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BOTTOM BED REGIMES IN A CIRCULATING FLUIDIZED BED BOILER

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Abstract—This paper extends previous work on the fluidization regimes of the bottom bed of circulating fluidized bed (CFB) boilers. Pressure measurements were performed to obtain the time-average bottom bed voidage and to study the bed pressure fluctuations. The measurements were carried out in a 12 MW_{th} CFB boiler operated at 850°C and also under ambient conditions (40°C). Two bubbling regimes were identified: a "single bubble regime" with large single bubbles present at low fluidization velocities, and, at high fluidization velocities, an "exploding bubble regime" with bubbles often stretching all the way from the air distributor to the surface of the bottom bed. The exploding bubble regime results in a high through-flow of gas, indirectly seen from the low average voidage of the bottom bed, which is similar to that of a stationary fluidized bed boiler, despite the higher gas velocities in the CFB boiler. Methods to determine the fluidization velocity at the transition from the single to the exploding bubble regime are proposed and discussed. The transition velocity increases with an increase in particle size and bed height. Copyright © 1996 Elsevier Science Ltd.

Key Words: circulating fluidized beds, fluidization regimes, bubbling beds, fluid dynamics, fluctuations

1. INTRODUCTION

Previous studies have shown that a major part of the bed material in a circulating fluidized bed (CFB) boiler, operated under normal conditions, is present in a dense bottom bed (e.g. Svensson *et al.* 1993; Johnsson *et al.* 1995). In spite of fluidization velocities which may substantially exceed the terminal velocity of the bed particles, such a bottom bed has the major characteristics of a bubbling fluidized bed. There is no transition to turbulent fluidization. Instead, with an increase in velocity, there is an increase in gas flow through the bubbles and the bubble dynamics show fluctuation frequencies similar to those of a bubbling bed ($\sim$1 Hz).

The bottom bed is the entrance zone for primary combustion air, fuel and limestone for sulphur capture. In order to control harmful emissions, e.g. NO_x and SO_2, the mixing of air, fuel and limestone in the bottom bed is important. Also, the character of the bottom bed is linked to the flow pattern in the upper parts of the combustion chamber (Johnsson & Leckner 1995). Therefore it is important to study the bottom bed and attention is paid to its properties in this article as well as in previous work by Johnsson *et al.* (1992); Svensson *et al.* (1993); Svensson *et al.* (1996).

Different bubbling regimes

The large cross-sectional area of a commercial CFB boiler in combination with the desire to minimize the pressure drop of the gas supplied by the primary air fans, results in a bed height which is significantly lower than the bed width (bed-height to width $H_x/L < 1$). Furthermore, the pressure drop across the air distributor should be as low as possible.

In non-slugging, freely bubbling fluidized beds there are two types of bubbles reported in the literature: rather small "conventional" bubbles and, at higher velocities with coarse particles (Geldart B and D), exploding bubbles (Fitzgerald 1980; Johnsson *et al.* 1991; Johnsson *et al.* 1992; Svensson *et al.* 1993). In a previous investigation on the influence of the air-distributor pressure drop on the bottom bed behaviour in a cold CFB, a third type of bubble was found (Svensson *et al.* 1996). This third type of bubble occurred in the bed one at a time and was called "single bubble" in contrast to conventional bubbles which were termed "multiple bubbles". The multiple

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bubble regime was observed during operation at low gas velocities (U) with high distributor pressure drops (Δp_{dist}), whereas the single bubble regime occurred at low U and low Δp_{dist}. The single and the exploding bubble regimes, but not the multiple bubble regime, were observed in the CFB boiler investigated. This is interpreted to be a consequence of the low pressure drop of the CFB boiler distributor. The gas flow in both the single and the exploding bubble regimes is discontinuous since, after a bubble erupts at the bed surface, there is a pause before a new bubble enters the bed. The bubble regimes found in non-slugging beds (the cold CFB and the CFB boiler) are summarized with respect to Δp_{dist} in figure 1.

In the single bubble regime, the bubble frequency was about 1–2 Hz. This frequency could be calculated by a simple model, considering the bed as a mass of particles supported by a gas volume (Davidson, 1968) assuming the bed to oscillate around an equilibrium. The pressure drop of the distributor is not taken into account. The model shows that the frequency of the single bubbles is equal to the natural frequency of the oscillating bed supported by the gas volume of the air supply system. When the velocity was increased and the bed shifted to the exploding bubble regime, frequency spectra of in-bed pressure fluctuations moved to a dominant frequency of 0.5 Hz. Within the parameter range studied, this frequency was independent of gas velocity and bed height.

In the exploding bubble regime, the gas velocity is high enough to result in bubbles which are limited in size by the bottom bed height H_x. The bed does not become slugging. Fitzgerald (1980) made gas tracer measurements in a 1 m^2 fluidized bed operated under ambient conditions. There was a large through-flow of gas through the bed during operation in the exploding bubble regime. They explained this to be: "... direct experimental evidence of gas channelling which starts at the moment of bubble explosion". Svensson *et al.* (1996) visually observed that exploding bubbles caused large eruptions when they broke through the bed surface. These bubbles extended, for a moment, all the way from the air distributor to the surface of the bottom bed.

Fitzgerald *et al.* defined the velocity of transition from bubbling regime to exploding bubble

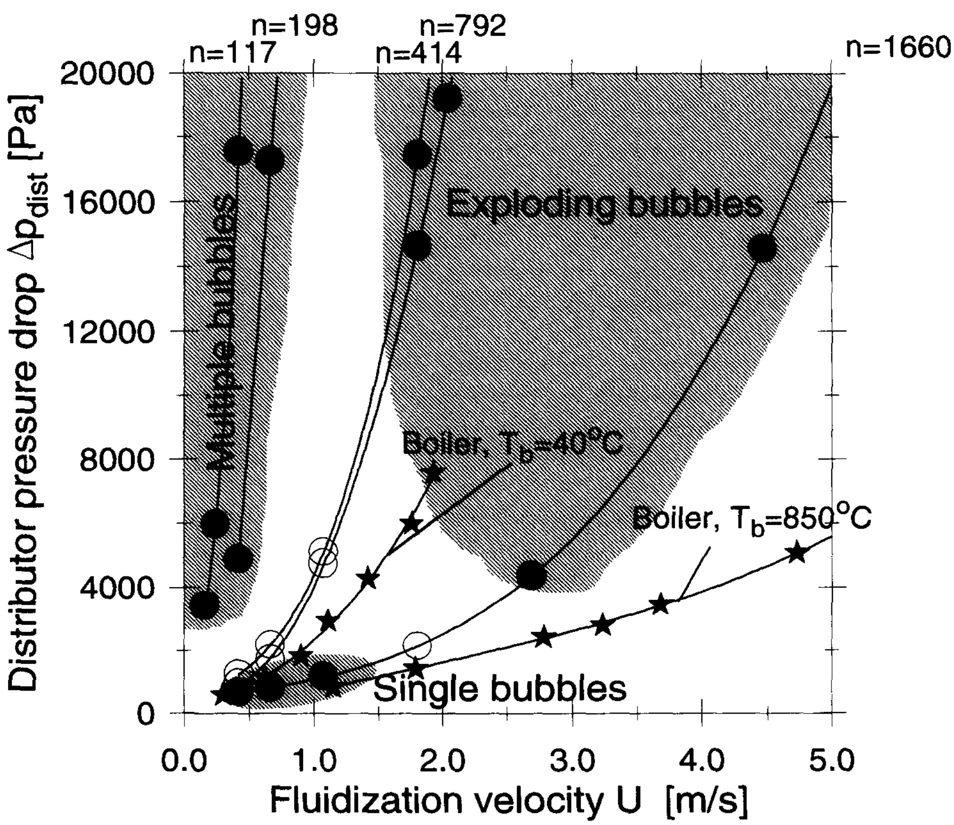


Figure 1. Bubbling regimes in the cold CFB and in the 12 MW CFB for different velocities and air distributor pressure drops. The cold rig was operated with the air distributors listed in table 2. The symbol "●" marks well defined regimes and "○" marks unspecified regimes (transitions).

regime as the velocity where the amplitude of bed pressure fluctuations starts to increase from a constant value. The excess gas velocity ($U - U_{mf}$) of transition was 0.45 m/s (sand, $d_p = 1.3$ to 4 mm, $U_{mf} = 0.73$ to 1.83 m/s). In the case treated by Svensson *et al.*, the transition velocity, estimated from a change in the dominant frequency of the pressure spectra, was about 1 m/s for the fine particles (sand, $d_p = 0.19$ mm, $U_{mf} = 0.01$ m/s). It was possible to maintain the exploding bubble regime in a dense bed up to fluidization velocities amounting to several times the terminal velocity of an average-size bed particle.

