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HYDRODYNAMICS AND EXTERNAL HEAT EXCHANGE OF A SPOUTING FLOWING BED

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Results of investigation of the hydrodynamics and the external heat exchange of a spouting flowing bed in a continuous-action apparatus are presented. The phase diagrams, heat-exchange characteristics, and rational arrangement of the heat exchangers in this apparatus were determined. The results obtained have been brought up to the level of new technical solutions and tested in the industry.

Keywords: *spouting flow layer, hydrodynamics, external heat exchange, loose material, fluidization number, heat-transfer coefficient, near-grid zone, gas distributor, free cross-sectional area.*

Introduction. The spouting-bed technique is used in the treatment of loose coarse-grained materials to provide an active hydrodynamics and an intensive heat and mass transfer in them. A spouting flowing bed is used as a rule in continuous-operation apparatus with uninterrupted feed and removal of the material being treated. Presently, the hydrodynamics of this bed and the heat transfer in it, especially the heat exchange between the bed and the surrounding surfaces, are not thoroughly understood.

The present work is a continuation of the previous investigations of the author in this direction [1–6]. The aim of this work is to investigate the hydrodynamics and the external heat exchange of a spouting flowing bed of carbamide with particles of diameter 1–2 mm [1, 2] on the basis of new data on these phenomena [3, 4], including those obtained in the last few years [5] for coarse-grained materials — wheat, pea, millet, and complex mineral fertilizers of the type of nitroammofoska — and, using the results obtained, to develop rational designs of apparatus in which such a bed is used.

The investigations were carried out in rectangular models of the indicated apparatus with a working chamber of size $0.6 \times 0.16 \times 1.0$ m or $1.5 \times 0.5 \times 2.0$ m. The flow rate of the material being treated was determined by the weighing method and was varied from 0.06 to 0.55 kg/sec. The fluidization number was changed from 1 to 3.

The main hydrodynamic feature of the flowing bed being investigated is the existence of very narrow zones of internal and flow-through spouting in it, which follows from the phase diagrams obtained in [1].

Discussion of Results. Our investigations provide support for the view that the jet-fluidization theory [7] takes no account of the flow-through feature of the bed being considered and, therefore, narrows the zone of active action of jets, i.e., the regions of internal spouting (IS) and through spouting (TS), in it. The indicated feature is due to the substantial suppression of the torches by the moving bed. Because of this, the parametric criterion H_s/H_w for the jet regime of a flowing bed is decreased to 0.3 unlike a nonflowing bed for which this quantity comprises 0.6 and larger value according to the existing theory, and the local-spouting regime is realized in a nonflowing bed within the range $0.6 \le H_s/H_w \le 0.85$ according to this theory and at $0.3 \le H_s/H_w \le 0.5$ in a flowing bed according to the data of our investigations.

The particles in an apparatus with a jet fluidization follow complex trajectories. Because of this, of practical interest is determination of the dependence of the velocity of their longitudinal movement in such an apparatus of definite design on the velocity of the air flow at a definite flow rate of the material blown through by the jets. Even though the particle velocities obtained in this way are averaged over the length of the apparatus, they allow one to determine the time of dwell of particles in the apparatus fairly accurately.

The dependence of the velocity of the particles W in the indicated apparatus on the velocity of the air flow V was determined for four flow rates of a material by the known method using labeled particles (Fig. 1).

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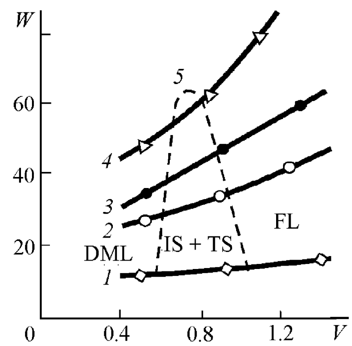


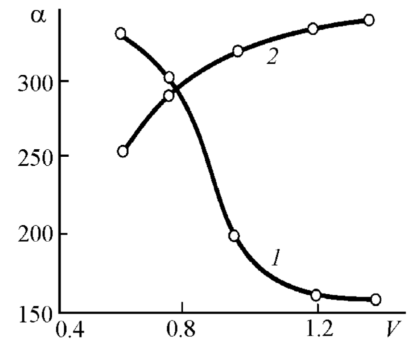


Fig. 1. Dependence of the velocity of horizontal movement of particles in a flowing carbamide bed on the velocity of the air flow at different flow rates of the particles: G = 0.062 (1), 0.215 (2), 0.318 (3), and 0.550 kg/sec (4); 5) region of the IS and TS regimes, dense moving layer (DML), and fluidized layer (FL). V, m/sec; W, mm/sec.

