Am. Chem. Soc., 77, 3427 (1955).
———, D. Z. Lippmann, R. F. Curl, C. M. Huggins, and D. E. Peterson, "The Volumetric and Thermodynamic Properties of Fluids. II. Compressibility Factor, Vapor Pressure and Entropy of Vaporization," *ibid.*, 3433 (1955).
Plank, R., and L. Riedel, "Ein neues Kriterium für den Verlauf der Dampfdruckkurve am Kritischen Punkt," *Ing.-Arch.*, 16, 255 (1948).
Rankine, W. J. M., "On the Centrifugal Theory of Elasticity, as Applied to Gases and Vapors," *Phil. Mag.*, 2, 509 (1851).
Reid, R. C., and T. K. Sherwood, *The Properties of Gases and Liquids*, McGraw-Hill, New York (1966).
Reynes, E. G., and G. Thodos, "Application of a Reduced Vapor Pressure Equation to Nonhydrocarbon Substances," *Ind. Eng. Chem. Fundamentals*, 1, 127 (1962).
Riedel, L., "Eine neue universelle Dampfdruckformel Untersuchungen über eine Erweiterung des Theorems der übereinstimmenden Zustande," *Chem.-Ing.-Tech.*, 26, 83 (1954).
Thek, R. E., and L. I. Stiel, "A New Reduced Vapor Pressure Equation," *AIChE J.*, 12, 599 (1966).
Thodos, G., "Vapor Pressure of Normal Saturated Hydrocarbons," *Ind. Eng. Chem.*, 42, 1514 (1950).
Waring, W., "Form of a Wide-Range Vapor Pressure Equation," *ibid.*, 46, 762 (1954).
Young, S., "The Vapour Pressures, Specific Volumes and Critical Constants of Normal Heptane," *Chem. Soc. J.*, 73, 675 (1898).
Zia, T., and G. Thodos, "A Generalized Vapor Pressure Equation for Hydrocarbons," *Can. J. Chem. Eng.*, 52, 630 (1974).

Manuscript received March 29, 1977; revision received June 20, and accepted June 29, 1977.

Gas Backmixing, Solids Movement, and Bubble Activities in Large Scale Fluidized Beds

H. V. NGUYEN
A. B. WHITEHEAD
and
O. E. POTTER

Department of Chemical Engineering
Monash University
and
Division of Chemical Engineering
CSIRO
Clayton, Victoria, Australia

This paper illustrates the interrelation between bubbles, solids movement, and gas backmixing in large fluidized beds by presenting experimental data obtained with a 1.22 m square bed. The results demonstrate a close relationship between gas backmixing, solids movement, and bubble coalescence, with the behavior altering markedly as bubble coalescence patterns change.

SCOPE

Fluidization is attracting a high degree of interest, but detailed research into gas and solids mixing within the bed is lacking, particularly on a large scale.

Some progress, however, has been recently achieved with small scale equipment. In beds up to 30 cm in diameter, the phenomenon of gas backmixing has been experimentally studied in both reactive and nonreactive systems (Nguyen and Potter, 1974, 1975; Fryer and Potter, 1976). These studies reveal the important role that bubbles play in the overall gas and solid movement in the bed through their coalescence, size variation, and distribution. Experimental data showed that the countercurrent backmixing model provides an acceptable description of a fluidized bed. This model, incorporating the bubble wake, represents a significant forward step from other earlier one- or two-phase models which assume perfect mixing or plug flow in the dense phase.

Nevertheless, as reviewed by Potter (1971), information on systems of industrial size is sketchy and incomplete, uncomfortably so at this time when fluidization is assuming major importance in coal combustion and liquid fuel processes.

This paper reports experimental data on gas backmixing in a large fluidized bed operated at two fluidizing velocities. Together with information on solids circulation patterns and bubble distribution, these data provide a much more complete picture of the overall gas and solids movement within the bed than has hitherto been available.

CONCLUSIONS AND SIGNIFICANCE

Tracer studies in the large fluidized bed employed exhibit backmixing at about the level predicted by the countercurrent backmixing model. The downward movement of solids which is the dominant feature of the model certainly occurs and does so at about the expected order of magnitude.

The solids movement is shown to be clearly related to the bubble pattern and gas movement in the particulate phase related to the solids movement. Modeling of the bubble patterns, therefore, becomes considerably important. If the stable bubble pattern in the upper region of the bed is such as to leave a persistent central area relatively bubble free, then a pronounced solids downflow in this area, associated with strong gas backmixing, will occur. This condition was found to exist at 15.2 cm/s. At 24.4 cm/s, however, an unstable bubble pattern with concomitant variation in solids flow regimes was evident, this being reflected by short-term variation in the pressure profile recorded at the bed base. The data at the lower velocity are to some extent compatible with the Gulf stream mixing concept of Merry and Davidson (1973). However, at the higher velocity, the streams are more variable and difficult to define.

It is clear from these results that the overall patterns of gas and solids mixing in large fluidized beds are dependent on the fluidizing velocity. Another important indication is that sampling in large beds will not be reliable if it is not done systematically and extensively throughout the bed. A few sampling points in the bed may not reflect accurately the overall gas and solids mixing.

The results represent the first integrated explanation of the interrelation of gas and solids movement in a large system.

Gas backmixing in aggregatively fluidized beds is closely identified with solids and bubble movements. Rising bubbles carry along wakes of solids and associated gases, as demonstrated by Noble (1962), Rowe (1961), Woollard and Potter (1968), and release these materials on bursting at the bed surface. The released solids then flow downwards against the rising bubbles and their wakes. As the fluidizing velocity increases, the downflow also increases, and at a critical fluidizing velocity U_{CRIT}, the velocity of downflowing solids exceeds the interstitial velocity so that interstitial gas is carried downwards, or backmixed, as described by Stephens et al. (1967), Latham et al. (1968); and Nguyen and Potter (1974, 1975). In these papers, it has been assumed that the downflow is uniform. Since bubble coalescence is not a random process, preferred bubble paths exist as demonstrated by Whitehead et al. (1967*a*), utilizing photography of the bed surface as well as an internal probe. For example, one condition observed was that in the top region of the 1.22 m of bed, there were four bubble tracks disposed in the four corners, leaving a region in the center where downflow would be encouraged. Whitehead et al. (1970) have also made measurements of solids movement using dyed sand or labeled spheres as solids tracers. This work offers some very useful information on solids movement, demonstrating in some cases a strong downflow in a small area at the center of the bed. Werther (1974) and Schmalfeld (1976), working with beds up to 1 m in diameter, reported the existence of a solids circulatory pattern under certain conditions. Other studies on a comparatively large scale, as reviewed by Potter (1971), have been concerned with cracking catalyst, which has properties rather distinct from sands and mineral particles (Geldart, 1970; Thiel and Potter, 1977).

