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Fluidization Characteristics of Circulating Fluidized Bed Boilers

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Dedicated to Prof. Dr.-Ing. Joachim Werther on the occasion of his 80th birthday

The fluidization in the furnace of circulating fluidized bed (CFB) boilers is described. Data have been obtained from a 12 MW_{th} CFB boiler and from literature. The bottom bed is bubbling rather than fast or turbulently fluidized as is often assumed. There is an extended splash zone above this bed, in which particles are thrown up by the bubbles and fall back onto the bottom bed. Other particles are carried further into the transport zone, initially as a saturated gas flow, but some particles move into the wall layers and the density decays towards the exit. The density of the transport zone is low, and the circulation rate is relatively seen small. Therefore, this flow is in a state of dilute-phase transport and not in the fast fluidization regime, such as is often claimed.

Keywords: Boiler, Circulating fluidized bed, Combustor, Fluidization, Fluidization regimes

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1 Introduction

This is a description of important fluidization characteristics of circulating fluidized bed (CFB) boilers. It relies on, discusses, and completes previous publications on the same topic, Leckner [1, 2]. The focus is on the bottom bed, its fluidization state, and on the fluidization state of the bed in the upper riser.

There is a problem in obtaining information form CFB boilers because of their size. Examples of dimensions are presented in Fig. 1. Furthermore, the combustion chambers are bounded by membrane-tube walls, and the few available inspection holes are not sufficient for systematic measurements. Moreover, there is a danger for erosion if dedicated measurement holes are opened in the wall. Therefore, few investigations have been carried out on commercial furnaces and the information must be taken from theory or from measurements in laboratory risers. These investigations are generalized but rarely directly referring to the fluidization conditions prevailing in CFB furnaces.

The research CFB boiler at Chalmers University of Technology is an exception from these size limitations. A research boiler of sufficient dimensions for representative measurements is available, Fig. 2. The figure shows the boiler, its air-supply (22), and its flue-gas recirculation system (23). Hoppers for fuel and additives (20 and 17–19) are also seen. The circles on the furnace body are measurement holes. The device at the lower right is a cement pump for sewage sludge. During the years (now 32 years) many research projects have been carried out involving fluid dynamics, heat transfer, and combustion-related topics. The cement pump is now removed, and a gasification unit is added in connection with the loop seal. For the present interest in the fluidization regimes, it is important to know that there are many pressure taps installed with transducers for pressure recordings from the bottom to the top of the furnace on the furnace side opposite to that seen on Fig. 2. The boiler is 13.5 m high, its cross-section dimensions at the refractory-lined bottom part are about 1.47 × 1.42 m. Usually, silica sand (density 2600 $kg\,m^{-3}$) with a size distribution predominantly between 200 and 400 μm is used, diluted by fuel ashes. Fig. 3 gives an example of pressure-measurement results.

Fig. 3a shows pressures measured along the height of the riser during a typical full load run with coal (bed temperature 850 °C, air ratio 1.2) with all air entering from the bottom. The bed material is distributed over four regions: the bottom bed, the splash zone, the transport zone, and the exit zone. Fig. 3b shows the same data as in Fig. 3a, plotted in a semilogarithmic scale to illustrate the exponential character of the axial density distribution. The bottom bed is also seemingly logarithmic. However, this is in the low-argument linear range of the exponential, where it is uncertain

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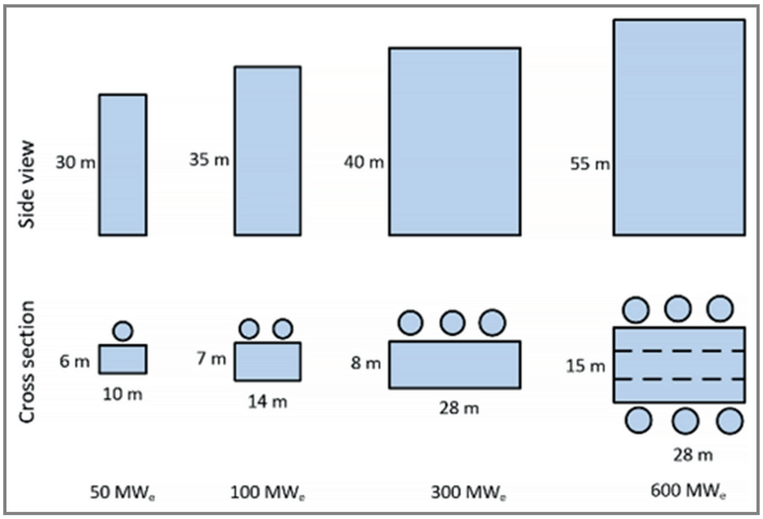


