Powder Technology, 42 (1985) 67 - 78

The Design of Distributors for Gas-Fluidized Beds

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SUMMARY

The performance of the gas distributor often determines the success or failure of a fluidized bed and although much more is known now than 20 years ago, there are still many pitfalls for the designer.

Particle and gas properties play a key role in successful design together with the critical pressure drop ratio, and hole size, geometry and spacing; these strongly influence jet penetration, dead zones, particle sifting, attrition and mixing.

The current state of the art is reviewed in the light of recent research and industrial experience.

INTRODUCTION

Inadequate design of gas distributors, or their malfunction in operation, is responsible for a substantial proportion of the difficulties encountered in fluidized bed processes.

In applications which involve the processing of solids, the major concerns are to achieve rapid dispersion of the solids feed and to prevent segregation and settling out of the larger or denser particles on the distributor. The second situation is particularly serious, as it can cause variable temperatures and rapid defluidization of the entire bed, particularly if the solids go through a sticky stage.

In applications where high conversions of the reacting gases are required, careful design of the distributor so as to give uniform gas distribution and small bubbles at the grid can improve performance.

Distributors must also have sufficient strength to resist deformation under operating conditions and to support the static bed. They must be able to withstand stresses induced by thermal expansion, operate for long periods without blocking and be easy to unblock, prevent backflow (sifting) of solids into the windbox, avoid promoting erosion of the plate and attrition of the particles and operate at as low a pressure drop as possible in order to minimize power consumption. Not all these requirements are compatible and their relative importance may change with the process requirements.

How and why bed performance is influenced by the gas distributor is now much better understood, though much work remains to be done, and in this paper we review the current state of the art and present design equations.

Some examples of distributors in common use are shown in Fig. 1.

BASIC EQUATIONS FOR DISTRIBUTOR PRESSURE DROP

It is well known that if a gas distributor gives a pressure drop which is too low, the result is poor fluidization; that is, some parts of the bed will receive much less gas than others, and may be temporarily or permanently defluidized, whilst in other parts the gas forms semi-permanent spouts or channels.

In order to understand the relative importance of the design variables, and their influence on the distributor pressure drop, let us consider a flat plate with drilled holes (often known as a multi-orifice distributor) as shown in Fig. 2.

The pressure drop across the bed is

$$\Delta p_{B}(\mathrm{N/m^2}) = \frac{M_g}{A} = \frac{\rho_p(1-\epsilon_{mf})AH_{mf}g}{A} \qquad (1)$$

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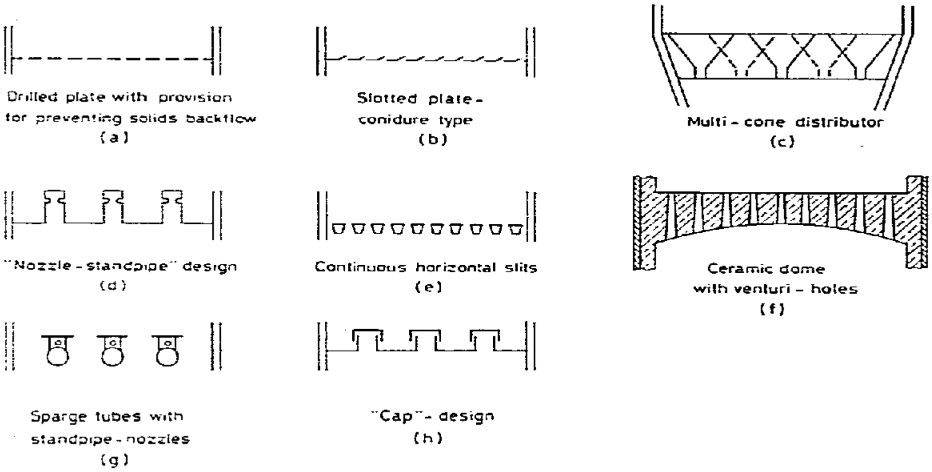

Fig. 1. Examples of distributors in common use.

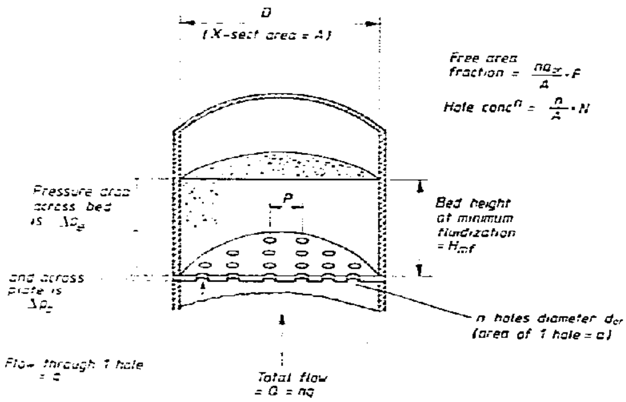

Fig. 2. Multi-orifice distributor.

The fractional free area F is

$$F = \frac{na_{or}}{A} \tag{2}$$

$$F = Na_{or} \tag{3}$$

By Bernoulli's equation, the flow rate through one hole is

$$q = C_d a_{or}\left(\frac{2\ \Delta p_D}{\rho_g}\right)^{1/2} \tag{4}$$

and across the entire plate

$$Q = C_d n a_{or}\left(\frac{2\ \Delta p_D}{\rho_g}\right)^{1/2} \tag{5}$$

Dividing by A and substituting from eqn. (2),

$$U = C_d F\left(\frac{2\ \Delta p_D}{\rho_g}\right)^{1/2} \tag{6}$$

where U is the superficial velocity of the gas. Rearranging,

$$F = \frac{U}{C_d\left(\frac{2\ \Delta p_D}{\rho_g}\right)^{1/2}} = \frac{U}{U_{or}} \tag{7}$$

or

$$\Delta p_D = \frac{\rho_g U^2}{2C_d{}^2F^2} \tag{8}$$

In many applications the windbox and the bed are at different temperatures and it is more convenient to use the mass flux of gas, G.

$$G = \rho_g U = \rho_{gw} U_w \tag{9}$$

where the subscript w refers to conditions in the windbox.