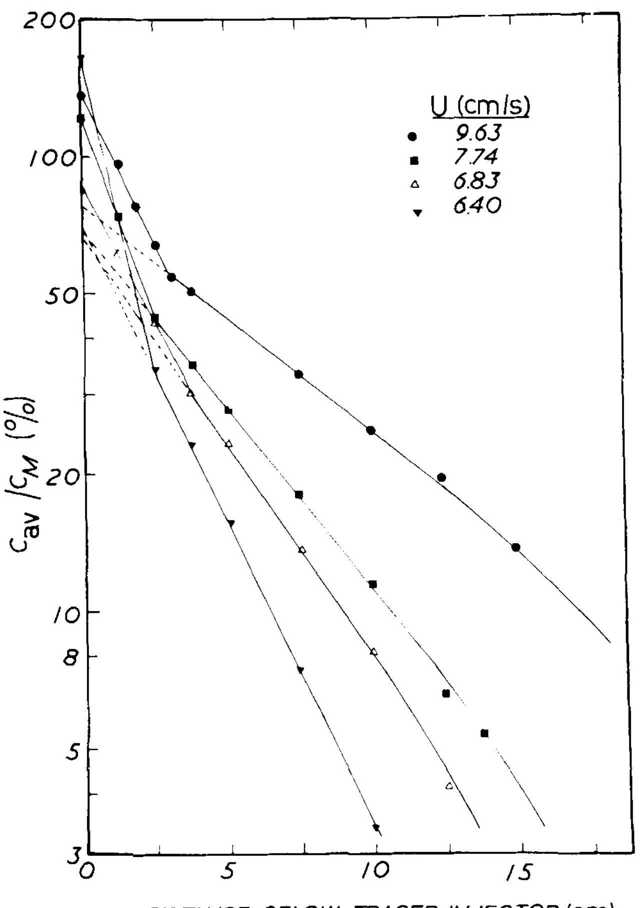


Fig. 1. Axial mean concentration profiles below plane of injection of tracer in a backmixing experiment. (Bed diam. = 30 cm, particle size = 145 μm, U_{mf} = 2.5 cm s^{-1}, helium tracer.) Data of Nguyen (1975).

GAS MIXING IN BEDS FROM 15 TO 30 CM

The model of gas and solids movements proposed by Stephens et al. (1967) was investigated by a number of workers. Latham and Potter (1970), Hamilton et al. (1970), and Nguyen and Potter (1974, 1975) investigated backmixing of tracer gas injected across a plane near the top of the bed. In these studies, the concentration of tracer across a section was averaged, and the logarithm of the average concentration was plotted against the distance below the plane of injection, Figure 1. Note how gas backmixing increases with fluidizing velocity. If the volumetric rate of gas exchange per unit volume of bubble is less for large bubbles than for small bubbles, then backmixing will be stronger when large, rather than small, bubbles are present. Obviously, if the rate of exchange among the phases is infinite, then no backmixing will be observed. A computation, in terms of the theoretical development as expounded by Nguyen and Potter (1974, 1975), assuming uniformity of backmixing across a section, has been made and is compared with the data taken in the large beds in Figure 5. Experimental data obtained in the 30 cm diameter bed, however, showed that gas backmixing was not uniform across the bed section. Figure 2 presents a set of radial tracer concentration profiles to illustrate this point. At the velocity employed, gas backmixing was at its maximum at the center of the bed, went through a minimum at larger radii, and slightly recovered at the wall. This pattern was found compatible with known bubble and solids movement. Bubbles were observed to form regular tracks in the area between center and wall, after an initial period of rapid coalescence near the distributor. The downflow stream of solids and backmixed gas was evidently forced to flow predominantly in the central region for the specific situation illustrated.

BUBBLE AND SOLIDS MOVEMENT DATA IN BEDS FROM 60 TO 122 CM

Studies of solids mixing and bubble distribution were performed by Whitehead and et al. (1970, 1976 *b*, 1977) in a 1.22 m bed loaded with sands of various sizes, using solids tracers and light sensor techniques. Labeled plastic spheres, arranged in a 61 cm pattern situated at the middle and just above the bed surface, were dropped into the bed (Whitehead et al., 1976 *b*). Fluidization was stopped after a known time and the stagnant bed sectioned, the positions of the spheres being recorded. Alternatively, colored sands were released at various locations in the bed, and after a predetermined time, fluidization was stopped and the movement of colored sands was traced. These tracer techniques revealed approximately the size, location, and velocity of the solids streams in the bed. Bubble distribution was determined by the use of arrays of light sensors to detect the appearance of bubbles at each bed level (Whitehead et al., 1967 *b*). From these findings, a complete picture of bubble and solids movement was presented. Bubbles, by virtue of their coalescence and consequent establishment of bubble tracks, influenced the formation of solids streams. Solid circulation occurred with downflow at areas unoccupied by bubble tracks and upflow in the wakes moving with the bubbles.

In the same bed, Whitehead et al. (1970) and Whitehead and Auff (1976*a*) measured the mean differential pressure at the distributor of one tuyere at the corner and one at the center of the bed. It was found that a stable pressure profile with a maximum at the center and a minimum at the four corners of the bed was always associated with a strong downflow of solids in the central region.

This paper presents gas backmixing data obtained at the same operating conditions employed by Whitehead et al.

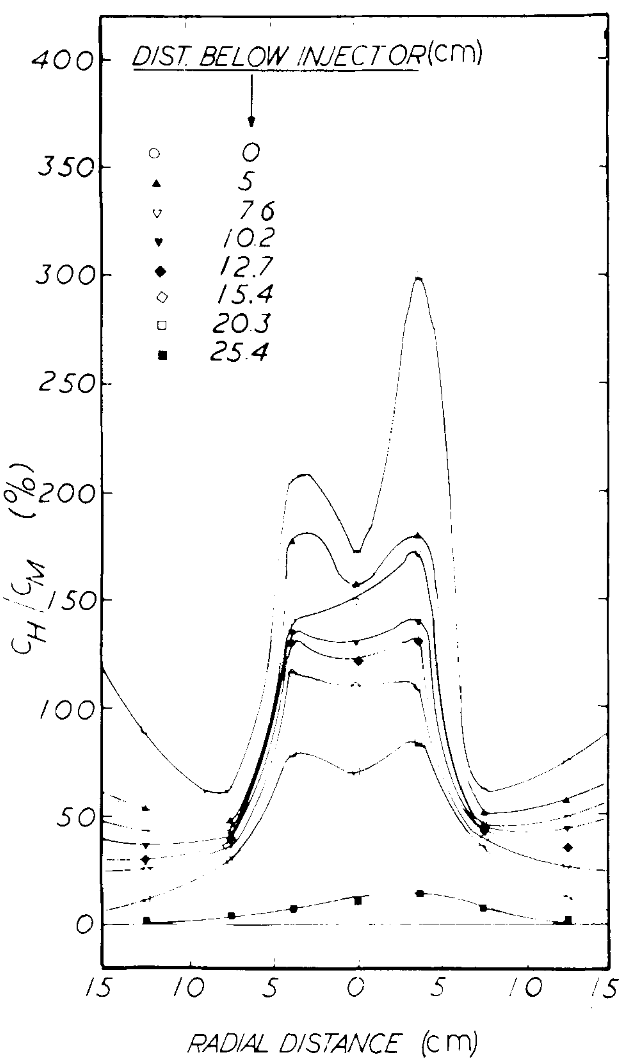


Fig. 2. Radial concentration profiles at different depths below plane of injection of tracer in a backmixing experiment. (Bed diam. = 30 cm, particle size = 85 μm, U_{mf} = 0.8 cm s^{-1}, helium tracer.) Data of Nguyen (1975).