Average bed voidage

The increase in average voidage of the bottom bed, $\bar{\varepsilon} = 1 - [(-dp/dh)/g]/\rho_p$, is an expression of bed expansion. Thus, dp/dh corresponds to the slope of the pressure gradient below the height H_x of figure 2. H_x defines the height of the dense bottom bed (e.g. Johnsson *et al.* 1991). Bed expansion is also often related to the bed height (e.g. defined as H_x/H_{mf}), but under CFB conditions, the bed height ratio cannot be used as a measure of the bed density, because the bed material is present, not only in the bottom bed, but also in the rest of the riser and on the cyclone side, and H_x is lower than H_{mf} for a given amount of bed material in the CFB loop. To avoid misunderstanding, it is therefore preferable to use the term "average bed voidage" instead of "bed expansion".

The present investigation focuses on the average bed voidage in non-slugging beds at high fluidization velocities ($U > U_t$) along with variations in bed height, particle size, temperature and air-distributor pressure drop. Most studies of average bed voidage reported in the literature have been carried out at low fluidization velocities ($U \ll U_t$) or in narrow reactors ($H_x > L$). There is, however, some work which has treated the bed voidage in non-slugging beds at intermediate velocities (maximum U is about U_t), e.g. Canada *et al.* (1978); Johnsson *et al.* (1991). For higher velocities ($U > U_t$) some data have been presented in two previous publications by Svensson *et al.* (1993); Johnsson *et al.* (1995). Together, during a rise in velocity from low to high, these results show a substantial increase in bed voidage at low velocities, a moderate increase in an intermediate range of velocities and an almost constant bed voidage at high velocities in the exploding bubble regime. Exploding bubbles cause a high through-flow of gas, so, in spite of the high gas velocity, the bed voidage is not necessarily greater than for a conventional bubbling bed.

Aim of this work

Our aim is to study the bottom bed voidage for a wider range of parameters than previously covered (Johnsson *et al.* 1991; Svensson *et al.* 1993). This is of interest, since little is known about the influence of bottom bed height and distributor pressure drop on the bed voidage at high

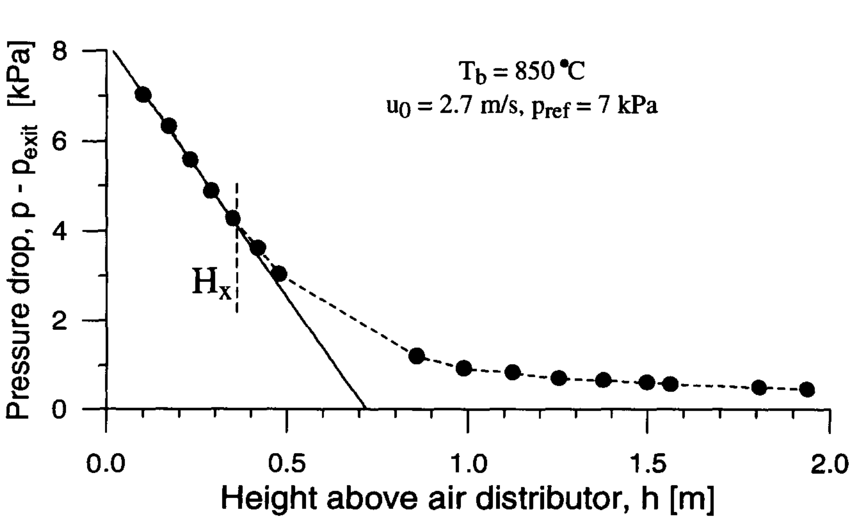


Figure 2. Determination of the bottom bed height, H_x, from the pressure distribution in the lower part of the boiler furnace.

Table 1. Experimental conditions

Experimental unit	Cold rig	Boiler
Cross section of bottom bed, m × m	0.12 × 0.7	1.47 × 1.42
Height of the unit, m	8.5	13.5
Volume of the air supply system (plenum and air pipes), m^3	2.88	5.17
Distributor	Four different perforated plates, table 2	Bubble cap type, table 2
Bed temperature, T_b, °C	40	40 and 850
Fluidization velocity, U, m/s	0.4 to 1.8	0.3 to 1.9 at 40°C and 1.2 to 6.5 at 850°C
Bed material	Silica sand	Silica sand
Bed material density, ρ_b, kg/m^3	2600	2600
Average particle diameter, $d_p = 1/(\Sigma x_i/d_{pi})$, mm	0.32 (value from a typical particle size distribution, figure 1)	0.15, 0.50 at 40°C and 0.19, 0.32, 0.43 at 850°C
Minimum fluidization velocity, U_{mf}, m/s	0.07	0.01, 0.18 at 40°C and 0.01, 0.03, 0.06 at 850°C
Terminal velocity of an average-sized particle, m/s	2.3	1.0, 3.2 at 40°C and 1.0, 2.1, 3.3 at 850°C
Bed height, H_x, m	0.22 to 0.84	0.21 to 0.79

fluidization velocities. The aim is also to locate the gas velocity of transition from the single bubble regime to the exploding bubble regime.

2. EXPERIMENTS

Tests were carried out in a 12 MW_{th} circulating fluidized bed (CFB) boiler operated at 850°C. Corresponding measurements were also made at 40°C in the boiler bed and in a cold CFB. The average bed voidage, $\bar{\varepsilon}$, and the height of the bottom bed, H_x, were obtained from pressure drop measurements, as shown in figure 2. The pressure drop due to hold-up of the gas is small and can be neglected and so is the influence of friction between particles and wall. Also the pressure drop due to acceleration of the bed particles [$= G_s(u - u_t)$] is neglected here, since the recirculation flow, G_s, is low for boiler conditions ($G_s \sim 10$ kg/m^{2}s). The bed conditions were characterized by the amplitude (standard deviation) and the frequency distribution of the pressure signal.

The 12 MW_{th} CFB boiler was operated from 1.2 to 6.3 m/s at 850°C and 0.3 to 1.9 m/s at 40°C. All air was introduced as primary air through the bottom of the boiler. The boiler is described by Leckner *et al.* (1991), but some data are given in table 1 together with the experimental conditions. Data for the air distributor, which is of a bubble cap type, are given in table 2. Silica sand with mean particle sizes (d_p of 0.15, 0.19, 0.32, 0.43 and 0.50 mm were used as bed materials. The particle size distributions are shown in figure 3. There was a slight influence on the size distribution when the bed material was exposed to combustion conditions.

Table 2. Data on air distributors

Unit	Type	Number of holes	Hole diameter (mm)	Bed area (m^2)	Open area (%)	Distributor thickness (mm)
Cold CFB	Perforated plate, triangular pitch	117	2	0.084	0.44	1.5
Cold CFB	Perforated plate, triangular pitch	198	2	0.084	0.74	1.5
Cold CFB	Perforated plate, triangular pitch	414	2	0.084	1.55	1.5
Cold CFB	Perforated plate, triangular pitch	1660	2	0.084	6.21	1.5
12 MW_{th} CFB boiler	Bubble cap type, square pitch	144	20†	2.03	2.26†	10†
16 MW_{th} SFB boiler‡	Bubble cap type, square pitch	627	13†	9.86	0.84†	90†

†Minimum area of gas passage in bubble cap.
‡Used by Johnsson *et al.* (1991).

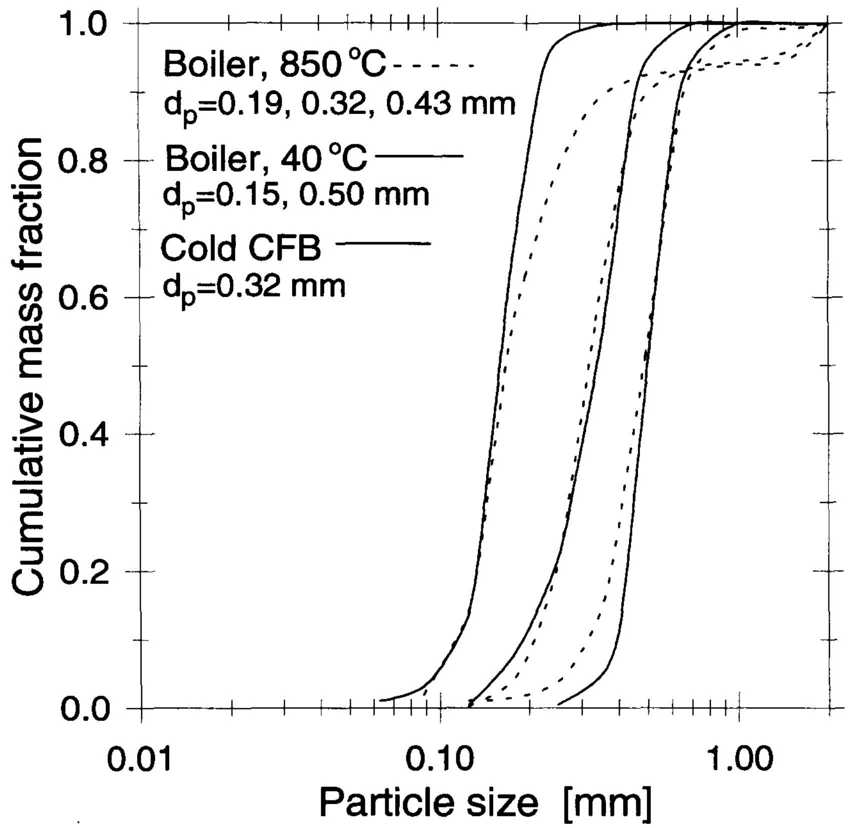


Figure 3. Particle size distribution of the bottom bed materials in the boiler and the cold CFB.

