LITERATURE CITED

1. S. S. Zabrodskii, Hydromechanics and Heat Transfer in Fluidized Beds [in Russian], Moscow–Leningrad (1963).
2. Fluidization [in Russian], Moscow (1974).
3. J. Botterill, Heat Transfer in Fluidized Beds [Russian translation], Moscow (1980).
4. V. A. Chlenov and N. V. Mikhailov, Vibrationally Fluidized Beds [in Russian], Moscow (1972).
5. W. J. Thomas and R. J. Melcher, Ind. Eng. Chem. Fundamentals, 14, No. 3, 140-153 (1975).
6. R. Eldson and C. J. Shearer, Chem. Eng. Sci., 32, No. 10, 1147-1153 (1977).
7. O. A. Myazdrikov, Electrodynamic Fluidization in Dispersed Systems [in Russian], Leningrad (1984).
8. V. V. Pushkov, Z. R. Gorbis, and M. K. Bologa, Inzh.-Fiz. Zh., 27, No. 2, 253-259 (1974).
9. Yu. E. Tetelya, V. V. Vishnevskii, V. P. Usenko, et al., EOM, No. 4, 61-64 (1984).
10. M. K. Bologa, V. V. Pushkov, and Yu. E. Tetelya, Proceedings of the 3rd All-Union Conference on Electron-Ion Technology Applications [in Russian], Tbilisi (1981), p. 16.
11. B. G. Sapozhnikov and N. I. Syromyatnikov, Izv. Vyssh. Uchebn. Zaved., Energ. No. 5, 116-119 (1969).
12. M. K. Bologa, F. P. Grosu, and I. A. Kozhukhar', Electroconvection and Heat Transfer [in Russian], Kishinev (1977).
13. P. W. Dietz and J. R. Melcher, Trans. ASME, C97, No. 3, 429-434 (1975).
14. M. K. Bologa and A. B. Berkov, Inzh.-Fiz. Zh., 53, No. 1, 77-84 (1977).
15. A. B. Berkov, Accelerating Heat Transfer and the Motion of Electrically Conducting Particles in a Gas-Fluidized System in the Presence of an Electric Field: Ph. D. Thesis [in Russian], Kishinev (1985).
16. M. K. Bologa, V. V. Pushkov, and A. B. Berkov, Izv. Akad. Nauk SSSR, Energ. Transport, No. 4, 109-116 (1980).
17. S. S. Zabrodskii, I. L. Zamnius, S. A. Malyukovich, et al., Inzh.-Fiz. Zh., 14, No. 3, 448-453 (1968).
18. M. K. Bologa, V. V. Pushkov, and A. B. Berkov, Int. J. Heat Mass Transfer, 28, No. 7, 1245-1255 (1985).
19. Z. R. Gorbis, Heat Transfer and Hydromechanics for Dispersed Through Flows [in Russian], Moscow (1970).

FLUID DISTRIBUTION IN THE GRANULAR LAYER FOR A DESCENDING GAS-LIQUID STREAM

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The flow of a fluid in an irrigatable granular layer is investigated experimentally.

In catalytic reactors with a fixed gradular layer in which a descending gas-liquid stream is used, the liquid distribution over the layer is of great value. The appearance of stream inhomogeneities results in a reduction in reactor productivity and causes additional difficulties in the control and regulation of the process [1] as well as in its mathematical description.

At this time there is not a well-founded quantitative description of the gas-liquid stream flow inhomogeneities in a layer because of the inadequacy of experimental data. Annular coaxial liquid collectors mounted at the exit from the layer that are used to determine the velocity profile along the radius of the apparatus [2-4] are ordinarily utilized in experimental investigations. But as is shown in [5], significant local stream inhomogenei-

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ties exist in an irrigatable granular layer, whose scale is much greater than the particle size and which are randomly distributed over the layer section. The influence of the gas flow-rate on the uniformity of the liquid flow was not investigated in [5].

EXPERIMENTAL METHOD AND CONDITIONS OF THE EXPERIMENT

Experiments were performed in a column with 20 cm inner diameter for a 20-100 cm granular layer height. Glass balls of 3.2 mm diameter; cylindrical activated charcoal granules AG-3 of 2 mm diameter and 3 mm length; the same charcoal granules hydrophobized by a polyfluoroethylene resin with 3 and 10% content by weight; granules of polypropylene of almost global shape with a 3.5 mm diameter were used as charges. The air velocity varied between 0-60 cm/sec limits while the fluid velocity was 0.35-1.3 cm/sec.

