THE HEIGHT OF EXPANSION IN BUBBLING FLUIDIZED BEDS

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A theory of bubble growth due to coalescence in gas fluidized beds was employed to predict the bed expansion. Using the semi-empirical correlation of bubble size by Mori and Wen, an equation for the height of expanded beds has been developed and verified by experiment. Predictions of the proposed formula were compared with the predictions of recent bed expansion correlations.

The performance of a fluidized bed reactor is, in general, strongly influenced by the hydrodynamic behaviour of the bed. A bubbling gas fluidized bed is usually considered as a two-phase system: an emulsion (or continuous or dense or particulate) phase which contains virtually all the particles plus interstitial gas and a bubble (or lean or discontinuous) phase made up of more or less discrete rising bubbles containing few particles. The formation and behaviour of the bubbles are essential factors in interpreting many aspects of the fluidized bed such as mass transfer, heat transfer and mixing.

Most reactor models for fluidized beds are based on the presence of bubbles in the bed[1-4]. The application of these models requires knowledge of the bubble size in reactors or, furthermore, information on the variation of bubble size along the height of bed.

In previous work[5-10], we have investigated the fundamental properties of the fluidized bed of lime and inert particles for removal of sulphur dioxide from flue gas. Fluidization phenomena, such as the minimum fluidization velocity[5-7], the terminal velocity[8] and pressure fluctuations within the bed[9-10] have been measured both at ambient and high temperatures.

The objective of the present work was to develop and verify by experiment a correlation of expanded bed height for the ceramsite particles. This inert material is being employed in our measurements on the performance of the diluted bed of lime particles in removal of sulphur dioxide from a gas stream[11].

THEORETICAL

In the beds of particles belonging to group B of Geldart's classification[12], bubbles are formed immediately beyond the point of minimum fluidization and the porosity of dense phase does not change with gas velocity. For such a bed we can express the rising velocity of bubbles at height h as

$$u_b = \frac{Y(U - U_{mf})}{\varepsilon_b}, \tag{1}$$

where Y denotes the ratio of the actual visible bubble flow rate to the gas flow rate given by $(U - U_{mf})\, A_t$ and ε_b is the local bubble fraction within the bed.

The simple two-phase models assume that all gas in excess of that required for incipient fluidization passes through the bed as a visible bubble flow, *i.e.*, $Y = 1{\cdot}0$. Geldart[13] and Werther[14] recommend for Y values smaller than unity. The measurements of Chavarie and Grace[15] suggest that the visible bubble flow rate increases with distance above the distributor. Rowe[16] and Yacono *et al.*[17] examined the two-phase model and developed a comprehensive theory to describe the division of gas between the bubble and interstitial phases.

The overall bubble volume fraction within the bed is obtained by integrating Eq. (*1*) from the bottom ($h = 0$) to the top of bed ($h = H$):

$$\bar{\varepsilon}_b = \frac{Y(U - U_{mf})}{H} \int_0^H \frac{dh}{u_b}. \tag{2}$$

The overall fraction of the bed occupied by bubbles can be related to the bed expansion if the dense void fraction is assumed to remain constant at ε_{mf}. This common assumption is supported by measurements of Lockett and Harrison[18].

A schematic drawing in Fig. 1 shows the bubbling fluidized bed with multi-orifice gas distributor.

On substituting

$$\bar{\varepsilon}_b = \frac{H - H_{mf}}{H} \tag{3}$$

we can write for the height of expanded bed

$$H = H_{mf} + Y(U - U_{mf}) \int_0^H \frac{dh}{u_b}. \tag{4}$$

The velocity of the rising bubbles, u_b, in Eq. (*4*) can be approximated using the rising velocity of a single bubble with the aid of a commonly accepted relationship proposed by Davidson and Harrison[19]

$$u_b = U - U_{mf} + 0{\cdot}711(gD_b)^{1/2}\,. \tag{5}$$

Inserting this relationship in Eq. (*4*) we obtain

$$H = H_{mf} + Y \int_0^H \frac{dh}{1 + \dfrac{22{\cdot}27}{U - U_{mf}}(D_b)^{1/2}}\,. \tag{6}$$

The experience shows that the bubble size can be influenced by the vessel dimension. The bubble growth restrictions are usually assumed when bubble size is close to a half of the bed diameter. When the ratio of bubble size to bed diameter is about 2/3, the bubble regime approaches the onset of slugging.

