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Axial Solids Distribution and Bottom Bed Dynamics for Circulating Fluidized Bed Combustor Application

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In this study simple models are used to describe both the axial solids distribution of Circulating Fluidized Bed (CFB) risers and the dynamic characteristics of the bubbles and "slugs" present in the bottom bed Model results are compared with experimental data obtained in CFB risers of different size, all operated with 0.30 mm silica sand (i.e Group B solids) at gas velocities typical for combustors (U_g= 2÷4 m/s). The wide risers (D_t> 0.4 m) have flow conditions representative for combustors, while the narrow risers have not. The bottom bed solids concentration is well described and the calculated bubble dynamics are in agreement with Kolmogorov entropy and frequency obtained from pressure fluctuations measured in the bottom bed of tile larger CFBs The comparison of experiments and models indicates that the major part of the gas (> 95 %) passes the bed other than as visible bubble flow. The decay constant a for the splash zone above the bottom bed increases significantly when riser size is increased. The decay constant a in narrow risers shows a strong dependency on riser solids holdup (and thus on solids mass flux) at constant superficial gas velocity

Circulating Fluidized Bed (CFB) combustors are typically operated with Group B solids at superficial gas velocities which exceed the terminal velocity of an averaged sized bed particle but are not much higher than the terminal velocity of the largest size fraction of the particles. At these conditions there is considerable elutriation of solids, although solids mass fluxes are typically below 40 kg/m^2/s and much lower than common in FCC riser applications. Generally, CFB combustors are operated with a bottom bed with a height typically less than 0.5 m.

The dynamics of the bottom bed in wide risers (D_t>0.4 m) are of the exploding bubble type. Bubbles are large, compared to bubbles in the bubbling bed regime at low superficial gas velocities and explode vigorously at the bed surface, throwing solids into the splash zone. Since the bubbles are limited in size by the height of the bottom bed they are smaller than the diameter (or width) of the facility [1], *i.e.* slugging does not occur. A large amount of gas by-passes the bed, because when an exploding bubble has grown to its final size, it may reach from the gas distributor to the bed surface and a significant gas flow through the bubbles occurs. Hence, the bottom bed voidage is low with a value of about 0.6. Exploding bubbles do not have a well-defined shape as the bubbles in the bubbling bed at low superficial gas velocities, but can be seen as large voids of gas.

The dynamics of the bottom bed in narrow risers (D_t<0.4 m) operated with a bottom bed at relatively high gas velocities (*i.e.* circulating conditions) are of the "slugging" type: the size of the voids is limited by the size of the cross-section of the facility. A similar phenomenon is well known for non-circulating conditions, for which wall limitations causes a bed to develop slugs, while the wide bed is still bubbling for similar bed solids holdup and superficial gas velocity. For circulating conditions in risers smaller than 0.4 m ID, Arena *et al.* [2] observed typical "plug-slug" structures in the presence of a bottom bed. These "plug-slug" structures can be considered as a result of wall-limitations and were found to have a high rise velocity (> 5 m/s). The gas "slugs" push the solids "plugs" with high velocity through the riser. In the transition zone the plugs loose solids to the wall zone, break up and finally the "plug-slugs" cannot be observed anymore in the transport or dilute zone. The dynamics of the bottom bed in narrow risers is significantly different from the dynamics of the dense bottom bed in wide risers, as shown by Zijerveld *et al.* [3] and the bottom bed voidage is about 0.8. These flow conditions in the narrow risers are not representative for combustors.