Figure 1. Typical CFB furnace dimensions.

to distinguish between a linear and an exponential curve, as seen by comparison of the two diagrams. The splash and transport regions have been described by exponential decays in the model of the axial density distribution by Johnsson and Leckner [3], and this agrees well with this profile as well as with profiles in other boilers, as shown by Djerf [4]. However. there are irregularities in the measurements, and complications can be seen, such as the effect of the T-exit of the boiler, accumulating bed material in the top above the exit (the exit configuration is seen in Fig. 2), as investigated by Johnsson et al. [5], expressing a "back-flow ratio" of particles.

The formation of a splash zone is related to the existence of bubbles, causing outbursts of particles when they reach the surface of the dense bed. In case the riser is operated without a bottom bed there are no bubbles. Then this splash zone disappears and is replaced by a smaller transition zone, as shown by Djerf [4].

Based on the vertical density distribution of Fig. 3, and if the conditions in the lower furnace (Region 1 and 2) do not change, only the height of the transport zone (Region 3) is extended to the height of a 43 m tall furnace, such as seen in Fig. 1, under the assumption of a rather commonly used total pressure drop of 12 kPa. This would result in the density distribution with height shown in Tab. 1. The data in Tab. 1 should be understood as a characteristic distribution of bed material in a furnace where the total quantity of bed material, the particle size distribution, the fluidization velocity, and the pressure drop across the height of the furnace are chosen to give a bottom bed height with its splash zone, and a riser particle concentration that is in the typical range for CFB boilers, cf. Johansson et al. [6] where some data are included.

2 The Distribution of Bed Material in a Circulating Fluidized Bed Riser

The control system of a CFB boiler acts to maintain the desired pressure drop across the furnace constant during operation under constant conditions: either bed material is

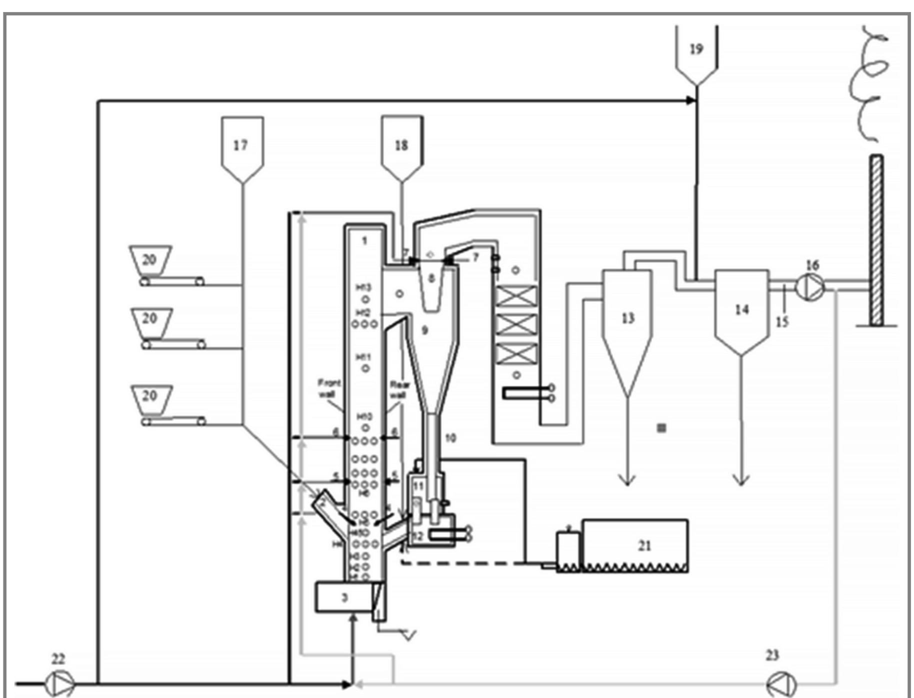


Figure 2. The Chalmers 12 MW CFB boiler with its fuel (20) and air (3–6) feed systems.