In a CFB boiler, pressure measurements, for practical reasons, are carried out at the walls of the combustion chamber. In order to find out if the pressure profile measured at the walls gives a correct value of the average bed voidage, simultaneous measurements were made in the centre of the combustion chamber in the low temperature case. A vertical beam with pressure taps was inserted so that the pressure taps protruding from the beam were at the centre line of the combustion chamber.

Figure 4 shows that the pressure gradient, i.e. the average bed voidage, is equal at the wall and in the centre of the bed. This was the case for all gas velocities and for the two particle sizes investigated under ambient conditions. However, as seen in figure 4, the pressure drop over the entire bottom bed, Δp_b, is somewhat lower in the centre than at the wall. This difference increases with gas velocity and is in agreement with results obtained by Johnsson *et al.* (1991). They explain the difference in Δp_b by a larger observed bubble flow in the centre than at the walls, which leads to a horizontal redistribution of bed material. This effect is not considered further here. The conclusion is that wall pressure measurements can be used to evaluate the average voidage of the bed.

The cold CFB has a cross-section of 0.12×0.70 m and a height of 8.5 m. The experimental conditions are summarized in table 1. Four perforated plate air-distributors were used; these had orifices of the same diameter but different numbers of orifices. The data for the distributors are given in table 2.

The pressure fluctuations in the bottom bed were measured as gauge pressure (i.e. related to ambient pressure) at 0.1 m above the distributor in the CFB boiler and 0.2 m above the distributor in the cold CFB. A gauge pressure sensor may be influenced by various disturbances from the system, and this is why a gauge pressure sensor was chosen in this case when the interaction between bed and windbox below the bed should be studied. As shown by a previous work (Johnsson *et al.* 1992) under circulating conditions there was no important difference between gauge pressure and

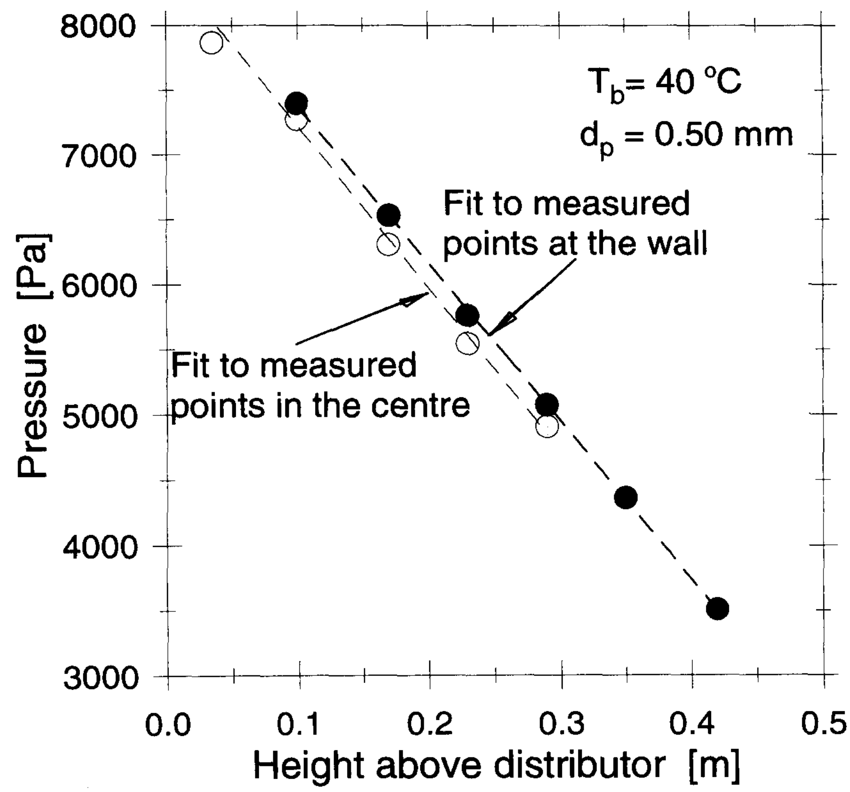


Figure 4. Pressure at the wall and in the centre of the combustion chamber. The measurements were made in the boiler at ambient temperature: $d_p = 0.50$ mm, $H_x = 0.7$ m and $U = 1.6$ m/s.

differential pressure measurements in the characterization of bed dynamics: the pressure fluctuations in the bed are the dominant ones.

Frequency analysis was used to evaluate the bubble frequency. The sampling frequency was 20 Hz, and to ensure sufficient accuracy in the statistical analysis, 32,768 samples were taken. All of the frequency spectra are based on an average of 32 spectra, each of which consists of 1024 samples. The time constant of the pressure transducers is 0.001 s.

The amplitude of the pressure fluctuations is a measure of the relative bubble size. Since both gas velocity and bed height affect the pressure fluctuations, the bed height H_x should be kept constant when the gas velocity is increased. In order to illustrate this, two methods of operation of the boiler were compared; one in which the bed mass was kept (approximately) constant and one in which the total particle mass in the system was constant while the amount of bed material changed freely with fluidizing velocity:

- The first method, **the Δp_{ref} case**, consists in maintaining an almost constant bottom bed height (usually <0.8 m) as the gas velocity was increased by operating the boiler with a constant reference pressure drop, Δp_{ref}, measured between 0.1 m and 1.6 m above the distributor. To achieve this, bed material was added from a hopper in order to compensate for the transfer of solids from the bottom bed to the upper parts of the combustion chamber.
- The second method, **the M_{tot} case**, consists in maintaining a constant total particle mass in the CFB system (riser and return leg), M_{tot}, as the gas velocity was increased; bed material was not added or removed. Three bed masses ($M_{tot,1}$, $M_{tot,2}$ and $M_{tot,3}$) were used. Their absolute values could not be determined.

For practical reasons the bed height H_x is only approximately constant in the Δp_{ref} case; a certain redistribution of bed material occurs from the dense bed to the space between the dense bed and the 1.6 m level as fluidization velocity changes. This effect increases with velocity and with decreased bed height.

3. RESULTS

Single and exploding bubble regimes

Figure 5 shows frequency spectra of pressure fluctuations measured in the bottom bed at fluidization velocities from $U = 1.2$ to 6.3 m/s. At $U = 1.2$ m/s (figure 5a), the spectrum has two peaks at approximately 0.5 Hz and 1.4 Hz. With an increase in gas velocity, the peak at the low frequency (0.5 Hz) is not affected, whereas the higher frequency peak (1.4 Hz) shifts towards lower frequencies. In the previous work by Svensson *et al.* (1996), a similar behaviour was found for beds with smaller particle sizes ($d_p = 0.15$ and 0.19 mm) at both 40 and 850°C. By simultaneous measurements of local bed voidage by optical probes (i.e. direct detection of bubbles) and pressure measurements, it was shown that the high frequency peak is caused by "single bubbles", while the low frequency peak corresponds to "exploding bubbles".

Figure 6 shows the dominant frequency of single and exploding bubbles during operation at 40 and 850°C. The frequency of the single bubbles increases with decreasing bed height. The simple model of the frequency of single bubbles (Davidson 1968) fits the data well, indicating that the single bubble frequency corresponds to the natural frequency of the oscillating bed mass supported by the air volume of the air supply system.

At higher velocities, the bed changes to the exploding bubble regime. The frequency of the exploding bubbles is independent, not only of the gas velocity (0.5 Hz peak in figure 5), but also of the bed height (figure 6b). The frequency of the exploding bubbles cannot be explained at the present state of knowledge.

Turbulent fluidization regime?

Figure 7 shows the bottom bed height, H_x, at a constant bed mass (M_{tot}) and constant reference pressure drop (Δp_{ref}). In the three M_{tot} cases the bottom bed height decreases with the rise in gas velocity. The slope of the curves is almost independent of the value of M_{tot}. Hence, the relative reduction in bed height becomes smaller at greater bed heights. In the Δp_{ref} case (open symbols) the decrease in H_x is smaller than in the M_{tot} case and is caused only by redistribution of bed material below 1.6 m height. The velocity at which the bottom bed disappears in figure 7 depends on the amount of bed material.