The gas passing through the holes is usually at the temperature of the windbox so ρ_{gw} should be used. Equation (7) becomes

$$F = \frac{G}{C_d\sqrt{2\rho_{gw}\ \Delta p_D}} \tag{7a}$$

and eqn. (8) becomes

$$\Delta p_D = \frac{G^2}{2\rho_{gw} C_d{}^2F^2} \tag{8a}$$

When dealing with a multi-orifice distributor it is usually convenient to express eqn.

(8a) in terms of the ratio of pitch/hole diameter p/d_{or}.

For square pitch $N = 1/p^2$ and for triangular pitch $N = 2/(\sqrt{3}p^2)$. Substituting in eqn. (3) for triangular pitch,

$$F = \frac{\pi}{2\sqrt{3}}\left(\frac{d_{or}}{p}\right)^2 \tag{10}$$

Substituting in eqn. (8a) for F,

$$\Delta p_D = \frac{6G^2}{\pi^2 C_d{}^2 \rho_{gw}}\left(\frac{p}{d_{or}}\right)^4 \tag{11}$$

Equations (8a) and (11) show clearly how sensitive the distributor pressure drop is to F, p/d_{or} and gas flow rate

THE PRESSURE DROP RATIO $\Delta p_D/\Delta p_B$

The next step is to specify what the pressure drop should be to ensure 'satisfactory' bed operation. It should be understood that the fact that equal areas of a gas distributor deliver, on average, the same flow rate of gas does not necessarily mean that the bubble flow is spatially or at all times uniform.

It is well known (*e.g.* see Ref. 1) that solids circulation patterns develop, especially in large beds, which influence the bubble flow and that holes may not operate continuously. 'Satisfactory' or 'stable' bed operation implies [2, 3] that all holes or tuyères are in continuous operation. In the case of any individual operational tuyère, the gas flow through it oscillates randomly in response to local pressure changes due to solids flow and bubble detachment. During operation in the unstable regime, some tuyères or holes stop bubbling for periods of several minutes or even permanently. Non-operational tuyères or holes are surrounded by immobile solids and although they deliver gas to the bed it is not sufficient to produce bubbles: consequently, both the gas flow rate and local pressure differential are steady. It should be emphasised that such holes are not necessarily mechanically blocked.

Agarwal *et al.* [4] proposed that Δp_D should be approx. 10% of the bed pressure drop and never less than about 3400 N/m^2 (35 cmH_2O), whilst other authors [5, 6, 7] propose minimum values for c, the ratio of distributor to bed pressure drop ($\Delta p_D/\Delta p_B$) ranging from 0.02 to 1, with 0.3 as a widely quoted value [8]. In order to appreciate how these apparently different values have come to be used, we must examine what the response of the system is to a disturbance.

Following the approach of Hiby [9], consider Fig. 3A, in which all the holes are operating when supplied with a flow of gas Q, superficial gas velocity U_o.

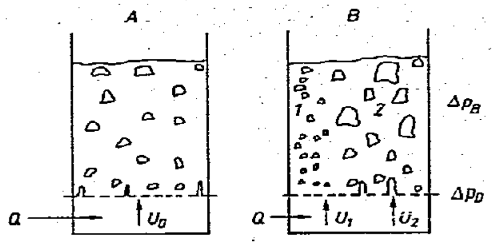

Fig. 3. Bubbling bed with (A) even bubbling (B) maldistribution.

Now assume that the system is disturbed so that some of the holes cease to operate but the flow remains at Q overall. The velocity in region 1 falls to U_1 and that in region 2 rises to U_2. Solids are displaced from section 2 to section 1 and $\Delta p_{B2} < \Delta p_{B1}$. However, because the velocity has risen in section 2, $\Delta p_{D2} > \Delta p_{D1}$. If the rise in the local value of the distributor pressure drop is less than the fall in the same local bed pressure drop, the perturbation will be damped out, *i.e.* if $\Delta p_{B2} + \Delta p_{D2} > \Delta p_{Bo} + \Delta p_{Do}$.

If the new total pressure drop in that region is less than the original value, the maldistribution will get worse. Clearly, the rate of change of Δp_D with velocity, $d(\Delta p_D)/dU$, is the controlling factor, as illustrated in Fig. 4. A distributor having a 'high' pressure drop at the operating velocity U_o has a larger value of $d(\Delta p_D)/dU$ than a distributor with a 'low' pressure drop. Hiby's analysis led to a criterion for c of 0.15 at low values of U/U_{mf} and 0.015 at high velocities.

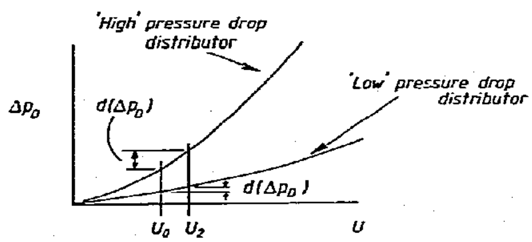

Fig. 4. 'High' and 'low' pressure drop distributors.

