HYDRODYNAMICS AND EXTERNAL HEAT TRANSFER IN GRANULAR BEDS WITH ROTATING CYLINDERS PLACED IN THEM

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The results are presented from an experimental investigation of the effect of rotation of a cylinder placed horizontally in fixed and fluidized granular media on the structure of the layer adjacent to the wall and on the intensity of external heat transfer.

It is known that the heat transfer over the perimeter of tubes placed horizontally in a fluidized bed [1] is quite nonuniform. This is caused by its structural nonhomogeneity: by a layer of stationary particles lying above the tube and by a gas cavity below [2] which arise during the flow of the disperse medium around the solid body. One of the possible methods of eliminating the stagnant zones and equalizing the structures of fluidized systems over the perimeters of the tubes is the rotation of the tubes.

The purpose of the present paper is to explain the effect of rotating a cylinder placed horizontally in a fluidized bed on the structure of the layer adjacent to its surface and on the intensity of external heat transfer.

The experiments were carried out in a laboratory apparatus an outline diagram of which is shown in Fig. 1. In the granular bed apparatus a copper cylinder of diameter 35 mm was placed at a height of 50 mm from the perforated gas distribution grid and mounted in slip bearings. An electric heater was placed inside the cylinder, and the heat which was liberated was transferred to the bed. The temperature of the surface of the cylinder was measured by thermocouples, the signals from which were passed to a digital voltmeter through mercury current slip-rings. The cylinder was rotated by an electric motor whose rate of rotation could be smoothly adjusted. In the experiments the rate of rotation of the cylinder was varied from zero to 10 revolutions per second and this was measured simultaneously by an electronic frequency counter and by a digital electronic tachometer operating in conjunction with a photoelectric device. Monofractions of particles of corundum, glass, and fireclay of diameters 0.18, 0.32, 0.44, 0.55, and 0.74 mm were used as the solid phase of the bed. The fluidizing agent was room temperature air.

The picture of the flow of the disperse medium around the rotating cylinder and the structure of the layer close to the surface of the cylinder were investigated by visualization observations and photography using a glass apparatus of square cross section 6 × 6 cm and also by the use of high-speed motion-picture photography with a rate of 1200 frames/sec.

Experiments carried out in fixed beds with no gas flow and with gas flows at filtration velocities less than the critical velocity showed that the rotating cylinder entrained with itself the first layer of particles adjacent to its surface. As a result of cohesion forces a second row of particles was drawn into motion, but at a smaller velocity, and this in turn transmitted some momentum to a third row, and in this way a loosened bed of particles or packets of particles moving in the direction of rotation of the cylinder (Fig. 2a) was formed near the surface. The thickness of this layer became stabilized and increased from $3d_p$ to $7d_p$ as the rate of rotation varied from 1 to 10 rev/sec. Within the rotating bed there was a continuous replacement of particles with others drawn from the stationary boundary.

Figure 2a shows the picture of the particle movement in a fixed bed with no gas flow with rotation of the cylinder in a counterclockwise direction. Before the beginning of the exposure the upper surface of the bed was flat. The tracks which arise from the motion of the particles during the exposure time clearly show the direction of their entry into the

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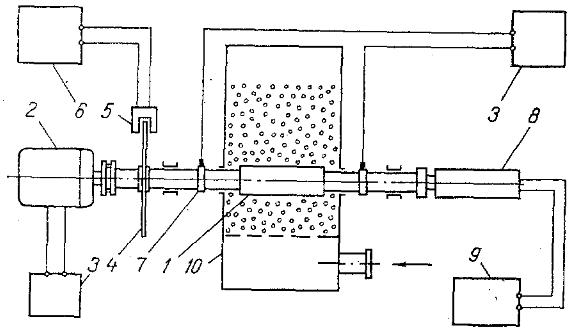

Fig. 1. Block diagram of the experimental equipment: 1) cylinder-calorimeter; 2) electric motor; 3) autotransformer; 4) disk with openings; 5) photodiode probe; 6) digital electronic tachometer; 7) current-carrying slip-ring; 8) mercury current collector; 9) digital electronic voltmeter; 10) apparatus with fluidized bed.

rotating layer adjacent to the surface and their departure from it. The particles were drawn in mainly in the upper part of the cylinder surface in the zone ~10-45° to the right of the vertical axis (a funnel formed in the surface layer at this position), while particles were ejected by centrifugal force over practically the entire remaining perimeter of the rotating layer. The particles escaping from the layer exerted a pressure on the stationary particles, compressing them (in the photograph the motion tracks of the particles in all directions from the cylinder can be clearly seen), causing the bed to move in the direction of the free outer surface, so that a distension of the bed is observed there to the left and the right of the cylinder. In this way a continuous transfer of particles occurred, together with a replacement of the particles in the zone adjacent to the wall.