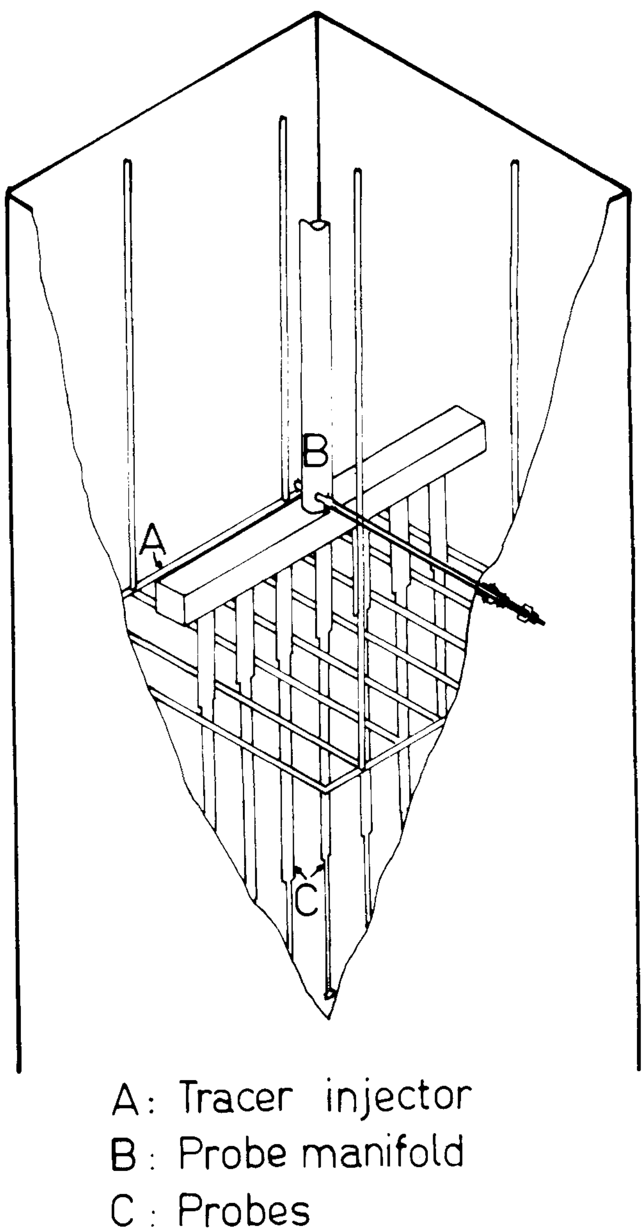
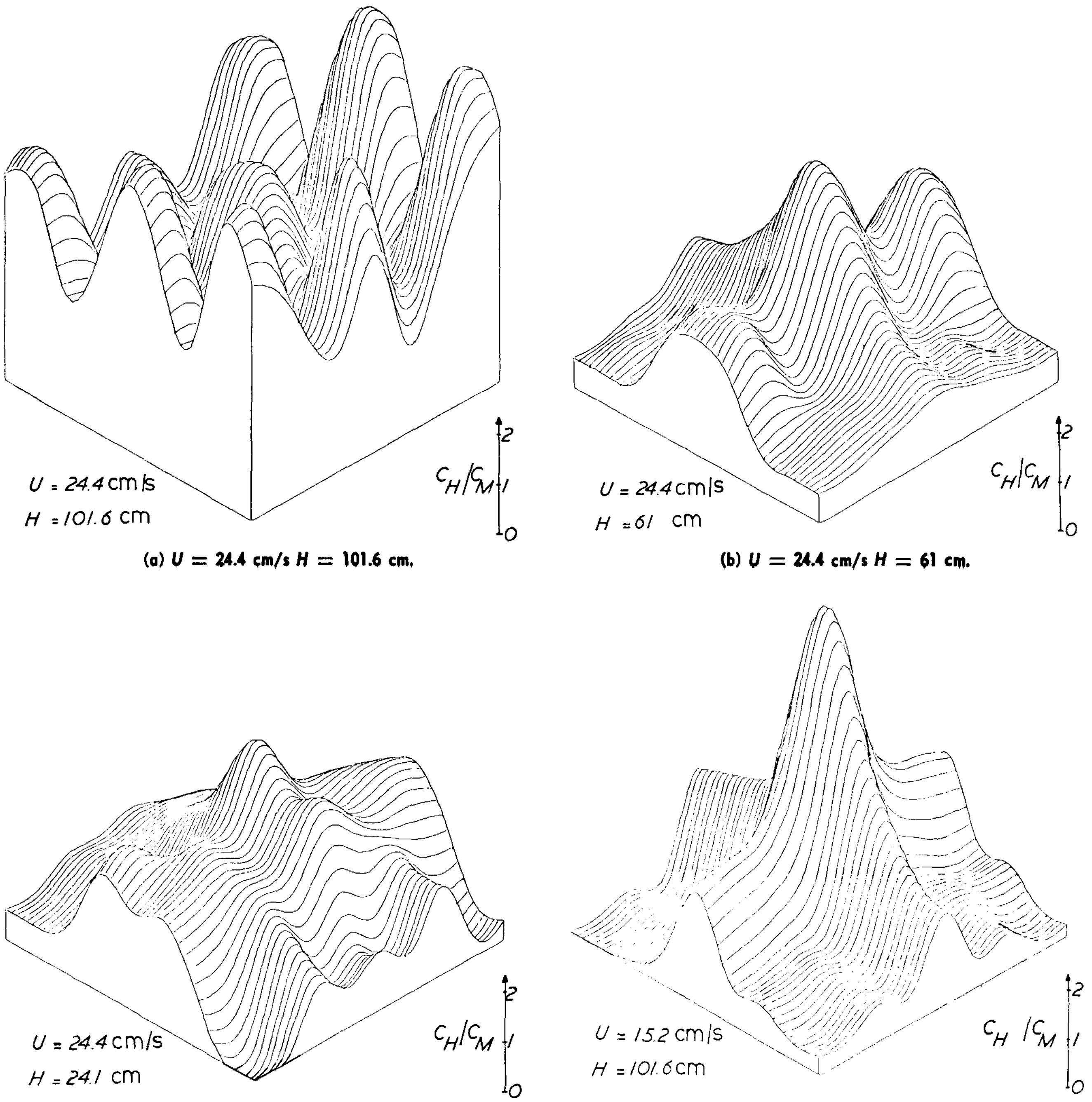


Fig. 3. Isometric cutout view of the 1.22 m of bed.



(a) $U = 24.4$ cm/s $H = 101.6$ cm.

(b) $U = 24.4$ cm/s $H = 61$ cm.

(c) $U = 24.4$ cm/s $H = 24.1$ cm.

(d) $U = 24.4$ cm/s $H = 14$ cm.

Fig. 4. Isometric representation of tracer concentration at various bed levels in a 1.22 m of

EXPERIMENTAL

Apparatus

The equipment is illustrated in Figure 3.

Fluidized Bed. The 1.22 × 1.22 m of bed used in this work has been described elsewhere (Whitehead et al., 1967 *a*).

The distributor had thirty-six tuyeres with 10.2 cm diameter caps arranged in a 20 cm of pitch. The design of tuyere is available elsewhere (Whitehead et al., 1967*a*, 1970). Pressure drop across the distributor plate was kept at 38 cm water with the use of interchangeable orifices of 12.7 and 15.75 mm for the two velocities employed. The bed was loaded with 1.5 m (static height) of silica sand (S1) having an incipient velocity of 2.5 cm/s.

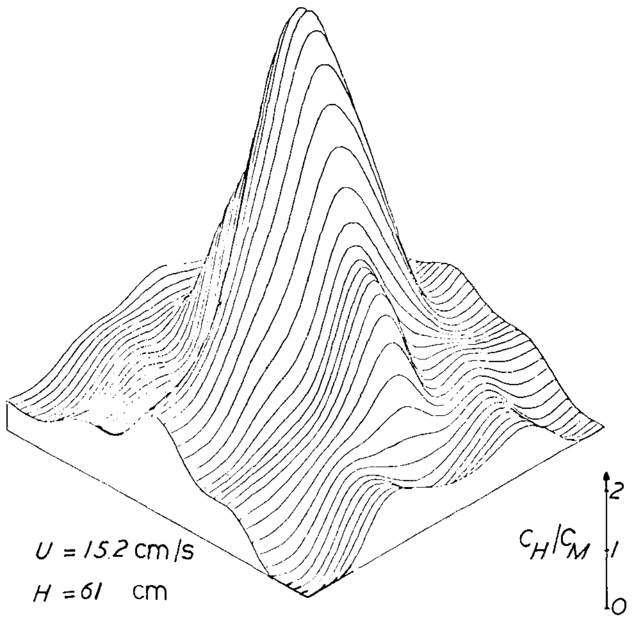
Tracer injector and sampling probes. The tracer injector consists of eight parallel arms, arranged at 7.6 cm from the wall and at 15.2 cm intervals. Each arm has ten evenly spaced injecting holes covered with sintered brass disks. Four downcomers are used to supply the mixture of air and tracer gas to the injector.

