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Particle Population Balances in a Refuse Derived Fuel Fired Circulating Fluidized Bed Combustor

Kai Redemann* Ernst-Ulrich Hartge† Joachim Werther‡

*Technical University Hamburg-Harburg, redemann@tuhh.de
†Technical University Hamburg-Harburg, hartge@tuhh.de
‡Technical University of Hamburg-Harburg, werther@tu-harburg.de

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Kai Redemann, Ernst-Ulrich Hartge and Joachim Werther
Institute of Solids Process Engineering and Particle Technology,
Technical University Hamburg-Harburg,
D-21071 Hamburg, Germany

ABSTRACT

With a population balance model the effects of changes in the operation of a circulating fluidized bed combustor on the bed inventory are simulated. Feeding, discharge and recirculation of solids as well as separation effects in the combustion chamber and in the cyclone are considered. The model predictions are compared with measurements at the refuse-derived fuel incineration plant Neumuenster. The simulation shows that by sieving of the ash withdrawn from the bottom of the combustion chamber and recycling the fine fractions to the bed the non-elutriable fraction of the ash in the bottom zone of the circulating fluidized bed (CFB) can be kept at a low level.

INTRODUCTION

For the operation of a circulating fluidized bed the particle size distribution (PSD) is of major interest. It affects for example the vertical solids distribution and the pressure profile, the externally circulating solids mass flux, the heat transfer bed/wall and in an external bed cooler, the loss of bed material through the cyclone overflow and the state of fluidization in the bottom bed. The particle size distribution in the circulating fluidized bed combustor (CFBC) is affected by a multitude of factors. First of all the fuel will during combustion form ash particles which may fragment and will be attrited during their residence time in the combustion system (1). Furthermore classification processes are taking place in the combustion chamber and in the cyclone.

Keeping the PSD within prescribed limits is particularly important in the field of waste combustion, where the widely varying composition of the fuel often leads to coarse ash particle formation, which may accumulate in the bottom bed of the CFB and may lead to poor fluidization or even defluidization of the bed. The present work has been carried out in close cooperation with Stadtwerke Neumünster GmbH, who commissioned in 2005 a plant which incinerates refuse-derived fuel (RDF) produced from the household waste of the city of Neumünster and surrounding districts (2).

THEORY

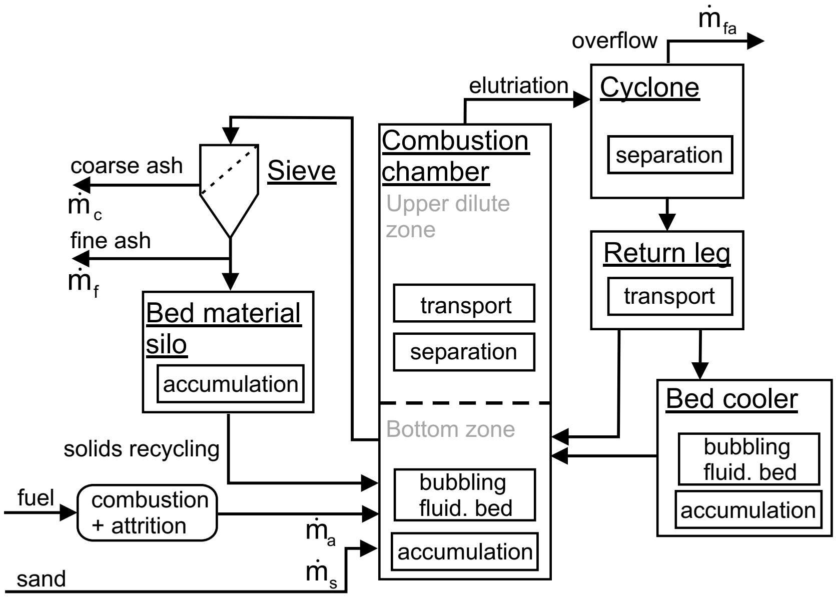

Figure 1: Simplified population balance model of a CFB including a fluidized bed cooler and solid recirculation.