The main feature of the hydrodynamics of the bed close to the surface of the rotating cylinders compared with a stationary cylinder in a fluidized medium (Fig. 2a, b) is that when the solid rotates the immobility of the layer of particles adjacent to its outer surface is disrupted.

In the case when the cylinder moves in the fluidized bed, as in the case of the stationary bed, the formation of an intermediate layer of particles is observed at the surface of the solid body which moves in the direction of rotation of the cylinder (Fig. 2c). However, both the thickness of this intermediate layer and the concentration of particles in it are nonuniform over the perimeter of the cylinder. When the cylinder rotates counterclockwise, the maximum thickness of the intermediate layer (on the order of 10-15d_p) is observed as a rule on the right, and the minimum thickness (1-2d_p) on the left in the equatorial zone of the cylinder. This is obviously related to the fact that the directions of the velocity vector for the motion of the cylinder and that of the gas stream coincide on the right-hand side, while on the left-hand side they are opposed. In the upper zone of the cylinder the thickness of the layer of moving particles adjacent to the wall amounts to 3-4d_p. Against the lower part of the cylinder a gas cavity is formed, the thickness of which becomes smaller while the extent over the perimeter becomes larger and the concentration of particles in it becomes considerably larger than in the case of the stationary cylinder. The latter is explained by the fact that the particles moving in the path of rotation of the cylinder are thrown into the gas cavity. The momentum causing the motion of the particles with the cylinder is transferred during the "collapse" of the particles adjacent to the surface. This behavior was observed for fluidized beds of particles of irregular and circular shapes.

Thus, while in the zone adjacent to the wall of a stationary cylinder perpendicular to which a fluidized medium flows it is possible to clearly distinguish three characteristic zones (see Fig. 2b) which differ in their structural and hydrodynamic parameters, namely: periodically collapsing air cavities below the cylinder, a "cap" of stationary particles lying on top the cylinder, and a randomly moving disperse stream in the equatorial zones of the cylinder, in the case where the cylinder rotates (Fig. 2c) it is possible to distinguish only two zones: a pulsating gas cavity filled with particles below the cylinder, and the remainder of the surface which is wetted by an intensively moving two-phase stream. The greatest uniformity of the structure at the surface of the rotating cylinder was observed at W = 1.05-1.20, when the size of the gas cavity reached a minimum and the concentration of particles in it passed through a maximum.

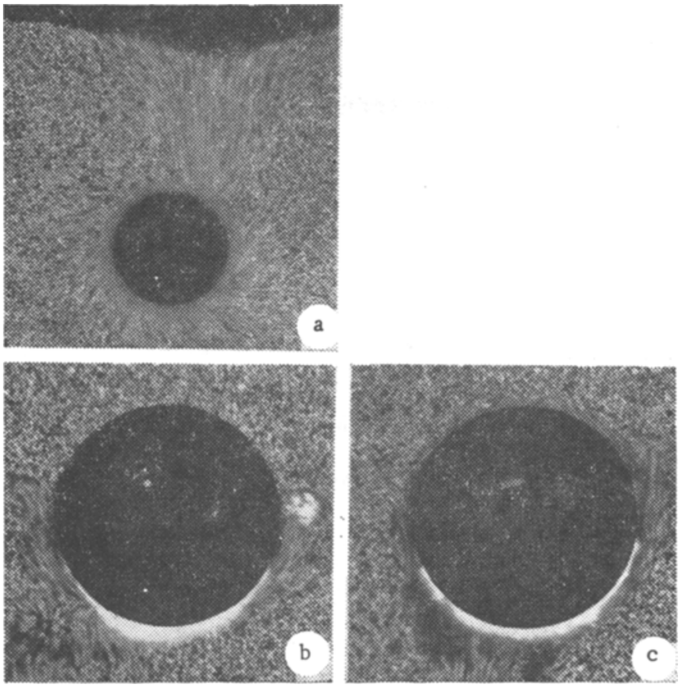

Fig. 2. Pictures of the structural and hydrodynamic conditions arising close to the surface of the cylinder: a) cylinder rotating in a fixed bed of fireclay particles of diameter d_p = 0.74 mm with no gas flow at n = 5 rev/sec, at an exposure time of 30 sec; b) stationary cylinder in a fluidized bed of fireclay particles of d_p = 0.74 mm, W = 2, exposure time 1/125 sec; c) cylinder rotating in a fluidized bed of fireclay particles of d_p = 0.74 mm at n = 9 rev/sec, W = 2, with an exposure time of 1/125 sec.