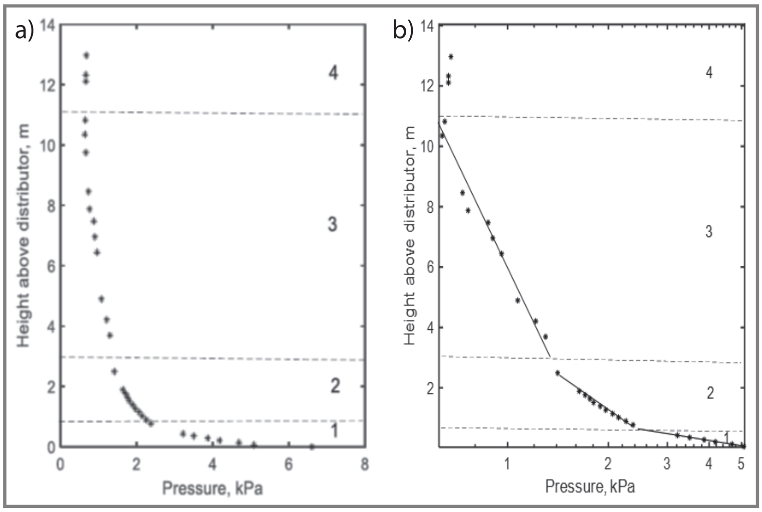


Figure 3. Pressure measurements along the height of the Chalmers CFB boiler. a) linear and b) semi-logarithmic scale. 1) Bottom bed, 2) splash zone, 3) transport zone, and 4) exit zone.

added, or it is withdrawn, to maintain the bed content. The control strategies for load changes depend on the operation strategy chosen. Here, let us assume that the amount of bed material in the boiler system, in the furnace plus the return system, is kept constant. Fig. 4 shows an example of the redistribution of bed material as a function of fluidization velocity under such conditions during operation with primary air only.

Fig. 4a illustrates how the bed material is redistributed in the furnace during increasing load, that is fluidization velocity. The bed height, defined here as the part of the dense bed having a constant pressure drop with height, dp/dh, where p is pressure and h height, decreases with velocity, since bed material is redistributed to the other regions of the circulation system at the same time as the splash zone increases because of the increased bubble activity in the bed. Gradually, material is entrained, and the portion of particles in the transport zone increases. Fig. 4b shows how the bed height decreases while the boiler load increases. The resulting bed height depends on the amount of bed material in the system $M_{\mathrm{tot},3} > M_{\mathrm{tot},2} > M_{\mathrm{tot},1}$. If the content of bed material is too small, then, at sufficiently high velocities, the bottom bed may disappear. Then, also the splash zone disappears. During operation of the boiler as a combustor, the bottom bed is needed, as it fulfils a significant role in accommodating the fresh fuel injected into the furnace, and its content is maintained constant by the boiler's control system, but in laboratory tests it appears from publications that at some occasions circulating systems are operated without a bottom bed without clearly accounting for it.

Table 1. The distribution of bed material in a 43 m tall furnace.

Part of bed	Region in Figure 3	Height	Pressure drop	Particle concentration	
		H [m]	[kPa]	$1-\varepsilon$	C [kg m^3]
Bottom bed	1	0.5	4	0.30	780
Splash zone	2	2	1	0.02	52
Transport zone	3 + 4	40	7	0.007	18
Total	–	43	12	–	–

3 The Bottom Zone

Fig. 3 shows a linear pressure drop in the bottom zone (Zone 1) with a height, typically up to between half a meter and a meter, characterising a dense fluidized bed, but so far it cannot be determined whether the bed is bubbling or turbulent.

Bubbles are formed at the distributor and increase in size, influenced by the fluidization velocity and bed height, as described by correlations [8, 9]. The bubble size changes in the bed due to coalescence, and it may be reduced due to bubble splitting. There are three options for the development of bubbles with increasing fluidization velocity, eventually leading to turbulent fluidization according to Bi [10]: 1) Bubbles in beds of Group A particles are significantly influenced by splitting and attain a maximum size. 2) Bubbles of both Group A and B also may attain a maximum size because of limitations in the size of the fluidization vessel, causing slugging, and with further increase in velocity, bubble splitting. 3) Bubbles of Group B in wide beds, having a low bed height compared to its width, are supposed to split because of the penetration of gas jets from the distribution nozzles like in a spouted bed.