Both in wide and narrow risers a splash or transition zone characterized by strong solids backmixing is present above the bottom bed. The solids concentration in this zone may be described with an exponential decay (*e.g.* Kunii and Levenspiel, [4]; Johnsson and Leckner [5]). Although the axial solids profile in both narrow and wide risers may be characterized in a similar way (*i.e.* by a bottom bed and an exponential decay), they are different from a *dynamic* point of view [3], so phenomena in the transition or splash zone are not necessarily similar. Above the splash zone a transport zone may be present, which is characterized by solids backmixing mainly at the riser wall. This zone is included in the model of Johnsson and Leckner [5]. The solids concentration may be described with an exponential decay, with a lower value of the decay constant than in the splash zone. Kunii and Levenspiel [4] model this zone with a constant solids concentration in case of low superficial gas velocities ("bubbling or turbulent fluidized bed") or this zone lacks in case of high superficial gas velocities ("fast fluidized bed"). This difference between the models of Kunii and Levenspiel [4] and Johnsson and Leckner [5] causes that the decay constant for the transition or splash zone in CFBs may have different values, depending on the model used to estimate it, although in both models this decay constant is indicated with the same symbol a.

The observation that the bottom bed in wide combustors is of a bubbling type encourages the use of a conventional bubbling bed model [6,7] extended with a description of the splash and transport zone [5] to model the hydrodynamics of CFB combustors. Here the model of Werther and Wein [7] is included for the bottom bed modelling, since they showed that void sizes and velocities found experimentally under conditions (U_g and Group B solids) typical for combustors, were in agreement with calculated bubble diameters and rise velocities.

The present work investigates whether it is possible to use existing models [e.g. 4,5,7] to predict the axial solids distribution of CFBs of different size operated at typical combustor gas velocities with the same Group B solids. Furthermore, it is investigated whether the bubble characteristics of the bottom bed predicted by the model are consistent with dynamical measurements carried out in the larger CFBs.

EXPERIMENTS

Experimental facilities

Table 1 shows the main characteristics of the five CFB facilities used. The 12 MW_{th} boiler was operated at 850 °C and the other four facilities were operated under ambient conditions. The 1.2x0.8 m facility is equipped with water manometers and the other facilities are equipped with pressure transducers along the riser height to obtain the axial pressure profile.

Experimental methods

The average axial voidage profile is calculated from the axial pressure profile neglecting friction effects and acceleration of the gas-solids suspension. From voidage and pressure profiles the bottom bed can be detected and its voidage determined. For all risers the solids holdup is directly obtained from the pressure profile; the pressure drop between the lowest pressure tap in the riser and the distributor is estimated by means of an extrapolation of the pressure profile assuming constant voidage in the bottom bed.

Measurements of pressure fluctuations were carried out at the wall in the bottom zone of two of the risers: at 0.35 m in the centre of the wall of the 12 MW_{th} boiler and at 0.28 m in the centre of the 1.2 m wall of the 1.2x0.8 m riser. The fluctuations of the pressure with respect to its local average were measured with piezoelectric transducers (Kistler, type 7261).

Average and peak frequency of the pressure fluctuations are determined by means of spectral analysis. The cycle frequency is calculated as the number of times that the pressure signal crosses its average value per time unit divided by two. Kolmogorov entropy is estimated by means of non-linear analysis. It is a measure for the predictability and is large for very irregular dynamics (*e.g.* turbulence) and small for more regular dynamics (*e.g.* slugging). The method used to estimate the Kolmogorov entropy from measured time series is described elsewhere [8].

Table 1 Characteristics of CFB facilities

Facility	Location	Distributor	Cross-section Size [m]	Cross-section Shape	Height[m]
12 MW_{th} boiler	Chalmers	144 bubble caps	1.47x1.42	Rectangular	13.5
1.2x0.8 m CFB	Cerchar	54 bubble caps	1.20x0.80	Rectangular	9.0
0.70 x 0.12 m CFB	Chalmers	6000 holes	0.70x0.12	Rectangular	8.5
0.12 m ID CFB	Naples	various	0.12 ID	Circular	5.8
0.083 m ID CFB	Delft	porous plate	0.083 ID	Circular	4.0

Operating conditions

The silica sand used is characterized by d_p=0.30 mm, U_t=2.5 m/s (25 °C) and U_t=1.9 m/s (850 °C). Each run is characterized by superficial gas velocity (U_g) and riser solids holdup, the latter quantified by riser pressure drop (ΔP_{riser}). Superficial gas velocity ranges from below to beyond the terminal velocity and a bottom bed could always be identified.