Mori and Wen[20] proposed a semi-empirical correlation of the bubble size (See Eq. (7)) which, aside from the effect of distributor design, also incorporates the influence of bed diameter.

$$D_b = D_{bm} - (D_{bm} - D_{bo}) \exp\left(-\frac{0{\cdot}3}{D_t}h\right). \tag{7}$$

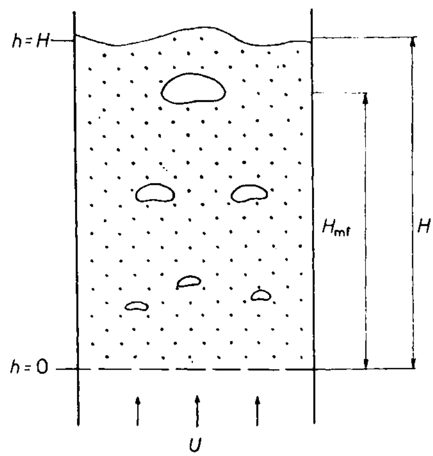

FIG. 1

Schematic drawing of an expanded bubbling bed

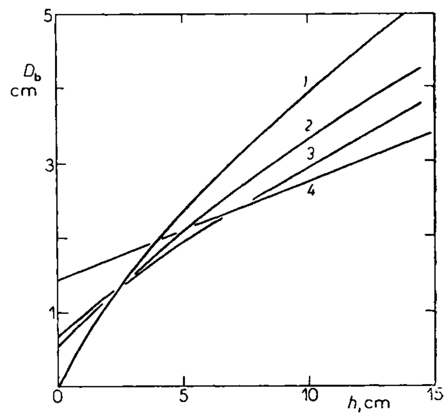

FIG. 2

Bubble size predicted as a function of distance above the gas distributor. $U - U_{mf} = 15\ \text{cm s}^{-1}$; *1* Rowe[21], $h_0 = 0$; *2* Mori and Wen[20], $D_t = 8{\cdot}5$ cm, $n = 226$; *3* Darton *et al.*[22], $A_0 = 0{\cdot}2511\ \text{cm}^2$; *4* Werther[14]

D_{bm} denotes the maximum attainable bubble diameter due to the total coalescence of bubbles which can be viewed as a fictitious bubble diameter for a bed of large dimensions.

$$D_{bm} = 0{\cdot}652[A_t(U - U_{mf})]^{0.4} \tag{8}$$

D_{bo} is the initial bubble diameter formed at the surface of the gas distributor. The value of D_{bo} for perforated plate distributors can be evaluated from

$$D_{bo} = 0{\cdot}347\left[\frac{A_t}{n}(U - U_{mf})\right]^{0.4}. \tag{9}$$

The set of the semi-empirical Eqs (7)–(9) was developed by statistical treatment of about 400 data points from different investigators which covered a wide range of operating conditions. Eqs (7)–(9) fit the observed bubble diameters with the mean deviation 54%. As shown in Fig. 2 the predictions of Eqs (7)–(9) lie in the midst of the region bounded by predictions provided by the recent correlations of Rowe[21] and Werther[14].

