The Effect of Packing on a Fluidized Bed

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A study has been made of the effect of fixed packing on the properties of a gas-fluidized bed, including minimum fluidization velocity, pressure drop, and bed expansion. Experiments using a range of glass beads as fluidizing solid with smooth uniform spheres as packing indicate that both packing size and the ratio of particle to packing diameter are the main variables in correlating the results. Other solids of varying density and shape have also been used. In addition to smooth spherical packing, rough spheres and variously shaped packings such as Raschig rings, Berl saddles, and a cylindrical, open-ended screen packing have received preliminary study. With the screen packing, which occupies only 5% of the column volume, it has been possible to operate a fluidized bed at a much higher gas throughput without slugging than is possible with a conventional bed.

A preliminary study has also been made of heat transfer rates, and the results indicate that the same factors are significant. With spherical packing, values of *h* of the order of 70% of that in a conventional bed have been obtained, while with screen packing values ranging up to 100% or greater have been observed.

Possible applications of this technique and its limitations are also discussed.

The fluidization of particulate material is a technique which has many industrial applications. Good mixing and high rates of heat and mass transfer are generally associated with its use. However, the nature of conventional gas-solid fluidization imposes certain restrictions on the design of fluid-bed reactors. For example, there is a limit to the length-to-diameter ratio which can be used without serious slugging. Also, if high gas flow rates are desired, the reactor depth must include a large freeboard to allow for fluctuations in bed height.

As a result, various methods of improving the quality and range of applicability of gas-solid fluidization have been proposed. Beck (*1*) suggested a helical rotating screw to permit the use of high L/D_t ratios in reactor tubes. Hall and Crumley (*2*) inserted a series of slightly convex gauze disks to produce smooth fluidization of Fischer-Tropsch catalyst. Other suggestions have included tapered beds to reduce slugging (*3*) and the use of the tray type of internals to break up a reactor into a series of shallow beds (*4*). More recently, Romero and Johnson (*5*) found that the insertion of concentric and helical large mesh screens improved fluidization quality. Volk, et al. (*6*) used vertically oriented internals to break up a large diameter bed into a number of smaller ones with favorable results. Another proposal has been the addition of nonfluidizing dump packings such as Raschig rings and Berl saddles (*7*). The effects and some of the limitations of these techniques are discussed briefly by Zenz (*8*).

Although many proposals have been made for expanding the useful range of gas-solid fluidization, very little quantitative information about their effect on the behavior of fluidized beds has been published. The data which are available, including an extensive study by Massimilla into the use of screen baffles (*9, 10, 11*) and a paper by Gabor, et al. on heat transfer rates in fluidized beds containing fixed spherical packing (*12*), indicate that the changes resulting from the insertion of these devices can be very marked. Reactor internals affect not only the overall quality of the fluidization but also the expansion and the flow patterns of both gas and solid. As a result, heat and mass transfer rates can deviate significantly from those occurring in conventional beds.

The work described in this paper is an introductory investigation into the effect of fixed packing on the properties of gas-fluidized beds. Most of the experimental work was performed with beds of packed spheres, but a limited number of experiments were done using packings of different shapes, in particular a cylindrical screen packing which was felt to be particularly suitable for use in fluidized beds. Results are reported for unit pressure drop and gas velocity at incipient fluidization and for bed expansion. A limited number of results showing the effect of packing on heat transfer rates are also given.

APPARATUS AND PROCEDURE

The equipment used was of conventional design. The column consisted of a 3-in. I.D. plastic tube mounted on a 60 deg. cone. The supporting grid, which was placed between the column and the cone, was of porous stainless steel. The pressure drop was measured against atmosphere by means of a manometer from a tap below the cone and corrected for pressure drop across the supporting grid and packing where necessary.

The packing consisted of closely sized glass spheres of three sizes (0.750, 0.580, 0.238 in.) and smooth alundum spheres of 0.150-in. diam. Two sizes of rough alundum spheres (0.750, 0.262 in.) were also used. In addition, a limited number of tests were performed with other packing shapes such as Berl saddles, Raschig rings, and ½-in. × ½-in. open-ended 14-mesh screen cylinders (wire 0.020 in. in diam.).