The fluidized bed system, which is treated here, is shown in figure 1. The combustion chamber is separated into a bottom bed, which is modeled as a bubbling fluidized bed (BFB) and an upper dilute zone, where transport and separation according to the particles' terminal velocities occurs. The elutriated solids are separated in the cyclone; the fine fly ash leaves the cyclone in the overflow. The recycled solids are transported via the return leg, either directly into the combustion chamber or via the fluidized bed cooler (FBC), which contains a BFB. Bottom ash is withdrawn and fractionated with a sieve. The coarse fraction of the sieve is discharged for disposal, while the fine ash is totally or in part stored and recycled into the combustion chamber.

The fuel feed generates ash particles during combustion inside the system. The freshly generated ash will experience fragmentation and attrition (3, 4). However, in order to simplify the model the ash formation and the attrition process are taken out of the system. As a first approximation, it is assumed that the fuel combusts outside the system and only ash with a PSD that results after fragmentation and attrition is introduced into our model system. Besides this "fresh" ash with its "attrited ash particle size distribution" (AAPSD) and the recycled fine ash fraction sand is fed into the combustor for stabilization of the bed PSD.

The fluid mechanics of the dense bottom zone are described by the bubbling fluidized bed model by Werther and Wein (5), which provides means for the calculation of the height-dependent values of bubble size, bubble rise velocity, visible bubble flow and bubble volume fraction. The upper dilute zone can be described by an entrainment model with an exponential decay of the solids volume concentration with height h (6) combined with the entrainment flux correlation suggested by Tasirin and Geldart (7).

The two sub models for the dense bottom bed and the upper dilute zone have to be solved simultaneously in order to fulfil the total mass balance for each size class inside the CFB riser.

From the solids volume concentration at the top of the combustion chamber in combination with the gas flow in the combustion chamber, the entrainment into the cyclone can be calculated. Inside the cyclone the solids are separated from the gas phase. The separation effects are modeled by the Trefz and Muschelknautz model (8, 9). The solids are flowing from the cyclone into the return leg and are splitted into two streams. One stream enters directly the combustion chamber, while the other one passes through the external fluidized bed cooler. To keep the pressure drop in the combustion chamber constant, bed material is drained from the dense bottom zone of the combustion chamber. This drained bed material is sieved into a coarse and a fine ash fraction. To control the PSD in the combustion chamber some of the fine ash fraction is stored in a bed material bunker and later reintroduced into the combustion chamber.

The overall mass balance of solids the plant is (cf. Figure 1)

$$\dot{m}_a + \dot{m}_s - \dot{m}_c - \dot{m}_f - \dot{m}_{fa} = 0 \quad (1)$$

which also holds for the particle size class i,

$$\dot{m}_{a,i} + \dot{m}_{s,i} - \dot{m}_{c,i} - \dot{m}_{f,i} - \dot{m}_{fa,i} = 0 \quad (2)$$

In eqs. (1) and (2) $\dot{m}_{a,i}$ is the mass flow of the AAPSD. Since samples of the sand, coarse ash, fine ash and fly ash can be taken at the plant, their PSDs can be measured and their mass flows can be determined from the operation of the plant. Eqs. (1) and (2) allow us then to determine the unknown mass flow and PSD of the attrited fuel ash.

The model described above is solved by a dynamic calculation, so time-dependent effects can be calculated as well. The whole population balance model of the CFB system contains further mass balances and descriptions of the particle behavior in the return leg, FBC and the bed material silo.

EXPERIMENTAL

The simulation was applied to the CFBC of the Stadtwerke Neumünster GmbH in north Germany. A sketch of this plant is shown in figure 2.

Samples of the ash material leaving the plant and of the sand have been taken and analyzed on its PSD by sieving and a laser diffractometer. Their mass flows have been determined as average values by evaluating the operating data of the plant for an operation period from April 1 until June 30, 2006. The calculated AAPSD was used to validate the fluid dynamic model.