Using Eq. (7) for the equivalent bubble diameter, then Eq. (6) can be arranged to give

$$H = H_{mf} + Y\int_0^H \frac{dh}{1 + (A - B\,.\,e^{-mh})^{0.5}}, \tag{10}$$

where

$$A = \left(\frac{22{\cdot}27}{U - U_{mf}}\right)^2 D_{bm} \tag{11}$$

$$B = \left(\frac{22{\cdot}27}{U - U_{mf}}\right)^2 (D_{bm} - D_{bo}) \tag{12}$$

$$m = \frac{0{\cdot}3}{D_t}. \tag{13}$$

The integral in Eq. (10) can be transformed by introducing a new variable

$$(A - B\,e^{-mh})^{0.5} = y \tag{14}$$

into the form

$$\int_0^H \frac{dh}{1 + (A - B\,e^{-mh})^{0.5}} = \frac{2}{m}\int_{(A-B)^{0.5}}^{(A-B\,e^{-mH})^{0.5}} \frac{1}{1 + y}\,.\,\frac{y}{A - y^2}\,dy \tag{15}$$

for which

$$\frac{dh}{dy} = \frac{1}{m} \frac{2y}{A - y^2}. \tag{16}$$

An elementary function of the integrand on the right hand side of Eq. (*15*) was found by the decomposition to partial fractions and has the form

$$\int \frac{y}{(1 + y)(A - y^2)} dy = \frac{1}{2(A^{0.5} - 1)} \ln (A^{0.5} + y) -$$

$$- \frac{1}{2(A^{0.5} + 1)} \ln (A^{0.5} - y) + \frac{1}{1 - A} \ln (1 + y) + \text{const.}. \tag{17}$$

Having introduced the new limits of integration occurring in Eq. (*15*), we obtain after some arrangements the final equation for the average height of an expanded bed:

$$H = H_{mf} + \frac{Y}{m} \left[\frac{1}{A^{0.5} - 1} \ln \frac{A^{0.5} + (A - B\,e^{-mH})^{0.5}}{A^{0.5} + (A - B)^{0.5}} - \right.$$

$$\left. - \frac{1}{A^{0.5} + 1} \ln \frac{A^{0.5} - (A - B\,e^{-mH})^{0.5}}{A^{0.5} - (A - B)^{0.5}} + \frac{2}{1 - A} \ln \frac{1 + (A - B\,e^{-mH})^{0.5}}{1 + (A - B)^{0.5}} \right]. \tag{18}$$

The above expression describes the expansion of bubbling fluidized beds as a function of the excess gas velocity, $U - U_{mf}$, characteristics of the distributor, A_t/n, and diameter of the bed, D_t. Eq. (*18*) can easily be solved for H by iteration using H_{mf} as an initial guess. The experience shows that convergence is rapid. The difference between two consecutive values of H was less than 1% after 3–5 iterations.

RESULTS AND DISCUSSION

Measurements of bed expansion have been conducted in a 14 cm i.d. glass column fitted with a perforated plate distributor of gas. The distributor of free area $\varphi = 6.93\%$ contained 151 orifices with the diameter $d_o = 0.2$ cm. The flow rate of dried air was measured by a calibrated rotameter.

Ceramsite particles of mean diameter 1·0 mm (0·9–1·1 mm) were used in the experiments. This material is made by mechanical treatment and subsequent calcination of a claystone cover from lignite mines. The bulk density of ceramsite amounted to 1·6 g cm^{-3}. The minimum fluidization velocity measured by the usual method was 25·8 cm s^{-1} at ambient temperature and pressure.

The bed height was determined by direct visual observations as a mean level of the fluctuating surface of the bed. For a given gas flow rate, the observations were generally repetead several times by two observers.

The average values of measured bed heights are plotted in Fig. 3 against the superficial excess velocity. As can be seen curve *1* computed from Eq. (*18*) is quite close to the experimental data points and follows very well their trend.

Curve *1* also fits very well experimental results of Fryer and Potter[22]. These workers investigated the expansion of a bed of fine sand particles (H_{mf} = 11 cm, d_p = 0·117 mm, ϱ_p = 2·65 g cm^{-3}, U_{mf} = 1·7 cm s^{-1}). Their fluidized bed reactor was 22·9 cm in diameter and was equipped with a bubble cap distributor.