The fluidized solids consisted of four size ranges of glass beads (0.0035, 0.0047, 0.0110, 0.0190 in. mean diam. and 90% within the range $\pm 0.1\ d$) and four sizes of angular silica (0.0049, 0.0070, 0.0106, 0.0195-in. mean diam.). Nickel, aluminum and alumina powders were also used. Dry air metered by a rotameter was used as the fluidizing gas.

In a typical run, the packing was poured into the column to a height slightly greater than that of the particulate bed at incipient fluidization. Then a known weight of bed was added, followed by sufficient packing to allow for a three- or fourfold expansion. A ratio of $H_{mf}/D_t \approx 1.5$ was maintained for most runs.

To measure the pressure drop and expansion, the gas flow was increased until fluidization was definitely occurring and then it was decreased incrementally to zero flow with pressure drop and bed height being noted. The gas flow rate was then increased to fluidization level and again increased incrementally to maximum conditions. This procedure was adopted to avoid excessive pressure drop in the quiescent bed range

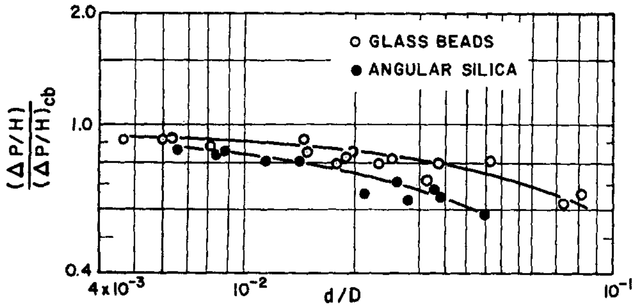

Fig. 1. The effect of smooth spherical packing on unit pressure drop at incipient fluidization.

because of settling and to minimize the effect of any solids holdup in the packing over the fluidization range.

PRESSURE DROP

Fluidization occurs when the pressure drop across a particulate bed becomes equal to the effective bed weight, that is,

$$\Delta P = H_{mf}\,(1 - \epsilon_{mf})\,(\rho_s - \rho_f)$$

Further increases in gas flow will not affect the pressure drop as the bed will expand to facilitate its passage. The unit pressure drop ($\Delta P/H_{mf}$) at minimum fluidization is thus dependent on ϵ_{mf}, the bed voidage. In turn, ϵ_{mf} depends to some extent on the diameter of the tube, although in conventional beds d/D_t is sufficiently small for the effect to be negligible. If, however, the fluidization occurs in the interstices of a packed bed, this effect can no longer be ignored. As d/D increases the value of ϵ_{mf} will also increase until the limiting conditions are reached under which the particles cannot pass through the points of constriction in the packed bed. The maximum value of d/D depends of course on the shape and the configuration of the packing. For beds of uniform spheres $(d/D)_{\max} = 0.414$ for simple cubic arrangement and 0.154 for close-packed hexagonal. Since randomly packed beds display incomplete hexagonal configuration (*13*), $(d/D)_{\max} = 0.154$ will apply to them also. In actual practice $(d/D)_{\max}$ would be lower as the particles require a finite clearance to fluidize (*14*). The magnitude of the clearance depends on the shape and roughness of both particles and packing (*15*).

Values of $\Delta P/H_{mf}$ are reported in Tables 1-8*. Typical results are shown in Figure 1. For glass beads fluidized in smooth spherical packing $(d/D)_{\max} = 0.096$ based on mean particle size and 0.120 based on maximum particle size. For angular particles $(d/D)_{\max}$ was 0.055 based on mean screen size. Rough packing lowered the value of $(\Delta P/H_{mf})$.It can be seen from the reported data that in the presence of packing $(\Delta P/H_{mf})$ is always lower than for conventional fluidization. This fact indicates the existence of gas channeling and is supported by calculation of ϵ_{mf} based on bed dimensions and weight. $(\Delta P/H_{mf})$ is approximately 10% lower than $(1 - \epsilon_{mf})$ $(\rho_s - \rho_f)$ for glass beads in the presence of spherical packing.

