SOME PROBLEMS OF HEAT TRANSFER AND HYDRODYNAMICS IN A FLUIDIZATION BED WITH SMALL-VOLUME PACKING

A. I. Tamarin, D. M. Galershtein, S. S. Zabrodskii, R. R. Khasanov, and V. P. Borisenko

UDC 536.244:66.096.5

A study was made concerning the effect of four kinds of small-volume packing on the expansion of a fluidization bed, on its effective thermal diffusivity, and on the heat transfer between the bed and an immersed surface. The test results for two materials have been generalized in terms of empirical formulas.

The fluidization technique has found many uses in chemical engineering for catalysis, adsorption, and other processes which involve some interaction between a gas and a disperse solid material. It is suitable for small grain sizes, for narrow temperature ranges, and for fast heat removal from the reaction zone. At high gas filtration velocities, however, the system becomes inhomogeneous and gas cavities or bubbles appear with contact between phases. These effects become magnified as the apparatus is scaled up.

The inhomogeneity of a fluidization bed is due to the instability of the disperse phase suspended in the gravitational field by the gas stream. If stationary surfaces are spaced uniformly inside such a system, their local interaction with the gas stream will produce a more stable bed structure where the gas cavities and the inhomogeneity centers will collapse. A judicious design of these surfaces, their location, and geometry, can make it possible to control the hydrodynamics and the structure of a fluidization system and to optimize it for each specific technological application. This is the reason for the recently growing interest in packed beds [1-15], but information about them in the technical literature is very scattered.

We will present here the results of experimental studies concerning the hydrodynamics and the heat transfer in a fluidization bed with small-volume packing. Tests were performed with four different kinds of packing, the characteristics of which are given in Table 1. The disperse phase in these tests was quartz sand or silica gel with the average paticle size 0.23 and 0.19 mm, respectively. The height of the loose charge was about 30 cm. Most tests were performed with a column 30 cm in diameter (packing type 2, 3, and 4). Part of the tests was performed with a column 15 cm in diameter (packing type 4 and 5).

TABLE 1. Packing Characteristics

Packing type	Number	Diameter of packing element D_p, (cm)	Equivalent packing size
Bundle of vertical tubes	2	2	8,1
	3	5,5	3,0
Bunch of wire spirals	4	2	0,52
	5	1	0,34

In the first test series the aim was to establish the effect of packing on the bed porosity. The latter was determined from the static gas pressure transmitted through a segment of a given height. The measurements were made over a range of initial charge heights within which the porosity remained almost uniform throughout [16, 17]. The test results with silica gel are shown in Fig. 1. The relative porosity has been plotted here along the ordinates, with the porosity at the start of fluidization ε_0 as the scale unit. This porosity was measured along the height of the bed after

Institute of Heat and Mass Transfer, Academy of Sciences of the Belorussian SSR, Minsk. Translated from Inzhenerno-Fizicheskii Zhurnal, Vol. 23, No. 4, pp. 635-640, October, 1972. Original article submitted December 14, 1971.

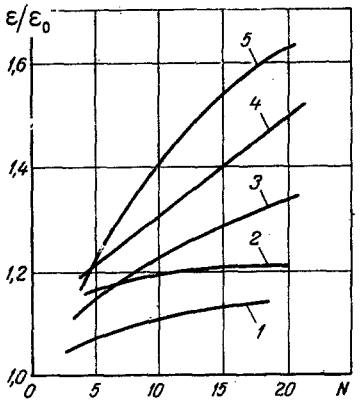

Fig. 1

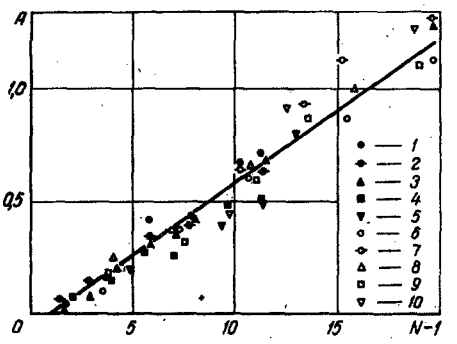

Fig. 2

Fig. 1. Relative porosity of a fluidization bed as a function of the fluidization number, for silica gel as the disperse phase: 1) free bed; 2, 3, 4, 5) packed bed with the respective packing type.

Fig. 2. Correlation curve for the effective thermal diffusivity of the fluidization system, with sand as the disperse phase: 1) free bed; 2, 3, 4, 5) packing types according to Table 1; with silica gel as the disperse phase: 6) free bed; 7, 8, 9, 10) packing types according to Table 1. $A \equiv a_{eff}/u_0H_0 \cdot \exp(0.46/l_p)$.