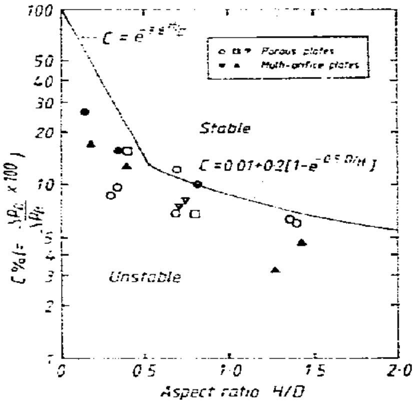

Fig. 5. Data of Geldart and Kelsey [3] for stable operation of distributors.

Geldart and Kelsey [3], experimenting in a two-dimensional bed, concluded from their data that the bed height/bed width — the aspect ratio — influenced the critical value of c required to ensure stable fluidization, that is for all holes to be in operation.

Quereshi and Creasy [10] reviewed the literature for data on successful and unsuccessful commercial fluid beds and also concluded that aspect ratio has an influence. They proposed that for stable operation

$$c \geqslant 0.01 + 0.2[1 - \exp(-0.5D/H_{mf})] \qquad (12)$$

Geldart and Kelsey's [3] data (see Fig. 5) broadly agree with this, except for low aspect ratio beds, for which eqn. (12) gives values which are, in our opinion, too low.

A more conservative approach is to use eqn. (13) for values of $H_{mf}/D < 0.5$.

$$c \geqslant \exp(-3.8H_{mf}/D) \qquad (13)$$

The finding that c depends on aspect ratio explains the range of criteria found in the literature; workers who used shallow, large-diameter beds in general recommend high values of c, whilst those who used small-diameter, fairly deep laboratory equipment, recommend lower values of c. For a powder with a bulk density of 1000 kg/m^3, a static bed depth of 0.2 m would require, for stable operation, distributor pressure drops of 4.4, 13.6 and 16.5 cmH_2O in beds 0.5, 1.2 and 4 m diameter respectively. A distributor pressure drop of 35 cmH_2O at the operating velocity as advocated by Agarwal *et al.* [4] should be regarded as conservative for almost all beds up to about 4 m diameter.

There is no evidence that there is any advantage to be gained from increasing the distributor pressure drop above the minimum given by eqns. (12) and (13); indeed, there may be disadvantages, since high pressure drops require larger values of p/d_{or} and U_{or} with the attendant risk of dead areas and particle attrition.

It should be pointed out that, since for multi-orifice distributors $\Delta p_D \propto G^2/\rho_{gw}$ (eqn. (8a)), a distributor designed to give an adequate pressure drop at the operating velocity and temperature may fail to do so if the mass flux of gas is reduced and/or the windbox temperature falls; the possibility of turn-down must therefore be considered carefully when designing the distributor.

The other parameter which determines Δp_D is the discharge coefficient C_d.

Kunii and Levenspiel [11] present a chart of C_d *versus* the approach Reynolds $\rho_{gw}U_wD/\mu$, taken from Perry [12]. For distributors having $F < 0.1$, operating at $Re > 3000$ and with the plate thickness/orifice diameter, $t/d_{or} < 0.1$, $C_d \approx 0.6$. For square-edged circular orifices with $t/d_{or} > 0.09$, Quereshi and Creasy [10] give

$$C_d = 0.82(t/d_{or})^{0.13} \qquad (14)$$

Assuming that Δp_D has been specified and C_d has been estimated, the fractional free area F can be calculated for the normal operating velocity using eqn. (7a). Equation (10) shows that this also fixes p/d_{or}, but knowing p/d_{or} is not sufficient to define the distributor since, for a given value of p/d_{or}, there could be many small holes close together or a few large ones widely spaced. It is therefore necessary to specify p or d_{or}, or at least define their limits.

SIZE AND SPACING OF HOLES

Practical considerations

A number of generalizations can be made about the pitch and diameter of the multi-orifice distributor.

(a) Distributors having holes smaller than 1 mm diameter are expensive but are available commercially down to 0.1 mm in plates 0.75 mm thick for applications such as fluid bed dryers.

(b) If holes are larger than about 5 $\overline{d_p}$, the bed will drain into the windbox when defluidized. However, this limitation can be avoided by the use of mesh under the plate or by using bubble caps, or tuyères with horizontal holes.
(c) Porous plates give the smallest bubbles, highest chemical conversion, no dead areas at the plate level, and have the advantage that particles cannot fall through. However, there is often considerable variation in porosity (giving uneven fluidization) and the danger of blockage from the underside. Segregation is also more likely because the bubbles are too small to lift the solids and mix them.
(d) If N exceeds 1000 holes/m^2, a few centimetres above the plate bubble sizes are virtually the same as those from a porous plate [13]. This suggests that from the point of view of chemical reactions, there is normally little point in making p smaller than about 3 cm. However, if N is too small (large p and d_{or}), problems of jet penetration and inadequate particle movement may occur and these are discussed below.