When packings of other shapes were used, channeling was again observed at low flows. In addition, the ΔP vs. u values showed a lack of reversibility as indicated in Figure 2. This decrease in ΔP in the fluidizing region is owing to the holdup of particulate material in packing interstices. When angular silica was used as the particulate bed, smooth fluidization could not be developed in the presence of Raschig rings and Berl saddles. When the gas flow was reduced from maximum values virtually no bed contraction was observed. Screen packings interfered very little with the fluidization of both glass beads and angular silica. Channeling was very slight and slugging was completely eliminated. A few runs were performed using nonspherical aluminum powder and spherical nickel. Results were similar to those obtained with the spherical and nonspherical materials already discussed.

VELOCITY OF MINIMUM FLUIDIZATION

The minimum fluidization velocity was taken as the point at which the pressure drop across the bed became constant. For a conventional bed this point was well defined, but for fluidization in packing interstices this was not always true. In the event that the pressure drop vs. gas flow curve did not show a sharp discontinuity u_{mf} was taken to be at the point of intersection of projections of the two parts of the curve.

Suggested correlations for determining u_{mf} in terms of particle and fluid properties show large deviations, part of which is no doubt owing to differences in definition of u_{mf} (*16*). The results obtained in the present experiments for conventional fluidization are about 8% greater than those obtained by Miller and Logwinuk (*17*), who defined u_{mf} in the same way.

Results are given in Tables 1-8 for the fluidization of silica and glass beads in a range of spherical packings as well as in packings of different shape. Gas velocities given have been corrected for packing voidage; tortuosity was ignored. Typical values are shown in Figure 3. A suggested relationship between G_{mf} (and of course u_{mf}) (*18*) and the properties of the fluidized solid is

$$G_{mf} = \frac{0.005\, d^2\, \rho_f(\rho_s - \rho_f)\, \phi_s^2}{\mu} \cdot \frac{\epsilon^3{}_{mf}}{1 - \epsilon_{mf}} \quad \text{for } Re < 10$$

For given gas-solid system therefore

$$u_{mf} = K\left(\frac{\epsilon^3{}_{mf}}{1 - \epsilon_{mf}}\right) d^2$$

As the measured values of ϵ_{mf} show a dependence on (d/D) it would then be expected that the relationship between u_{mf} and d would be similarly affected. However, the observed values of u_{mf} show only very slight dependence on (d/D) since with D as parameter the curves relating u_{mf} to d are very nearly parallel straight lines on a log-log plot. See Figure 3. From this it must be concluded that the value of ϵ_{mf} measured does not represent a true porosity and that the change in u_{mf} must be owing to a form of channeling combined with an incomplete use of the void space present in the packing configuration. An empirical equation of the form

$$u_{mfD} = \left(1 + \frac{0.169}{D}\right) u_{mf}$$

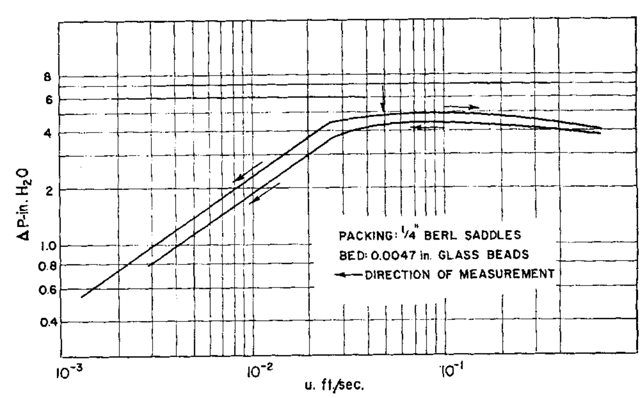

Fig. 2. The effect of packing on pressure drop reversibility.