MODEL

Table 2 shows the model equations used in this work. The bottom bed is modelled with equations (1) to (9), according to the model of Werther and Wein [7]. This part of the model calculates the bottom bed solids concentration $c_v(Z)$ and the bubble diameter $d_v(Z)$ at height Z above the gas distributor.

Equation (11) assumes the Kolmogorov entropy to be linearly proportional to the total number of bubbles that erupt at the bed surface per unit of time, $N_{bubble}(H_x)$, cf.[9]. Equation (12) assumes the bubbles to be spherical and equally sized. This part of the model takes the bottom bed dynamics into account.

The part of the model which calculates $c_v(Z)$ and the bubble diameter $d_v(Z)$ at height Z above the gas distributor is solved simultaneously with the part which calculates the bottom bed dynamics (N_{bubble} and K_{ML}). The proportionality constant ξ (the bubble impact factor [9]) in equation (11b) is systematically fitted with one value for all operating conditions and facilities.

The riser solids distribution above the bed is modelled according to *i*) Johnsson and Leckner [5], assuming a decay constant *a* for the splash zone and a decay constant *K* for the transport zone (equations (13) and (14)) and *ii*) Kunii and Levenspiel [4], assuming one decay constant *a* for the entire riser above the bottom bed (equation (15)). The riser pressure drop ΔP_{riser} can be obtained from an integration of equations (1) and (13) or (1) and (15).

Table 2 Model equations

$$c_{v_x} = (1-\varepsilon_b) c_{v_d} \qquad Z<H_x \tag{1}$$

$$\frac{c_{v_d}}{c_{v_{mf}}} = 1 - 0.14 N_{Re_p}^{0.4} N_{Ar}^{0.13} \tag{2}$$

$$\varepsilon_b = \frac{\varphi(u_g - u_{mf})}{u_b} \tag{3}$$

$$u_b = \varphi(u_g - u_{mf}) + 0.71\,\vartheta\sqrt{g d_v} \tag{4}$$

$$\varphi = 1.45 N_{Ar}^{0.18} \qquad 10^2 < N_{Ar} < 10^4 \tag{5}$$

$$\vartheta = \begin{cases} 0.63 & d_t < 0.1m \\ 2.0\sqrt{d_t} & 0.1m \le d_t \le 1.0m \\ 2.0 & d_t > 1.0m \end{cases} \tag{6}$$

$$\frac{d(d_v)}{dZ} = \left(\frac{2\varepsilon_b}{9\pi}\right)^{\frac{1}{3}} \tag{7}$$

$$d_{v0} = \gamma \frac{(u_g - u_{mf})^{0.4}}{g^{0.2}} \tag{8}$$

$$\gamma = 1.3 \left(\frac{A_{bed}}{N_{nozzle}}\right)^{0.4} \tag{9}$$

$$u_{slug} = \varphi(u_g - u_{mf}) + 0.35\sqrt{g d_t} \tag{10}$$

$$K_{ML} \propto \frac{\#\ bubble\ eruptions}{s} = N_{bubble} \tag{11a}$$

$$K_{ML} = \xi\ N_{bubble} \tag{11b}$$

$$N_{bubble}(H_x) = \frac{\varphi(u_g - u_{mf}) A_{bed}}{\frac{\pi}{6} d_v^3(H_x)} \tag{12}$$

$$c_v = (c_{v_x} - c_{v_{2,x}}) e^{-a(Z-H_x)} + c_{v_{exit}} e^{K(H_{exit}-Z)} \qquad H_x<Z<H_{exit} \tag{13}$$