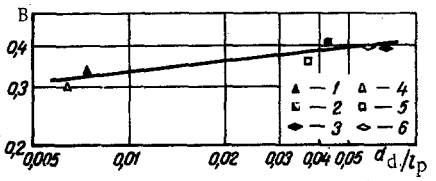

Fig. 3. Correlation curve for the heat transfer in the fluidization system, with sand as the disperse phase: 1, 2, 3) packing types 3, 4, 5 according to Table 1; with silica gel as the disperse phase: 4, 5, 6) packing types 3, 4, 5 according to Table 1. $B \equiv 1-\alpha/\alpha_0 + (\rho_d/\rho_G) \cdot 10^{-4}$.

stabilization by a gradual decrease of the gas velocity. The curves represent averages of the test series, with the deviation of actual values not exceeding 10%. The graph indicates clearly that packing types 3, 4, and 5 [18] change the bed porosity appreciably, the latter increasing almost proportionally to the gas velocity. Packing type 2 has only a small effect on the bed; the system expands here only slightly at higher values of the fluidization number. Curves 2 and 1 tend to converge within the range of high filtration velocities. A similar pattern is noted also in the tests with sand.

The porosity of a fluidization bed characterizes, to some measure, the homogeneity of the system. The higher the porosity is, i.e., the more the bed expands, the higher is its gas content. It is also well known that the buoying velocity of a bubble and thus its dwell time in the bed are proportional to the square root of its diameter [19, 20]. The higher the bed porosity is, therefore, the slower do bubbles of excess gas move through it and the smaller is their size, and the more homogeneous is the bed structure.

In the second test series we studied the effect of packings on the stirring rate of the solid phase in the system. The circulation of the solid phase was measured by the method of the instantaneous heat source [21, 22, 23], such a heat source being produced here by some hot particles at the upper bed surface.

The effective thermal diffusivity of a fluidization system is determined by the gas filtration velocity (u), the charge height (H_0), and the aerodynamic characteristics of the material (u_0). These three quantities can be combined into the dimensionless group a_{eff}/u_0H_0 and the dimensionless ratio $u/u_0 = N$. It will be expedient to generalize the test data into a function of these two dimensionless variables. A preliminary evaluation of the test data has shown that the values for a free bed in this system of coordinates are higher than the values for a packed bed. This indicates that the stirring rate of the solid phase in a bed with packing is lower than in a free bed. In this system of coordinates, furthermore, the test points for all kinds of packing studied here fit on straight lines passing through the origin of coordinates. The slopes of these lines depend on the equivalent packing size (l_p), which is defined as the bed volume per unit area of charge surface. The slopes of the test curves decrease with increasing l_p.

The final evaluation of the test data is shown in Fig. 2, where values for the two materials and for the four kinds of packing have been plotted along with test values for a free bed. All points fit closely on a straight line and can be generalized by the equation

$$\frac{a_{\text{eff}}}{u_0 H_0} = 0.066\,(N-2)\exp\left(-\frac{0.46}{l_p}\right) \tag{1}$$

for the following range of variables: $2 < N \leq 20$, $0.33 < l_p < 10$ cm, $2 \leq u_0 \leq 6$ m/sec. The scatter of test data does not exceed 20%. The derived relation indicates that the stirring rate of the solid phase increases proportionally to the fluidization number.

In the third test series the aim was to obtain information about the effect of packings on the rate of heat transfer between the fluidization bed and an immersed surface. The tests were performed on an apparatus shown earlier with type 3, 4, and 5 packing in two materials (sand, silica gel).

The heat transfer was measured by the stationary method with a probe-heater. The probe was placed vertically at the center of the column, with its bottom edge 100 mm above the gas distributor mesh. The probe was made of copper tubing 20 mm in diameter, 100 mm long, and 5 mm wall thickness. Inside the probe was placed a heater of nichrome wire, fastened on a ceramic tube 4 mm in diameter. In order to minimize heat leakage through the end surfaces of the probe, the latter were covered with 20 mm thick Textolite stoppers. The temperature difference Δt was measured with a Chromel–Alumel differential thermocouple using 0.15 mm wire. One thermocouple bead was fastened to the outside surface of a copper cylinder in the center inside the probe. The other bead of the differential thermocouple read the temperature of the fluidization bed. During the tests we measured the heater power as well as the temperature difference between probe surface and bed. The heat transfer coefficient was measured in these tests within a 3% error. With increasing gas filtration velocity in a free bed as well as in a packed bed, the heat transfer coefficient passed through a soft maximum. Maximum heat transfer in a packed bed was attained at higher velocities than in a free bed [1]. Moreover, the maximum value of the heat transfer coefficient differed from one kind of packing to another. The maximum value of the heat transfer coefficient decreased with decreasing length l_p.