* Tabular material has been deposited as document 7575 with the American Documentation Institute, Photoduplication Service, Library of Congress, Washington 25, D. C., and may be obtained for $1.25 for photoprints or for 35-mm. microfilm.

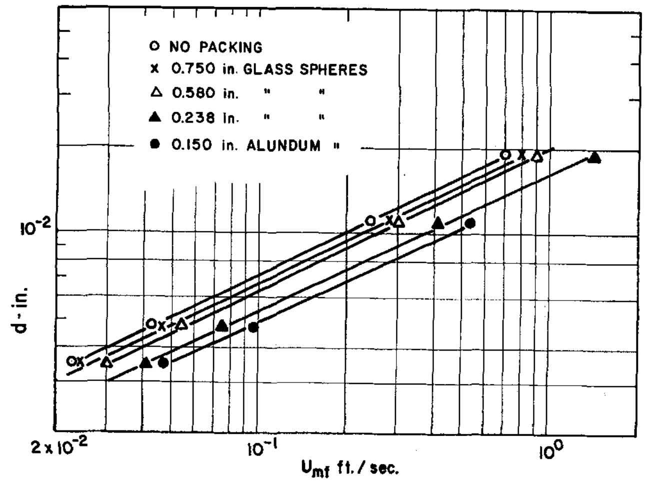

Fig. 3. The effect of smooth spherical packing on the minimum fluidization velocity of glass beads.

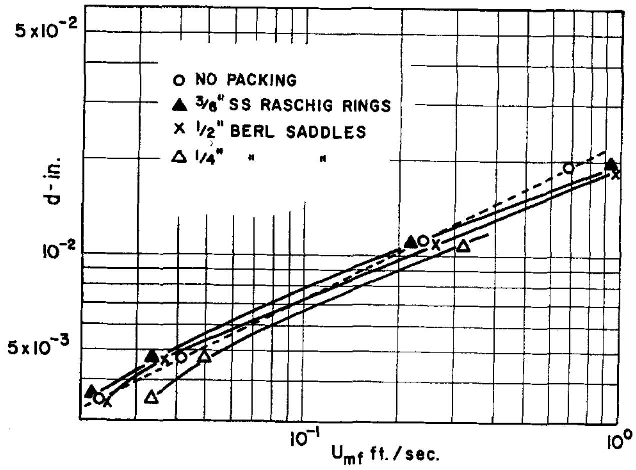

Fig. 4. The effect of packing shape on the minimum fluidization velocity of glass beads.

correlates the results obtained with smooth spherical packing and glass beads. A similar equation

$$u_{mf} = \left(1.15 + \frac{0.04}{D}\right) u_{mf}$$

is rather less satisfactory for the system of angular silica and smooth spherical packing. The one size of nickel powder investigated (which was in the form of rough spheres) was influenced by packing in the same way as glass beads, but showed slightly more dependence on the presence of packing.

Values of u_{mf} for glass beads fluidized in rough spheres are similar to those obtained with smooth spherical packing, except in the lower particle size ranges where channeling was increased. Values of u_{mf} for glass beads in the presence of nonspherical packings are given in Figure 4. It is noted that in some cases u_{mf} based on total packing voidage is lower than it is in the absence of packing; this indicates that for these packings only a fraction of the total voidage is available for gas flow. The ½-in. cylindrical screen packing increased the minimum fluidization velocity of gas in beds of glass beads by about 25%.

As yet no completely satisfactory explanation of the effect of dump packing on the minimum fluidization velocity has been found. The principal factor appears to be the characteristic dimension of the packing rather than the particle-to-packing size ratio. This fact suggests that at velocities just above u_{mf} the packing partially supports the bed and keeps it from settling into a more compact configuration as the velocity is reduced. Hence the pressure drop is reduced and fluidization stops at a higher velocity than normal. Since the amount of surface area presented by the packing varies inversely with packing diameter its effect would become greater as packing diameter decreases as is shown by the experimental results.