In a bed through which no flow occurred (under the experimental conditions), the heat from the hot surface of a stationary cylinder was passed to the particles in contact with it by heat conduction through the positions of their direct contact with the surface and through the air filling the spaces between the projecting roughnesses of the particles. During the rotation of the cylinder, the rolling, slipping, turning, and transverse movement of the particles and the gaseous medium entrained by them had a positive effect on heat transfer. The expansion of the layer adjacent to the wall increased the thermal resistance to heat transfer, particularly at large rates of rotation. The existence of these two factors, which have opposite effects on heat transfer, led to the fact that their effects partly cancel each other out, and the intensity of heat transfer at a rate of rotation of the cylinder of n = 1-2 rev/sec increased only insignificantly (by 7-11%), and for a further increase in the rate of rotation remained the same as for the stationary cylinder.

When air was passed through the stationary bed at a filtration velocity less than the critical velocity (W = 0.5-0.9), rotation of the cylinder had a positive effect on heat transfer, which increased with increasing rate of rotation, reached a maximum at n = 4-7 rev/sec, and then decreased again (Fig. 3, curve 5). The maximum increase in the intensity of heat transfer with the cylinder rotating compared to the case with the cylinder stationary was 30-50%.

The results of the experimental determinations of the heat transfer coefficients α from rotating cylinders to fluidized beds compared with the coefficients α_0 for stationary cylinders are given in Figs. 3 and 4.

The experiments showed (Fig. 3) that as the rate of rotation of the cylinder increased, the heat transfer coefficients increased. They reached maximum values at n = 4-7 rev/sec, after which they decreased again. The effect of the rate of rotation of the cylinder on the intensity of heat transfer is particularly marked at fluidization numbers close to unity. Thus, at W = 1.2 the maximum heat transfer coefficient from a cylinder rotating in a fluidized bed of corundum particles of d_p = 0.18 mm was about six times larger than in the case of the

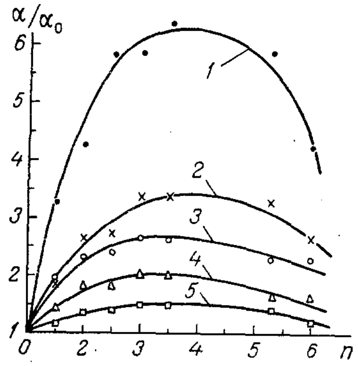

Fig. 3

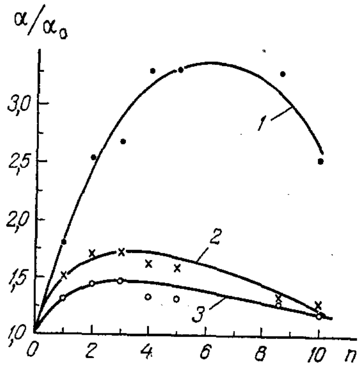

Fig. 4

Fig. 3. Dependence of α/α_0 on the rate of rotation of the cylinder n for a fluidized bed of corundum, d_p = 0.18 mm: 1) W = 1.2; 2) W = 1.5; 3) W = 2; 4) W = 3; 5) W = 0.8. The rate of rotation n is given in rev/sec.

Fig. 4. Dependence of α/α_0 on the rate of rotation of the cylinder n for fluidized beds of particles of various diameters (W = 1.75): 1) d_p = 0.18 mm; 2) d_p = 0.32 mm; 3) d_p = 0.74 mm.

stationary cylinder (curve 1), while at W = 3 it was twice as large (curve 4). This is explained by the structural and hydrodynamic features which occur under these conditions.

At W = 1.05-1.20 the value of the mean heat-transfer coefficient from the stationary cylinder is relatively small as a result of the fact that a large part of its upper surface is in contact with a stationary layer of particles, through which the intensity of heat transfer is minimal [1]. Under these conditions this part of the surface area of the rotating cylinder was wetted by a disperse medium in which small-scale pulsations are equally probable over its entire perimeter. The increased expansion of the bed in the zone adjacent to the wall of the rotating cylinder led to a decrease in the hydraulic resistance, which in turn led to an additional inflow of air from the core of the bed [3] and increased the gas velocity and turbulence of the particles close to the heat transfer surface [4]. In addition, as a result of the rotation of the cylinder, the linear surface velocity of which varied from 0.1 to 1 m/sec (n = 1-10 rev/sec), it was possible for a vortex motion of the gas to be set up in the layer adjacent to the wall.

