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EXPERIMENTAL INVESTIGATION OF TUBE FLOW OF A GAS-PARTICLE MIXTURE

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The results of an experimental investigation of the velocity distribution and the discrete-phase mass flow distribution in an ascending gas-particle tube flow are presented and partially generalized.

In two-phase flow the conditions of jet formation, determined by the characteristics of the accelerating device, have a much greater influence on subsequent flow development than in the case of single-phase flow. The simplest effective accelerating device is a length of round tubing. In this case the characteristics of the tube flow enter into the initial jet flow conditions. The existing publications on the tube flow of gas-particle mixtures are mainly concerned with questions of heat transfer and pneumatic transport, while the direct study of the mechanics of fine-dispersion flows has been held back by the lack of suitable measuring techniques. Recently developed methods of probing flows of the gas-particle type with lasers [1, 2] have opened up prospects of obtaining the necessary information.

The most important factor influencing the distribution of the discrete phase in a two-phase jet system is the phase slip velocity produced by the accelerating device. It is impossible to exclude the effect of the nonuniformity of the discrete-phase distribution, although it is difficult to estimate. Certain information concerning the distribution of particle concentration and mass flow on the particle size range of interest is provided by Soo in [3-5], who noted a maximum of the particle mass flow distribution on the flow axis. The particle velocity data were obtained by conversion of the results for the particle concentration and mass flow and it was noted that the mean particle velocity lagged behind the mean gas velocity.

We have studied the phase velocity distribution and the discrete-phase mass flow distribution in an ascending tube flow with reference to the particle size of the discrete phase, the velocity of the carrier phase, and the tube diameter. The stainless-steel tubes were 0.9 m long and had the following inside diameters: 7.8; 12.2; 15.6; and 25.8 mm. As the discrete phase we used narrow-fraction electrocorundum powders with a mean mass particle size of 23, 32, 70, or 88 μ and density of 3.9 g/cm^3. The carrier phase was air at 15-20°C. The air velocity was varied on the range 10-90 m/sec and the Reynolds number, based on the mean gas velocity, on the range from $5\cdot10^3$ to 10^5. In order to eliminate as far as possible the reaction of the discrete phase on the gas-phase velocity field, in almost all cases the mass flow concentration $\varkappa$ was kept at 0.3 or less [6], except at $\bar{v} \sim 10$ m/sec, when $\varkappa$ reached values of 0.8-1.0 kg·h/kg·h. The distribution of the discrete phase and its velocity were measured by optical laser methods at a distance of 3 mm from the tube exit. In order to exclude the possibility of electrization of the particles [7], the apparatus was grounded. A more detailed description of the measurements and the apparatus may be found in [8].

Under the experimental conditions (isothermal flow, $\varkappa \to 0$), steady-state flow of the two-phase system was primarily determined by forces of a hydrodynamic nature and a system of criteria can be constructed from the ratio of mean mass particle size to tube diameter d_s/D, the Reynolds number $\mathrm{Re} = \bar{v}D/\nu$, and the particle-gas density ratio ρ_p/ρ. In this case the adhesion properties of the particle—tube surface interaction can be defined in terms of the

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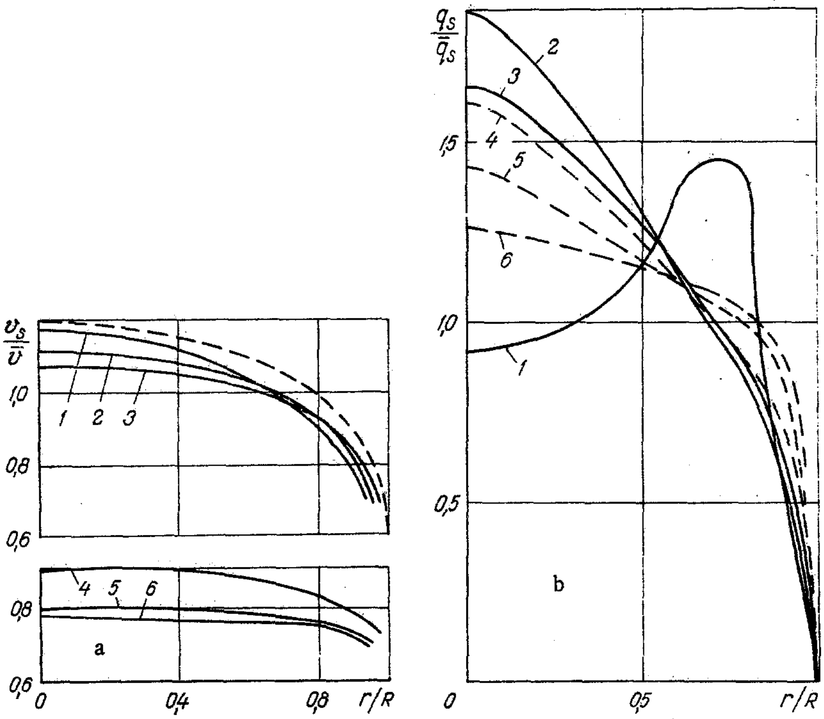