Uniform and nonuniform gas-liquid stream distributions were used at the entrance to the layer in the experiments. An installation of the type [6] with a separate gas and liquid delivery was selected for the uniform gas and liquid distribution. The liquid entered the layer through 132 tubes with 1.8 mm inner diameter, that projected 0.5 cm above the upper plane of the grating and 1 cm below the lower plane to eliminate liquid overflow over the lower plane. The gas entered by 37 tubes with 10 cm inner diameter that projected 15 cm above the upper plane of the grating. The tubes for the gas and liquid insertion were disposed uniformly over the grating section.

The stream nonuniformity at the entrance to the layer was produced by two methods: 1) the liquid was delivered through a single 15 mm diameter tube at the center of the layer, a metallic mesh was here superposed on the layer surface to avert layer particle dispersal by the liquid jets; 2) a rubber circular plate with a cutout segment in the 2/4 area of the section was placed on the layer surface under the distributive grating; in this case the uniformly distributed gas-liquid stream entered the layer only through the circular segment with 90° aperture cut out of the plate.

Measurements of the liquid flow-rate distribution were performed after a support grating which was a metallic mesh separated by thin sealed baffles into 37 cells in the form of a honeycomb with 1 cm wall height before and after the mesh, that hindered liquid progress along the mesh. This construction afforded the opportunity of measuring the inhomogeneity of the reduced local velocity of the liquid with a 3.2 cm scale during averaging of the liquid flow rate in a section of several tens of grains.

The pressure drop and the determination of the flow mode were measured simultaneously with the measurement of the liquid flow-rate distribution. The change of modes for a given gas flow rate and the change in the liquid flow rate were determined visually by the occurrence of liquid flow rate fluctuations at the exit from the layer or by the occurrence of pressure fluctuations in the layer.

The influence of surface tension which was measured by using surfactants (PAV) was investigated in part of the tests with glass balls with a uniform initial distribution. The PAV used was the laundry powder "Lotos" in the amount of 100 ml powder per 0.4 m^3 of water.

HYDRODYNAMIC MODES AND PRESSURE DROP

The diagram of the hydrodynamic modes for a descending gas-liquid stream in a layer is presented in Fig. 1 in conformity with [7]. Two modes were observed in tests: I) stratified flow of phases or weak interaction mode, II) piston flow mode of the gas-liquid stream or strong phase interaction mode.

Our data for a uniform stream distribution of glass balls or activated charcoal granules at the layer entrance agree with the transition boundary (straight line) obtained in [7]. We did not remark influence of either the layer height or the order of build-up of the liquid and gas mass flow rate on the boundary of the hydrodynamic modes.

The piston flow mode was not observed in the investigated range of variation of the gas and liquid mass flow rates for a layer of poorly wetted polypropylene and hydrophobized carbon particles (10% polyfluoroethylene plastic). What is the reason for this phenomenon? According to visual observation the piston flow mode occurs when there is the formation of liquid films overlapping the whole column section. It is known that the liquid film is spoiled if it is punctured by a dry needle, say, and remains whole during puncturing by a wetted needle. The poorly wetted particles of a granular layer apparently play the part of

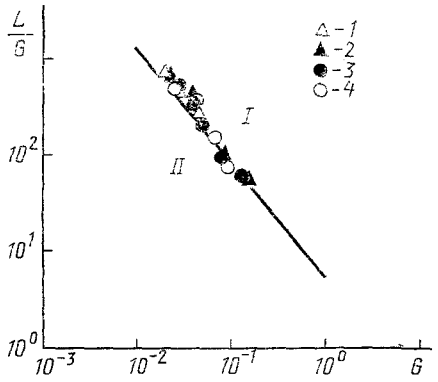

Fig. 1

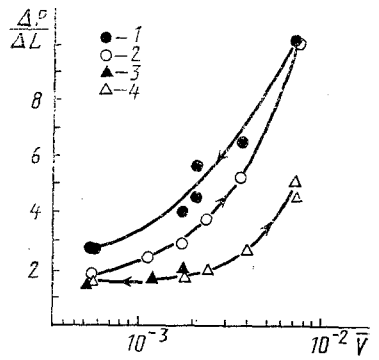

Fig. 2

Fig. 1. Mode pattern (I – strong interaction mode, II – weak interaction mode; the line is the boundary between the modes according to data in [7]): 1, 2) granular layer of hydrophobized charcoal granules (3% polyfluoroethylene resin), 3, 4) glass balls; 1 and 3 correspond to a 20.5 cm layer height; 2 and 4 to a 40.5 cm height (G, L are the gas and liquid mass flow rates, respectively, referred to the area of the column section, kg/(m²·sec)).