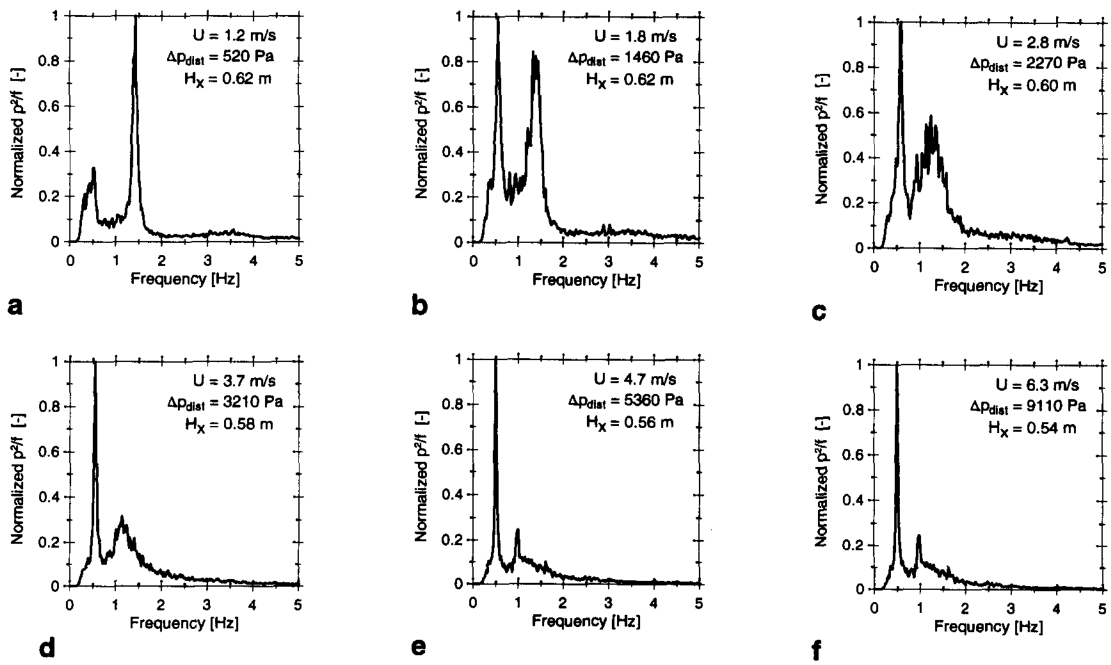
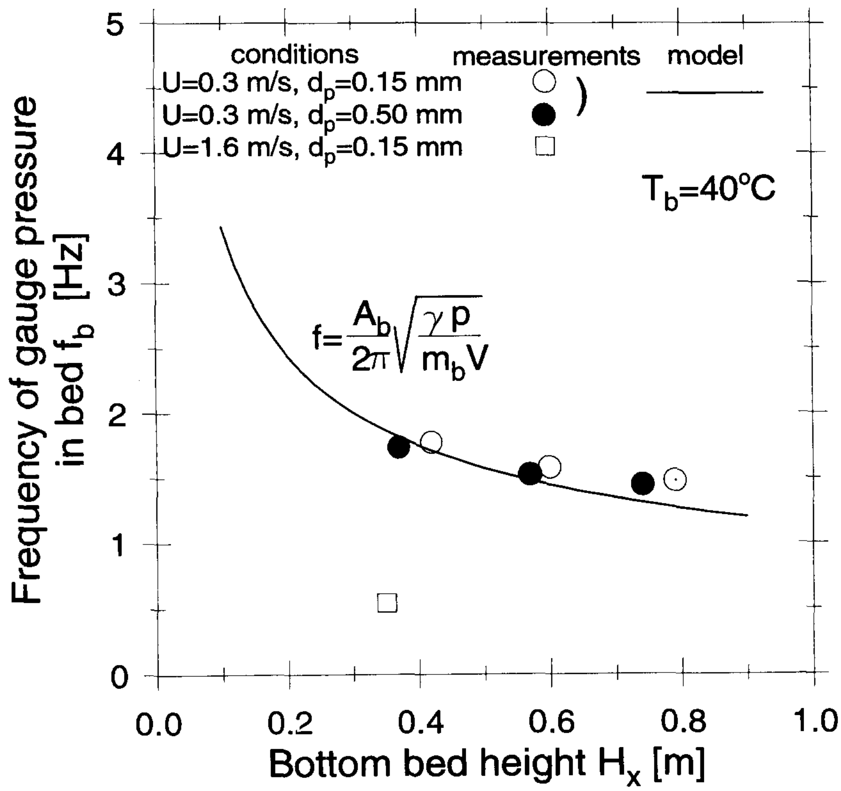


Figure 5. Frequency spectra of the gauge pressure signal measured in the bed of the boiler at different fluidization velocities. The boiler was operated at 850°C with a constant reference pressure drop of 6500 Pa; $d_p = 0.43$ mm.



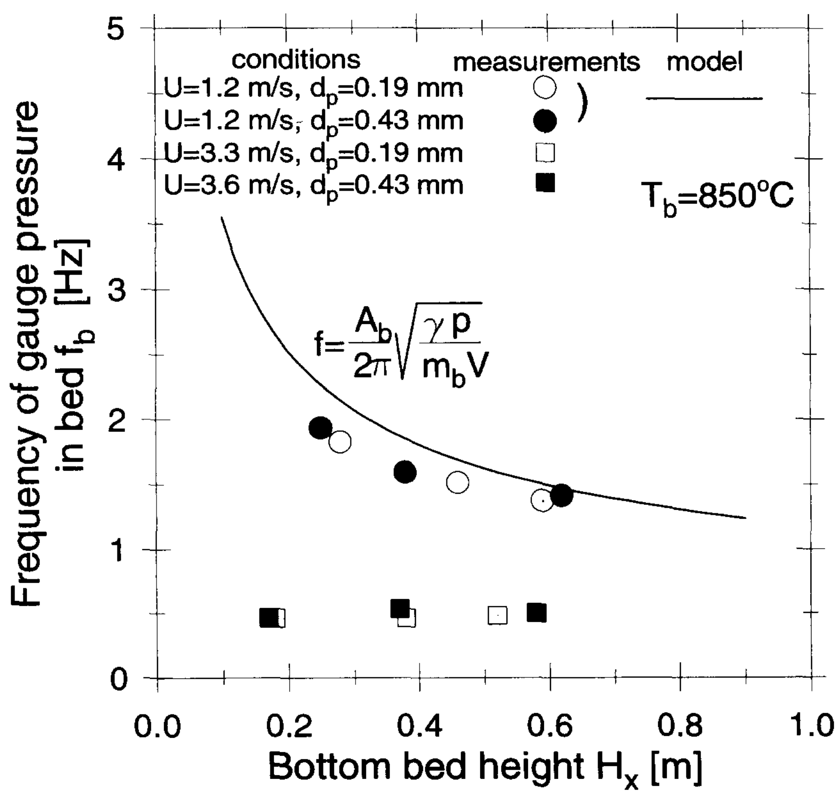


Figure 6. Dominant frequency of pressure fluctuations from single and exploding bubbles in the bottom bed as a function of bed height at 40 and 850°C. Two different sizes of bed material were used. The frequencies of single bubbles are compared with the model of the single bubble frequency.

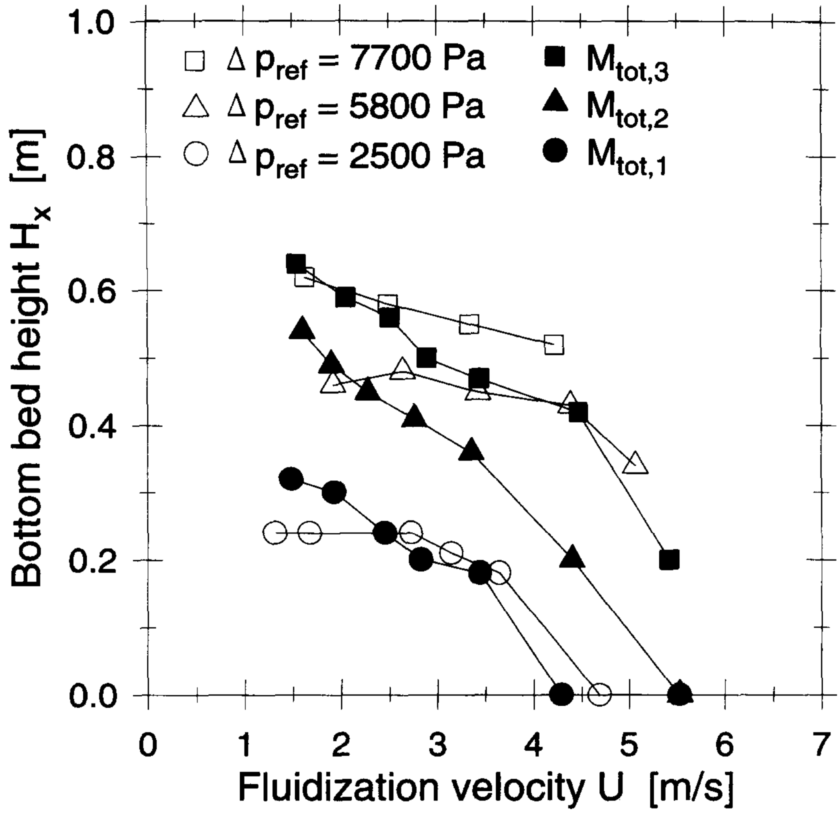


Figure 7. The height of the bottom bed vs fluidization velocity at three reference pressure drops Δp_{ref} and at three masses in the CFB-loop M_{tot}; $d_p = 0.32$ mm and $T_b = 850°C$.