In order to establish the relative effect of packings on the rate of heat transfer between a fluidization bed and an immersed surface, it is worthwhile to look for a functional relation between dimensionless ratios in the form:

$$1 - \frac{\alpha}{\alpha_0} = f\left(\frac{d}{l_p};\ \frac{\rho_d}{\rho_G}\right). \tag{2}$$

In Fig. 3 are shown test data for three kinds of packing in the two different materials. The test data can be accurately enough approximated by the relation

$$1 - \frac{\alpha}{\alpha_0} = 0{,}53\left(\frac{d}{l_p}\right)^{0.1} - 10^{-4}\,\frac{\rho_d}{\rho_G}. \tag{3}$$

This relation applies within the range $0.06 < d/l_p < 0.8$, $2.2 \cdot 10^3 > \rho_d/\rho_G > 0.9 \cdot 10^3$. The maximum scatter of test points is 8%. According to relation (3), each packing reduces somewhat the rate of heat transfer between bed and immersed surface. Under our test conditions this reduction did not exceed 30%.

Thus, small-volume type 3, 4, and 5 packings homogenize the bed, increase its expansion, and slow down the solid phase stirring. At the same time, there results a decrease in the rate of heat transfer between bed and immersed surface.

NOTATION

a_{eff}	is the effective thermal diffusivity;
u_0, u	are the starting fluidization velocity and gas filtration velocity, referred to the total bed cross section;
ε_0, ε	are the starting and running porosity of the fluidization bed;
H_0	is the height of the settled bed;
ρ_G, ρ_d	are the density of the fluidizing agent (gas) and of the disperse phase, respectively;
N	is the fluidization number;
d	is the diameter of the solid particles;
g	is the acceleration due to gravity;

l_p is the equivalent length of the packing;
α, α_0 are the heat transfer coefficient for a packed and for a free bed, respectively.

LITERATURE CITED

1. L. Massimilla and S. Bracale, Ricerca Scientifica, No. 2, 487-503 (1956).
2. L. Massimilla and S. Bracale, ibid., No. 5, 1509-1525 (1957).
3. S. Bracale, A. Cabella, and L. Massimilla, Chimicale Industria, No. 8, 621-627 (1958).
4. L. Massimilla and J. W. Westwater, AIChE J., 6, No. 1, 134-138 (1960).
5. W. Volk, C. A. Johnson, and H. H. Stotler, Chem. Eng. Progr., 58, No. 3, 44-47 (1962).
6. J. P. Sutherland, G. Vassilator, H. Kubota, and G. L. Osberg, AIChE J., 9, No. 4 (1963).
7. A. I. Tamarin, Candidate's Dissertation [in Russian], Inst. Heat and Mass Transfer, Acad. Sci. ByelSSR, Minsk (1963).
8. T. Ishii and G. L. Osberg, AIChE J., 2, No. 2, 279-287 (1965).
9. O. Winter, K. Schuegerl, F. Fetting, and G. Schiemann, Chem. Eng. Sci., 20, No. 9 (1965).
10. K. Kato, K. Imofuku, K. Hattori, and H. Kubota, Chem. Eng. Japan, 5, No. 1, 66-69 (1967).
11. B. H. Chen and G. L. Osberg, Canad. J. Chem. Eng., 45, No. 2, 4 (1967).
12. D. Kunii and O. Levenspiel, Fluidization Engineering, New York (1969).
13. M. G. Slin'ko and V. S. Sheplev, Kinetika i Kataliz, 2, No. 2 (1970).
14. J. R. Grace and D. Harrison, Chem. and Process Eng., No. 6, 127-130 (1970).
15. S. S. Zabrodskii, High-Temperature Apparatus with Fluidization Beds [in Russian], Énergiya, Moscow (1971).
16. P. J. Bakker and P. M. Heertjes, Brit. Chem. Eng., 3, No. 5, 240-245 (1958).
17. M. E. Aerov and O. M. Todes, Hydraulic and Thermal Operating Principles of Apparatus with a Stationary and a Fluidized Granular Bed [in Russian], Khimiya (1968).
18. A. I. Tamarin and V. D. Dunskii, Authors' Disclos. No. 242146, Byull. Otkryt. Izobret., No. 15 (1969).
19. I. F. Davidson and D. Harrison, Fluidization of Solid Particles [Russian translation], Khimiya (1965).
20. K. Godard and J. F. Richardson, Chem. Eng. Sci., 24, No. 4, 663-670 (1969).
21. V. A. Borodulya and A. I. Tamarin, Inzh.-Fiz. Zh., 5, No. 11 (1962).
22. A. I. Tamarin, ibid., 6, No. 7 (1963).
23. A. I. Tamarin and V. A. Bordulya, ibid., 6, No. 11 (1963).
24. I. M. Fedorov, Theory and Design of Desiccation Processes in Suspended State [in Russian], Gosénergoizdat, Moscow (1955).