d_e = 1.09$ mm, the lower limits referring to $H_0/D = 0.25$, and the upper—to $H_0/D = 1.5$ (intermediate values may easily be obtained by interpolation).

2. The quantity w_p is greater than w_K, since when $w_p = w_K$ the particle velocity equals zero.

3. The formula has been verified for a fluidized bed of spherical particles of a narrow initial class and a system air—aluminum-silicate catalyst.

It is obvious that, given the contemporary level of mathematical description of the fluidization process, it cannot be stated categorically that the chosen criteria have been rigorously assembled, i.e., that they fully describe the phenomenon of particle motion in a bed. The formula is empirical, and should not be regarded as being complete as regards structure.

$Ar \equiv \frac{gd_э^3}{\nu^2}\frac{\Delta\rho}{\rho}$ – Archimedes number; $Fr_т \equiv \frac{\bar{u}_т^2}{gd_э}$ – Froude number for a particle; $Re \equiv wd_e/\nu$ – Reynolds number; D – equipment diameter; d_e – equivalent particle diameter; F_μ – viscous force due to interaction of particle and gas; F_c – particle impact force; G – particle gravity force; H_0 – height of stationary bed; I – particle inertia force; i – ordinal index of particle in bed; P_f – particle form drag force due to gas flow.

REFERENCES

1. O. M. Todes and A. K. Bondareva, Khimicheskaya nauka i promyshlennost, no. 2, 1957.
2. A. K. Bondareva, Motion of Particles and Heat Transfer in a Suspended Bed [in Russian], Thesis, Moscow, 1958.
3. F. A. Rosenthal, "Heat and mass transfer in vaporization processes," Sb. trudov ENIN AN SSSR, Izd. AN SSSR, 1958.
4. J. C. Chu, Fluidization [Russian translation], ed. D. F. Othmer, Gostoptekhizdat, Moscow, p. 25, 1958.
5. M. Leva, Fluidization [Russian translation], Gostoptekhizdat, Moscow, 1961.
6. L. Massimilla and J. A. I. Westwater, Ch. E. Journ., 6, no. 1, 1960.
7. N. B. Kondukov, A. N. Kornilaev, I. M. Skachko, A. A. Akhromenkov, and A. S. Kruglov, IFZh, no. 7, 1963.
8. N. B. Kondukov, A. N. Kornilaev, A. A. Akhromenkov, I. M. Skachko, and A. S. Kruglov, IFZh, no. 7, 1964.
9. I. M. Fedorov, Theory and Calculation of the Drying Process in the Weightless State [in Russian], GEI, Moscow-Leningrad, 1955.

*The formula in [13] was given without allowance for the exponent of the parameter H_0/d_e, and it is therefore better to use our formula (5) for calculations.

10. N. N. Varygin, The Fluidized Bed as a Cooling Medium for Heat Treatment of Metals [in Russian], Author's abstract of thesis, Moscow, 1961.
11. S. S. Zabrodskii, Hydrodynamics and Heat Transfer in the Fluidized Bed [in Russian], GEI, Moscow, 1963.
12. A. A. Gukhman, Introduction to the Theory of Similarity [in Russian], Moscow, 1964.
13. A. N. Kornilaev and N. B. Kondukov, collection: Conference on the Use of the Fluidized Bed in the Chemical Industry [in Russian], Izd. Khimiya, Moscow-Leningrad, p. 18, 1965.

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