The separation efficiency curves of the screens used in the calculation are shown in figure 3. Screen A is the screen currently installed in the plant. Its separation efficiency curve has been calculated from mass flows and PSDs of the fine and coarse ash fractions. Its cut size is 1.7 mm. The curve of screen B is constructed to examine the impact of a screen with a smaller mesh size (cut size 300 µm) on the PSD in the plant.

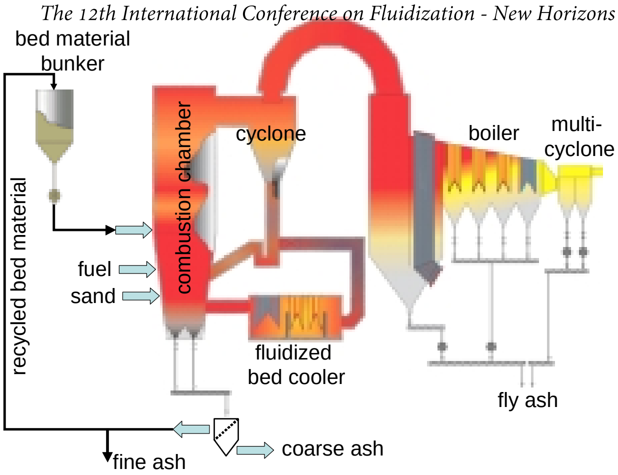

Figure 2: Flow sheet of the CFBC in Neumünster

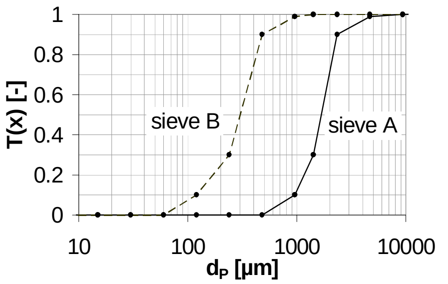

Figure 3: Separation efficiency curves for sieves utilized in the simulation.

The pressure drop in the combustion chamber was measured and was used as an input to the model. The calculation is performed with time steps of 0.5 seconds and is executed until steady state conditions are reached.

RESULTS AND DISCUSSSION

Determination of the "attrited ash particle size distribution" (AAPSD)

The mass flows of the solids entering and leaving the plant are summarized in table 1. Figure 4 shows the corresponding PSDs of the solids entering or leaving the CFBC in Neumünster. From the balance in eq. (1) follows the attrited ash feed rate which results as 2912 kg/h. The resulting AAPSD is very broad and this illustrates the difficulty with operating a waste incineration plant. 30% of the particle mass belongs to particles in the range of 1 – 25 mm or still 20% of the mass describes particles with sizes exceeding 6mm. This high percentage of coarse material indicates that there is a non-negligible risk of defluidization.

solid stream	mass flow [kg/h]
fly ash $(\dot{m}_{fa})$	- 1284
coarse ash $(\dot{m}_{c})$	- 968
fine ash $(\dot{m}_{f})$	- 1080
sand $(\dot{m}_{s})$	+ 420
fuel ash $(\dot{m}_{a})$	+ 2912

Table 1: Solid streams entering (+) or leaving (-) the plant, averaged over the operation period April 1 – June 30, 2006.

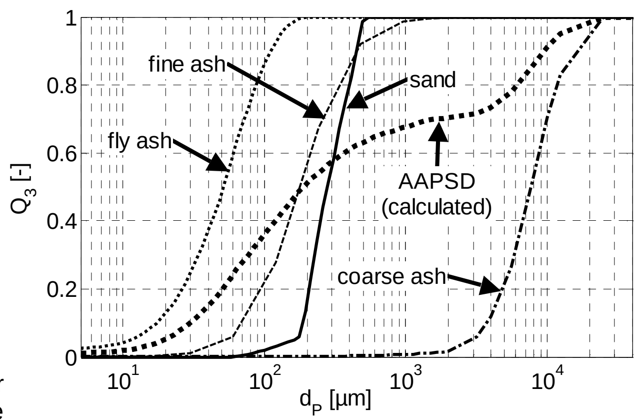

Figure 4: Particle size distributions of the solids entering or leaving the circulating fluidized bed in comparison with the calculated attrited ash particle size distribution (AAPSD).