EXPANSION

Since no generally applicable correlation for bed expansion in a gas fluidized bed is available (*19*), experimental results are tabulated as reduced expansion, $H/H_{mf} = R$, as a function of reduced gas velocity u/u_{mf}. The reduced expansions observed in the presence of packing can then be compared with those found in a conventional bed at the same reduced velocity (Figure 5). The value of u_{mf} used in calculating u/u_{mf} is the one obtained experimentally for the particular combination of particle and packing.

In all cases the amount of expansion observed is increased by the addition of packing. For a given particle, expansion increases as packing diameter decreases, and for a given packing the expansion increases with increasing particle size as in conventional fluidization. In addition, the magnitude of R/R_{cb} increases with increasing reduced velocity. In appearance, the fluidized bed containing packing has local regions of varying density depending on packing configuration, but the total expansion is free from marked fluctuations. On the basis of incremental pressure drop measurements the bed density is not a function of vertical position in the bed.

Although no quantitative description of the expansion of mixed fluid beds can be given at present, the trends observed in the data can be explained qualitatively. In gas fluidized beds much of the expansion is caused by the formation of voids or bubbles which rise through the bed at a rate dependent on their diameter (*20*). The presence of packing restricts the size of the void which can form and lowers the rate at which the particulate material can slip around the ascending voids, particularly at constriction points in the column. The net effect of these factors is to increase the amount of expansion. Obviously they will be of increasing significance as packing size decreases and the unit void spaces become smaller.

Another factor influencing the results as reported is the nature of the gas flow through the packing. Both u and u_{mf} are calculated on the basis of packing voidage in the column used, ignoring tortuosity. Thus, although the value of the ratio u/u_{mf} may be correct, the magnitude of the gas velocity will be considerably closer to the terminal velocity of the particles, resulting in a greater expansion. In addition, u/u_{mf} is based on u_{mf} for the given particle-packing combination. If the most meaningful value of u_{mf} is taken to be u_{mf} with no packing, then the values of u/u_{mf} will be increased for a given expansion. The effect of applying this second correction is shown in Figure 6. The effect of packing on expansion is considerably reduced, but is by no means eliminated. This net difference in expansion is considered to be a genuine packing effect.

Angular silica was affected by packing in a similar way. When nonspherical packings were used with glass beads, the values of R obtained were frequently misleading because of the difficulty of determining an appropriate value of the packing voidage. Cylindical screen packing behaved very much like spheres of the same diameter.

Closely related to the expansion of a fluidized bed is the phenomenon of slugging. Leva gives recommended

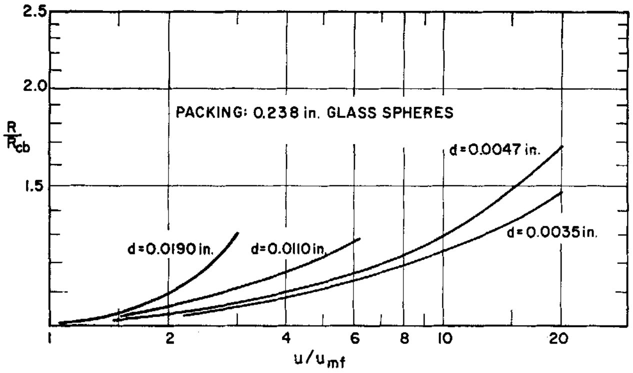

Fig. 5. The expansion of glass beads in spherical packing.

expansion limits for close cuts of solids which range from $R = 1.2$ to 1.3 for a wide range of particle diameters including those used here. For a particle of 0.019-in. diameter his correlation shows slugging beginning at $(u/u_{mf}) = 2$, and for $d = 0.0035$ in., at $(u/u_{mf}) = 9$ (21). If these values are accepted as limiting then the presence of packing permits the use of much higher gas velocities without noticeable fluctuations in R. For example, with 0.75-in. spherical packing a bed of 0.0035-in. glass beads was expanded to $R = 2.05$ at $u/u_{mf} = 70$, and a bed of 0.0190-in. glass beads to $R = 3.6$ at $u/u_{mf} = 7$. With an allowance for the tube volume occupied by the packing, these values represent an increase in gas velocity based on the empty tube of from 50 to 250%. For packings of high voidage such as the cylindrical screen packing, the increase in gas throughput may be as high as 200 to 500%.