The gas within the bed was sampled by seven 1.27 cm O.D. probes arranged at 15.2 cm distance on a common manifold. The probe tips are covered with porous glass disks to prevent the entry of fine sand particles. The sample probe system may be clamped rigidly in any desired position. The bed pressure was adequate to force a continuous flow of gas through a suitable metering jet into seven sampling balloons.

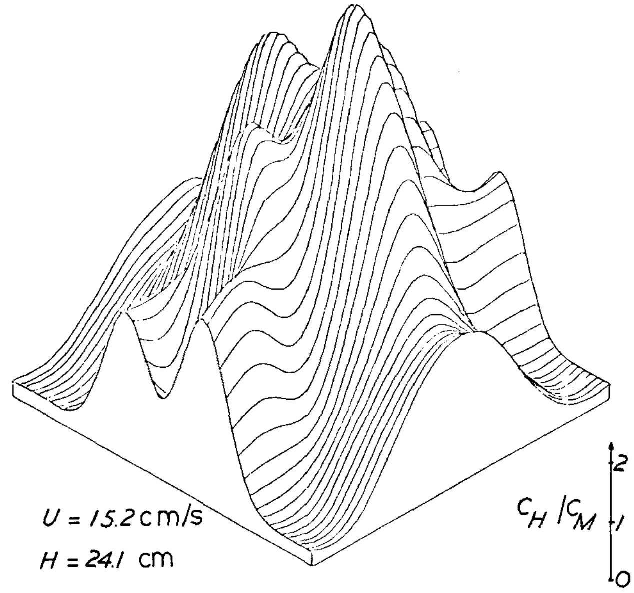
Analytical equipment. Carbon dioxide was selected as a tracer gas because of low cost, safety of operation, and ease of analysis. Gas samples were analyzed by four sets of thermoconductivity cells operated at constant pressure and cell block temperature. In a small scale experiment, backmixing of helium and of carbon dioxide were compared in order to ensure that adsorption effects were negligible. Adsorption of carbon dioxide appears to be negligible for the sands employed; hence, carbon dioxide may be used as a suitable nonadsorbing tracer

Experimental Procedure

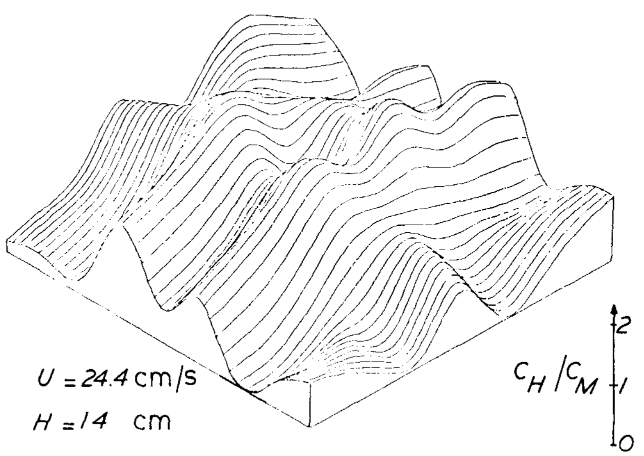
Preliminary work (same bed, same sand) indicated that backmixing would start at a critical fluidizing velocity U_{CRIT} of



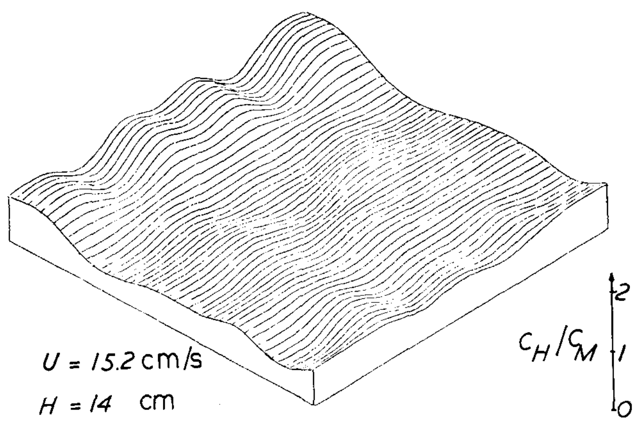
(e) U = 15.2 cm/s H = 101.6 cm.



(f) U = 15.2 cm/s H = 61 cm.



g) U = 15.2 cm/s H = 24.1 cm.



(h) U = 15.2 cm/s H = 14 cm.

bed at two different velocities, 24.4 cm s^{-1} and 15.2 cm s^{-1}.

6.4 cm/s; that is U_{CRIT}/U_{mf} = 2.6. In the preliminary investigation, no backmixing was observed until U_{CRIT} was reached. Two fluidizing velocities, 15.2 and 24.4 cm/s, were selected since bubble distribution and solids movement data for these velocities were available and known to differ considerably.

The tracer injector was positioned at a distance of 132 cm from the distributor. Time averaged samples were taken at forty-nine points forming a 15.2 cm of pitch at each bed level, starting at a distance of 122 cm and continuing to 3.8 cm from the distributor plate. Sampling rate was adjusted to be in the range of 300 to 1 000 ml/min. Since the probe must sample all three phases in the bed, namely, the bubble, cloud wake, and dense phases, in proportion to their volumetric fractions, it was necessary to sample at each position for a duration of 15 to 20 min for averaging purposes.

Carbon dioxide was used at the rate of 37 l/min. It was diluted with air to give a total injected volume of 280 l/min. This resulted in a mean mixed composition C_M, of 0.78 and 0.51% for U = 15.2 and 24.4 cm/s, respectively. It is noted that this mean mixed composition was above the normal carbon dioxide content of air. Since the entering air was also used as the reference gas for the analytical instrument, concentration readings from samples were directly related to the injected carbon dioxide.

RESULTS

Figure 4 presents three-dimensional tracer concentration profiles at selected levels for both fluidizing velocities. Concentration terms are expressed at C_H/C_M to allow direct comparison between the two runs.

In Figure 5, the average concentrations below the plane of injection are presented and compared with the values predicted by the countercurrent backmixing model (Nguyen and Potter 1974, 1975), for the two velocities employed. The model predicts strong backmixing, and this is confirmed by the experimental data. In a general sense, the agreement is surprisingly good.

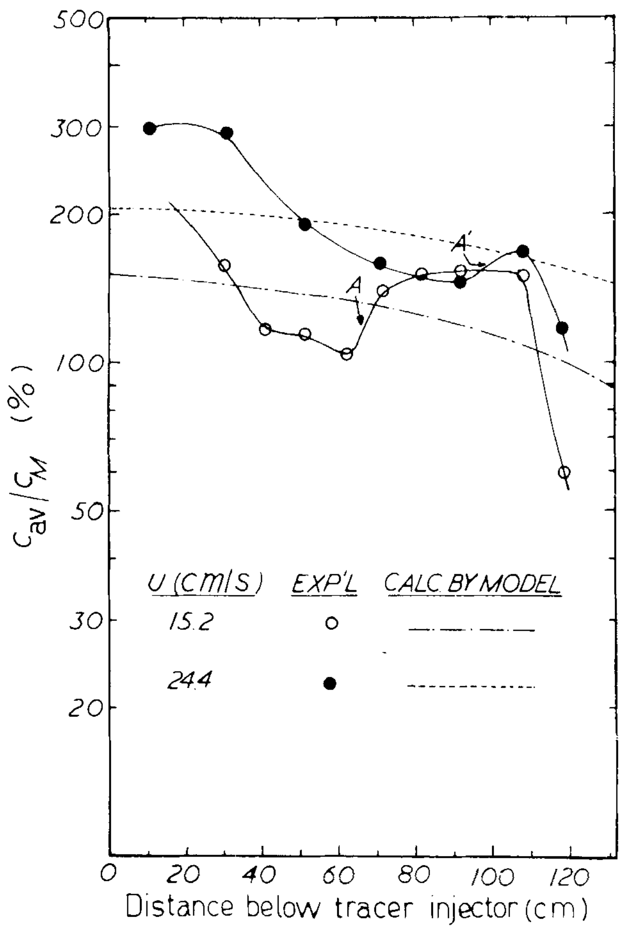


Fig. 5. Average vertical concentration profiles measured experimentally, compared with theoretical calculation of backmixing model, Nguyen and Potter (1975).