$$c_{v_{2,x}} = c_{v_{exit}} e^{K(H_{exit}-H_x)} \tag{14}$$

$$c_v = (c_{v_x} - c_v^*) e^{-a(Z-H_x)} + c_v^* \qquad H_x<Z<H_{exit} \tag{15}$$

RESULTS AND DISCUSSION

The upper part of Table 3 shows experimental and calculated c_{vx} for some characteristic runs carried out in the 1.2x0.8 m riser. Calculations are carried out with the φ and γ as suggested by Werther and Wein [7]. Clearly can be seen that the model with the values of φ and γ given by equations (5) and (9) is not able to describe the experimental c_{vx}. Similarly c_{vx} for the 12 MW_{th} boiler is not accurately described, which is not reported here. These discrepancies may exist, since the values for these parameters used were obtained for specific solids type and operating conditions. Therefore in this work two adjustments have been made. *i*) The value of the fraction of total gas flow in visible bubble flow, φ, has been lowered considerably. One value of φ is estimated in a systematic way for all operating conditions (U_g and H_x) and facilities. *ii*) The value of the

Table 3 Predicted and experimental bottom bed characteristics

	U_g	ΔP_{dist}	Model φ	γ	d_{v0}	c_{vd}	$c_{vx}(H_x)$	Exper c_{vx}
This work	2.0	2.6	0.344	0.259	0.212	0.43	0.34	0.45
param. W&W [5]	3.2	4.0	0.344	0.259	0.258	0.41	0.29	0.44
1.2x0.8 m CFB	4.0	5.3	0.344	0.259	0.283	0.39	0.26	0.42
This work	2.0	2.6	0.0175	0.325	0.266	0.43	0.42	0.45
param. W&W [5] adapted	3.2	4.0	0.0175	0.325	0.324	0.41	0.40	0.44
1.2x0.8 m CFB	4.0	5.3	0.0175	0.325	0.355	0.39	0.38	0.42
This work	2.0	1.3	0.0175	0.386	0.343	0.48	0.47	0.42
	3.2	2.4	0.0175	0.386	0.417	0.46	0.45	0.40
12 MW_{th} boiler	4.0	3.9	0.0175	0.386	0.457	0.45	0.44	0.39
This work	4.0	-	0.0175	★	★	0.39	0.32	0.32
0.083 m ID CFB	5.0	-	0.0175	★	★	0.38	0.30	0.23

★ Not applicable here, eq (10) is used instead of eq. (4)

constant γ in equation (8) as calculated with equation (9) (γ=0.259) has been modified in a systematic way. For each gas distributor (thus each facility) one value of γ is estimated for all operating conditions (U_g and H_x) to meet the experimental c_{vx} *and* bubble dynamics. Thus the characteristics of each different gas distributor with respect to the initial bubble formation is accounted for.

Bubble dynamics are in agreement with the experiments only with low values of the visible bubble flow; φ=0.0175 gives good results. They can be expected at these high gas velocities although it should be pointed out that the low values of φ obtained should be seen as an indication of high through-flow of gas rather than exact values. Bubble diameters grow to the order of magnitude of the bottom bed height, which favours the temporally channelling or short cut of gas [6]. This is in agreement with visual observations of the bottom bed in the 1.2x0.8 m riser, which contains bubbles which have a considerable upward directed flow of solids.

For the 1.2x0.8 m riser it is found that γ=0.325 gives good results for all operating conditions (U_g and H_x). At the high gas velocities of this work it may be doubted that the assumption underlying equations (8) and (9) (*i.e* each bubble cap acts as an individual bubble generator) is still valid. More than one bubble cap will contribute to an initial bubble formed. However, provided that γ is adjusted to meet the gas distributor characteristics, equation (8) may still be used to account for the dependency of the initial bubble size on the excess gas velocity U_g-U_{mf}.

