EXPANSION OF AN INHOMOGENEOUS FLUIDIZED BED

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The expansion of an inhomogeneous fluidized bed in a column 700 mm in diameter is determined experimentally. A calculated correlation is obtained on the basis of a two-phase model.

In designing large industrial installations incorporating fluidized beds it is essential to know the expansion of the bed in order to determine the working height of the apparatus and the positioning of the heat-transfer surfaces.

Published information relates mainly to homogeneous fluidization [4, 13]. Experimental data on the expansion of inhomogeneous layers (beds) is somewhat scattered and there are no generalizing correlations. A curve reflecting the expansion of a fluidized bed as a function of the diameter of the particles presented in [6] was based on experiments in small laboratory installations and was unsuitable for the design of large reactors. Only isolated information exists for column diameters of 300 [3, 5] and 512 mm [2].

The aim of the present investigation was to obtain information regarding the expansion of an inhomogeneous fluidized bed in a column 700 mm in diameter and to generalize the existing data on the basis of the two-phase theory of fluidization.

According to the two-phase theory of fluidization [7] the relative lift rate of the bubbles averaged over the bed is given by the equation

$$\bar{v}_{\text{B}} = \frac{u - u_0}{H - H_0} H_0. \tag{1}$$

It was shown in [1] that the relative velocity of the bubbles averaged over the cross section of the fluidized bed depended on the height measured to the gas-distributor lattice and was given by

$$v_{\text{B}} = q_0 [g(u - u_0) h]^{1/3}. \tag{2}$$

From Eq. (2) we obtain the following for $\bar{v}_{\text{B}}$:

$$\bar{v}_{\text{B}} = \frac{1}{H} \int_0^H v_{\text{B}} dh = q_1 [g(u - u_0) H]^{1/3}. \tag{3}$$

Substituting (3) into (1), we have

$$p - 1 = \frac{q_2}{p^{1/3}} \mathrm{Fr}^{1/3}. \tag{4}$$

Remembering the expansion of the free layer is slight ($1 < p < 1.3$) for $p^{1/3}$ we obtain the estimate $1 < p^{1/3} < 1.09$. Approximately putting $p^{1/3} \approx 1$ we may simplify Eq. (4):

$$p - 1 = q_2 \mathrm{Fr}^{1/3}. \tag{5}$$

We have thus obtained a mutual relationship between the relative expansion of the layer ($p - 1$) and the dimensionless excess gas-filtration velocity (Fr) for an inhomogeneous fluidized bed.

Special experiments were carried out in order to refine Eq. (5).

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TABLE 1. Experimental Conditions for Obtaining the Expansion of an Inhomogeneous Fluidized Bed

Literature source	Material	u_0, cm/sec	Particle diam., mm	Column dimen., mm	Stat. layer ht., mm	Measuring method
[2]	Quartz sand	3,1 7,1 13,8	0,15 0,26 0,35	512	430	from pressure drop
[3]	Quartz sand	1,4, wide fr. 1,37, narrow fr.	0,13 0,10	310	49—291 54—219	visually
[5]	Quartz sand	6,0	0,23	300	300	from pressure drop
[9]	Glass spheres	7,1	0,38	100	166	visually
[10]	Quartz sand	6,0	0,23	100	170	visually
[11]	Glass spheres Al catalyst	0,73 0,73	0,074 0,041	75—150	370	x rays
[12]	Glass spheres	8,0	0,19	90	110	capacitive transducer

The experiments were executed in a column of diameter 700 mm. The gas-distributing lattice was made of two perforated sheets with apertures 10 mm in diameter. The step between the apertures (25 mm) ensured an active cross section of 11.5%. A layer of felt 9 mm thick was compressed between the perforated sheets. As dispersed material we used quartz sand with a wide spread of particle size averaging 0.23 mm in diameter (u_0 = 4.6 cm/sec, ε_0 = 0.45). For a mesh cell size of 315, 200, 160, and 100 μ the weight residue on the mesh was 10.3, 58.2, 24.6, and 6.9%. The quantities u_0 and ε_0 were determined subject to a smooth reduction in gas velocity, using the standard method. The height of the stationary filling varied from 40.7 to 136 cm. Fluidization was effected by air at t = 3-10°C. The air-flow rate was measured with a diaphragm and differential manometer to an error of no greater than 3%.