Fig. 2. Pressure drop per unit length of the layer ΔP/ΔL (kPa/m) as a function of the mean mass flow rate of the liquid $\bar{V}$ (m/sec); arrows indicate the measurement sequence; liquid distribution at entrance to the layer: 1, 2) uniform; 3, 4) liquid delivery at the center (gas velocity is 60 cm/sec).

dry needles destroying the liquid films, and they thereby hinder formation of the piston flow mode. It is not possible to make a more detailed analysis of the piston mode formation within the framework of this paper.

Hysteresis of the pressure drop in the layer is noted in our tests. As the liquid mass flow rate increases for a given gas mass flow rate the pressure drop is less than the pressure drop for the same mass flow rates, but obtained for a diminution in the liquid mass flow rate (see 1 and 2 in Fig. 2). The appearance of hysteresis is apparently associated with the different wetness of the granular layer which depends on the sequence of liquid and gas mass flow rate build-up as well as on the liquid distribution. The pressure drop hysteresis vanished (3 and 4 in Fig. 2) upon delivery of the liquid by a single jet (first method).

The addition of the PAV for the gas-liquid stream through a glass ball layer resulted in an increase in the pressure drop. In this case the transition to the piston flow mode was observed at lower liquid mass flow rates, i.e., in the domain of weak phase interaction without the PAV. The pressure fluctuations in the piston mode were apparently less because of frothing of the liquid.

The substantial influence of the method of the original organization of the gas-liquid stream on the domain of hydrodynamic mode and pressure drop existence should be noted. Thus, in the case of liquid delivery by a single jet (first method) the piston flow mode was not observed. The pressure drop diminished 1.5-2 times (Fig. 2) as compared with the uniform distribution. In the case of gas-liquid stream delivery through part of the entrance section (second method), the passage to the piston mode was expressed weakly and was observed in the domain of mass flow rates corresponding to the existence of the piston mode for a uniform distribution.

LIQUID DISTRIBUTION IN A GRANULAR LAYER

For a uniformly distributed stream at the entrance the characteristic form of the liquid local velocity distribution at the layer exit is similar to the distribution displayed in Fig. 3a. Qualitatively, this depends slightly on the particle material, the layer height,

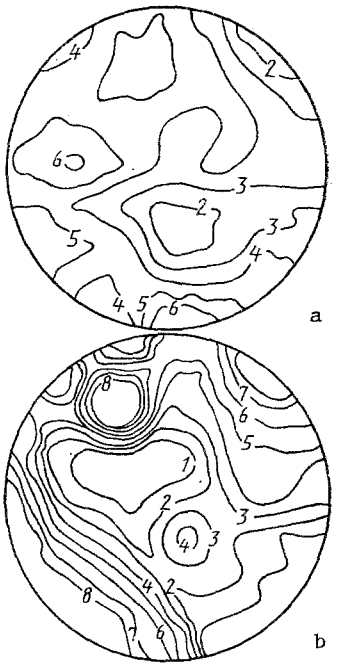

Fig. 3. Liquid velocity distribution at the layer exit upon the addition of PAV: a) for a single-phase stream; b) for a gas-liquid stream with a mean gas velocity of 14 cm/sec (the increase in the number of isolines per unit corresponds to the increase in liquid velocity by 0.09 cm/sec; the liquid velocity on isoline No. 4 equals the mean mass flow rate of 0.45 cm/sec).

and the gas and liquid mass flow rates in the investigated range of their variation. The measured liquid velocities in the different cells had a noticeable spread around the mean velocity. This spread was not associated with measurement error since it was reproduced well upon duplication of the tests. The large-scale inhomogeneities of the layer porosity as well as the presence of suspended gas bubbles in the granular layer could be indicated as the reason for the nonuniformity of the liquid flow in the layer. The inhomogeneity of the liquid distribution in the layer was estimated quantitatively by the empirical dispersion [9] of the relative liquid velocities over all the measuring cells:

$$S^2 = \frac{1}{36}\sum_{k=1}^{37}(1 - V_k/\bar{V})^2.$$

The dependence of S on the gas mass flow rate for different test conditions is represented in Fig. 4. For a uniform gas and liquid distribution at the column entrance S would be practically independent of the kind of charge, the layer height and the phase of the mass flow rate. For the flow of one liquid S = 0.25 ± 0.03 and as the gas mass flow rate increases S would diminish to the value 0.21 ± 0.03 (1 and 3 in Fig. 4).