Figure 8 shows the influence of fluidization velocity on the standard deviation of the bottom bed pressure fluctuations in the M_{tot} cases (for which H_x is shown in figure 7). In all cases, except for $M_{tot,1}$, the amplitude of the pressure fluctuations reaches a maximum at a certain velocity, U_c. For $M_{tot,1}$ this velocity is less than 1.5 m/s, which was the lowest possible operating velocity.

The maxima originate from two phenomena acting in opposition. First, an increase in gas velocity leads to larger voids (bubbles), i.e. larger fluctuations. Second, the increase in gas velocity reduces the height of the bottom bed, and this leads to smaller bubbles, i.e. smaller fluctuations. The relative reduction in the bed height is less for a high bed than for a shallow bed. Consequently, the effect of velocity dominates for high beds and the effect of the bed height dominates for shallow beds. The dominant frequency of these pressure fluctuations was low, about 0.5 Hz, and independent of fluidization velocity. This, together with visual observations made in the cold bed, shows that no transition to a turbulent fluidization regime occurred, in spite of the maxima of pressure fluctuations.

As long as the height of the bottom bed can be kept constant, there is no evident maximum in the amplitude of the bottom bed fluctuations, as shown for the Δp_{ref} case in figure 9. In the M_{tot} case, where there is a dramatic reduction of the height of the bottom bed, there is a maximum, but there is only a tendency to a slight fall in the amplitude in the Δp_{ref} case at the highest velocity.

The conclusion is that, provided the bottom bed height can be kept constant, there is no clear maximum in the relationship between the amplitude of the pressure fluctuations and the gas velocity. If a maximum is seen, it is an effect of a redistribution of bed material and as illustrated by figure 9, of the influence of the bottom bed height. The amplitude is a relative measure of the bubble size.

Bottom bed height

For the runs under ambient temperature (at 40°C) the boiler bottom bed height (H_x) was almost independent of fluidization velocity, although the velocity was (for the 0.15 mm particles) increased

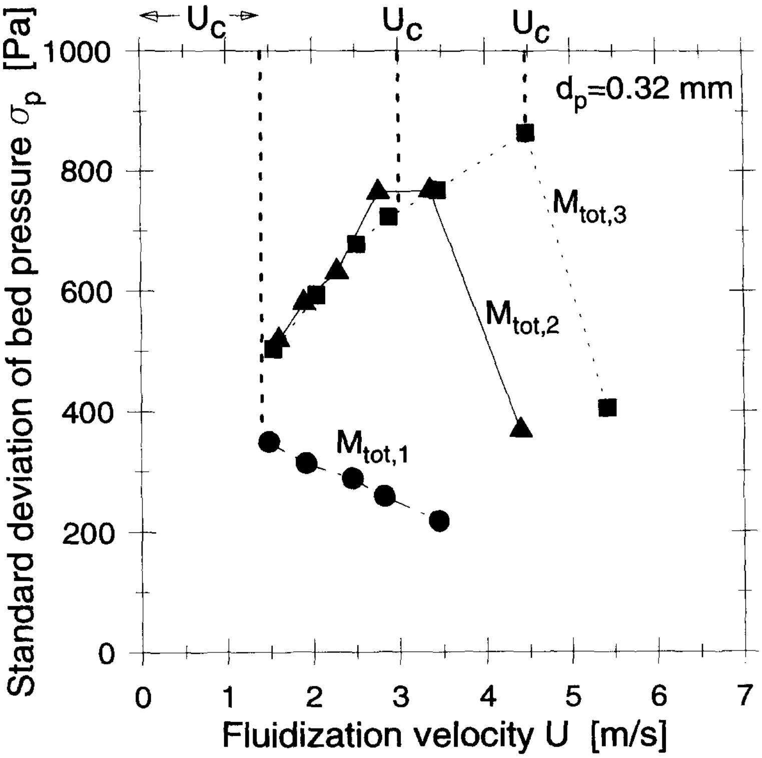


Figure 8. The standard deviation of bed pressure vs fluidization velocity for three masses M_{tot}. The velocities for the maximum pressure fluctuations, U_c are indicated. The corresponding bed heights are shown in figure 7.

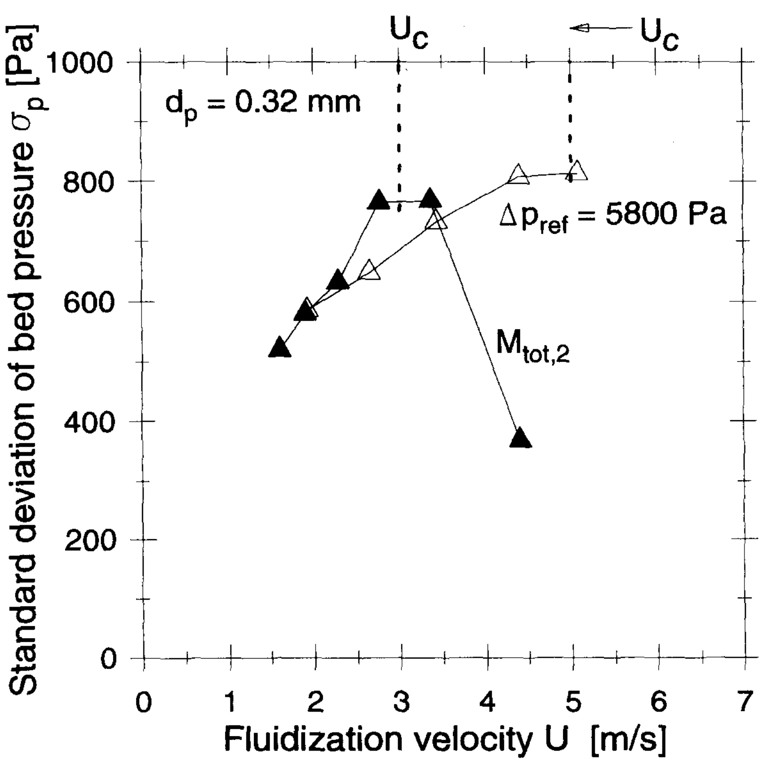


Figure 9. The standard deviation of bed pressure vs fluidization velocity for a constant M_{tot} and a constant Δp_{ref}. The velocities for the maximum pressure fluctuations, U_c, are indicated. The corresponding bed heights are shown in figure 7.

up to about twice the terminal velocity of the bed particles. At 850°C higher velocities were attained and figure 10 shows the bottom bed height as a function of gas velocity for the particle sizes 0.19 and 0.43 mm. The relative reduction in bottom bed height was slightly greater for a bed of small particles than for coarse particles. For both particle sizes the bottom bed was maintained up to several times the terminal velocity of an average size bed particle. The bed containing 0.43 mm particles was maintained at the maximum available gas velocity, whereas with the 0.19 mm particles the bed became unstable and could disappear very rapidly at fluidization velocities above 3 m/s, corresponding to $U/U_t \approx 3$. When the bed disappeared, the boiler had to be stopped. Johnsson *et al.* (1992) showed that the disappearance of the bed has not primarily to do with the particle size, but with the geometry of the CFB unit, and especially with the area of the return duct which impeded efficient recirculation of particles. In the 0.12 × 0.7 m cold CFB, on the other hand, a bottom bed of 0.15 mm particles could be maintained up to 7 m/s at a given inventory of solids (i.e. to about 7 U_t), thus indicating a smaller overall resistance of the external solids recirculation duct. In the Δp_{ref} cases of the boiler runs the system limitations of the recirculation of solids is compensated for by addition of solids.