As the rate of filtration of the gas through the bed increased there was an increase in the large-scale pulsations as a result of the appearance of gas bubbles of large volume, and structural nonhomogeneities occurred close to the rotating cylinder which also decreased the influence of the rotational effect on the intensity of heat transfer.

As the diameter of the particles in the bed increased, the intensity of heat transfer decreased (Fig. 4). For a fluidized bed of corundum of d_p = 0.18 mm the maximum value of α/α_0 at W = 1.75 was ~3.5 (curve 1); for d_p = 0.32 mm it was about 1.6 (curve 2), and for d_p = 0.74, it was only 1.4 (curve 3).

By approximating the experimental values of the maximum coefficients of heat transfer occurring at n = 4-7 rev/sec it was possible to arrive at the equation

$$\frac{\alpha_{\max}}{\alpha_0} = 12{,}7W^{-0,9}\mathrm{Ar}^{-0,16},$$

which is valid for W = 1.5-4.0 and Ar = 150-37,000. The relative mean-square deviation of the values of α_{max} calculated by this equation from the experimental values is 14%. The fluidization number W was determined with respect to the narrow cross-section, and the Archimedes number is defined as Ar = $[gd_p^3/\nu^2][(\rho_p - \rho_g)/\rho_g]$.

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GAS DISTRIBUTION IN A PACKED BED WITH JET GAS INJECTION

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This article examines the distribution of gas flows in the vicinity of a jet entering a packed bed. A qualitative explanation of the character of particle motion around the cavity is given.

The rate of heat and mass transfer processes in units with a packed bed is determined by the hydrodynamic situation in the bed – particularly by the distribution of gas-phase velocity [1]. Several studies have modeled the gas distribution in a granular bed [2-4]. However, these studies focused either on the so-called filtration regime of flow – when the gas flow does not form a cavity free of solid particles in the bed [2] – or on the spouting regime of jet injection [3]. The study [4] solved the problem of the gas distribution in a fixed granular bed as a special case of the gas distribution in a fluidized bed with jet gas injection and a bed aeration rate equal to zero. This approach to modeling the gas distribution in a fixed granular bed does not reflect the specifics of the problem or its differences from the problem for a fluidized bed, and it can be used only as a first approximation.

It was established experimentally [5] that the character of change in static pressure inside the cavity formed with jet gas injection into a fluidized or fixed granular bed is different in these two cases. In the jet in the fluidized bed, static pressure increases sharply only over a short initial section of the jet and remains nearly constant for most of its length. Thus, the assumption of a constant pressure in the jet entering a fluidized bed which was adopted in [4] can be considered valid. In a jet entering a fixed granular bed, static pressure increases over the entire length of the jet on the jet axis in accordance with a law which is close to linear, and it reaches its maximum value near the end of the cavity. Thus, the assumption of constancy of pressure inside the cavity cannot be adopted for a fixed bed.

Here we propose a model of gas distribution in the vicinity of a plane or axisymmetric jet in a fixed granular bed which is based on the assumption of a linear change in static pressure inside the jet channel.

Let a granular bed of height H be located in a plane or cylindrical unit of width or diameter 2R (Fig. 1). The gas jet is injected in the plane of the base of the bed $y = 0$ through a slit or circular opening of width or diameter 2a which is coaxial with the walls of the unit. The jet forms a cavity of height b in the bed, the boundary of this cavity being described by the equation $x^2/a^2 + y^2/b^2 = 1$. The chosen form of cavity is close to the shape actually seen in practice and differs from the shape assumed in [4]. It makes it possible to consider not only the height of the cavity, but also the size of the opening for entry of the gas. The latter is particularly important in the case of low gas velocities, when the height of the cavity becomes commensurate with the diameter of the nozzle. We will assume that the static pressure on the top boundary of the bed is constant and equal to zero, while inside the cavity it changes in accordance with a linear law with the coefficient K:

$$p = Ky \quad \left(\frac{x^2}{a^2} + \frac{y^2}{b^2} \leqslant 1\right). \tag{1}$$

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