A c_{vx} is calculated for the 12 MW_{th} boiler with the same φ as used for the 1.2x0.8 m facility, see table 3. Good model results for several superficial gas velocities are obtained when γ=0.386. The 12 MW_{th} boiler has more bubble caps per unit bed surface (see table 1) and a lower gas distributor pressure drop and hence intial bubble formation is different. For the gas distributor of the 12 MW_{th} boiler larger bubbles are calculated than for the 1.2x0.8 m riser at similar superficial gas velocities. This is in agreement with experimental observations at circulating conditions made in previous work [3].

Similarly, the bottom bed concentrations was modelled for the 0.083 m ID riser. Initial bubbles in this riser are of the same size as the riser diameter, so the slug rise velocity (equation (10)) has been used instead of the bubble rise velocity (equation (4)). Bottom bed concentrations are calculated which are considerably lower than those obtained in both large size facilities, which is in line with the trend in the experimental data, see table 3. Equation (10) is not capable to describe slug rise velocities at high superficial gas velocities and therefore we may expect the difference between the calculated and experimental bottom bed concentration at 5 m/s.

Figure 1 shows the calculated and the experimentally determined Kolmogorov entropy for the 1.2x0.8 m riser and 12 MW_{th} boiler for runs at different superficial gas velocities. The proportionality constant ξ in equation (11b) was optimized for all operating conditions and facilities to be 8.9. Model results of Kolmogorov entropy are in agreement with experimental results for the different superficial gas velocities, different riser solids holdups (*i.e.* bottom bed heights H_x) and the two facilities. The N_{bubble} is in between the peak frequency in the spectrum and the cycle frequency.

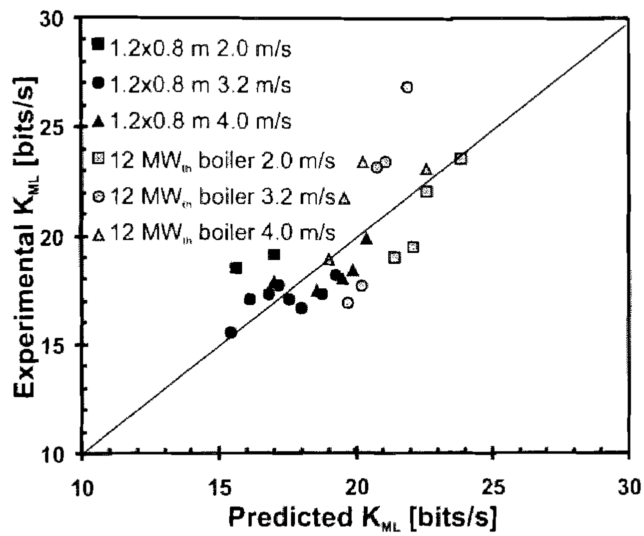

Figure 1 Experimental K_{ML} as a function of the predicted K_{ML}.

Model results of the dynamics of the bottom bed of the small size risers could not be made, because an adequate slug length relation at these high gas velocities is lacking.

Figure 2 shows the decay constant a obtained with equation (13) from experimental voidage profiles of the five CFB facilities in this work, see table 1. The decay constant a decreases when the gas velocity increases, in agreement with previous observations [4,5]. The decay constant a increases when the width or diameter of the riser increases, which is in contradiction with previous observations [4]. Furthermore, the decay constant a decreases significantly with the riser pressure drop (and thus with solids mass flux [10]) at constant U_g in the 0.083 and 0.12 m ID risers. For the 0.083 m ID riser a is calculated both with equation (13) and (15). The decay constant a calculated with equation (13) is higher than the one calculated with equation (15) for low riser pressure drop and the decrease of the decay constant a with an increase of riser pressure drop is found for both equation (13) and (15). For the 12 MW_{th} boiler and the 1.2x0.8 m riser equation (15) gives a poor fit of the axial solids profile above the bottom bed, since the transport zone is significantly present in these risers [1]. Therefore, it is preferred not to use equation (15) for these large risers.