The third option is that of present interest. There are no limiting walls, and the bubbles continue to grow with fluidization velocity without reaching any observable maximum bubble size, Glicksman et al. [11], unless the bubble size approaches the bed height, which then limits the size. The pressure fluctuation in the bed is related to the size of a bubble and so to the bed height, as shown by Fig. 5, taken from tests carried out by Svensson et al. [7].

Fig. 6a gives an example of a bubble at a fluidization velocity of about $2\,\mathrm{m\,s^{-1}}$ above U_{mf} at room temperature with sand, 0.79 mm of size, in a freely bubbling bed ($0.07\times0.68\times3.5$ m). The appearance of the bubble is far from the typical bubbles observed at lower velocity. It is quite irregular and extends across the height of the bed. It appears as if there were a jet penetrating the bubble. Is this a turbulent bubble because of its irregular form and the abundance of particles inside? No, the corresponding turbulent bed is seen in Fig. 6b, in a bed operating under the same condition as that of Fig. 6a, except

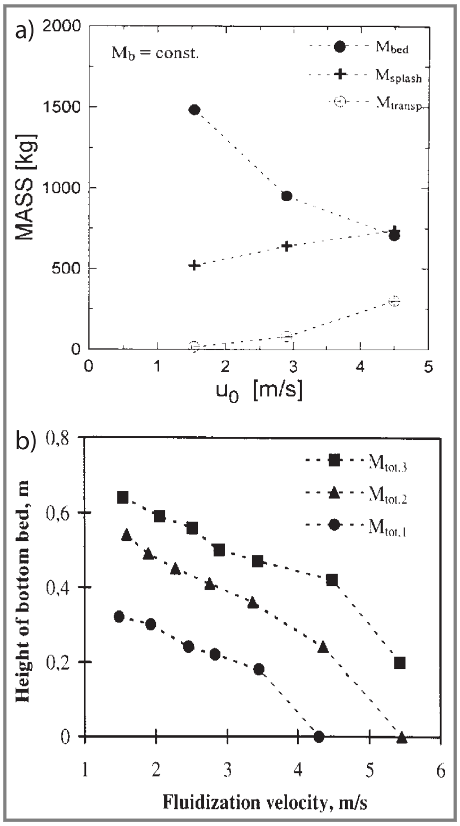


Figure 4. a). Distribution of bed mass in bottom bed, splash zone, and riser in the Chalmers boiler. The abscissa, fluidization velocity u_o, is proportional to load. b) The change of bottom bed height for three different initial charges of bed material in the boiler. Replotted from Svensson et al. [7].

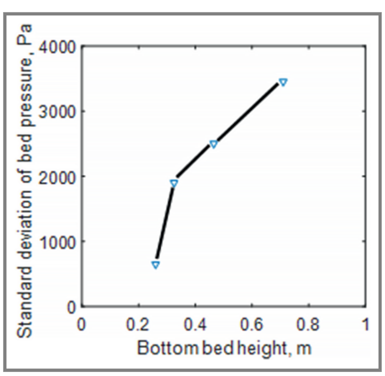


Figure 5. The standard deviation of the pressure fluctuations in a sand bed versus bed height at a fluidization velocity of about $2\,m\,s^{-1}$ and a particle size of 0.32 mm (sand). Replotted from Svensson et al. [7].

that it is provoked by disturbance elements in the bed in the form of tubes; the circular shapes seen in the figure. It does behave like most descriptions of a turbulent bed, for instance by Yerushalmi and Crancurt [12]. This picture is distinctly different from bubbling beds with tubes at lower velocities when the bubbles "peacefully" pass around the tubes. Also, the resemblance with a spouted bed is weak. Certainly, the bubble growth is affected by the inlet nozzles. However, unlike the nozzles in a spouted bed, these nozzles are usually bubble cap nozzles where the gas velocity is directed downwards before the gas is mixed in the bed and then turns upwards. The effect should be less severe than the that of a jet in a spouted bed. In any case, there is a strong through-flow of gas through the bubble, and the effect of this is maybe what is seen inside the bubble in Fig. 6a.