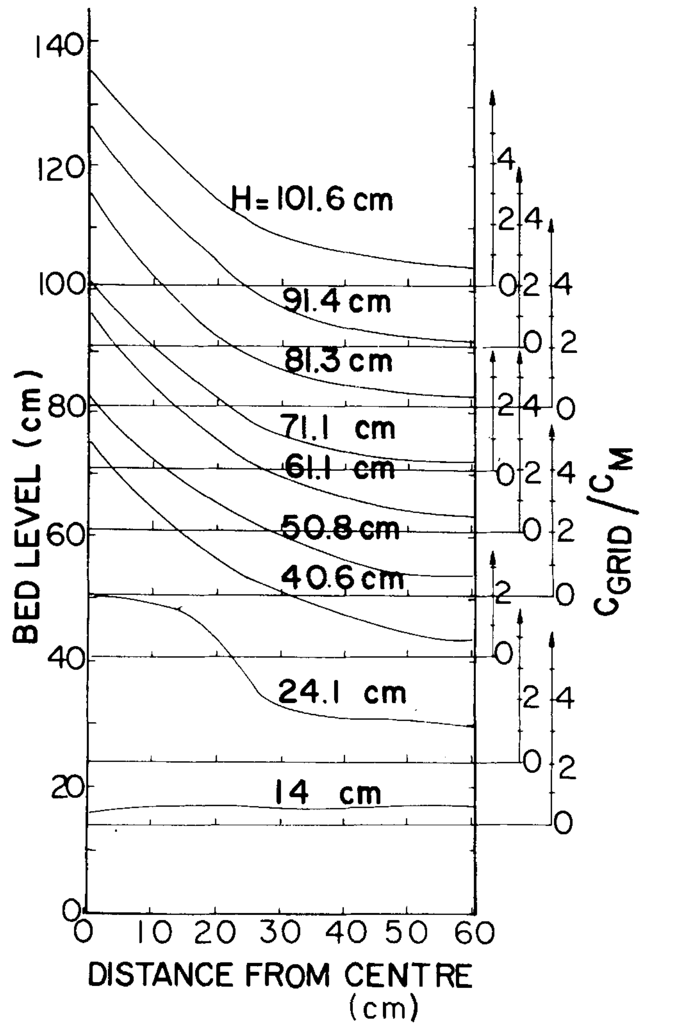


Fig. 7. Smoothed radial tracer concentration profiles at different heights in bed, $U = 15.2$ cm/s.

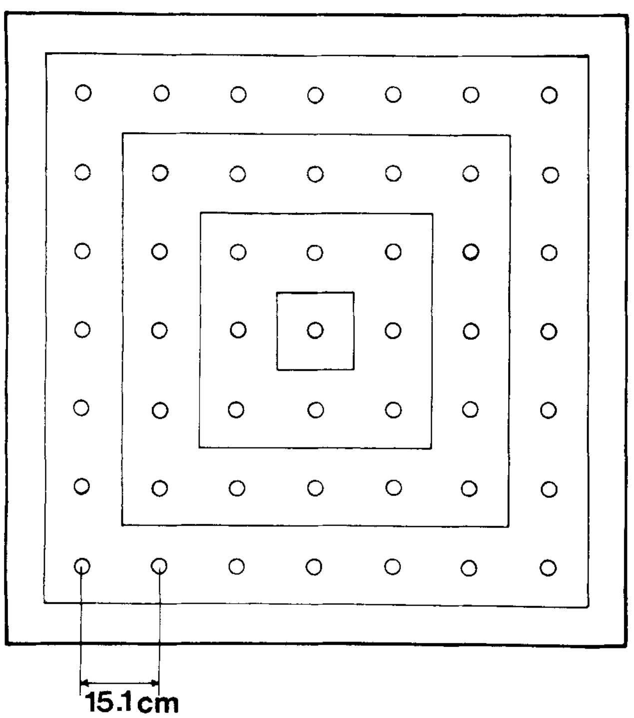


Fig. 6. Sampling positions and individual grid areas over the bed cross section.

However, there are indications that nonuniformity is greater in the larger beds than those of Potter and coworkers. The data in the large bed show, points A and A′ in Figure 5, a rise in average concentration. This rise requires some close consideration. One may visualize this phenomenon as the result of the downflow of solids and associated gas, branching out into lateral flow as the distributor is approached. The backmixed gas is purged out as soon as the vertical component of the velocity becomes less than the interstitial velocity. Newly formed bubbles then take up the solids in their wakes. Below this level, little tracer gas is detected, and whatever is found is the result of random mixing due to the vigorous coalescence of bubbles near the distributor. This cutoff in backmixing was earlier reported in a 30 cm diameter bed (Nguyen, 1975), where it was shown that gas backmixing ceased at about 4.5 cm from the distributor. The bed aspect ratio (H/D) was approximate by the same as in this case.

In the 30 cm bed, the radial concentration profiles were not uniform, as shown in Figure 2. These data were believed to be due to a greater downflow at the center, but solids movement data were lacking and hence the explanation was speculative.

In the large bed, the backmixing was also found to be nonuniform. Each bed level was divided into sections as shown in Figure 6, and the average value C_{GRID} of all point concentrations within each section is plotted in Figures 7 and 8. The presentation is designed to overcome the square geometry of the bed cross section. The data at

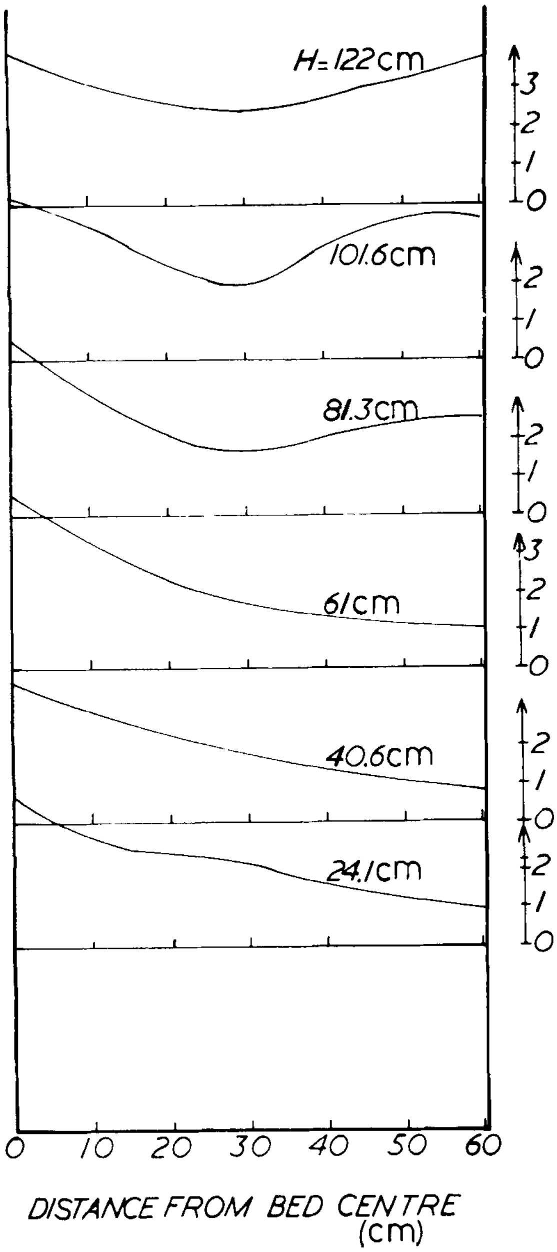


Fig. 8. Smoothed radial tracer concentration profiles at different heights in bed, $U = 24.4$ cm s^{-1}.