Since it was reported that baffles in a fluidized bed alter its characteristics from aggregative to particulate (9), an attempt was made to correlate the results in terms of the equations proposed by Richardson and Zaki (22). This did not prove to be successful, and one may conclude that the expansion is not truly particulate even though slugging is eliminated. Rather, packing controls the size of the gas voids and attenuates their effect on the bed.

HEAT TRANSFER

A preliminary investigation into the effect of fixed packing on the rate of wall-to-fluid bed heat transfer in a fluidized bed has also been made. The apparatus consisted of a steam-jacketed brass tube 4-in. I.D. and 3½ in. long mounted on a porous metal grid. Above the jacketed section was a 4-in. diam. transite tube, electrically wired to eliminate upward or outward heat loss. Fluid temperatures were measured by a network of thermocouples in the bed. Heat transfer coefficients were calculated by the equation $q = hA\Delta T$. The heat input was determined from the rise in temperature of the fluidizing gas as it passed through the bed, and ΔT was taken as the mean difference between the wall temperature and the mean bed temperature at the same level in the bed over the whole jacketed section. The bed height at incipient fluidization was normally about 4 in.

Again, a comparison between the value of h obtained in a conventional fluidized bed and that found in a packed fluidized bed was of chief interest. However, the values of h reported for the conventional bed are in reasonable agreement with values given by Leva (23).

Typical results for the relationship of h to u/u_{mf} are given in Figures 7 and 8 which show the effect of spherical and other packings on the heat transfer coefficient. With smooth spheres an increase in sphere diameter resulted in higher heat transfer coefficients, while for a fixed packing size a decrease in particle size produced an increase in the maximum value of h. The value of u/u_{mf} for which h was a maximum decreased with increasing particle size but was slightly increased by the addition of packing.

The effect of spherical packing on heat transfer shows the influence of packing on the movement of solids in the bed. As expected, with small packing particle movement is more severely hindered with the result that h_{max} is reduced. This effect is no doubt owing to the number and size of the constriction points in the bed. Also, at each point on the wall which is touched by the packing a dead area exists which cannot be reached by the fluidizing solids. Preliminary attempts at correlation of the data indicate that h is dependent on both D and d/D. In view of the limited number of data and the small heat transfer area used further attempts to treat the data were not made.

The effect of other packing shapes is shown in Figure 8. Berl saddles and Raschig rings permit better heat transfer than do spheres, but all three packings substantially reduce h_{max}. However, with cylindrical screen packing h was not appreciably lower than in a conventional bed at high gas rates. This effect is considered to be owing to the better solids circulation occurring in this type of fluidized packed bed and the absence of slugging.

Similar experiments were performed with nickel and alumina powders, and the results were consistent with those reported for glass beads. As a check on the effect of particle shape a few runs were performed using a nonspherical aluminum powder. Values of h observed were 60% of those reported for glass beads; this indicates that shape factor affects the movement of solids.

CONCLUSIONS

The addition of fixed packing to a gas-fluidized bed has a marked effect on its properties. Slugging and the consequent fluctuations in bed level are eliminated permitting smooth operation at gas flows as high as 75% of the terminal velocity of the bed. The packing must be sufficiently large to permit rapid flow of solids through the interstices. For smooth spheres, the limiting ratio of particle-to-packing diameters is about 0.090. As the angularity of the bed particles increases this ratio decreases until for very irregular particles it is not possible to develop fluidization in the presence of packing.

The unit pressure drop at incipient fluidization is lower than that for a conventional bed; this indicates that the bed is kept in a more open structure. The change in $\Delta P/H$ is dependent on the ratio of the particle-to-packing diameters. However, for spherical packing, the velocity of minimum fluidization has been shown to be primarily a function of packing diameter; this suggests that the packing promotes a form of channeling at these velocities.