The expansion of the bed was determined from the equation

$$p=\frac{H}{H_0}=\frac{1-\varepsilon_0}{1-\varepsilon}\,. \tag{6}$$

For a fluidized bed we have [13]

$$\Delta p=\rho_p\, g\,(1-\varepsilon)\,\Delta x. \tag{7}$$

Equation (6) may thus be rewritten

$$p=\frac{\Delta p_0}{\Delta p}\,. \tag{8}$$

Equation (8) was used for the experimental determination of the expansion of the bed. The pressure drop was measured with a cup-type manometer. Chokes were placed in the pulse lines leading to the manometer, and the time constant of the measuring system was of the order of 1 min. This enabled us to determine the average pressure drops in the core of the fluidized bed to an error of no greater than 2%.

In order to choose the disposition of the pressure-measuring points we measured the pressure drops between points 1-2, 2-3, and 3-4 for H_0 = 136 cm. The vertical step between the points was 200 mm. Point 1, closest to the gas distributor, lay at a distance of 120 mm from the latter. The experiments showed that for all gas flow rates the pressure drops between the points indicated could not be reliably distinguished from one another. In subsequent experiments the pressure measurements were therefore carried out at points 1-2 for H_0 = 40.7 cm and at points 2-3 for H_0 = 66 or 85 cm.

The relative error in determining p from Eq. (8) was no greater than 4%.

Using Eq. (5), the experimental data were plotted in a logarithmic coordinate system, with the relative expansion of the bed (p − 1) in the vertical direction and the dimensionless excess gas velocity (Fr) in the horizontal direction. The experimental points lay satisfactorily on inclined straight lines with a slope of 1/3. Experiments carried out at different H_0 are grouped around several equidistant straight lines. This indicates a dependence