The deviation of the local velocities from the mean mass flow velocity for points equidistant from the center was random in nature for a uniform or axisymmetric distribution at the entrance. Consequently, the local velocities were averaged over the measuring cells equidistant from the column axis. The values obtained, referred to the liquid mean mass flow velocity are presented in Fig. 4. The velocity profiles have the characteristic form described in the literature [4, 8] for single-phase streams; the maximal liquid velocity was observed at the wall. As the liquid mass flow rate increases, the near-wall effect diminishes. An increase in the liquid mass flow rate for two-phase streams resulted in a qualitative change in the velocity profile (Fig. 5), the velocity at the center became greater than at the wall although S did not here change in practice (see 1-3 in Fig. 4).

The coefficient of surface tension exerts substantial influence on the liquid velocity

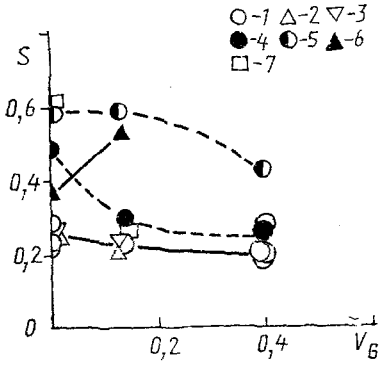

Fig. 4

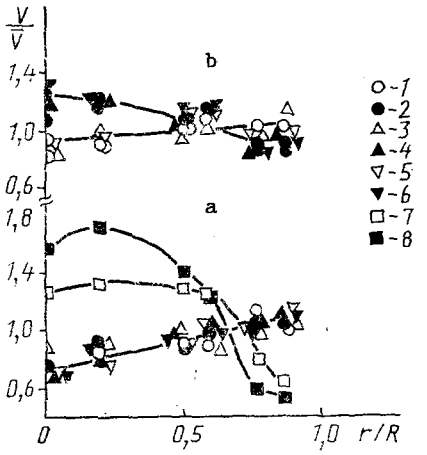

Fig. 5

Fig. 4. Dependence of S on the gas mean mass flow velocity V_G, m/sec: 1-3, 7) for a uniform gas and liquid distribution at the entrance; 4, 5) for gas-liquid stream insertion through $^1/_4$ section of the column; 6) liquid insertion at the center of the apparatus; 7) liquid with PAV additions; granular layer: 1, 4, 5, 7) glass balls; 2, 6) charcoal granules; 3) polypropylene granules; liquid mean mass flow velocity: 1, 2, 3) 0.35-1 cm/sec; 5, 6, 7) 0.35 cm/sec; 4) 1 cm/sec.

Fig. 5. Liquid velocity distribution at the layer exit over the column radius: distribution at entrance: 1-6) uniform; 7, 8) liquid delivery at center; granular layer: 1, 2) glass balls; 3, 4) polypropylene granules; 5-8) charcoal granules: liquid velocity: a) 0.35 cm/sec; b) 1 cm/sec; gas velocity: 1, 3, 5, 7) 0; 2, 4, 6, 8) 14 for a) and 46 cm/sec for b).

distribution. It was shown in [3] in which a single-phase liquid stream is investigated in a layer that a reduction in the coefficient of surface tension results in degradation of the liquid distribution over the layer. In our tests the addition of a PAV degraded the liquid distribution significantly for a single-phase stream even for a uniform initial distribution (see Fig. 3b). The quantity S here grew from 0.25 to 0.61 (1 and 7 in Fig. 4, $V_G = 0$). Addition of a PAV for a gas-liquid stream had practically no effect on the liquid distribution (Fig. 3a and 1 and 7 in Fig. 4, for $V_G = 0.14$ m/sec).