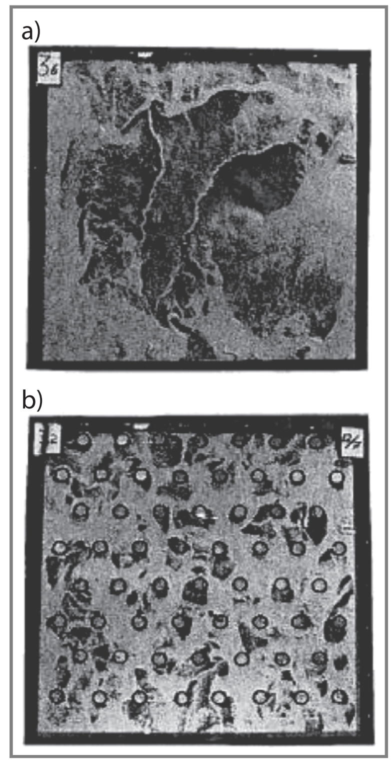


Figure 6. a) A bubble in a two-dimensional bed at $U - U_{mf} = 2\,m\,s^{-1}$, 0.79 mm sand at room temperature, Andersson et al. [13]. b) Same conditions as in Fig. 6a, but tubes are inserted (the circular disturbance elements are seen). Andersson et al. [13].

The size of a bubble is related to the pressure fluctuation in the bed. This feature is normally used to determine the onset to turbulent fluidization, which coincides with the maximum pressure fluctuation at a velocity, usually denoted U_c. At $U > U_c$, the bed approaches turbulent fluidization and the fluctuations caused by the bubbles become smaller. This was shown to be valid for the cases in Fig. 6, Andersson et al. [13], where it was a marked difference between the fluctuation behavior in Figs. 6a and b.

It was questioned [2] if turbulent fluidization takes place in a bottom bed of a commercial CFB operated at a fluidization velocity that may range between 2 and $6\,m\,s^{-1}$ at 850 °C. The doubt about a transition to turbulent fluidization is based on measurements in Chalmers 12 MW_{th} CFB boiler where no clear maximum pressure fluctuation was observed despite velocities of up to 5 m/s in the bottom bed. Because of the increase of pressure across the bottom nozzles with velocity, this is the maximum possible velocity in that boiler with only primary air.

The explanation to why the bed remained in the bubble regime at high velocity [2] was tentatively assumed to be the great throughflow of gas through the bubbles. This is perhaps what Bi [10] called jets, penetrating the bubble? Another explanation was provided by Wang et al. [14] who showed that the tendency of coarser bed material in the bottom made the bottom bed operate at $U < U_{tr}$ where U_{tr} is the transport velocity, although this is not a direct proof of a bubbling bed because $U_c < U_{tr}$ and a transition to turbulent fluidization could have taken place before U_{tr} is attained. A similar but more intuitive conclusion was forwarded by researchers from the Tsinghua University, Yue [15] who found that the bottom beds in Chinese CFB boilers tended to accumulate coarser bed material than that circulating and, thus, that it operates under bubbling bed conditions.

A further proof of a bubbling bottom bed was provided by Sekret and Nowak [16] who made a unique research installation in a 670 MW_{th} CFB boiler in mounting enough pressure-taps in the bottom region to be able to capture the height of the thin dense bottom bed and its fluctuations. They found a weak maximum fluctuation, indicating a turbulent bed. They called it a "bubbling-turbulent bed". However, the results can be interpreted as follows: the amplitude of the pressure fluctuation is related to the maximum bubble size and the height of the bed, as seen in Fig. 5. When the fluidization velocity increases, the measured bed height decreases (cf. Fig. 4). So, the maximum pressure fluctuation observed was a result of the increase of bubble size during the initial rise in velocity and the subsequent fall of the amplitude of the pressure fluctuation at higher velocities because of the decreasing bubble size following from the lower bed height at higher velocity. The argument is solid, while the bed height can be followed; it was measured by Sekret and Nowak.