CONCLUDING REMARKS

The axial solids distribution in large size risers and the dynamic characteristics of the bubbles present in the bottom bed can be described with simple models. The model simulations indicate that the major part (>95 %) of the gas passes the bed other than as visible bubble flow. The gas distributor plays an important role in initial bubble formation, which needs to be investigated in more detail. The decay constant a for the splash zone is significantly dependent on riser solids holdup in narrow risers and increases significantly with riser width.

ACKNOWLEDGEMENT

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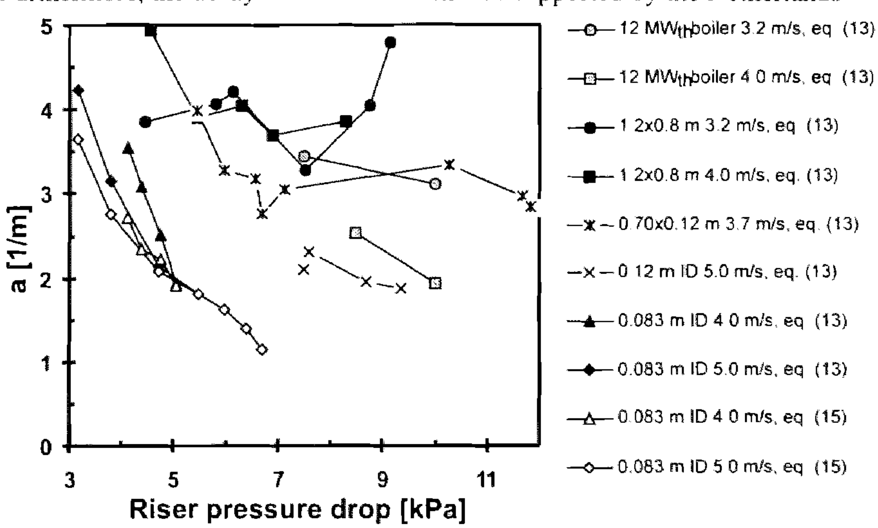

Figure 2 Decay constant a as a function of ΔP_{riser}.

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NOTATION

a	decay constant, 1/m
A_{bed}	bed surface, m^2
c_s	solids volume concentration, -
c_{s2}	solids volume conc. from dispersed phase, -
c_{sd}	solids volume concentration dense phase, -
c_{sexit}	solids volume concentration at exit, -
c_{smf}	solids volume concentration at U_{mf}, -
c_{sb}	solids vol. concentration bottom bed, -
c_s^*	solids volume concentration at saturated carrying capacity, -
d_v	bubble volume equivalent diameter, m
d_{v0}	initial d_v, m
d_p	solids mean diameter, m
d_t	equivalent bed diameter, m
g	gravity acceleration, m/s^2
G_s	solids mass flux, kg/(m^2 s)
H_{riser}	riser height, m
H_b	bottom bed height, m
K	decay constant transport zone, 1/m
K_{ML}	Kolmogorov entropy, bits/s
N_{Ar}	Archimedes number, -
N_{bubble}	number of bubble eruptions, 1/s
N_{nozzle}	# of distributor orifices, -
N_{Re}	Reynolds number $=(U_g-U_{mf})d_p\rho_g\mu_g$, -
U_b	bubble rise velocity, m/s
U_g	superficial gas velocity, m/s
U_{mf}	minimum fluidization velocity, m/s
U_{slug}	slug rise velocity, m/s
U_t	terminal velocity, m/s
Z	height above the dist. plate, m

Greek symbols

ΔP_{dist}	distributor pressure drop, kPa
ΔP_{riser}	riser pressure drop, kPa
ϵ_b	bubble volume fraction, -
γ	parameter defined in eq. 9, -
μ_g	gas viscosity, $N*s/m^3$
ξ	parameter defined in eq. 11, -
ϑ	parameter defined in eq. 6, -
ρ_g	gas density, kg/m^3
ρ_s	solids density, kg/m^3
φ	parameter defined in eq. 5, -

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