The method of organizing the gas-liquid stream at the entrance exerted substantial influence on the spreading of the fluid over the layer and therefore also on the local velocity distribution. For a nonuniform liquid distribution at the entrance the gas stream can result in both an increase in the liquid stream in homogeneity in the layer and to its diminution as well.

The velocity profile at the exit from a 40 cm layer is shown in Fig. 5a (7 and 8) for liquid delivery at the center (second method). In this case the gas stream tightened the liquid jet to the center and the nonuniformity of the liquid velocity along the radius was greater for the gas-liquid stream than for the single-phase stream. The quantity S also was increased (see 6 in Fig. 4). As was mentioned above, the pressure drop here is 1.5-2 times at the layer as compared with the uniform distribution. Consequently, the pressure drop can be a qualitative integrated characteristic of the liquid distribution over the layer.

In the case of stream organization by the second method when the uniformly distributed gas-liquid stream was delivered through a segment, an increase in the gas mass flow rate would result in diminution of the quantity S, i.e., to a better liquid distribution, as does an increase in the liquid mass flow rate (see 4 and 5 in Fig. 4).

NOTATION

V, liquid velocity averaged over cells equidistant from the column axis; $\bar{V}$, V_G, mean mass flow velocities of the liquid and gas, respectively; V_k, liquid local velocity in the k-th measuring cell obtained by dividing the liquid mass flow rate through this cell by its area; S^2, empirical dispersion.

LITERATURE CITED

1. V. M. Kurganov, Z. U. Novikova, A. I. Vaseiko, et al., Hydrogenized Processes to Obtain Motor Fuels [in Russian], Moscow (1978), pp. 70-88.
2. K. E. Porter, V. D. Barnett, and J. J. Templeman, Trans. Inst. Chem. Eng., 46, No. 2, 69-94 (1968).
3. E. Dutkai, Ruckenstein, Chem. Eng. Sci., 23, No. 11, 1365-1373 (1968).
4. J. Buchanam, Chem. Eng. Sci., 26, No. 5, 746-747 (1971).
5. P. J. Hock, J. A. Wesselingh, and F. J. Zuiderweg, Chem. Eng. Res. Des., 64, No. 6, 431-439 (1986).
6. J. Levec, A. E. Saez, and R. G. Carbonell, AIChE J., 32, No. 3, 369-380 (1986).
7. G. Tosun, IEC Proc. Dev., 23, No. 1, 29-35 (1984).
8. J. Prchlik, J. Soukup, V. Zapletal, et al., Collection Czech., Chem. Comm., 40, No. 10, 3145-3152 (1975).
9. S. Wilks, Mathematical Statistics, Wiley, New York (1962).

TOWARD A THEORY OF TRANSPORT PROCESSES IN BROWNIAN SUSPENSIONS

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Fluctuations in dispersed phase concentration increase the particle gradient diffusion coefficient and the effective viscosity of the suspension.

A large range of literature has been dedicated to the definition of effective properties of concentrated suspensions. A review of studies of the rheology of such suspensions can be found in [1]. The influence of collective effects on the gradient diffusion coefficient of Brownian particles in weak concentration suspensions was evaluated in [2-5], while in [6] the theory of [2] was generalized to concentrated suspensions. In all studies known to the present author the concentration of the dispersed phase was taken as a deterministically specified quantity, and the possibility of its fluctuation over time and space was neglected. However in suspensions in which the particles perform intense Brownian motion such fluctuations do occur and their amplitude increases with increase in external noise disturbance. Moreover, from the well known analogy between processes of colloid coagulation and molecular gas condensation [7] one can expect intense increase in fluctuations and their correlation radius in colloidal systems upon approach to the critical coagulation point, which within the framework of the indicated analogy corresponds to the critical gas-liquid phase transition point [8-10].

Below we will evaluate the effect of fluctuations in dispersed phase concentration on the effective diffusion coefficient of suspension particles and effective viscosity. A constructive description of the macroscopic behavior of suspensions can apparently only be given within the framework of the continuum approximation, which becomes inadequate if the amplitude of the fluctuations in porosity becomes comparable to the mean value of that quantity, while the correlation radius becomes comparable to the dimensions of the region occupied by the suspension. Therefore below we will assume the value of the fluctuations to be low in comparison to the mean concentration of the dispersed phase. We note that this limitation of smallness of the fluctuations is caused not only by the requirements of the continuum ap-

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