Experimental data of Best and Yates[23] are also plotted in Fig. 3. These authors measured the bed height in a 10 cm i.d. reactor fitted with a porous plate distributor. In contrast to the majority of fluidization studies, Best and Yates[23] worked at an elevated temperature of 410°C. They used larger, alumina spheres (H_{mf} = 12 cm, d_p = 0·81 mm, ϱ_p = 1·27 g cm^{-3}) the minimum fluidization velocity of which was 15·7 cm s^{-1} at 410°C. As can be seen in the figure, these data fall about 10% below the predicted line *1*. In view of the difficulties inherently involved in the determination of expanded bed heights, the agreement between computed and experimental data in Fig. 3 seems to be satisfactory.

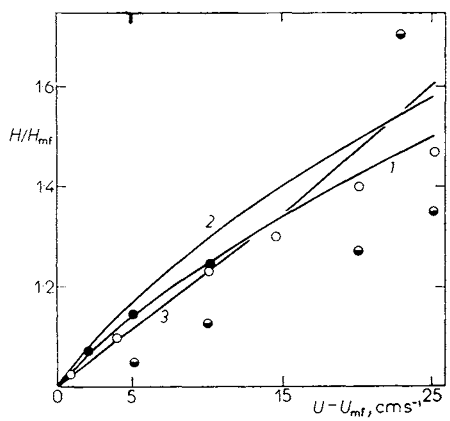

FIG. 3

Comparison of the experimental and computed values of bed expansion ratio. H_{mf} = = 13 cm, $\bar{d}_p$ = 0·1 cm; ○ this work; ◒ data of Best and Yates[23]; ● data of Fryer and Potter[24]. The solid lines show the bed expansion ratio predicted by different equations for Y = 1. *1* Eq. (*18*) in this work; *2* Xavier *et al.*[25]; *3* Darton[26]

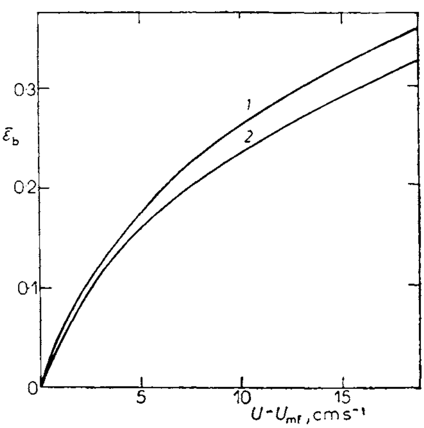

FIG. 4

Dependence of the overall fraction of bed occupied by bubbles on the excess gas flow velocity for different heights of bed. Computations carried out for a 8·5 cm i.d. reactor fitted with a perforated plate distributor with 226 orifices, Y = 1. *1* H_{mf} = 8 cm; *2* H_{mf} = = 14 cm

The predictions provided by the equation of Xavier *et al.*[25] are slightly higher than those from Eq. (*18*). As also shown in Fig. 3, curve *3* representing the predictions of the Darton correlation[26], changes somewhat its trend at highes gas velocities and intersects both curves *1* and *2*.

In order to explore the influence of major operating variables such as the excess gas flow velocity and height of bed at incipient fluidization, we made systematic computations of the overall fraction of bed occupied by bubbles. The results plotted in Fig. 4 indicate a minor effect of H_{mf} on the expansion of bed.

The proposed equation predicts the height of an expanded fluidized bed with reasonable accuracy. Unlike previous correlations, this formula also incorporates the influence of bed diameter on its expansion. Bed heights predicted by the developed equation are slightly lower than those determined using the correlation of Xavier.