Fig. 2. Change in α in the main (1) and near-grid (2) zones of the flowing wheat bed. α, W/(m^2·K); V, m/sec.

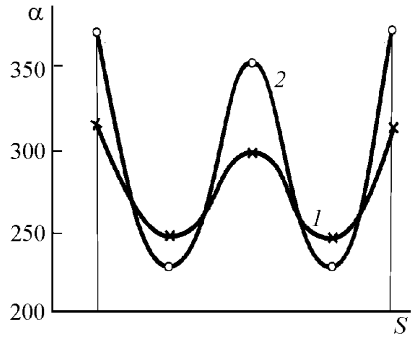
In Fig. 1, the region of IS and TS regimes is bounded by the dashed line in accordance with the phase diagram of a flowing bed constructed in [1]. The data obtained on the velocities of the particles in the apparatus allowed us to determine the time of their dwell in the working chamber. It has been established that, in this case, of main importance is the flow rate of the material and the velocity of the gas (air) flow: with increase in these parameters, the time of dwell of particles in the working chamber decreases because their velocity increases. Of certain importance are also differences between the hydrodynamic regimes of flowing of the material at one and the same rate of the air flow (spouting regimes or a developed fluidized bed).

Our investigations have shown that the velocities of longitudinal movement of particles in the apparatus being investigated range from 0.01 to 0.1 m/sec. According to the results of industrial testing of coolers with a flowing bed of nitroammofoska, a product temperature lower than 313 K, required in the summer period, is provided at a distance of 3–4 m from the feed hopper after cooling of the product at a specific rate of air flow of order of 2 m^3/kg. The coolers were equipped with new gas distributors [2], the longitudinal step of rows of whose holes exceeds the transverse one by 2–4 times. We recommend using gas distributors providing an efficient regime of internal spouting for cooling of granules.

The results of the present work were used to advantage for cooling of complex mixtures of mineral fertilizers and are protected by patent.

Long use of the above-indicated gas distributors has shown that, due to the sharp decrease in their blockage, the working resistance of them is of the same order as the resistance of the "classical" gas distributors. The main advantages of the reconstructed gas distributors are as follows: the period of their stable work without cleaning is increased by three times as compared to the "classical" gas distributors, and the flow rate of the fluidizing agent can be decreased by 20–30% due to its local blow, which makes it possible to correspondingly decrease the electric power expended by the gas-blowing facilities.

As applied to a number of heterogeneous processes of treatment of coarse-grained materials, a rational method of heat supply or heat removal is the use of surface heat exchangers immersed in a spouting bed. In this case, direct contact of the heat-transfer agent with the bed is excluded, and the apparatus can work at lower fluidization numbers due to the intense conductive heat exchange. Therefore, the implementation of the process being considered provides economic efficiency and decreases environmental hazard. Our investigations [3–5] have shown that the external heat exchange in the continuous-operation apparatus with a flowing bed is intense and homogeneous. The investigations were carried out in unit hydrodynamic cells of rectangular models of such apparatus. The length of a cell was equal to the step of the transverse slots (rows of holes) of the gas distributor, and its width was equal to the step of the longitudinal slots.



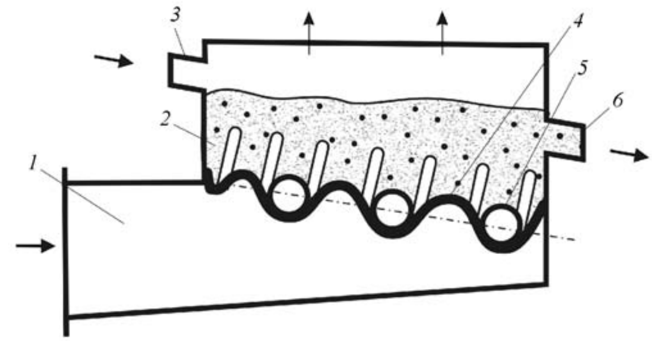


Fig. 3. Change in α in the flowing (1) and nonflowing (2) pea beds at S = 20–160 mm. α, W/(m^2·K).

Fig. 4. Diagram of an apparatus with a spouting flowing bed: 1) inlet pipe for feed of air under the grid; 2) rectangular working chamber; 3) inlet pipe for feed of the material being treated; 4) corrugated grid of the gas distributor; 5) heat-exchange pipes in the cavities of the corrugations; 6) outlet pipe for removal of the product.

For a flowing bed fluidized by the jets of a gas distributor, of chief interest is the character of change in the dependence of α on the main hydrodynamic parameters — the velocity of the fluidizing-agent flow V and the flow rate of the material G. An analysis of the results obtained in [3–5] has shown that α takes maximum values in the regime of a moving dense filter bed. In this case, α is increased in the zones located at a distance greater than 1/3 of the moving-bed height from the gas distributor. In the near-grid zone, α is lower by 15–20%. However, when the velocity of the gas flow increases, α in this zone increases and α in the main zone located at a larger distance from the gas distributor decreases (Fig. 2).