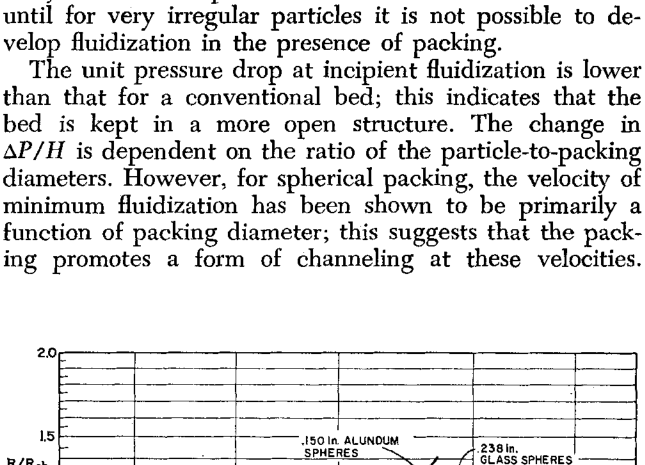

Fig. 6. The reduced expansion of 0.0035-in. glass beads in spherical packings based on U_{mf-cb}.

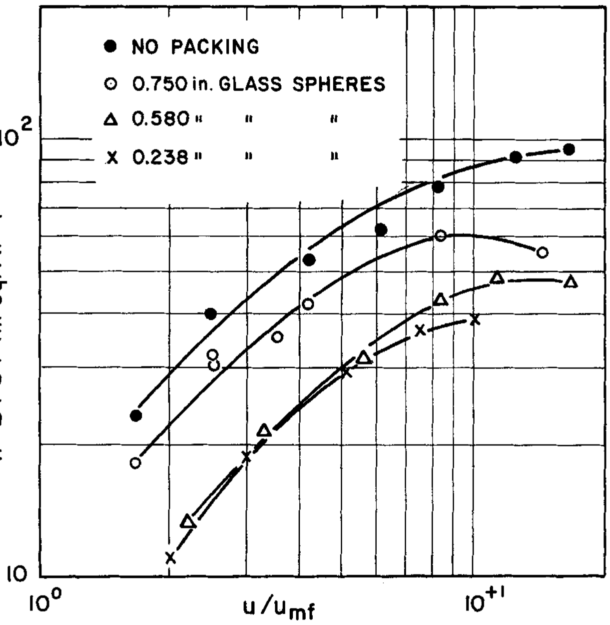

Fig. 7. The effect of packing diameter on heat transfer in a bed of 0.0047-in. glass beads.

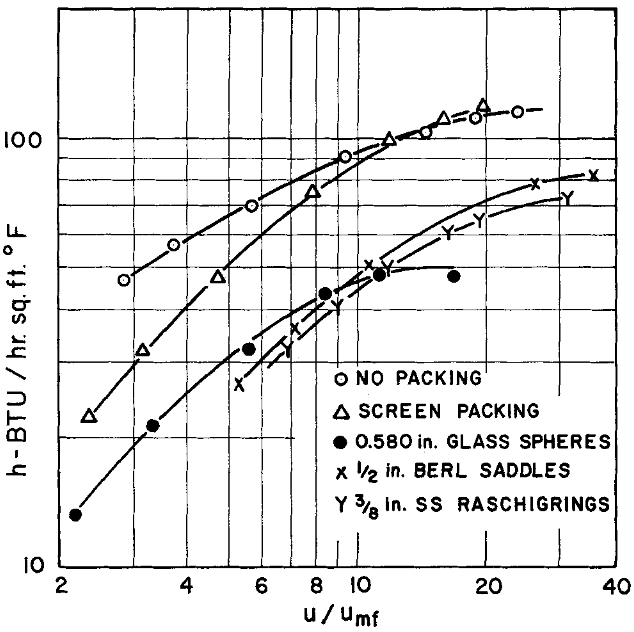

Fig. 8. The effect of packing shape on heat transfer in a bed of 0.0047-in. glass beads.