Jet penetration and initial bubble size

There is still discussion as to whether drilled plate or tuyère distributors give rise to bubbles or jets. As Clift [14] has pointed out, all turbulent jets are unstable and present a wavy appearance which can be interpreted either as jets or rapidly coalescing bubbles. At low values of $U - U_{mf}$ bubbles are probably formed directly; at high values of $U - U_{mf}$ jetting occurs prior to bubble formation. Even in the second situation, the eventual initial bubble size can still be calculated from the Davidson–Schüler [15] equation for the volume of a bubble formed by a flow rate of gas q into a liquid.

$$V_b = \frac{1.14q^{1.2}}{g^{0.2}} \tag{15}$$

For a multi-orifice distributor,

$$q = \frac{U}{N} \quad \text{and} \quad V_b = \frac{\pi d_{eq}{}^3}{6}$$

so that

$$d_{eqi} = 1.3\left(\frac{q^2}{g}\right)^{0.2} = 1.3\left(\frac{U^2}{gN^2}\right)^{0.2} \tag{16}$$

or

$$d_{bi} = 1.43\left(\frac{q^2}{g}\right)^{0.2} \tag{17}$$

where d_{eqi} is the diameter of the equivalent volume sphere at the distributor and d_{bi} is the frontal diameter of the bubbles formed initially. (Some workers prefer to use $U - U_{mf}$ instead of U in eqn. (16), arguing that the gas needed to incipiently fluidize the bed leaks away rapidly before bubbles are formed.) The gas/solid contact near the grid is an important factor in obtaining good chemical conversion and the smaller the bubbles, the larger their surface area and the better the conversion [16].

Whether or not there are jets or rapidly coalescing bubbles, it is certain that if surfaces such as heat-exchanger tubes are positioned too close to the distributor, erosion by the particles can occur. If the bed is shallow, breakthrough of jets to the surface may occur and this is usually undesirable. It is therefore important to be able to estimate the height over which particle acceleration effects extend. However, published correlations for the jet penetration length L_j present problems of interpretation.

Knowlton and Hirsan [17] define three lengths (see Fig. 6). The largest, L_b, is the distance reached by bubbles before their momentum is dissipated.

Some workers take an average of the lengths reached by the fluctuating dilute

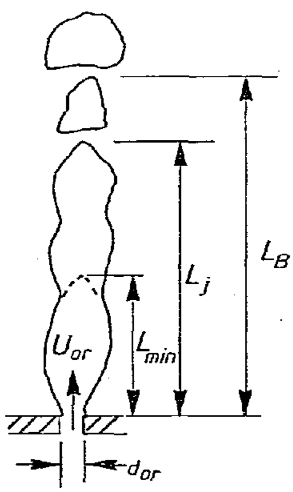

Fig. 6. Jet penetration lengths.

phase region, others, the maximum of the fluctuating region.

Some workers used secondary gas to fluidize the bed [17, 18]. Most did not [8, 19, 20]. Generally, the dilute phase portion of the jet penetrates further when there is no secondary gas.

Few workers varied all the parameters included in their correlation. For example, although the gas density is in all the correlations, usually only air at ambient conditions was used. Predictions for jets or hot gases at high pressures, or gases containing solids, should be made with considerable caution. Knowlton and Hirsan [17] and Yang [21] are the only workers to vary pressure.

For systems in which there is no secondary supply of gas, *i.e.* all the gas enters through the holes, the correlation of Merry [19] gives reasonable agreement with data deduced from erosion patterns in commercial units.

$$\frac{L_j}{d_{or}} = 5.2\left(\frac{\rho_g d_{or}}{\rho_p d_p}\right)^{0.3}\left[1.3\left(\frac{U_{or}^2}{g d_{or}}\right)^{0.2} - 1\right] \qquad (18)$$

The correlation by Yang [21] for the penetration of large jets into beds fluidized at U_{cf} by a separate supply of gas makes use of data by Knowlton and Hirsan [17]. U_{cf} is the velocity at which the bed is completely supported and Knowlton and Hirsan have proposed that it can be estimated from

$$U_{cf} = \sum X_i U_{mfi} \qquad (19)$$

The correlation, which is claimed to be applicable to pressures up to 53 atm, is

$$\frac{L_j}{d_{or}} = 7.65\left(\frac{1}{R_{cf}}\,\frac{\rho_g}{\rho_p - \rho_g}\,\frac{U_{or}^2}{g d_{or}}\right)^{0.472} \qquad (20)$$

where $R_{cf} = \dfrac{U_{cf} \text{ at pressure}}{U_{cf} \text{ at 1 atm}}$.

Once F has been specified for a particular flow rate, then U_{or} is fixed (eqn. (7)). Selection of a jet penetration length which avoids impingement on internal surfaces then allows an orifice size d_{or} to be calculated from eqns. (18) or (20), depending on whether or not secondary gas is used. Since $F \propto (d_{or}/P)^2$, the pitch is now also fixed and must be checked against other criteria such as dead areas (see below). Note that these equations relate to *vertical* jets. Horizontal jets emerging from nozzles penetrate to much smaller distances, but few data are available [22].

Particle movement at the plate

If the solids to be fluidized have a tendency to stickiness or go through a sticky stage, it is usually important to ensure that there are no dead areas on the plate. The particle movement induced by the gas issuing from a hole depends on the flow properties of the fluidized solids (partially characterised by U_{mf}) and the gas flow rate per hole. At low gas flow rates there is little movement, although with porous plates bubbles occur randomly, aerating all the solids at the plate. However, the small bubbles produced at a porous plate do not have sufficient energy to cause the vigorous movement required to remix larger particles which may have segregated or even defluidized. On the other hand, if drilled plates having large holes are used, the large distance between them (for a given pitch/diameter ratio) allows solids to settle out (Fig. 7). Baeyens [23] made a study of particle movement caused by bubbles and suggested that for particle movement over the entire plate,

$$p < \lambda d_{eqi} \qquad (21)$$

where λ is a function of the particle mobility and d_{eqi} is the initial bubble size produced at the distributor. Geldart [24] used this approach to derive eqn. (22), which gives the minimum gas velocity required to avoid dead areas between holes:

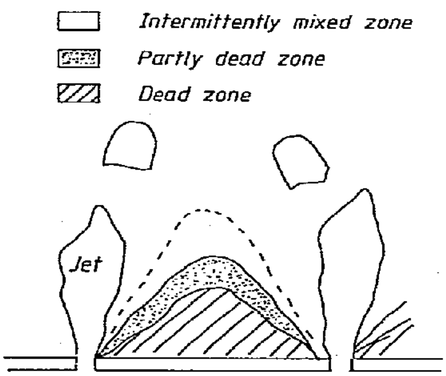

Fig. 7. Dead zones on a multi-orifice plate (after Wen *et al.* [26]).

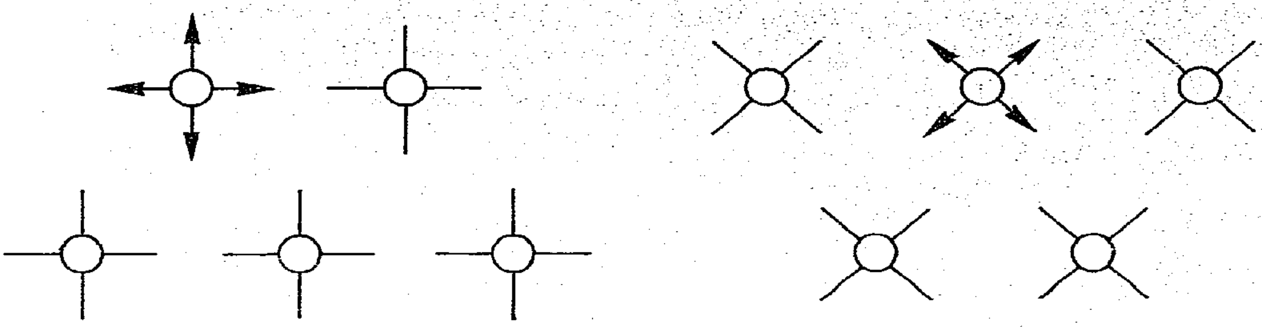

Fig. 8. Orientation of holes in tuyères. B is better than A in reducing dead zones.

$$U - U_{\mathrm{mf}} > \frac{(gp)^{0.5}}{2.45\lambda^{1.78}} \tag{22}$$

For conservative design, $\lambda \approx 1$ for Geldart [25] Group B and D solids (typically sandlike materials) and 1.5 for Group A (typically cracking catalyst). For fluidization by air at ambient conditions, a powder is in Group A if

$$\rho_{\mathrm{p}}^{0.934}(\overline{d_{\mathrm{p}}})^{0.8} < 1 \tag{23}$$

Wen *et al.* [26] used thermistors to detect the movement of particles in Groups B and D at distributor plates and proposed that

$$U - U_{\mathrm{mf}} > \frac{30}{p}(p - d_{\mathrm{or}})^{0.716 d_{\mathrm{p}}^{0.205}} \tag{24}$$

Units are cm/s and cm, with d_{p} in microns. Dead zones are easier to eliminate using tuyères with multiple horizontal holes or conical-top bubble caps (sometimes called 'chinese hats'). The orientation of the holes influences the dead zones and Wen *et al.* found arrangement B in Fig. 8 better than arrangement A.

Additional problems arise when cohesive Group C powders are fluidized. Powders with a mean size less than about 30 μm fall into this category and because the interparticle forces are large compared with hydrodynamic drag forces, the powder fluidizes poorly even with very uniform high pressure distributors because irregular gas channels form from distributor to bed surface. Stirrers positioned close to a porous plate are sometimes used and a combined stirrer/distributor has been used with some success [28]. For high-temperature applications, a screw plate is commercially available [29].

OTHER FACTORS INFLUENCING DESIGN

Proportion of holes in operation

The minimum pressure drop required across a multi-orifice distributor to ensure that all holes operate can be found from eqns. (12) and (13), but some workers have adopted an alternative approach.

At what velocity U_{m} should a distributor already built operate to ensure that all holes operate? If it operates at a velocity lower than U_{m}, what fraction of the holes will be in operation? Whitehead and Dent [1] and Fakhimi and Harrison [30] addressed these questions and Yue and Kolaczkowski [31] developed the ideas further. Their equations below can be rearranged to provide the minimum distributor pressure drop ratio c required to ensure uniform hole operation. Unfortunately, the resulting equations are not explicit in Δp_{D} and have to be solved by an iterative procedure. Equations (25) and (26) are of more value in giving an estimate of U_{m}, the superficial velocity needed by a given distributor to ensure all holes operate, and the fraction of the holes in operation when the flow is turned down to U, a lower velocity.

$$U_{\mathrm{m}} = \left[\frac{\Delta p_{\mathrm{B}}}{\rho_{\mathrm{g}}}\left(\overline{\epsilon_{\mathrm{b}}} + 0.36\frac{L_{\mathrm{j}}}{H_{\mathrm{mf}}}\right)2C_{\mathrm{d}}^{2}F^{2} + U_{\mathrm{mf}}^{2}\right]^{1/2} \tag{25}$$

The fraction of holes operating at a superficial velocity U less that U_m is

$$\frac{n_{op}}{n_{tot}} = \frac{U - U_{mf}}{U_m - U_{mf}} \tag{26}$$

$\overline{\epsilon_b}$ is the fraction of the bed occupied by bubbles and can be estimated from

$$\overline{\epsilon_b} = \frac{Q_b/A}{\overline{u_{bs}}} \tag{27}$$

Q_b is the visible bubble flow rate, and by the modified two-phase theory

$$\frac{Q_b}{A} = (U - U_{mf})(1 + 2\overline{\epsilon_b}) \tag{28}$$

According to Bar-Cohen *et al.* [32], this is satisfactory for large (Group D) particles. However, it probably overestimates Q_b for the smaller Group B and A powders and a better approximation is

$$\frac{Q_b}{A} = Y(U - U_{mf}) \tag{29}$$

where Y is a function of particle properties [23] and $0.5 < Y < 1$. This is in broad agreement with the data of Werther [33], who proposed $Y \approx 0.67$ for Group B solids.