Another of the few observations from CFB bottom beds were made by Werther and Wein [17] who measured a voidage of about 0.6 in the dense phase between the bubbles at velocities up to $3\,\mathrm{m\,s^{-1}}$ in a bed of 0.5×0.5 m cross section, operated at velocities of up to $4\,\mathrm{m\,s^{-1}}$. However, it was uncertain whether the bottom bed was bubbling or turbulent.

In conclusion, much evidence indicates that CFB boilers are operated with a bottom bed even when exposed to a high fluidization velocity, and this bottom bed is a bubbling bed with great through-flow of gas, forming a substantial splash zone above. The bubbles that could also be called voids, are irregular and do contain a through-flow of gas and particles, but they do not conform to the feature called "turbulent fluidization".

4 The Transport Zone

The splash zone is delimited from the transport zone by the Transport Disengaging Height (TDH) [18]. Above the TDH, particles are carried by the gas at its superficial velocity, presumably equal to the maximum or saturation dilute-phase particle-carrying capacity, G^*. Both concepts, TDH and G^*, are extremely important, but it has been difficult to determine them with great certainty. Cahyadi et al. [19] investigated 25 published correlations on TDH and found a deviation of several orders of magnitude between them. It can be assumed that the difficulty in finding well-defined correlations is because of the influence of particle-size distributions, which presumably yield several TDHs, each depending on its particle size fraction. In the Chalmers boiler, under the operation conditions shown on Fig. 3, the TDH is found in the intersection of the exponentials describing Regions 2 and 3, the splash and transport zones. The particle flow above this point could be characterized by the "saturation carrying capacity", namely the quantity of particles that the gas can carry upwards, expressed as a flux, particle flow per cross-section area of the riser, G^*. In a recent publication, Breault et al. [20] gives a definition of G^*: "the solids flux when a riser has a dense bed at the bottom and the solids flux into the riser equals the solids flux out of the riser". The latter definition does not apply to the conditions in the boiler in Fig. 3 where particles are removed continuously to the wall layers falling downwards as the gas-solids suspension ascends in the riser, and the concentration decays, as seen in Fig. 3. So, the flow grows under-saturated above the intersection between Regions 2 and 3. Nevertheless, G^* gives an indication on the state of the transport zone and in that sense, it can be treated further.

The saturation carrying capacity, influenced by the fluidization velocity, separates the fast fluidized regime on the low-velocity side of the saturation curve from dilute phase transport on the high-velocity side, such as has been determined in many investigations, for instance, by Yang [21]. In Ref. [2] a correlation of G^* from Bi and Fan [22] is used to compare with boiler data from the Tsinghua University, Yue et al. [23]. This diagram is shown in Fig. 7 where also data from Breault et al. [20] are inserted for comparison. The boiler data, called "Operation data", taken from [23], fall on the side of the dilute-phase flow and not in the fast-fluidized region as generally claimed. The recent survey of Breault et al. [20] for Group B particles shows that the correlation of Bi and Fan [22], based on data from several investigations with particles of Group A and risers of < 0.2 m diameter, performs relatively well also for Group B particles, but it ignores the size of the riser that could have an effect. The correlations are presumably carried out with data from room temperature just as most investigations of this kind. The diagram of Fig. 7 includes recalculated correlations for higher temperatures to illustrate the possibility of extrapolating to a typical furnace temperature.

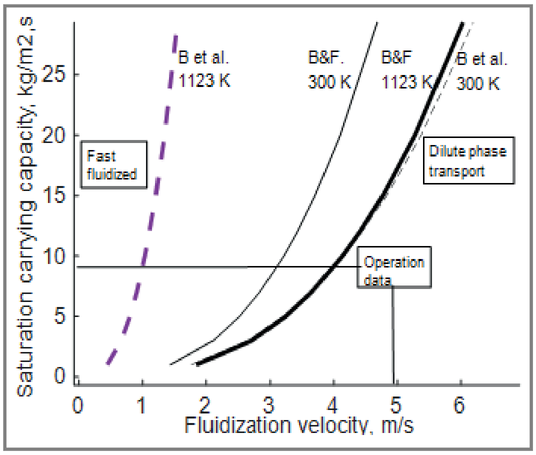


Figure 7. Saturation carrying velocity according to Bi and Fan [22] and Breault et al. [20], evaluated for 0.2 mm sand-like particles. The operation area is taken from the Tsinghua state diagram, Yue et al. [23].