Average bottom bed voidage

Figure 11 shows the bottom bed voidage of the CFB boiler during "operation" at 40°C. The bed voidage increases with an increase in gas velocity and with a decrease in bed height. It is higher for the 0.15 mm particles than for the 0.50 mm particles. These results are in accordance with observations at low gas velocities ($U < U_t$) published by e.g. McGrath and Streatfield (1971); Cranfield and Geldart (1974). At high velocities, corresponding to the exploding bubble regime, the influence of bed height is smaller. This is expected, since a major part of the gas passes the bed as through-flow, not increasing voidage. The increase in voidage with a decrease in particle size is also in agreement with previous results for small particles, $d_p < 1$ mm, e.g. Hilligardt and Werther (1986); Johnsson *et al.* (1991).

Figures 11 and 12 show qualitatively the same voidage at a bed temperature of 850°C as at 40°C for velocities below 2 m/s, with only a slightly greater voidage at 850°C. It is not possible to detect an influence of particle size in the high temperature case due to problems with maintaining equal Δp_{ref} for both particle sizes. Figure 12 shows an increase in bed voidage which levels off and becomes constant at high velocities.

Figure 13 gives the average bed voidage as a function of bed height measured in the cold CFB at a low gas velocity. Four distributors with different numbers of holes, n = 117–1660, were used. The bed voidage decreased slightly for the three distributors with 117 to 414 holes (high Δp_{dist}), and more for the distributor with 1660 holes (low Δp_{dist}). The larger change in the latter case is explained by the difference in bubbling conditions present with the 1660-hole distributor. The bed contained many bubbles in the "multiple bubble regime" for 117 to 414 holes, but with 1660 holes it was bubbling with large single bubbles.

In Figure 14 the bottom bed voidage in the CFB boiler is compared with the bed voidage in a 16 MW_{th} stationary fluidized bed (SFB) boiler, see table 2. The bed voidage is higher in the SFB boiler than in the CFB boiler at similar bed heights and particle sizes. It is possible that this also is an effect of the air distributor design. The air distributor of the SFB boiler has a smaller number of holes per bed area and a smaller area of gas passage (which makes the distributor pressure drop higher) than those of the CFB boiler (see table 2). Results from the cold CFB, included in figure 14 (also figure 13), support this hypothesis: the low pressure drop distributor gives a lower bed voidage than the one with a higher pressure drop (but at the velocity of 1.8 m/s the pressure drops have increased and the bed voidage becomes similar for the two distributors).

4. DISCUSSION

Average bed voidage

The reason is not known for the variation in the bed voidage with different bed heights or particle sizes, at a constant gas velocity. It could be due to a change in the voidage of the dense phase or in the bubble density. Wein (1992) used a capacitance probe to measure the voidage of the dense

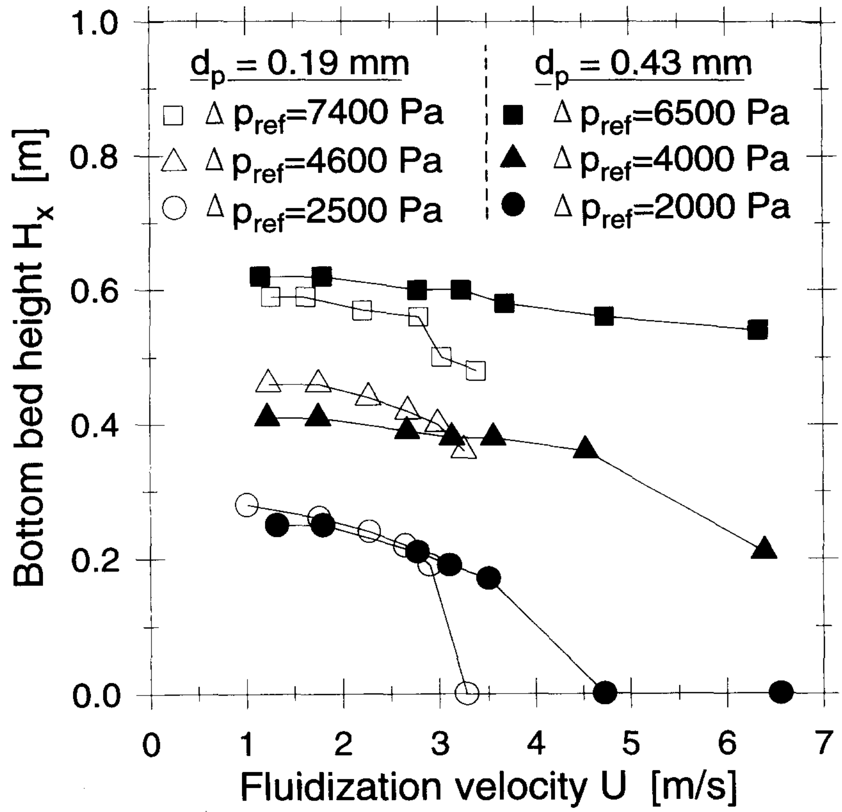


Figure 10. The height of the bottom bed as a function of fluidization velocity for two particle sizes $d_p = 0.19$ mm and $d_p = 0.43$ mm. Three Δp_{ref} for each particle size were investigated; $T_b = 850°C$.

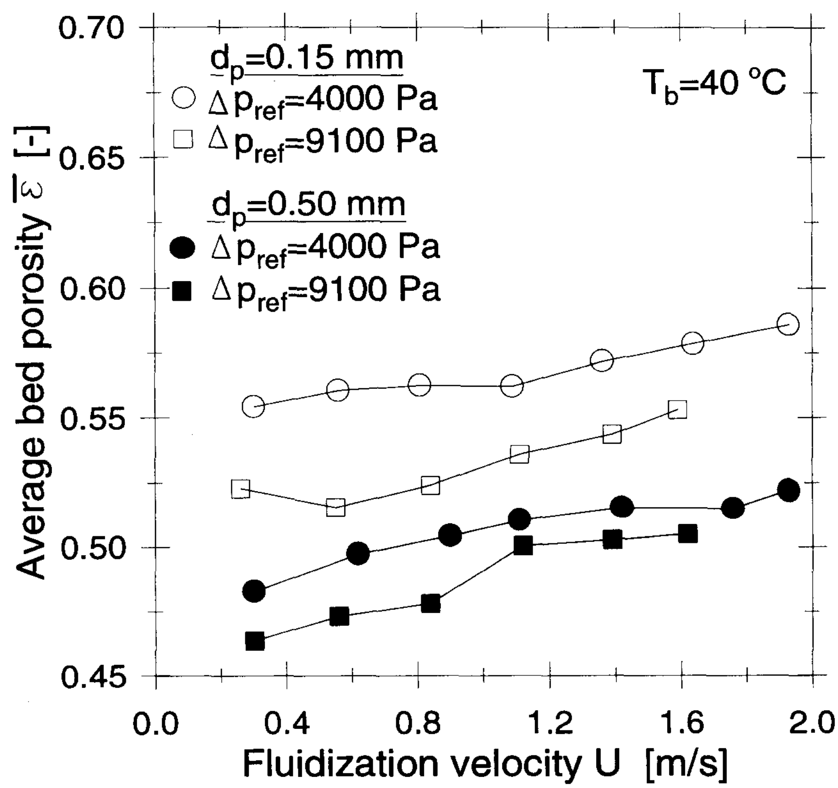


Figure 11. Average bed voidage as a function of fluidization velocity for two Δp_{ref} and two sizes at 40°C.

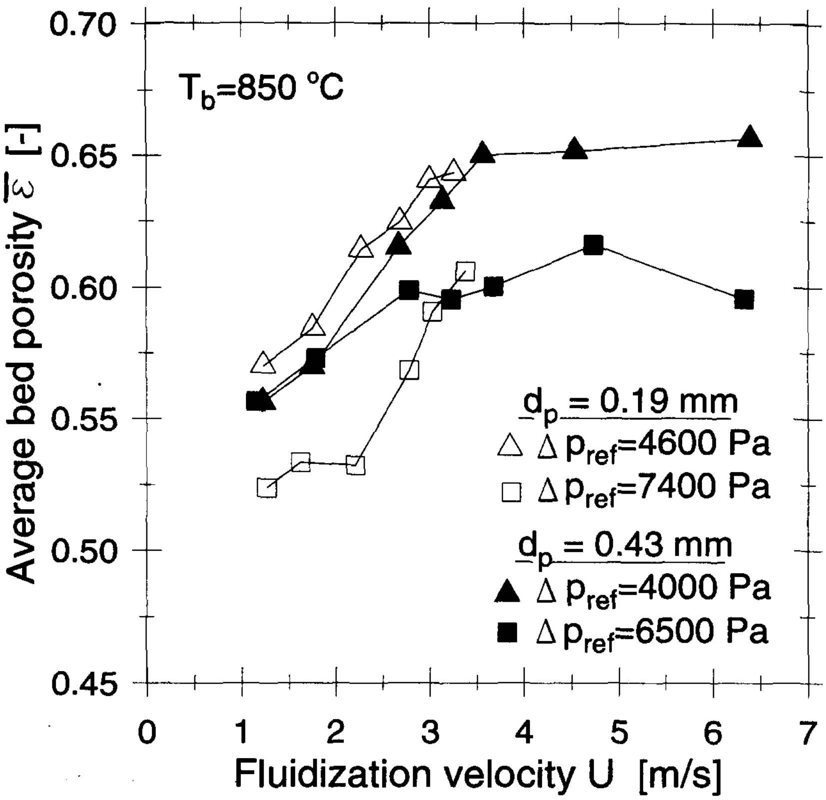


Figure 12. Average bed voidage vs fluidization velocity for two Δp_{ref} and two sizes of bed material at 850°C. The corresponding bed heights are shown in figure 10.