The other term in eqn. (27) is the average bubble rise velocity $\overline{u_{bs}}$:

$$\overline{u_{bs}} = U - U_{mf} + 0.71\sqrt{g\overline{d_{eq}}} \tag{30}$$

$\overline{d_{eq}}$ is the average bubble size in the bed and can be estimated from one of the bubble size equations, for example that of Darton *et al.* [34].

$$\overline{d_{eq}} = \frac{0.54}{g^{0.2}}(U - U_{mf})^{0.4}\left(h + 4\sqrt{\frac{1}{N}}\right)^{0.8} \tag{31}$$

It is usually sufficiently accurate to use $h = 0.5H_{mf}$, but note that neither eqn. (31) nor any of the other bubble size equations can be applied if the bed is likely to be slugging, that is, if d_{eq} in the upper part of the bed $> D/3$.

Solids backflow

The backflow of solids through the distributor into the windbox can occur
(a) during shut-down of the bed (when it is known as *dumping*);
(b) during normal operation (when it is called *sifting* or *weeping*).
The former, (a), can lead to such large amounts of solids in the windbox or sparge

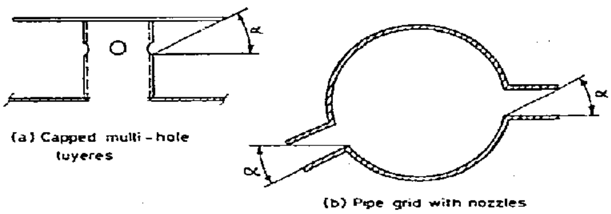

Fig. 9. Designs to prevent solids backflow on shut-down (Group B and D solids).

pipes that start-up is difficult or impossible; the latter, (b), is one of the main causes of hole erosion.

For Group B and D solids, the simple constructions shown in Fig. 9 can be effective in preventing dumping; the angle α is made smaller than the angle of repose of the solids. However, these designs are much less effective for Geldart [25] Group A powders, which de-aerate slowly and so retain their flowability for many seconds after the gas has been shut off. As most reactor systems operate for many months between shut-downs, this is not usually a problem, but where beds of Group A powders are shut down frequently, bubble caps may be required.

Weeping occurs even when the velocity in the holes is far in excess of the terminal velocity of the particles. This is due sometimes to the pressure pulsations in the bed and sometimes to the design of the tuyère or hole which gives rise to velocity distribution and secondary circulating flows. Serviant *et al.* [35] used 60 μm cracking catalyst at superficial gas velocities up to 0.3 m/s. Results from his single-orifice experiments indicate that weeping decreases with increasing orifice velocities, decreases with increasing length of the inlet nozzle and is relatively insensitive to bed height.

Multi-orifice tests indicate that weeping increases with increasing p/d_{or} values. Contrary to the single-nozzle results, it was found that excessive weeping occurs if the nozzle length exceeds a critical length l_c.

$$l_c = \alpha \frac{\rho_g}{\rho_B} \frac{1}{2g} U_{or}^2 \tag{32}$$

α combines empirical correlation factors as indicated in the original paper. This l_c value is important for example when designing a thick refractory-domed distributor with simple vertical nozzles.

Briens *et al.* [36] determined that weeping of cracking catalyst through an orifice plate is correlated as

$$F_{wp}(\text{kg}/(\text{m}^2\,\text{s})) = 6.34 \times 10^9 U_{or}^{-7.82} \qquad (33)$$

This equation is based on the active holes for $U < U_m$ or on all the holes if $U > U_m$. The exponent of U_{or} corresponds to the slope of Serviant's results.

Experiments by Petrie and Black [37] for cap distributors and by Gregory *et al.* [6] for baffled slot plates resulted in weeping fluxes which are a lot smaller than for orifice plates and virtually negligible for $U_{or} > 30$ m/s.

Solids attrition

Whilst high velocities in the holes may reduce or eliminate weeping and give good gas distribution due to the large distributor pressure drop, they can cause attrition of the particles. Attrition in fluidized beds has recently attracted more attention but it is not easy to characterize because much depends on the particles. Standard tests are used in the petrochemical industry and these are useful for comparative purposes but cannot be readily used to predict attrition rates in a fluidized bed [38, 39].

Using plates with multiple drilled holes, Petrie and Black [37] found that for alumina particles attrition remained low at orifice velocities below 50 m/s, whereas at 100 m/s excessive fines were produced. Bergougnou [40] also suggests that velocities >90 m/s should be avoided. For standpipe or cap designs, the high velocity required to give an adequate pressure drop can be reduced before it contacts the powder as shown in Fig. 10.