This change in the coefficient of heat transfer in a flowing bed is explained by the increase in the conductive heat exchange caused by the substantial increase in the concentration of particles in the upper part of the bed at a constant feed and the existence of only internal near-grid torches. The later structural feature of the bed makes it possible to minimize the carry-over of the upper dense part of the flowing bed, serving as a grain filter.

As our investigations have shown, the mass flow rate of the material being treated does not exert a substantial influence on the intensity of the heat exchange in the flowing bed.

As a result of the analysis and processing of the data obtained for the flowing and nonflowing beds, we obtained a series of curves reflecting the change in the heat-transfer coefficient α over the width of a hydrodynamic cell. These curves are qualitatively similar for both types of beds; however, the difference between the extremum values of α in the flowing bed is substantially smaller than that for the nonflowing bed. In our opinion, this feature can be explained by the fact that, in the case of longitudinal mixing of a material in the working chamber of an apparatus, the velocities of its particles relative to the heat-exchange surface are equalized. Thus, the distribution of the heat-transfer coefficient in the flowing bed is more homogeneous that that in the nonflowing bed (Fig. 3).

An analysis of the results obtained allowed us to determine the optimum zones for arrangement of heat-exchange surfaces for provision of a maximum heat exchange and a minimum disturbance of the hydrodynamic structure of the bed by the pipes positioned in it. Our investigations have shown that the design of the gas distributor providing a jet fluidization should correspond to the design of the surface heat exchanger. An extended rectangular body is recommended for the continuous-operation apparatus with a flowing bed; in this body the heat-exchange and manifold pipes should be positioned in the near-torch zones of the grain layer in parallel with the gas-distributor slots. Of interest is also the heat-exchanger design in the form of pipes positioned in the bed on the gas-distributor grid that should be corrugated, and the pipes should be installed in the cavities of the corrugations [6] (Fig. 4).

The results obtained are evidence in favor of the assumption that the external heat exchange in the flowing bed being considered is intense (250–650 W/(m^2·K)). The heat exchanger proposed has a developed surface and provides intense heat exchange, which makes it possible to decrease the overall dimensions of the apparatus with a flowing bed by 2–3 times as compared to a purely convective one. The expenses for the use of this apparatus are decreased because it is convenient in service and in repair of its units positioned under the bed.

The results of our investigation also favor the use of a corrugated gas-distributor grid. They show that the region of disposition of the corrugation bulges coincides with the zone of possible stagnation of a material being treated if the distance between the gas-distributor slots is large (200–500 mm); therefore, a stagnation of the material is not a threat even in the case of the most inhomogeneous fluidization. Such a design admits a large step between the gas-distributor slots (without stagnation), which makes it possible to substantially decrease the number of perforations, i.e., the apparatus itself and its fabrication become cheaper. The free cross-sectional area of such gas distributors comprises 1.5–6% and not 10–20% as in their analogs. According to our tests, the service life of the indicated gas distributors is increased by three times due to the decrease in their blockage. Because of these, the design of the apparatus proposed offers substantial advantages over the analogous existing apparatus.

CONCLUSIONS

1. In a spouting flowing bed, the zone of active action of the IS and TS jets is very narrow, which is due to the substantial suppression of the torch zones moved by the bed.

2. The velocities of longitudinal movement of particles of a flowing bed fall within the range 0.01–0.1 m/sec.

3. The coefficient of external heat exchange between the surface immersed in a flowing bed and the bed depends substantially on the conductive heat exchange in the bed because of the high concentration of particles in its upper part.

4. The dense upper part of a flowing bed serves as a grain filter minimizing the carrying away of the dust.

5. The coefficient of external heat exchange in a flowing bed is fairly large (α = 250–650 W/(m^2·K) and is more homogeneously distributed over the volume of the bed as compared to that of a nonflowing bed.

6. In a spouting flowing bed there are zones optimum for disposition of heat-exchange surfaces, which makes it possible to develop a heat exchanger and a gas distributor of corresponding design that increases the economic efficiency of the apparatus proposed and decreases the environmental hazard.

NOTATION

d, diameter of the particles, mm; G, weight rate of the material flow, kg/sec; H_s, height of a spout, mm; H_w, working height of the bed, mm; S, step between the jets, mm; V, velocity of the gas; W, velocity of the particles, mm/sec; α, coefficient of external heat exchange, W/(m^2·K). Subscripts: w, working; s, spouting.

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