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RESEARCH, DESIGN, CALCULATIONS, AND OPERATING EXPERIENCE

CHEMICAL PLANT

HYDRODYNAMICS OF VESSELS WITH A THREE-PHASE AIR-FLUIDIZED BED USED FOR CLEANING GAS FLOWS

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The search for ecologically acceptable, effective, and economic methods of cleaning industrial effluents is urgent under modern conditions whereby anthropogenic effects on the environment are significant.

When gaseous effluents are cleaned of harmful impurities, large volumes of waste gases with low concentrations of removable substances are subjected to processing, as a result of which small liquid loads appear. High flow rates of gaseous phase make it necessary to increase gas velocities in the tower, while low flow rates of liquid phase result in nonuniform distribution of the gas-liquid layer across the section of the vessel.

In these cases, vessels with a three-phase air-fluidized bed may provide for stable conditions under which gas flows are cleaned [1]. In these vessels, use of a chemically stable packing with a developed surface, and at the same time, with a low hydraulic resistance, for example, a packing formed from a polyethylene mesh (flexible cellular hose) fabricated by the MGUIÉ is of particular interest.

In studying the hydrodynamics of vessels with a three-phase fluidized bed and a packing formed from a polyethylene mesh, we selected a certain vessel-operating range, for example, L/G = 0.15–0.9 (where L and G are the loads with respect to the liquid and gaseous phases), which is characteristic of the absorptive cleaning of readily soluble gases.

The hydrodynamics of a vessel with a three-phase air-fluidized bed was investigated on an experimental semi-industrial bench in a tower with a diameter of 400 mm on an *water–air* system with a downcomer type of support-distribution lattice with openings 8 mm in diameter and a free section of 12%.

The tower was built of acrylic plastic, which made it possible to observe the three-phase air-fluidized bed, and also to measure the height of the dynamic gas-liquid layer. During the course of the investigations, moreover, we measured the flow rates of gas and liquid, the hydraulic resistance of the plate with the three-phase air-fluidized bed, and the volume of liquid retained in the bed.

Since the hydraulic resistance of the three-phase air-fluidized bed depends heavily on the mass of the packing, we employed a principally new lightweight packing twisted from a polymeric mesh: the outside diameter of one element of the packing was 50 mm, and the height of an element of the packing was 38 mm.

Two basic hydrodynamic operating regimes were isolated as a result of processing of experimental data: the start of fluidization, and complete fluidization.

Observations indicated that in the initial-fluidization regime, the near-wall layers of the packed bodies are set in motion, and begin to move into the core of the gaseous flow, but their motion is slowed at the walls.

Moscow State University of Engineering Ecology (MGUIÉ). Translated from Khimicheskoe i Neftegazovoe Mashinostroenie, No. 8, pp. 3–5, August, 2004.

TABLE 1. Basic Characteristics of Polyethylene-Mesh Packing

Specific surface, m^2/m^3	367
Free volume	0.875
Weight of single packing, g	7.6
Number in 1 m^3	12500
Bulk density, kg/m^3	95

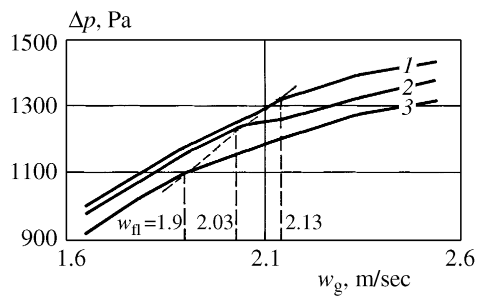

Fig. 1. Dependence of hydraulic resistance Δp of three-phase air-fluidized bed on gas velocity w_g for different specific irrigation densities L_{sp}: *1*, *2*, *3*) L_{sp} of 3.9, 3, and 2.14 $m^3/(m^2 \cdot h)$, respectively.

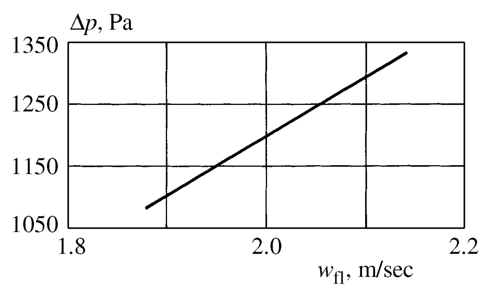

Fig. 2. Dependence of velocity w_{fl} at start of fluidization on hydraulic resistance Δp of three-phase air-fluidized bed of irrigated packing.

The pattern changes with increasing gas velocity: the amount of packed bodies, which go over into the air-fluidized state, and which are concentrated in the central portion of the plate increases, whereupon the amount of liquid retained in the gas-liquid layer increases, and, as a result, its height is raised.