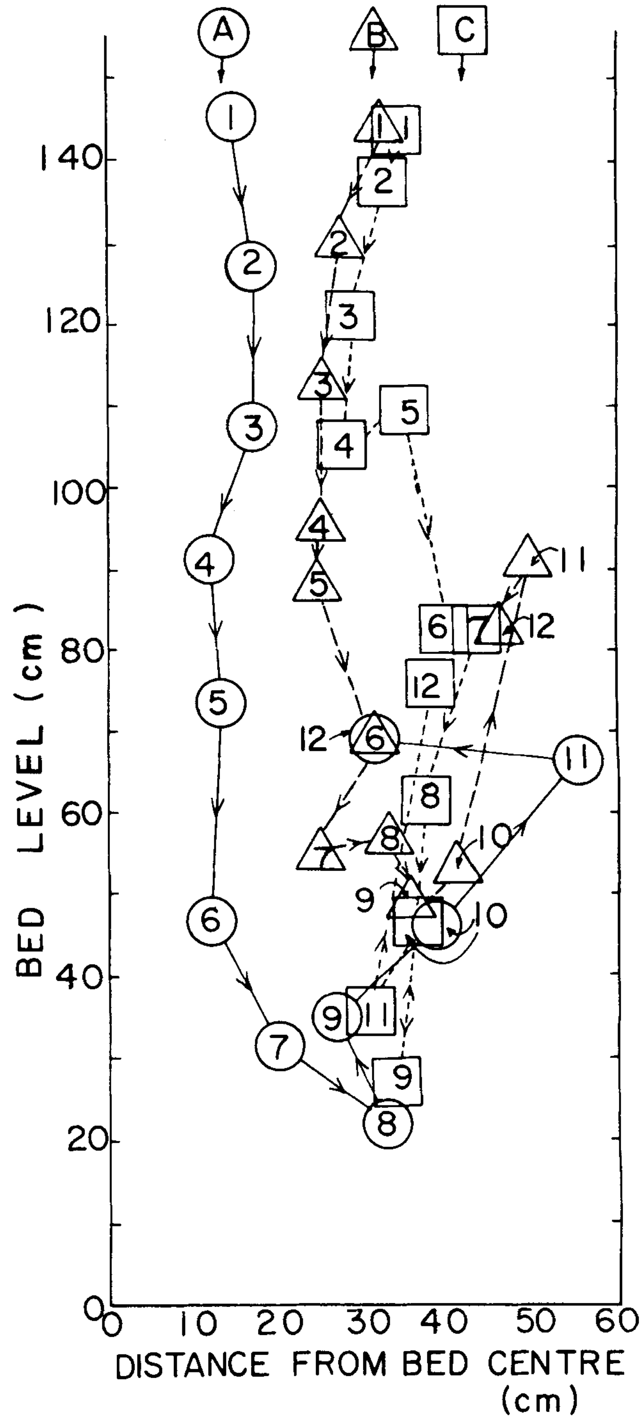


Fig. 9. Movement of tracer particles in the 1.22 m of bed. Tracer particles released from above bed positions shown in Figure 10 follow on average the paths shown. Numbers indicate time (in seconds) after the moment of release). Tracer data from Whitehead et al. (1976b). $U = 15.2$ cm/s.

15.2 cm/s show, similar to Figure 2, a central downflow of tracer gas. However, this central flow will not always exist, since at 24.4 cm/s the downflow was much more evenly distributed. The availability of the data of Whitehead and coauthors on bubble patterns and solids mixing makes a comparison possible between these and the observed backmixing. For this purpose, the data collected in the determination of solids movement in the central area of the bed operated at 15.2 cm/s, using sixteen labeled plastic spheres as tracers, are presented in Figure 9. Each path A, B, C, represents the average position of spheres labeled by the same letter as shown assembled in Figure 10 just before their release.

An alternative method of representing the data is shown in Figures 11 and 12, whereby the average concentration C_{GRID} is plotted against the incremental area of the bed, together with data on bubble distribution over the entire bed.

Whitehead et al. (1976 *b*) have pointed out that the mean pressure differential at the base of the bed is closely related to bubble distribution and solids movement. A pressure profile with a maximum at the center of the bed is associated with a strong and stable central downflow of solids, while a more uniform pressure profile results in the dispersal of this stream. Figure 13 shows the mean differential pressure between a group of four central tuyeres and four at the four corners for the sands employed. Values in Figure 13 are the average of 20 000 individual readings, each being the pressure differential between one central tuyere and the corresponding tuyere at the corner.

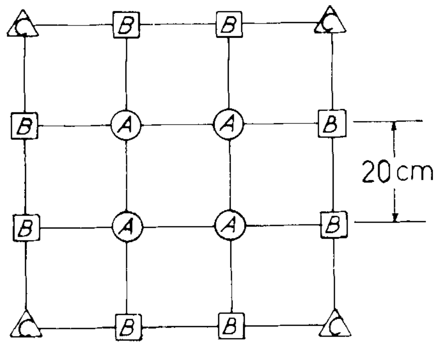


Fig. 10. Positions of labeled spheres prior to their release into the bed.

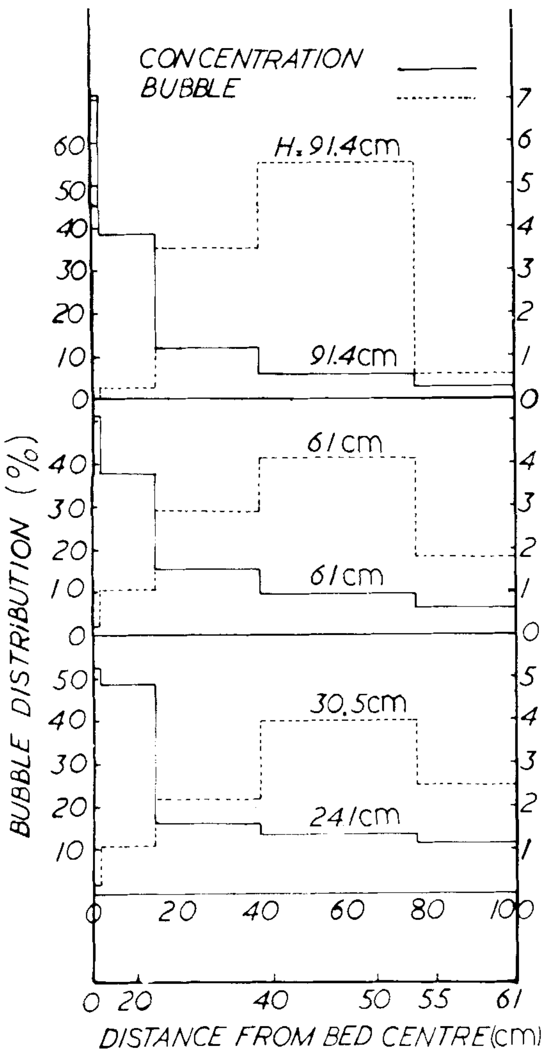


Fig. 11. Comparison of gas backmixing and bubble distribution at different bed heights. A high concentration results from strong local backmixing. The bubble distribution (%) is the fraction percent of bubble volume rising in a particular region. $U = 15.2$ cm s^{-1}.