The dependence on riser size was treated by Breault et al. [20], but unfortunately the diameter of the risers included is limited to 0.3 m, and it would be daring to extrapolate to 10 m to reach the size of a furnace. However, the trend shows decreasing G^* with increasing size of riser diameter between 0.1 and 0.3 m.

In Fig. 7 the correlation of Breault et al., "B et al." on the diagram, is compared to that of Bi and Fan, "B&F" at two temperatures: 300 K where most test results are obtained and 1123 K, which is an average furnace temperature. The success of the comparison is moderate. Particularly the temperature dependence of the data of Breault et al. is opposite that of Bi and Fan. It can be seen from the formula derived by Breault et al. that there is no gas density involved. Could that be an explanation for the behaviour seen in Fig. 7? Probably, Breault et al. [20] only had the intention to correlate data at room temperature.

There is some other information that can be used to assess the fluidization regime of the transport zone, namely the fluidization regime charts that have been established during the years: for instance, those of Li and Kwauk [24], Yerushalmi and Avidan [25], Squires et al. [26], Grace [28], Bai et al. [29], and Sun and Zhu [30]. In these regime charts, the boiler data according to Tab. 1 with an assumed circulation flux of 10–20 kg m^{-2}s^{-1} are outside of the range of fast fluidization and in the range of dilute phase transport, similar to the case in Fig. 7. Perhaps, the reason is that the diagrams are not as general as they pretend to be and are only valid for chemical reactors with Group A particles and at room temperature? However, this is usually not explained. Only in one case, that of Sun and Zhu, CFB combustors are mentioned, albeit with parameters outside of the ranges known for commercial CFB boilers. Finally, information might be available in the ranges of densities presented in certain publications, Tab. 2.

Also compared with the data of Tab. 2 the transport zone of Tab. 1 is outside of the fast fluidization regime, but the splash zone has a density that falls into that regime. The limit between fast fluidization and dilute phase transport is around 0.05, a number that agrees reasonably well with the regime charts and is well above the density in the upper part of the CFB riser presented in Tab. 1.

5 Discussion

Could an uneven air supply have influenced the bubble formation and the bubble behavior in the bed? The answer is "no" because the pressure drop across the bottom nozzles of the air distributor is always chosen high at full load, and then the air distribution over the surface of the distributor is even. This has to do with the design of a boiler. The boiler should be capable of load reduction, which means reduction of the gas supply and the corresponding velocity. This is not a simple task for a designer because the pressure drop is influenced by about the square of the velocity, and the reduction of the pressure drop is considerable during a load reduction. Two measures can be taken to allow an acceptable performance under such circumstances: 1) the full load pressure drop is chosen sufficiently high despite high fan power; 2) while reducing load, first the secondary air supply is reduced, maintaining the bottom air supply at as high level as possible. The general conclusion is that at high load (the cases investigated here), the pressure drop of the

Table 2. Limits of solids hold-up ($1-\varepsilon$) in CFB risers.

Source	Bubbling fluidization	Turbulent fluidization	Fast Fluidization	Transport
Bi, Grace, Zhu [30]	–	–	–	< 0.01
Bi, Grace, Zhu [30]	–	–	–	0.073 to $0.002 > 1-\varepsilon$
Various sources				
Bi and Grace [31]	–	$0.4 > 1-\varepsilon > 0.2$	$0.2 > 1-\varepsilon > 0.05$	$0.05 > 1-\varepsilon > 0.01$*
Bai et al. [32]	$(1-\varepsilon)_{mf} > 1-\varepsilon > 0.35$	$0.35 > 1-\varepsilon > 0.15$	$0.15 > 1-\varepsilon > 0.05$	$0.05 > 1-\varepsilon$
Grace et al. [33]	$0.6 > 1-\varepsilon > 0.35$	$0.35 > 1-\varepsilon > 0.2$	–	$0.03 > 1-\varepsilon$

*Core-annular dilute transport, dilute transport $0.01 > 1-\varepsilon$.

distributor is sufficiently high not to contribute with irregularities on the phenomena studied.