Under complete fluidization, which sets in at a certain (for each irrigation density) gas velocity, the near-wall layer of the packing is broken-up, all packed bodies go over into the air-fluidized state, and the gas flow is distributed virtually uniformly over the entire section of the vessel. Moreover, good mixing of the liquid and gas bubbles is observed in the volume of the moving bed.

It is known [1] that for spherical packings, the rate of conversion to complete fluidization decreases with increasing irrigation density. In conducting the hydrodynamic investigations of the new cellular packings, we analyzed the rate of conversion of the packing to complete fluidization. Results of the investigations are presented in Fig. 1, from which it is apparent that the rate w_{cr} of conversion to fluidization increases with increasing specific liquid load L_{sp}. This is obviously explained by the fact that accumulation of liquid occurs within the volume of the packed body as the irrigation density increases; this results in charging of the packing, and its conversion to the air-fluidized state at a higher gas velocity.

As is apparent from Fig. 1, values of the rate of conversion of an irrigated packing formed from polyethylene mesh to a complete-fluidization regime (w_{fl}) falls on a single straight line (– – –).

Let us construct the plot of the $w_{fl} = f(\Delta p)$ relationship (see Fig. 2) in order to obtain an equation for determination of the velocity at the start of fluidization of the irrigated packing:

$$w_{fl} = 0.0011\Delta p + 0.74. \tag{1}$$

Equation (1) makes it possible to estimate only the velocity at the start of fluidization.

The following sequence of analysis is proposed.

Let us first determine the rate of conversion of the three-phase air-fluidized bed to a complete-fluidization regime on the basis of gas flow rates known from the material balance. For this purpose, we processed experimental data from the $w_{fl} = f(L_{sp})$ curve, and obtained the following empirical equation for an *air–water* system:

$$w_{fl} = 1.64 L_{sp}^{0.19}. \tag{2}$$

According to this equation, it is possible to fix the boundaries of the two operating regimes: start of fluidization, and complete fluidization, which is recommended as the working regime for absorption processes.

The hydraulic resistance of the vessel, which is equal to the sum of the hydraulic resistances of the irrigated plate and packing, is evaluated in the second step:

$$\Delta p = \Delta p_{d.p} + \Delta p_p. \tag{3}$$

The hydraulic resistance $\Delta p_{d.p}$ of a downcomer plate is well understood; it is of interest, therefore, to determine the hydraulic resistance Δp_p of the packing.

Investigations were conducted on a downcomer plate with no packing, and on a plate with a cellular packing subjected to the same gas and liquid loads; this made it possible to isolate the hydraulic resistance of the irrigated cellular packing. The data obtained are presented in Fig. 3.

The following equations were derived as a result of processing of the experimental data:

$$\Delta p_p = 268.9 w_{cr}^{1.81} L_{sp}^{-0.47} \tag{4}$$

for the start of fluidization, and

$$\Delta p_p = 540.3 w_{cr}^{0.35} L_{sp}^{-0.18} \tag{5}$$

for complete fluidization.

The equations obtained enable us to calculate the hydraulic resistance of a vessel with a three-phase air-fluidized bed of the new cellular packing with sufficient accuracy for systems close to an *air–water* system.

It is known that the amount of liquid retained in the bed and the gas accumulation (gas content), which is the ratio of the volume occupied by the gas located in the bed to the total volume of the gas–liquid layer, exerts a major influence on the mass-exchange process in the gas and liquid phases:

$$\varphi = \frac{H_f - h_{st}}{H_f}. \tag{6}$$

The gas content φ is linked to the relative density of the foam by the following relationship:

$$k = 1 - \varphi = \frac{h_{st}}{H_f}, \tag{7}$$

where H_f is the height of the gas–liquid layer, or foam in mm, and h_{st} is the height of the static layer of liquid in mm.

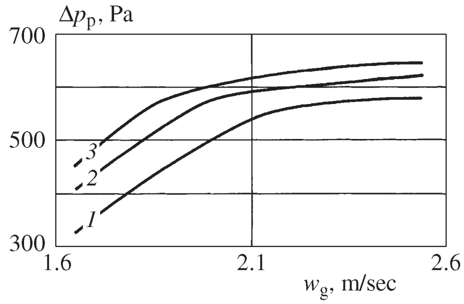

Fig. 3. Dependence of hydraulic resistance Δp_p of irrigated packing on gas velocity w_g for different specific irrigation densities L_{sp} (*1–3* – same as in Fig. 1).

Since a cellular packing differs in principle from previously investigated types of packings (packings with a closed internal recess were primarily used), it is impossible to use the equations cited in the literature to determine the gas content of a three-phase air-fluidized bed. We therefore conducted investigations on both a downcomer plate with no packing, and on a plate with the cellular packing.