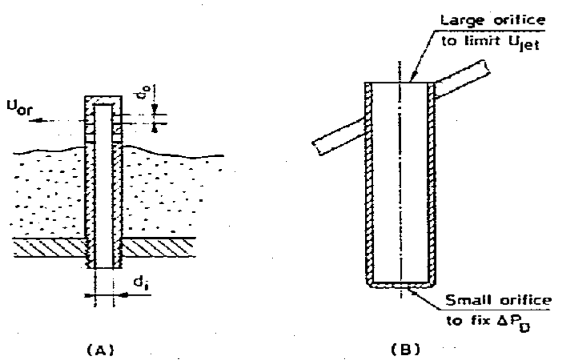

Fig. 10. (A) Multi-hole tuyère. (B) shrouded tuyère.

With a powder which is easily damaged, pressure drop (and highest velocity) should occur at the inlet, that is $d_i^2 \ll n'd_o^2$. If, however, the particles are hard and prevention of sifting back is of primary importance, it may be advisable to ensure that $n'd_o^2 < d_i^2$ (n' = no. of holes/nozzle). This approach is also adopted in the design patented by Monsanto [41]. Wherever possible, testing on a pilot scale should be done.

Erosion

Although erosion is not generally a problem, it is reported to have occurred at localised positions in one of the following forms:

(a) Local erosion of distributor parts, bed walls or in-bed surfaces is mainly due to the *direct impact* of the exit gas jets and entrained particles. It can be avoided by limiting the jet length.

(b) Erosion in the nozzle or orifice is often associated with solids weeping. It can be limited by careful selection of the operating velocities.

(c) Erosion of distributor parts (caps, orifices, ...) can also be due to secondary circulation. The examples below illustrate the phenomenon and show that no general rule applies. Only large-scale testing can give evidence.

— In Fig. 11, the sloping cap design caused secondary circulation patterns, permitting solids to enter the cap. The solids were attrited and the caps were perforated by erosion. The problem was solved by the redesigned nozzle shown [42].

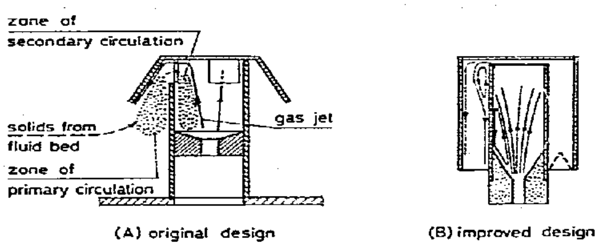

Fig. 11. Shell Chlorine Process nozzles — erosion was considerable in (A) and negligible in (B).

— In Fig. 12, cap erosion was observed due to particles drawn underneath the cap within vortices behind the support and spacer legs. The cap design was adapted to

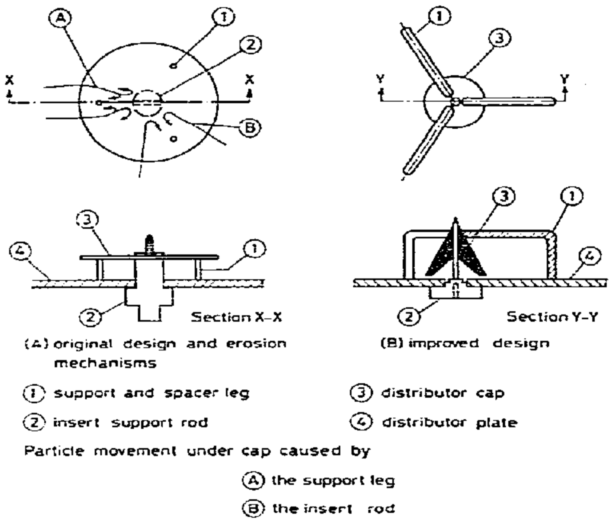

Fig. 12. Erosion in waste calciner distributor plates [37].

minimize solids inflow and using 'outside' spacer legs.

(d) Standpipes with multilayer identical orifices as illustrated in Fig. 13 can suffer erosion due to direct impact of particles upon the roof and within the upper orifices. These particles are sucked within the standpipe through the lower orifices due to venturi effects [43]. Pressure equalization can be achieved by variation of orifice size with height or by using special designs such as Dorr–Oliver's [44].

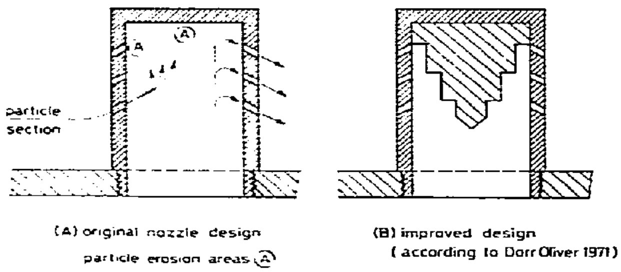

Fig. 13. Erosion in multi-orifice tuyère [43].

Special distributor effects

Although high pressure drop distributors ensure uniform distribution of the gas flow at the base of the fluidized bed, higher up, uniformity might not be achieved because of gross circulation patterns which develop. Deliberately non-uniform gas flow can produce more uniform bubbling patterns as illustrated in Fig. 14.

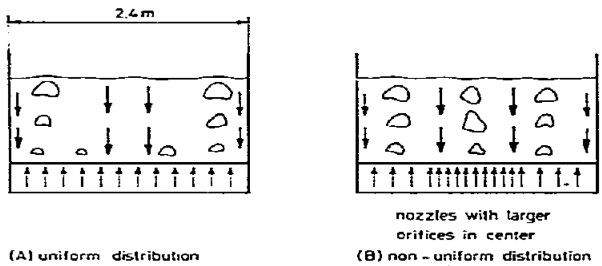

Fig. 14. Bubbling distribution in large shallow beds. More uniform bubbling can be promoted by a non-uniform distributor [1].