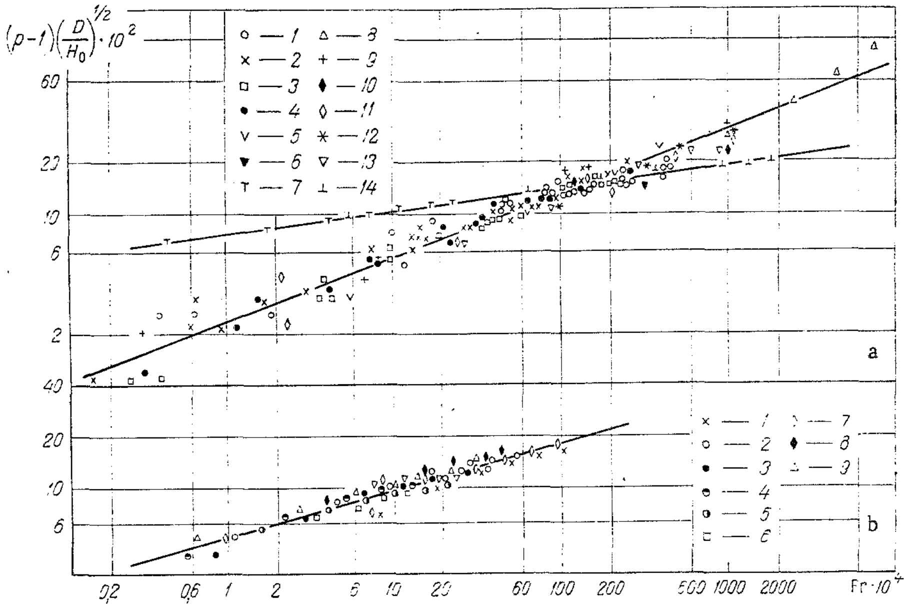

Fig. 1. Relationships obtained for the complex $(p-1)(D/H_0)^{1/2}$ [1–4) results of the present investigation for H_0 values of 40.7, 65, 87, 136 cm; 5) results of [10]; 6) [9]; 7) [3] (narrow fraction, H_0 = 21.95 cm, see Table 1); 8) [12]; 9, 10) [11] for glass spheres and an aluminum catalyst, respectively; 11-13) [2] for fractions 1, 2, and 3, respectively (Table 1); 14) [5]] (a) and for the relative expansion of the bed $(p-1)\cdot 10^2$ [1-6) [3] (wide fraction for H_0 = 4.9, 9.65, 14.55, 19.4, 24.2, 29.1 cm, respectively); 7-9) [3] (narrow fraction for H_2 = 5.4, 10.95, 16.45 cm, respectively, see Table 1)] (b) as functions of the Froude number Fr.

of the relative expansion of the bed on the geometrical characteristic of the system H_0/D, taking the following form: $(p-1)/\mathrm{Fr}^{1/3} \sim (H_0/D)^{1/2}$.

The results of a generalization of the experiments are shown in Fig. 1a. The figure also contains the results of other authors [2, 9, 10, 11, 12], the corresponding experimental conditions being shown in Table 1. The experimental points are correlated by the equation

$$p-1=0.7\left(\frac{H_0}{D}\right)^{1/2}\mathrm{Fr}^{1/3} \tag{9}$$

with a mean square deviation of 15%.

The resultant relationship shows that the relative expansion $(p-1)$ of the bed depends on the gas flow rate (Fr) and the geometrical characteristic of the bed as a whole (H_0/D). The generalization obtained is valid for relatively thick beds $(0.5 < H_0/D < 2)$. For narrow beds $(H_0/D < 0.5)$ [3] (Table 1) we seen from Fig. 1b that the relative expansion of the bed does not depend on this characteristic, and the experimental data are generalized by the following equation:

$$p-1=0.56\mathrm{Fr}^{0.27}. \tag{10}$$

Figure 1a also shows the results of experiments carried out in columns of diameter 310 and 300 mm [3, 5] for an H_0/D of the order of 1 (see Table 1). These data correspond to a smoother dependence of the expansion of the bed on Fr. We note that the experimental points obtained from [5] and corresponding to comparatively high gas-filtration rates lie quite close to the established relationship [Eq. (9)].

Thus for thin beds $(H_0/D < 0.5)$ the relative expansion of the inhomogeneous fluidized bed does not depend on the geometrical characteristic of the system (H_0/D) but is determined solely by the dimensionless excess gas-filtration velocity (Fr). For $0.5 < H_0/D < 2$ as indicated by an analysis of the experimental data $p-1 \sim (H_0/D)^{1/2}$. In the case of thick beds $(H_0/D > 5)$ in which the "piston" mode is operative, the expansion of the layer again becomes independent of the complex (H_0/D) [8].

In conclusion, we note that Eq. (9) may be recommended for calculating the height (thickness) of an inhomogeneous fluidized bed in installations with a uniform gas distribution at the inlet for $0.02 \leq Re \leq 3.2$.

NOTATION

D, column diameter; D_B, diameter of the gas bubble; d, diameter of particles; H, H_0, heights of the expanded and stationary beds, respectively; h, height above the gas-distributing lattice: $p = H/H_0$, expansion of the bed; $p - 1$, relative expansion of the bed; g, gravitational acceleration; $Fr = (u - u_0)^2/gH_0$, modified Froude number; q_0, q_1, q_2, dimensionless coefficients; Δp, pressure drop in the core of the bed at a distance Δx; Δx, distance between the pressure measuring points in the vertical direction; u, u_0, filtration velocity and velocity at the onset of fluidization, respectively; v_B, $\bar{v}_B$, relative rise velocity of the gas bubbles and the same averaged over the bed; ρ_p, density of particles; ε, ε_0, porosity of bed at u and u_0, respectively; ν, kinematic viscosity of fluidizing medium; $Re = u_0 d/\nu$, Reynolds number.

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