Downcomer plates are rather well understood. Ospanov [2] investigated in detail the effect of different parameters on the gas content of a dynamic dual-phase bed, and derived a computational equation for determination of the relative foam density, which takes on the following form when the International System of Units is employed:

$$k = 1.81 h_{st}^{0.16} (w_{cr}\sqrt{\rho_g})^{-0.33} \mu_L^{0.16} \sigma^{0.12}, \tag{8}$$

where ρ_g is the density of the gas in kg/m^3, μ_L is the viscosity of the liquid in Pa·sec, and σ is the surface tension in N/m.

Ospanov's investigations of a downcomer plate [2] were conducted under assigned gas (F-factor = 0.4–1.0) and liquid (L_{sp} = 3.37–9.16 m^3/(m^2·h)) loads; it is of interest, therefore, to examine how relationship (8) will be reproduced when the following experimental data are used: F-factor = 1.6–2.6, and L_{sp} = 2.14–3.9 m^3/(m^2·h). As is apparent from Fig. 4, the results obtained for the gas content of a downcomer plate correlate well with the data cited in [2].

Since data on the height of the static layer of liquid correlate well with Eq. (8) for a downcomer plate, it was decided not to conduct experiments to determine the effect of various physicochemical parameters on the gas content of the air-fluidized bed of an irrigated packing, but to assume $k \sim \mu_L^{0.16}$ and $\sigma^{0.12}$.

A downcomer plate with a three-phase air-fluidized bed of a cellular packing was investigated under the indicated gas and liquid loads. The extent of the effect of the F-factor and amount of liquid retained in the bed relative to the area of the plate (i.e., the height of the static layer of liquid) on the relative foam density k was determined in processing the experimental material obtained.

The computational relationships assumed the following form for the entire set of experimental data:

$$k = 0.019 h_{st}^{-1.40} (w_{cr}\sqrt{\rho_g})^{1.51} \mu_L^{0.16} \sigma^{0.12} \tag{9}$$

for the start of fluidization, and

$$k = 0.820 h_{st}^{-0.64} (w_{cr}\sqrt{\rho_g})^{-0.99} \mu_L^{0.16} \sigma^{0.12} \tag{10}$$

for complete fluidization.

The k values calculated from Eqs. (9) and (10) were compared with experimental values of the relative foam density; good agreement was obtained for the experimental and computed data (deviation of ±15%).

During operation of an absorber with a cellular packing, a certain amount of retained liquid, which will depend on a number of factors, and which will exert a major influence on the absorption process, always exists in the air-fluidized bed.

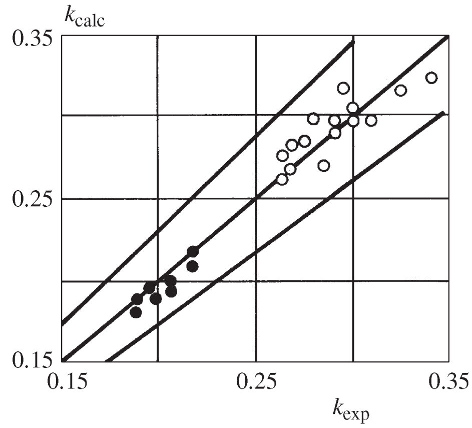

Fig. 4. Comparison of experimental and computed data on relative foam density of downcomer plates: ○) Ospanov's data [2]; ●) data derived by authors.

The amount of liquid retained is expressed as the volume of liquid in the bed covering a plate area of 1 m^2. Thus, the height h_{st} of the static layer of liquid can be treated as the height of the clear (not containing gas) liquid in the bed.

We investigated the effect of the following hydrodynamic parameters on the amount of retained liquid h_{st}: the gas velocity w_{op} in the openings, the irrigation density L_{sp}, and the relative foam density k.

The following computational relationships were obtained:

$$h_{st} = 5.28 w_{op}^{0.94} L_{sp}^{0.23} \tag{11}$$

for the start of fluidization, and

$$h_{st} = 10.86 w_{op}^{0.66} L_{sp}^{0.30} \tag{12}$$

for complete fluidization.

The h_{st} values calculated from formulas (11) and (12) were also compared with experimental values of the amount of liquid retained in the bed; the deviation amounted to ±15%.

Thus, the data derived from processing of the results of the hydrodynamic investigations of vessels with the new type of air-fluidized packing made it possible to obtain computational equations for characteristics of a three-phase gas-liquid bed (static layer of liquid and relative foam density, or gas content), and also to assess the increase in the hydraulic resistance of the cellular packing when it is used for a downcomer plate in the complete-fluidization mode. As is apparent from Fig. 3, the value of Δp_p does not exceed 650 Pa in the region of F-factor values from 1.9 to 2.6, and L_{sp} values to 4 $m^3/(m^2 \cdot h)$.

Use of the packing improved the structure of the gas-liquid bed on the plate; this should enhance the efficiency of the mass-exchange process.

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