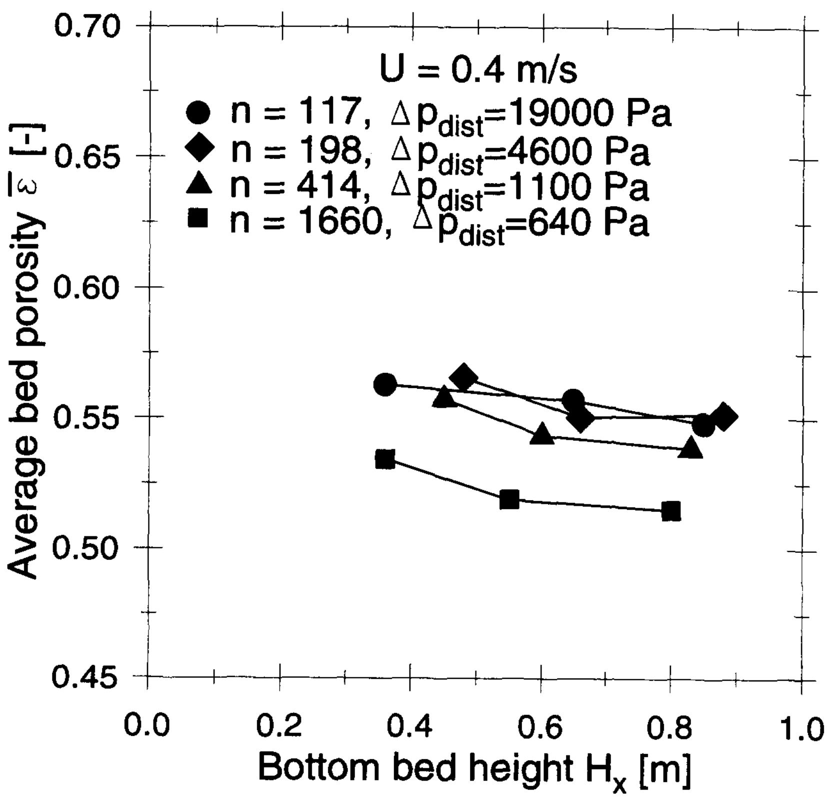


Figure 13. Average bed voidage as a function of bottom bed height for different numbers of holes in the air distributor. The measurements were made in the cold CFB (T_b = 40°C); d_p = 0.32 mm.

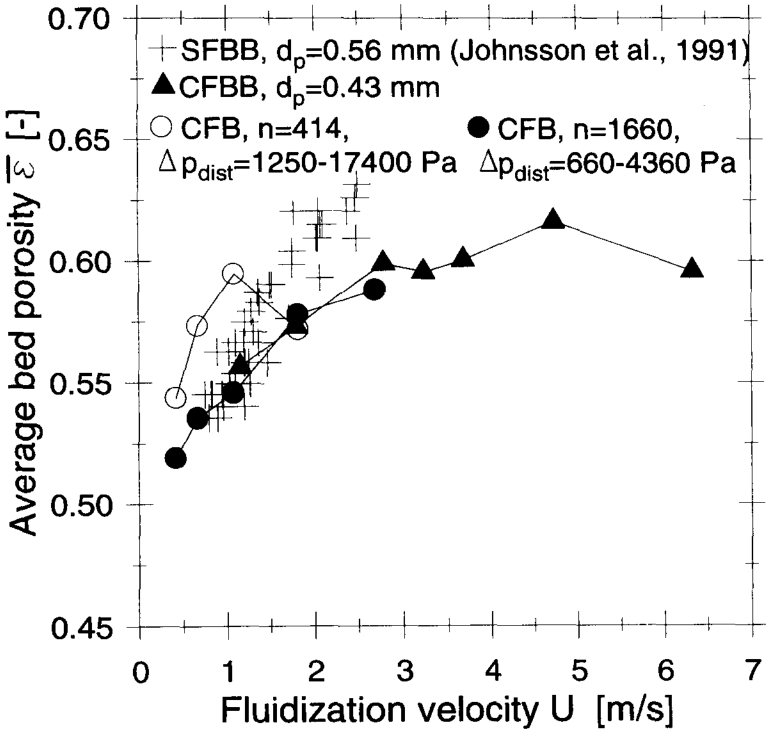


Figure 14. Average bed voidage vs fluidization velocity for the CFB boiler and a SFB boiler at 850°C. The boiler results are compared with the cold CFB where two different numbers of holes in the air distributor were investigated; d_p was 0.32 mm in the cold CFB.

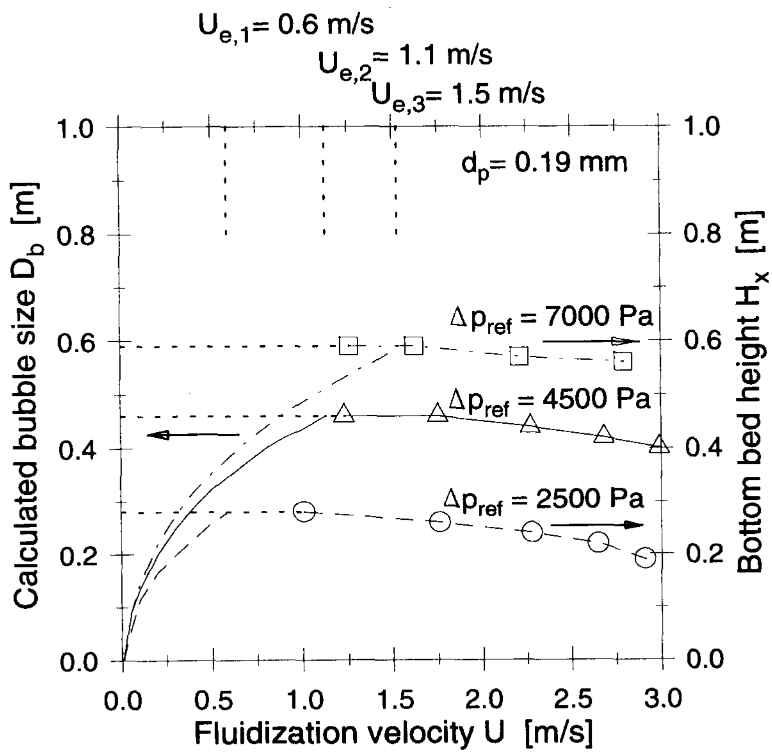


Figure 15. Bubble size calculated for the model by Choi *et al.* (curves) compared with the measured bed height for reference pressure drops similar to those in figure 7 (symbols), d_p = 0.19 mm. The transition to exploding bubbles U_e, when $D_b = H_x$ is indicated.

phase in a bubbling bed. He studied two sizes of bed material, 0.12 mm (ash), 0.24 mm (sand) and found that the particle size did not influence the voidage of the dense phase. Another observation was that the voidage of the dense phase increased slightly, about 5%, when the fluidization velocity was increased. Wein's results infer that the change in the average bed voidage does not originate from the dense phase; it is mainly an effect of a change in the visible bubble flow, which is a result of a different through-flow of gas through the bed.

The decrease in visible bubble flow, i.e. in bed voidage, coincides with an increase in mean bubble size when the bed height, H_x, increases or the distributor pressure drop, Δp_{dist}, decreases. Bubbles grow with height and the mean bubble size in a high bed is larger than in a low bed. Also, a low Δp_{dist} produces larger bubbles than a high Δp_{dist} (Geldart & Kelsey, 1968; Svensson *et al.* 1996). The conclusion is that larger bubbles promote through-flow of gas which reduces average bed voidage.

Transition from the single bubble regime to the exploding bubble regime

Although the transition from the single bubble regime to the exploding bubble regime is gradual and takes place over a range of velocities, as seen in figure 5, one single velocity, U_e, is introduced here to determine the influence of the operating conditions on the transition. Two methods for defining U_e are discussed.