LIST OF SYMBOLS

A	parameter defined by Eq. (*11*)
A_t	cross-sectional area of the fluidized bed, cm^2
A_0	catchment area for a bubble stream at the distributor plate, usually the area of plate *per* orifice, cm^2
B	parameter defined by Eq. (*12*)
d_o	diameter of plate orifices, cm
d_p	mean particle size, cm
D_b	equivalent spherical bubble diameter having the same volume as that of a bubble, cm
D_{bm}	maximum bubble diameter due to total coalescence of bubbles, cm
D_{bo}	initial bubble diameter at the distributor plate, cm
D_t	diameter of the fluidized bed, cm
g	acceleration due to gravity, 980·7, cm s^{-2}
h	length measured from the distributor plate in the vertical direction, cm
h_0	constant characterizing the distributor, cm
H	average height of expanded bed, cm
H_{mf}	height of bed at the point of incipient fluidization, cm
m	parameter defined by Eq. (*13*)
n	total number of orifices on the plate
u_b	mean absolute rising velocity of bubbles, cm s^{-1}
U	superficial gas velocity, cm s^{-1}
U_{mf}	superficial gas velocity at incipient fluidization, cm s^{-1}
y	quantity defined by Eq. (*14*)
Y	ratio of the actual visible bubble flow rate to the gas flow rate given by $(U - U_{mf}) A_t$
ε_b	fraction of bed volume occupied by bubbles at the level h
ε_b	overall fraction of bed volume occupied by bubbles
ε_{mf}	voidage fraction at the point of minimum fluidization
φ	free area of distributor
ϱ_p	particle density, g cm^{-3}

REFERENCES

1. Yates J. G.: Chem. Engineer *1975*, 671.
2. Horio M., Wen C. Y.: AIChE Symp. Ser. *73*, 9 (1977).
3. van Swaaij W. P. M.: *5th Internat. Symp. Chem. React. Eng., Houston 1978.*
4. Levenspiel O., Baden N., Kulkarni B. D.: Ind. Eng. Chem., Process Des. Dev. *17*, 478 (1978).
5. Pata J., Hartman M.: Ind. Eng. Chem., Process Des. Dev. *17*, 231 (1978).
6. Svoboda K., Hartman M.: Ind. Eng. Chem., Process Des. Dev. *20*, 319 (1981).
7. Svoboda K., Hartman M.: AIChE J. *27*, 866 (1981).
8. Pata J., Hartman M.: Ind. Eng. Chem., Process Des. Dev. *19*, 98 (1980).
9. Svoboda K., Čermák J., Hartman M., Drahoš J., Selucký K.: Ind. Eng. Chem., Process Des. Dev. *22*, 514 (1983).
10. Svoboda K., Čermák J., Hartman M., Drahoš J., Selucký K.: AIChE J. *30*, 513 (1984).
11. Hartman M., Svoboda K., Trnka O.: *8th Internat. Symp. React. Eng., Edinburgh, Inst. Chem. Eng. Symp. Ser. 87, 509* (1984).
12. Geldart D.: Powder Technol. *7*, 285 (1973).
13. Geldart D. in the book: *Fluidization Technology* (D. L. Keairns, Ed.), Vol. 1, p. 237. Hemisphere Publ. Corp., Washington 1976.
14. Werther J.: Ger. Chem. Eng. *1*, 243 (1978).
15. Chavarie C., Grace J. R.: Ind. Eng. Chem., Fundam. *14*, 75 (1975).
16. Rowe P. N.: ACS Symp. Ser., Chem. React. Eng. *65*, 436 (1978).
17. Yacono C., Rowe P. N., Angelino H.: Chem. Eng. Sci. *34*, 789 (1979).
18. Lockett M. J., Harrison D. in the book: *Proc. Int. Symp. Fluidization* (A. A. H. Drinkenburg, Ed.), p. 257. Netherlands Univ. Press, Amsterdam 1967.
19. Davidson J. F., Harrison D. in the book: *Fluidization* (J. F. Davidson and D. Harrison, Eds), p. 148. Academic Press, London 1971.
20. Mori S., Wen C. Y.: AIChE J. *21*, 109 (1975).
21. Rowe P. N.: Chem. Eng. Sci. *31*, 285 (1976).
22. Darton R. C., La Nauze R. D., Davidson J. F., Harrison D.: Trans. Inst. Chem. Eng. *55*, 274 (1977).
23. Best R. J., Yates J. G.: Powder Technol. *16*, 285 (1977).
24. Fryer C., Potter O. E.: AIChE J. *22*, 38 (1976).
25. Xavier A. M., Lewis D. A., Davidson J. F.: Trans. Inst. Chem. Eng. *56*, 274 (1978).
26. Darton R. C.: Trans. Inst. Chem. Eng. *57*, 134 (1979).

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