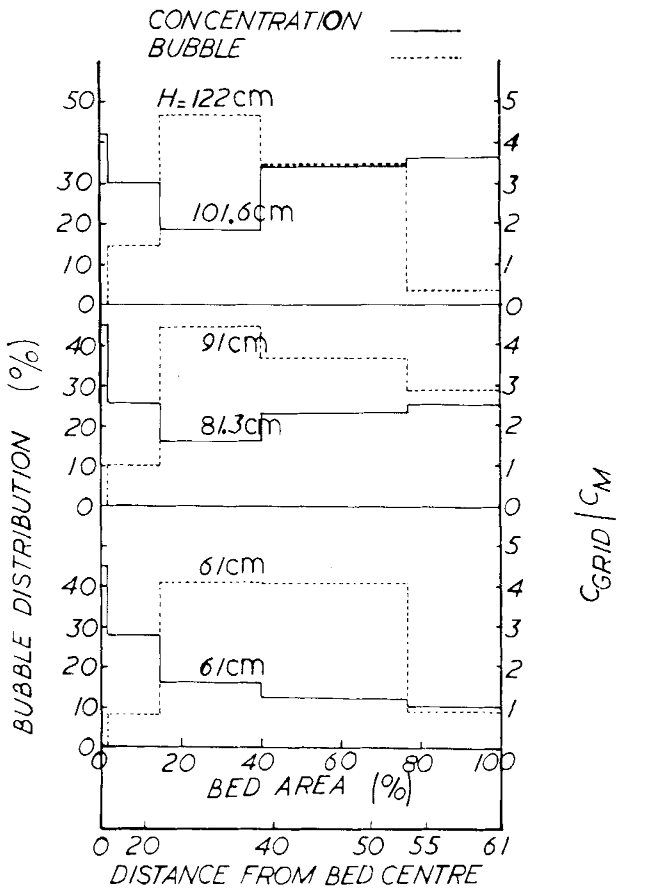


Fig. 12. Comparison of gas backmixing and bubble distribution at different bed heights. $U = 24.4$ cm s^{-1}.

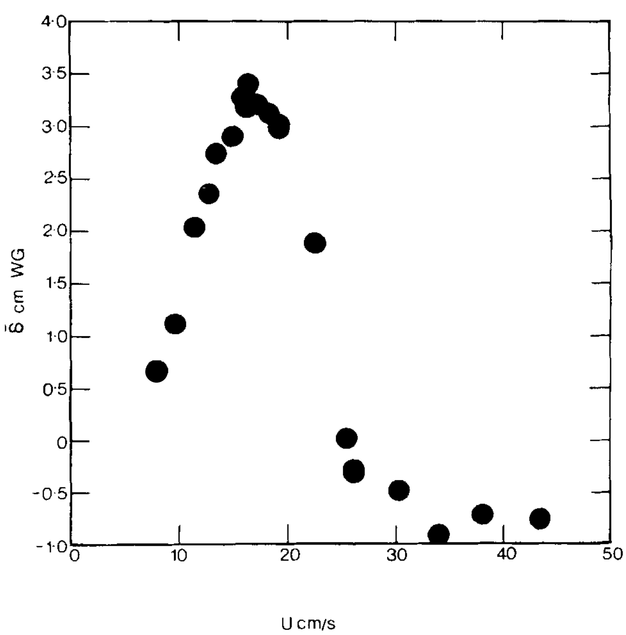


Fig. 13. Mean pressure differential at the base of the bed.

DISCUSSION

The average tracer concentration C_{av}, as shown in Figure 5 exceeded the mean mixed composition C_M at all bed levels except in the vicinity of the distributor. Behavior of this nature was predicted to occur in an idealized backmixing system, where uniformity in gas and solids mixing is assumed by van Deemter (1961). However, in the present work, gas backmixing was further enhanced by very strong localized solids downflows. The injected tracer gas was transported downwards very quickly with little interchange between dense and bubble phases. The bed was operated at ratios of U/U_{mf} of 6.08 and 9.76, compared with a critical ratio for backmixing of U_{CRIT}/U_{mf} of 2.6. At these high values of U/U_{mf}, the countercurrent backmixing model (Nguyen and Potter, 1974) predicts a significant degree of backmixing.

At 15.2 cm/s, gas backmixing was largely confined to a central stream with one maximum point concentration which reduced only slightly towards the distributor. Little tracer gas was detected near the walls of the vessel, and the ratio between the maximum and minimum point concentrations of any level was often higher than 20 (Figure 4). This behavior suggests that there was a strong central downflow of solids and backmixed gas, and mass transfer from this stream into the rest of the bed was limited. Whitehead et al. (1976 *a*), fluidizing the same sand at the same velocity, reported a downward velocity of about 17 cm/s in the stable coherent central stream, and this stream showed minimal disruption by mixing with the rest of the bed. At the wall, however, solids moved downward in slow streams, averaging only 4 cm/s, resulting in little or no gas backmixing. Figure 9 demonstrates the existence of this central stream; all spheres marked A released at 147 cm, reached a mean position at about 20 cm above the distributor after 8 s before beginning their return journey up the bed in the bubble wakes. Away from the bed center, the solids downflow velocity decreased rapidly; the mean positions of the B spheres was about 50 cm after 8 s, and C's had an oscillating position around 60 to 30 cm from the bed base. Data on bubble tracks (Figure 11), obtained with light sensors, indicated that there were four bubble tracks in the area of the bed, slightly favoring the corners, and the central region was practically devoid of bubbles. It was in this area that the solids were forced into a strong downflow.

At 24.4 cm/s, the gas backmixing profiles were more uniform with three maxima at bed levels higher than 60 cm from the distributor. This behavior is consistent with the absence of the persistent coherent central stream, as reported by Whitehead et al. (1976 *a*, and *b*). Some materials descended centrally at about 8 cm/s, but the overall flow pattern differed considerably from the 15.2 cm/s situation. As shown in Figure 12, the bubble tracks were still present, but there were some bubble activities in the center, and the positions of tracks were no longer well defined.

Figure 13 provides a useful resumé of bed behavior via the measurement of the pressure profile at the distributor level. It was shown that at 15.2 cm/s, a maximum pressure was present at the center of the bed and a minimum at the corners, but at 24.4 cm/s, the pressure at the base was almost uniform on a time averaged basis. The former pattern was found to be associated with a strong, stable central downflow, while the latter with the absence of this stream. The flow state at 24.4 cm/s was unstable with short-term variation in pressure profile, indicating transient operation with a central downstream for a short period, and perhaps flow down near the walls at other periods.

Examining available backmixing data on large scale equipment (Potter, 1971) and comparing these with the results presented in this paper, it is obvious that sampling in large beds will not be reliable if it is not done systematically and extensively throughout the bed. A few sampling points in a large bed do not reflect the overall gas and solids mixing, and conclusions derived from them may well be misleading. On the other hand, the interpretation of data on gas backmixing requires other information, such as bubble distribution and coalescence. Bubble therefore becomes the single most important parameter in the fluidized bed. This has been recognized by the incorporation of bubble size variation into the mathematical modeling of fluidization (for example, Toor and Calderbank, 1967; Kato and Wen, 1969; Nguyen, 1975; Fryer and Potter, 1976).

ACKNOWLEDGMENT

Figures 4 were plotted by Mr. J. C. H. McAdam, and Mr. D. C. Dent assisted in the design and operation of the equipment. This work has been supported, on the University side, by grants from the Australian Research Grants Committee to O. E. Potter.