The first method determines U_e from a comparison of a calculated bubble size, D_b, and the measured bed height, H_x, U_e is the velocity at which D_b equals H_x. Several models for the mean bubble size have been derived, e.g. by Mori and Wen (1975); Darton *et al.* (1977); Werther (1978); Choi *et al.* (1988). Here, the model

$$(U - U_{mf})(D_b - D_{b0}) + 0.474\, g^{1/2}(D_b^{3/2} - D_{b0}^{3/2}) = 1.132(U - U_{mf})h \qquad [1]$$

by Choi *et al.* is chosen for the comparison, since it was validated for data from a bed with a large cross-section, 0.6 m × 0.6 m, and for fluidization velocities up to 2 m/s ($d_p = 0.92$ to 1.2 mm). In agreement with Choi *et al.*, visual observations in the cold bed indicate that the most important region of bubble growth is just above the air distributor. For the conditions employed the bubble size almost immediately becomes of the same order as the bed height. This effect of bubble growth will then include horizontal bubble movements just above the air distributor. Since little is known concerning the bubble growth, the initial bubble size, D_{b0}, is calculated from an expression derived by Darton *et al.* (1977)

$$D_{b0} = 1.63[(U - U_{mf})A_0/g^{1/2}]^{2/5} \qquad [2]$$

which was previously shown to give reasonable bubble sizes in a non-slugging bed of similar particles (Andersson *et al.* 1989). In the present application the value of D_{b0} will not greatly affect the result. Figure 15 shows the measured bed height, H_x and the calculated bubble size, D_b, as a function of fluidization velocity. The bubble size is calculated at a height above the air distributor, h, equal to the measured bed height, H_x. Below the lowest operating velocity used, about 1.2 m/s, the bed height was assumed to be constant as indicated by the dashed horizontal lines in figure 15. The velocity U_e, at which the bubble size equals the bed height is rather low, about 1 m/s, but increases with bed height.

The second method assumes that the transition takes place at the velocity for which the two peaks in the power spectra have the same power (the area below the curve), see figure 5b. The power of the high frequency peak, representing single bubbles, falls and the low frequency peak, representing exploding bubbles, rises as the gas velocity is increased (figure 5). In figure 5b the areas below the two peaks are similar, indicating that the two bubble types appear to the same extent at this velocity, which is then the transition velocity, i.e. 1.8 m/s for the 0.43 mm particles. The validity of this method is supported by the increase in U_e with bed height, as illustrated by figures 16 and 17 in which the difference in area below the two peaks is reduced as the bed height increases, which implies that U_e is approached. A comparison with figure 16c shows that under the same operational conditions, but with 0.19 mm particles, the transition occurs at 1.2 m/s. Consequently, the transition velocity, U_e, is higher for coarse particles than for fine particles.

The velocities U_e as obtained by the two criteria proposed, from $D_b = H_x$ and from spectra, are summarized in table 3. The table shows that the velocity U_e is between 0.5 and 2 m/s, which agrees

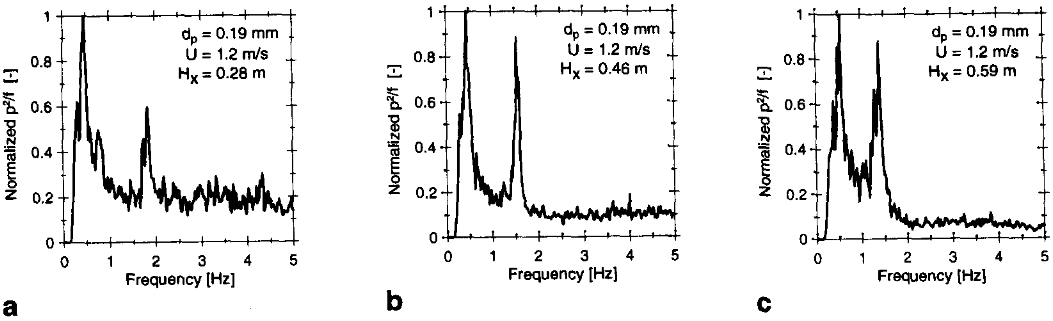


Figure 16. Frequency spectra for a pressure signal measured within the bottom bed (h = 0.1 m) at the fluidization velocity 1.2 m/s for different bed heights; d_p = 0.19 mm.

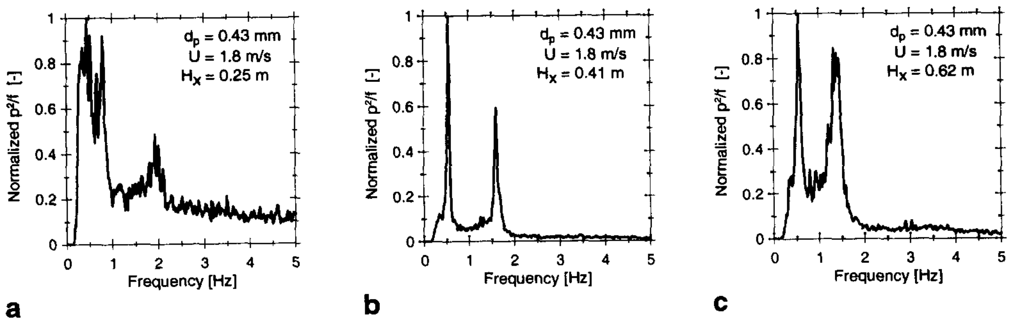


Figure 17. Frequency spectra for a pressure signal measured within the bottom bed (h = 0.1 m) at the fluidization velocity 1.8 m/s for different bed heights; d_p = 0.43 mm.

Table 3. Velocity (U_e) determined with two methods

	Transition velocity, U_e, [m/s]					
Particle size:	d_p = 0.19 mm U_t = 1.0 m/s		d_p = 0.32 mm U_t = 2.1 m/s		d_p = 0.43 mm U_t = 3.3 m/s	
Criteria: H_x (m)	$D_b = H_x$	From spectra	$D_b = H_x$	From spectra	$D_b = H_x$	From spectra
0.60–0.65			1.6–1.7	1.5–1.8	1.6–1.8	1.5–1.8
0.55–0.60	1.4–1.6	1.1	1.4–1.6	1.7		
0.45–0.50	1.1–1.25	1.1	1.1–1.3	1.7		
0.40–0.45			1.0–1.1	1.7	1.0–1.2	1.7
0.35–0.40			0.8–1.0	1.8		
0.25–0.30	0.5–0.7	0.8	0.5–0.7	<1.3	0.6–0.8	1.6

with results for large particles by Fitzgerald (1980) who used changes in amplitude of pressure fluctuations to determine the transition velocity. (The method used by Fitzgerald *et al.* is not applicable in this work, since the results presented here showed in all cases an increase in the amplitude of pressure fluctuations at low gas velocities, see figures 8 and 9.) The two methods used here reflect the predicted behaviour: an increase in U_e with increased bed height and particle size. Both methods are limited in accuracy: the first method because the models for bubble size are validated only for considerably lower velocities than the terminal velocity of an average-size bed particle, and the second method because the velocity at which the area of the peaks of the power spectra are equal is difficult to determine exactly.

5. CONCLUSIONS

The character of fluidization in the bottom bed of the 12 MW_{th} CFB boiler shows a gradual transition from a single bubble regime to an exploding bubble regime as the gas velocity rises from a low fluidization velocity (corresponding to a typical SFB boiler condition) to a typical high velocity CFB boiler condition. The single bubble regime is characterized by large single bubbles, which give a discontinuous gas flow, because when a single bubble erupts at the surface of the bed it takes some time before a new bubble is formed. A similar discontinuous gas flow was observed for the exploding bubble regime. The exploding bubbles can be described as large irregular voids, which often stretch from the air distributor to the surface of the bottom bed.

As a basis for an analysis of the influence of various boiler operating parameters on the transition, two methods for the determination of a transition velocity U_e are discussed. One method defines U_e as the velocity at which the maximum bubble size equals the height of the bottom bed. In the other method, U_e is equal to the velocity at which the power of the two dominant peaks of the frequency spectrum have the same magnitude. The transition velocity U_e, as defined by both methods, ranges from 0.5 to 2 m/s, and increases with bed height and particle size.

The average bed voidage in the CFB boiler was compared to the bed voidage in a 16 MW_{th} SFB boiler: the bed voidages were similar, in spite of the considerably higher gas velocities in the CFB boiler. The bed voidage in the CFB increased, as a result of an increased visual bubble flow, up to about 3 m/s. Above this velocity, in the exploding bubble regime, the bed voidage became almost independent of gas velocity. This is a consequence of the through-flow of all the additional gas through the bubbles, and because the visual bubble flow remains little affected.

At a constant gas velocity the bed voidage increased as a result of a rise in the visual bubble flow, caused by reduction of particle size or bed height, or with an increase in air-distributor pressure drop. In addition, the bed voidage was slightly greater for a 850°C bed than for a 40°C bed. At a constant gas velocity and within the ranges studied, these parameters affect the average bed voidage by less than 20%.

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