Fig. 1. Discrete-phase velocity profile $v_s/\bar{v}$ (a) and mass flow profile $q_s/\bar{q}_s$ (b) at outlet of 15.6-mm tube: 1-3) d_s = 23 μ; 4-6) 88 μ; 1, 4) $\bar{v}$ = 9.8 m/sec; 2, 5) 22.6; 3, 6) 42 m/sec.

coefficients of restitution of the normal and tangential particle velocity components upon collision with the walls. However, we still lack data for determining these coefficients, even in criterial form; accordingly, we are obliged to neglect the possible influence of this factor.

Part of the original data are presented in Fig. 1a, b. The measurements were made along the tube diameter, and the corresponding curves average the experimental points along both radii. The dashed curve in Fig. 1a represents the dimensionless gas velocity profile for all three of the indicated regimes, measured with a Pitot—Prandtl tube in the absence of a discrete phase. An analysis of the complete set of data (summarized in Tables 1 and 2) confirms the previous assumption concerning the decisive role of the criteria d_s/D and Re. As these increase, the velocity and mass flow profiles become fuller, the lag of the particle velocity behind the gas velocity on the tube axis increases, and in the wall region the particles acquire a greater velocity lead over the gas. It is worth noting an anomaly in the discrete-phase mass flow distribution: For the finest particles (Fig. 1b) the $q_s/\bar{q}_s$ profile changes from saddle-shaped to parabolic as the velocity increases from 9.8 m/sec to 22.6 m/sec (also observed at D = 25.8 mm). The data on the particle velocity lag are grouped in Tables 1 and 2, from which it is clear that in a number of cases the lag of the particle velocity on the tube axis behind the gas velocity amounts to tens of meters per second and considerably exceeds the critical velocity of the particles. This once more confirms that in this case gravity, in other words the Froude number, cannot play an important part. It is noteworthy that particle velocity measurements in an ascending tube flow at low flow velocities (up to 2.5 m/sec) [9] have shown that the particle velocity lag on the tube axis is close to the critical particle velocity, but that, despite the laminar nature of the flow, in the wall region the particles have a velocity lead over the gas.

In terms of d_s/D and Re the data on the particle velocity lag on the tuve axis can be generalized by means of the exponential expression

$$v_{s_0}/v_0 = \exp(-100\, n d_s/D), \quad (1)$$

where the coefficient n is a function of the Re number (n = 0.58; 0.78; 0.92; 1.10 for Re = $0.5 \cdot 10^4$; $2 \cdot 10^4$; $4 \cdot 10^4$; 10^5, respectively).

TABLE 1. Lag of the Discrete-Phase Velocity with Respect to Gas Velocity on the Tube Axis

D, mm	v_0, m/sec	v_{s_0}, mm/sec at various d_s, μ			
		23	32	70	88
7,8	11,6	12,0	8,6	7,2	6,0
7,8	22,4	20,2	16,4	14,4	12,2
7,8	43,9	34,6	27,1	21,3	19,7
7,8	96,5	69,1	53,2	36,2	36,9
7,8	96,5	—	—	38,4	—
12,2	11,9	—	10,5	8,4	—
12,2	27,5	—	21,7	18,5	—
12,2	49,6	—	38,0	27,5	—
12,2	106,0	—	71,7	52,6	—
12,2	106,0	—	72,3	—	—
12,2	19,4	—	—	—	12,5
15,6	11,9	11,2	11,4	10,0	8,8
15,6	26,9	25,2	24,4	19,4	17,8
15,6	49,6	45,8	43,6	30,7	—
25,8	11,9	—	11,5	—	9,3
25,8	26,6	—	26,3	—	20,7