NOTATION

C_{av} = average tracer concentration for each bed level
C_{GRID} = mean tracer concentration in the grids as defined by Figure 6
C_H = tracer concentration at each sampling point
C_M = mean mixed tracer concentration
D = bed diameter
H = bed height
U = fluidizing velocity
U_{CRIT} = critical fluidizing velocity, that is, velocity at which gas backmixing starts
U_{mf} = incipient fluidizing velocity
$\bar{\delta}$ = mean pressure differential at the bed base, $P_{\text{central region}} - P_{\text{wall region}}$

LITERATURE CITED

Fryer, C., and O. E. Potter, "Experimental Investigation of Models for Fluidized Bed Catalytic Reactors., "*AIChE J.*, **22**, 38 (1976).

Geldart, D., "The Size and Frequency of Bubbles in Two- and Three-Dimensional Gas Fluidized Beds," *Powder Technol.*, **4**, 41 (1970).

Hamilton, C. J., C. Fryer, and O. E. Potter, "Backmixing of Gas in Fluidised Beds," *Proc. of Chemeca '70*, Session 1, Australia, Melbourne and Sydney, Acad. Sci. and Inst. Chem. Engrs., 61 (1970).

Kato, K., and C. Y. Wen, "Bubble Assemblage Model for Fluidized Bed Catalytic Reactors," *Chem. Eng. Sci.*, **24**, 1351 (1969).

Latham, R. L., C. J. Hamilton, and O. E. Potter, "Back-mixing and Chemical Reaction in Fluidised Bed," *Brit. Chem. Eng.*, **13**, 666 (1968).

Latham, R. J., and O. E. Potter, "Backmixing of Gas in a 6-in Diameter Fluidised Bed," *Chem. Eng. J.*, **1**, 152 (1970).

Merry, J. M. C., and J. F. Davidson, "Gulf Stream Circulation in Shallow Fluidised Beds," *Trans. Inst. Chem. Engrs.*, **51**, 361 (1973).

Nguyen, H. V., "Effects of Adsorption on Backmixing in Fluidized Beds," Ph.D. thesis, Monash Univ., Australia (1975).

———, and O. E. Potter, "Gas Mixing in Fluidized Beds as a Function of the Adsorbency of the Solids for the Gas," *Chem. React. Eng.*, **11**, H. M. Hulburt ed., *Advance in Chemistry Ser. 133*, American Chemical Society, 290 (1974).

———, "Adsorption Effects in Fluidized Beds," *Fluidization Technology II*, p. 193, D. L. Keairns, ed., Hemisphere Publishing Corp., Washington, D.C. (1976).

Noble, P. J., "The Mechanism of Solid Mixing in Gas Fluidized Beds," *Mineral Dressing Res. Symp.*, Ann. Conf., Australasian Instn. Min. Met. (1962).

Potter, O. E., *Fluidization*, Chapt. 7, J. F. Davidson, and D. Harrison, ed., Academic Press, London, England (1971).

Rowe, P. N., "Drag Forces on a Hydraulic Model of a Fluidised Bed," *Trans. Inst. Chem. Engrs.*, **39**, 175 (1961).

Schmalfeld, V. J., "Feststoff-Bewegung in einem Gas-Feststoff-Fließbett," *V.D.I-Z*, **118**, 65 (1976).

Stephens, G. K., R. J. Sinclair, and O. E. Potter, "Gas Exchange Between Bubbles and Dense Phase in a Fluidised Bed," *Powder Tech. No. 1.*, **1**, 157 (1967).

Thiel, W. J., and O. E. Potter, "Slugging in Fluidized Beds," *Ind. Eng. Chem. Fundamentals*, **16**, No. 2, 242-247 (1977).

Toor, F. L., and P. H. Calderbank, "Reaction Kinetics in Gas-Fluidized Catalyst Bed: Mathematical Models," *Proceedings International Symposium on Fludization*, p. 373, A. A. L. Drinkenburg, ed., Netherland University Press, Amsterdam (1967).

Van Deemter, J. J., "Mixing and Contacting in Gas-Solid Fluidised Beds," *Chem. Eng. Sci.*, 13, 143 (1961).
Werther, J., "The Hydrodynamics of Fluidization in a Large Diameter Fluidized Bed," Preprint, paper presented at GVC/ A.I.Ch.E. Meeting, München (1974).
Whitehead, A. B., and D. C. Dent, "Behaviour of Multiple Tuyere Assemblies in Large Fluidised Beds," *Proceedings International Symposium on Fluidization*, p. 803, A. A. L. Drinkenburg, ed., Netherland University Press, Amsterdam, 803 (1967 *a*).
———, and A. D. Young, "Fluidization Studies in Large Gas-Solid Systems Part II: The Effect of Distributor Design and Solid Properties on Fluidisation Quality," *Powder Technol.*, 1, 149 (1967 *b*).
Whitehead, A. B., and A. Auff, "Fluidization Studies in Large Gas-Solid Systems Part IV: Similarity Criteria for Different Size Systems," *ibid.*, 15, 77 (1976 *a*).
Whitehead, A. B., D. C. Dent, and J. C. H. McAdam, "Fluidization Studies in Large Gas-Solid Systems, Part V: Pressure Instabilities and Some Effects of Immersed Vertical Surfaces," *ibid.* (to be published).
Whitehead, A. B., G. Gartside, and D. C. Dent, "Flow and Pressure Maldistribution at the Distributor Level of a Gas-Solid Fluidised Bed," *Chem. Eng. J.*, 1, 175 (1970).
———, "Fluidization Studies in Large Gas-Solid Systems, Part III: The Effect of Bed Depth and Fluidizing Velocity on Solids Circulation Patterns," *Powder Technol.*, 14, 61 (1976 *b*).
Woollard, I. N. M., and O. E. Potter, "Solids Mixing in Fluidised Beds," *AIChE J.*, 14, 388 (1968).

Manuscript received March 17, 1977; revision received August 1, and accepted August 8, 1977.

Model of the Constricted Unit Cell Type for Isotropic Granular Porous Media

The constricted unit cell model for isotropic granular porous media developed by Payatakes, Tien, and Turian (1973) is extended here to take in account the random orientation of the flow channels. In the proposed model, each unit cell corresponds to a pore (cavern) and has two coaxial constricted inlet and outlet ports (throats). The unit cells have random dimensions and orientations, the distributions of which can be determined from simple experimental measurements. The flow through a unit cell is assumed to be identical to that through a segment of the corresponding periodically constricted tube. The model is applied to the case of creeping Newtonian flow. The solution to the flow problem is obtained by a collocation method. Permeabilities of typical packings are calculated without use of any adjustable parameters (such as tortuosities, etc.) and are found in excellent agreement with experimental values.

A. C. PAYATAKES
and
MARIANO A. NEIRA
Chemical Engineering Department
University of Houston
Houston, Texas 77004

SCOPE

There are many important processes taking place in porous media for which the converging-diverging character of the porous space is of primary importance. This has been recognized for some time (Petersen, 1958); Houpeurt, 1959), but only recently have major advances been made in this direction. Payatakes, Tien, and Turian (1973*a*, *b*) proposed a model for granular porous media involving unit cells of the constricted tube type and solved the associated problem of Newtonian flow through periodically constricted tubes with a novel finite-difference method of the stream function-vorticity type, retaining the inertial terms of the equation of motion. Payatakes, Tien, and Turian (1974*a*, *b*) and Payatakes, Brown, and Tien (1977) used the above porous media model to model successfully deep-bed filtration of liquid-solid suspensions. It was shown that the converging-diverging character of the porous space is essential in explaining and analyzing the transient behavior of depth filtration. Sheffield and Metzner (1976) published an incisive analysis of flow of nonlinear fluids through packed beds and concluded that the experimentally observed behavior can be explained by taking into account the converging-diverging character of the flow. Recent fundamental studies in the mobilization,