Although the fluidized bed design should take into account particle mixing, there are alternatives to putting all the gas required through the distributor. There are advantages in promoting mixing and particle movement by other means so that the resulting lower gas velocity through the distributor can have a beneficial effect on chemical conversion and attrition/erosion. A draft tube is described by Decamps *et al.* [45] who used an internal airlift to prevent segregation at lower overall gas flow rates. Baxerres *et al.* [46] used a sloping distributor to form what they call the 'whirling' bed. Vibration of the plate can also be beneficial in promoting particle movement, especially for large sticky particles.

CONSTRUCTION OF DISTRIBUTORS

For cold experimental work, a simple and effective distributor can be made by sandwiching filter paper or fabric between commercially available punched plate having approximately 30% free area. The plate provides strength and the paper or fabric a controllable pressure drop. These distributors are useful for studying basic fluidization properties, but if the need is to get data for scale-up, distributors must be used which are full-size replicas of those intended for the full size unit.

For cold or low-temperature operations, some suitable commercial distributors are:

(a) Vyon — a sintered polythene or PVC formed into a flat disc.

(b) Rigimesh — a woven stainless steel. This can be variable in porosity, as can Vyon, and if blocked, it is difficult to clean.

(c) Conidur — precision-manufactured slotted holes; usually more suitable for high gas velocities because of the high free area (10%), though a wider range down to 0.01% is now available.

In some applications, although the bed is operated at high temperature, cold gases are supplied to the windbox. This simplifies material selection since the base plate and nozzles are partially cooled and thermal effects are reduced. In this case the 'nozzle standpipes' type of distributor (Fig. 1(d)) is commonly used. Since the air enters the bed from holes or slots at the top of the nozzle, bed material forms a static insulating layer between the hot fluidized zone and the base plate. Consequently, the base plate may be manufactured using mild steel without problems of thermal expansion, and only the nozzles need be of expensive heat-resistant material.

For units in which the bed is heated by fluidization with hot gases, the distributor and windbox must be designed to operate at the high temperature. The 'nozzle standpipe' design can still be used if the base plate is constructed as a sandwich, watercooled and insulated.

A better solution is to use a grid of pipes with multiple air outlets (sparge pipes) (Fig. 1(g)). The sparge pipes may have drilled holes in the underside or be fitted with nozzle standpipes. The advantages of pipe grids is that they can expand from the fixed end. If this construction is not possible, a metal base plate with a dished form can be used to accommodate the expansion or the plate may be fixed to the wall with expansion bellows (e.g. [47]).

Self-supporting ceramic domes (Fig. 1(f)) (with or without caps) are used in continuous operations. Resistance to thermal shock is low and therefore frequent start-up/shut-down procedures are not advisable.

CONCLUSIONS

The design of gas distributors for new fluidized bed processes and redesign to improve performance of existing equipment is now much more of a science than it was 20 years ago. Design equations have been presented and critical problem areas have been discussed. Nevertheless, if at all possible, testing of a full-size module or section of the commercial distributor is recommended.

LIST OF SYMBOLS

A	bed cross-sectional area, m^2
a_{or}	area of orifice, m^2
c	ratio of pressure drop through distributor/bed pressure drop, —
C_d	drag coefficient, —
D	bed diameter, m
d_{or}	diameter of orifice, m
d_b	bubble diameter, m
d_{eq}	diameter of the equivalent volume sphere, m
d_p	average particle size, m
F_{wp}	solids backflow flux, kg/(m^2 s)
F	fractional free area, —
G	mass flux of gas, kg/(m^2 s)
g	gravitational constant, 9.8 m/s^2
H	bed height, m
L_j	jet penetration length, m
l_c	critical nozzle length according to eqn. (32), m
M	mass of solids in the fluidized bed, kg
N	number of orifices per unit area of distributor, m^{-2}
n	number of orifices, —
p	pitch, m
Q	gas flow rate across the entire distributor, m^3/s
Q_b	volumetric bubble flow rate, m^3/s
q	gas flow rate through one hole, m^3/s
R_{cf}	correction coefficient for pressure (eqn. 20), —
t	plate thickness, m
$\overline{u_{bs}}$	average rise velocity of a single bubble, m/s
U	superficial gas velocity, m/s
U_m	superficial velocity needed to ensure that all distributor holes operate, m/s
V_b	volume of a bubble, m^3
X_i	weight fraction of particular size of solids d_i in a mixture, —
Y	correction coefficient for visible bubble flow rate (eqn. (29)), —

Greek symbols

α	correlation factor of eqn. (32), —
Δp_B	pressure drop across a fluidized bed, N/m^2

Δp_D	pressure drop across the distributor, N/m^2
ϵ_b	fraction of the bed occupied by bubbles, —
ϵ_{mf}	void fraction of the bed at minimum fluidization, —
λ	function of particle mobility (eqn. (21)), —
ρ_p	density of the particles, kg/m^3
ρ_g	density of fluidizing gas, kg/m^3

Subscripts

mf	condition at minimum fluidization
or	condition at orifice
g, p	related to gas or particle respectively
w	condition in the wind box
D	distributor
B	fluidized bed
eq	equivalent volume sphere
bi	initial bubble
cf	condition at which the bed is fully supported (eqn. (19))
mfi	at minimum fluidization for particular solids '*i*'

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