Another question concerns the influence of furnace size on the flow behavior in the transport zone. In a larger boiler the particles must be transported longer distances in the horizontal direction to reach the descending wall layers. Then the vertical decay would be influenced by the size of the cross section of the riser. The question can only be answered in an indirect way, as no direct measurements in boilers are available. The descending particle flow in the wall layers is a result of the migration of particles from the core to the walls. This results in a decay of particle concentration with height in the furnace, such as seen in Fig. 3. First, the thickness of the particle wall-layers δ is about independent of the horizontal dimension of the furnace D_e as seen by the correlation of Werther and co-workers [34],

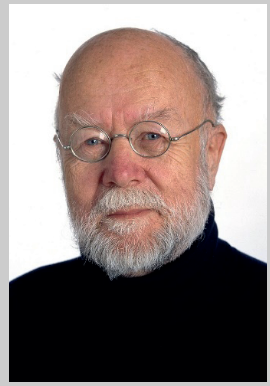
$$\frac{\delta}{D_e} = 1.1 Re^{-0.33} \left(\frac{H}{D_e}\right)^{0.68} \left(\frac{H-h}{H}\right)^{0.92} \tag{1}$$

The only remaining influences on the wall layer are those of the gas velocity U (hidden in $Re = UD_e/\nu$) and the furnace height H. The same publication [34] shows that the solids descent velocity in the wall layer only depends on the distance from the top $H-h$ for boilers between 12 MW_{th} to 235 MW_e. In conclusion, the effective horizontal transport of particles to the wall layer is independent of boiler width, within the limits of the accuracy of available measurements, and so are the vertical decay of particle concentration and the conditions in the core of the riser. A contributing factor to this result might be found in the rectangular shape of the cross section, shown in Fig. 1, where the width changes less with the boiler size that the length. However, dedicated studies are needed to verify this hypothesis.

6 Conclusions

An overview on the fluidization characteristics of the furnace of fluidized bed boilers has been based on data from the 12 MW_{th} CFB boiler at Chalmers University completed with what can be derived from literature.

The experiences indicate that the bottom bed is bubbling rather than fast or turbulently fluidized as is often assumed. Because of the bubbling activity, above the bottom bed, particles are thrown up by the bubbles and form an extended splash zone, in which particles group into clusters and fall back to a large extent into the bottom bed. The finer particles are carried further by the gas from the bottom bed and form the transport zone, initially characterized as a gas flow saturated by particles, but particles are deflected into the wall layers and the resulting particle density in the transport zone decays towards the exit, so the flow grows under-saturated. The suspension density of the transport zone is low, and the particle circulation-rate is relatively small. Therefore, this flow is in a state of a dilute phase transport, and it is too dilute to be in the fast fluidization regime according to the density ranges given by available definitions.



Bo Leckner graduated at Chalmers University in 1962 and presented his doctor thesis in 1972. He was appointed a professor in 1982 and has been at Chalmers University since then, except for one year spent at Moscow Energy Institute. Leckner has been working with various aspects of combustion of solid fuels and participated in the design of two 12 and 16 MW fluidized bed combustors for his research concerning fluid-dynamic, thermal, and chemical aspects (emissions) of fluidized bed conversion. Recently he has been engaged in reduction of greenhouse gases: biomass conversion and active CO_2 reduction from conversion of fossil fuels.

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Fluidization Characteristics of Circulating Fluidized Bed Boilers

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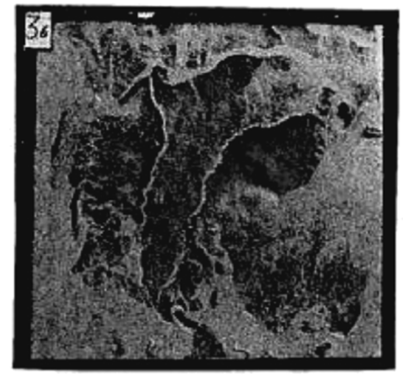

Review Article: Circulating fluidized bed boilers are difficult to access. Nevertheless, this is an effort to characterize the bottom part of the bed in a furnace, claiming that it is bubbling in contrast to the general opinion that states that it is turbulent or fast fluidized. In the upper part of the riser the bed is in a state of disperse transport and not fast fluidized because of the low particle concentration.