TABLE 2. Lag of the Mean Discrete-Phase Velocity with Respect to Mean Gas Velocity

D, mm	$\bar{v}$, m/sec	$\bar{v}_s$, m/sec at various d_s, μ			
		23	32	70	88
7,8	10,0	—	8,5	6,8	—
7,8	19,5	18,2	15,8	13,7	—
7,8	37,0	32,9	26,3	19,7	—
7,8	82,9	65,7	49,5	36,6	—
12,2	10,0	—	8,9	7,8	—
12,2	23,3	—	20,3	18,1	—
12,2	42,1	—	35,0	26,8	—
12,2	90,0	—	73,5	49,8	—
15,6	10,0	8,5	—	—	8,3
15,6	22,6	22,5	—	—	17,0
15,6	42,1	41,4	—	—	32,0
15,6	42,1	40,5	—	32,9	—

As noted in [3], these features of the velocity and mass flow distributions of the discrete phase can only be explained by a diffusion particle transport mechanism. The need to take lateral forces into account, as in [10, 11], is quite obvious. As a working hypothesis we may assume that under the influence of the transverse F_L and longitudinal F forces the particles execute oscillatory motions, the amplitude of the oscillations increasing with increase in the criteria d_s/D and Re, and the axis of the oscillations being correspondingly displaced toward the tube axis.

The exact analysis of this hypothesis is rendered difficult by the fact that, although the lateral migration effect has been established, we lack an expression for the transverse force at $Re_p > 1$, and it is precisely this regime that characterizes the experimental conditions. An approximate analysis of the relationship between F_L and F points to the feasibility of this mechanism of particle motion. In fact, using the relation obtained for the transverse force in [10], we arrive at $F_L/F = \omega_p d_p^2/24\nu$ at $Re_p < 1$. According to [12], the main source of particle rotation is collision with the walls, and $\omega_p \sim v_p/d_p$. Reducing to dimensionless criteria, we obtain

$$\frac{F_L}{F} \sim \frac{v_p d_p}{vD}\,\mathrm{Re}. \tag{2}$$

If for F_L we use the relation obtained in [11], we find

$$\frac{F_L}{F} \sim \frac{d_p}{D}\sqrt{\mathrm{Re}}. \tag{3}$$

Both expressions are valid only for Stokes flow. Assuming that the nature of the ratio F_L/F remains the same at large Re_p numbers, we find that the particle oscillation frequency and

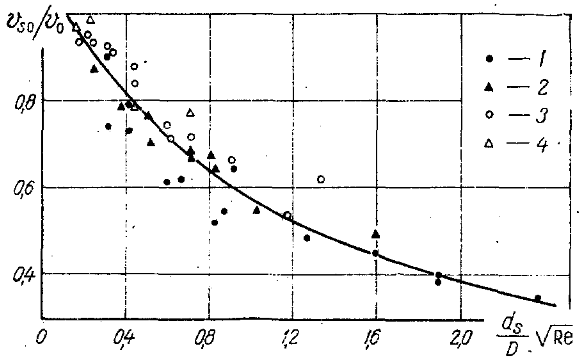

Fig. 2. Generalized graph of the particle lag on the tube axis: 1) D = 7.8 mm; 2) 12.2; 3) 15.6; 4) 25.8 mm.

hence the particle velocity lag is determined by the same simplexes; i.e., the expression for the particle lag should be sought in the form $v_{s_0}/v_0 = \Phi(d_s/D \cdot Re)$ in accordance with (2) or $v_{s_0}/v_0 = \Phi(d_s/D \cdot \sqrt{Re})$ in accordance with (3). A criterial analysis of the particle lag data leads to the dependence

$$\frac{v_{s_0}}{v_0} = \left(0.90 + 0.85 \frac{d_s}{D} \sqrt{Re}\right)^{-1}, \tag{4}$$