The presence of packing in a fluidized bed increases its expansion by an amount which is dependent on the reciprocal diameter of the packing.

The rate of wall-to-bed heat transfer is in most cases adversely affected by the addition of packing because of the restriction of particle movement within the bed and at the vessel walls. This was true for spheres, Berl saddles, and Raschig rings. However, cylindrical screen packing has been found to cause little or no reduction in maximum heat transfer rates because of its open structure. Its relatively small interference with particle movement, combined with its high porosity and ability to eliminate slugging, suggest that this type of packing should have a useful application in fluidized beds of free-flowing solids.

NOTATION

D = packing diam. in.
D_t = tube diam. in.
d = particle diam. in.
G = mass flow rate lb./hr. sq. ft.
H = bed height
h = heat transfer coefficient B.t.u./hr. sq. ft. °F.
L = tube length
ΔP = pressure drop cm. water
q = heat flow rate B.t.u. hr.
R = H/H_{mf}
ΔT = temperature difference °F.
u = superficial gas velocity ft./sec.
ϵ = average bed porosity
μ_f = viscosity
ρ = density
ϕ_s = shape factor

Subscripts

cb = conventional bed
mf = minimum fluidization
f = fluid
max = maximum
mfD = minimum fluidization with packing of diam. D
s = solid
t = terminal

LITERATURE CITED

1. Beck, R. A., *Ind. Eng. Chem.*, **41**, 1242 (1949).
2. Hall, C. C., and P. Crumley, *J. Appl. Chem.* (*London*), 2, Suppl. Issue I, S47-S55 (1952).
3. Lawroski, S., et al., *Reactor Fuel Process.*, **2**, 50 (1959).
4. Huff, L. C., *U. S. Patent 2,607,662.*
5. Romero, J. B., and L. N. Johnson, *Chem. Eng. Progr. Symposium Ser. No. 38*, **58** (1962).
6. Volk, William, C. A. Johnson, and H. H. Stotler, *Chem. Eng. Progr.*, **58**, 44 (March, 1962).
7. Matheson, G. L., *U. S. Patent 2,557,685.*
8. Zenz, F. A., and D. F. Othmer, "Fluidization and Fluid-Particle Systems," p. 273, Reinhold, New York (1960).
9. Massimilla, Leopoldo, and Sergio Bracale, *Ricerca Sci.*, **26**, 487 (1956).
10. ————, Antonio Cabella, *ibid.*, **27**, 1853 (1957).
11. Bracale, Sergio, Antonio Cabella, and Leopoldo Massimilla, *Chim. Ind.*, **40**, 621-7 (1958).
12. Gabir, J. D., W. J. Mecham, and A. A.Jonke, Paper presented at 46th national meeting of Am. Inst. Chem. Engrs. in Los Angeles (February, 1962).
13. Wadsworth, Jack, National Research Council of Canada, *Mechanical Engineering Report MT-41*, Ottawa, Canada (1960).
14. Atomic Energy Commission *Research and Development Rept. ANL-6145* (1960).
15. Zenz, F. A., and D. F. Othmer, "Fluidization and Fluid-Particle Systems," p. 168, Reinhold, New York (1960).
16. Leva, Max, "Fluidization," p. 62, McGraw-Hill, New York (1959).
17. Miller, C. O., and A. K. Logwinuk, *Ind. Eng. Chem.*, **43**, 1220 (1951).
18. Leva, Max, "Fluidization," p. 63, McGraw-Hill, New York (1959).
19. Davies, Geoffrey, and D. B. Robinson, *Can. J. Chem. Eng.*, **38**, 175 (1960).
20. Davidson, J. F., et al., *Trans. Inst. Chem. Engrs.* (*London*), **37**, 323 (1959).
21. Leva, Max, "Fluidization," p. 102, McGraw-Hill, New York (1959).
22. Richardson, J. F., and W. N. Zaki, *Trans. Inst. Chem. Engrs.* (*London*), **32**, 35 (1954).
23. Leva, Max, "Fluidization," p. 179-246, McGraw-Hill, New York (1959).

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