which is compared with the experimental results in Fig. 2. As in (1), the accuracy of generalization (4) corresponds to the accuracy of the experiments, which for the ratio v_{s0}/v_0 was 10%, although a certain stratification of the data within this error interval, depending on the tube diameter, was observed. This may be associated with the neglecting of the adhesion properties of the surface. The range of variation of the parameters generalized by expression (4) is: $d_s/D = (1.5\text{–}11.3) \cdot 10^{-3}$; $Re = (0.5\text{–}10) \cdot 10^4$; $\rho_p/\rho = 3.3 \cdot 10^3$. It is useful to note that in a recently published monograph [13] it is demonstrated, on the basis of an analysis of experimental data on the heat transfer and coefficient of friction for a fine-dispersion flow, that the controlling criterion is $\rho D^2/\rho_p d_s^2 Re$, which differs from ours only with respect to the factor ρ/ρ_p, the effect of which we disregarded. On the other hand, Eq. (4) can be written in the form

$$Re_{s_0} = Re_0 \frac{d_s}{D}\left[1 - \Phi\left(\frac{d_s}{D}\sqrt{Re}\right)\right] \tag{5}$$

where $\Phi(d_s/D \cdot \sqrt{Re})$ is the right side of (4) and the relation between the Re for tube flow and Re_{s_0} is explicitly indicated.

Our results demonstrate the importance of the interaction of the particles with the wall flow region and the need to take account of migration forces in describing the flow of a two-phase mixture of the gas-particle type in relatively narrow tubes. Another interpretation of the data obtained is that in convergent nozzles it is not possible to obtain a completely equilibrium two-phase flow.

NOTATION

d_s, mean mass particle size of discrete phase, m; d_p, particle diameter, m; R and D, tube radius and tube diameter, m; ρ_p and ρ, particle and gas densities, kg/m³; ν, kinematic viscosity of gas, m²/sec; v and v_s, streamwise velocities of gas and discrete phase, m/sec; v_p, streamwise particle velocity, m/sec; $\varkappa$, mass flow concentration, kg·h/kg·h; q_s, discrete-phase mass flow, kg/m²·sec; ω_p, particle angular rotation velocity, rad/sec. Indices: subscript 0 denotes values on the channel axis; a bar denotes the mean over the cross section.

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METHOD OF IMPROVING ENERGY EFFICIENCY OF VERTICAL PNEUMATIC TRANSPORT BY A RETARDED COMPACT LAYER

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The energy characteristic of pneumatic transport by a retarded compact layer is considered and recommendations are given on means of reducing its energy consumption.

A significant disadvantage of all kinds of pneumatic transport is high power consumption. There are little data in the literature on a comparison between the energy efficiency of different kinds of pneumatic transport, and the opinions of different authors on this question often diverge. Zabrodskii [1] considers that the pneumatic transport material is not efficient at low concentrations; transport in an ascending fluidized bed is much more advantageous. Reznikovich and Todes [2] carried out a theoretical investigation of the energy characteristics of vertical pneumatic transport and arrived at the deduction that the greatest energy efficiency can be achieved for either a low volume concentration of a two-phase stream ($\varepsilon \to 1$) or volume concentrations close to the concentration of the immobile layer ($\varepsilon \to 0.4$). Vel'shof [3] as well as Sandy, Daubert, and Jones [4] assert that the energy expenditure in pneumatic transport in a compact layer is much higher than in pneumatic transport with low concentration. Taking such a discrepancy in the estimation of the power consumption of different kinds of pneumatic transport into account, as well as the explicit inadequacy of the appropriate data on pneumatic transport by a retarded compact layer (RCL), an investigation of the energy characteristics of this kind of pneumatic transport is an important problem.

The efficiency

$$\eta = \frac{gG_t H}{LQ_0}. \tag{1}$$

is usually considered the principal energy characteristic of vertical pneumatic transport.

The efficiency of the compressor is not introduced into the equation since it is more convenient to take into account just the efficiency of the pneumatic lifter itself and not the efficiency of the pneumatic transport unit as a whole when comparing the energy expenditures of different kinds of pneumatic transport.

Let us clarify the dependence of the RCL pneumatic lifter on the parameters of the transportation process. The specific work of gas compression for an isothermal process is defined by

$$L = p_0 \ln \frac{p_f}{p_0}. \tag{2}$$

Consumption of the material is found from the formulas

$$G_t = us(1-\varepsilon)\rho_t. \tag{3}$$

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