Validation of the model

In the following model calculation, the fuel ash feed rate $\dot{m}_a$ and the AAPSD as well as the mass flow and the PSD of the sand are used as input data.

The calculation starts with the assumption that the CFB is operated with a sand bed. At the time t=0 the fuel ash feed is switched on. Figure 5 gives as a first result some information about the dynamic behavior of the plant. Plotted are the Sauter diameters in the bottom and top of the combustion chamber and in the external fluidized bed cooler. We see that it takes quite a time for the system to reach a steady state. Between a couple of days and one week are needed.

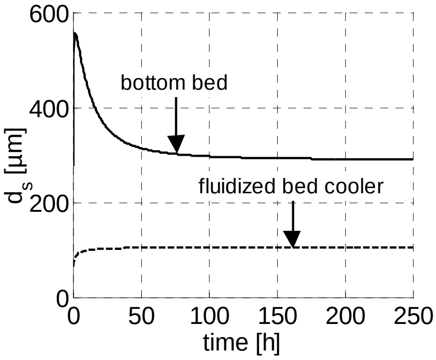

Figure 5: Time-dependent development of the Sauter diameter in the combustion chamber bottom bed and fluidized bed cooler with time.

In the operation period investigated 420 kg/h of fresh sand were fed into the combustion chamber. 840 kg/h of the fine fraction of the sieved bottom ash were fed into a bunker and were recycled into the combustion chamber. These data were used as input for the simulation model.

Figure 6 and 7 show the vertical profiles of solids volume concentration and pressure in the combustion chamber. The bottom bed with a high solids volume concentration can be clearly distinguished from the upper dilute zone. The calculated values correspond well to the measured ones.

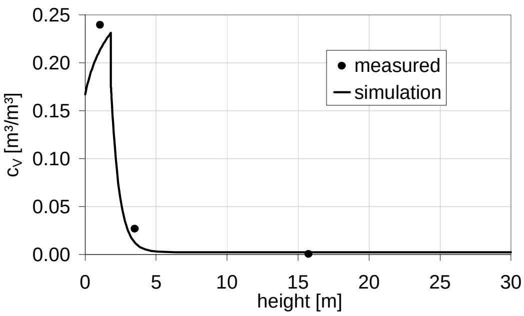

Figure 6: Measure and calculated solids volume profile

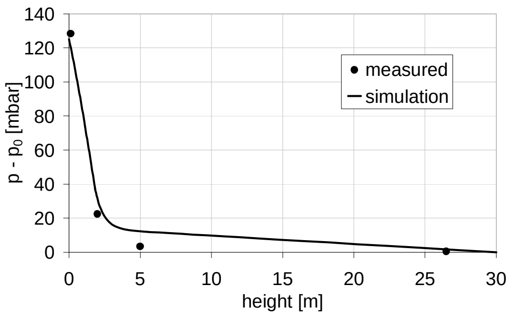

Figure 7: Measure and calculated pressure profile

A further information, which permits the judgment of the accuracy of the simulation, is the mass flow and the PSD of the fly ash. The operation data yield a fly ash mass flow of 1284 kg/h, whereas the simulation results in a massflow of 975 kg/h. The corresponding PSDs are compared in figure 8, The simulation yields a slightly finer fly ash. Since further measurments, e.g. of the external circulating solid mass flux or loading, are not available, it is not possible at present to judge whether the observed deviations of the mass flux and the PSD of the fly ash must be attributed to the cyclone separation model or to the fluid mechanics modelling in the combustion chamber.

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Adjustment of the PSD in the bottom bed

A major value of a simulation model is that it allows systematic investigations of the influences of different parameters on the operation behavior of a complex plant. In the present case, it is the control of the PSD in bottom bed which requires attention. On the one hand it is the content of coarse material, which should be kept below a certain level in order to avoid defluidization of the bed. On the other hand it is the content of elutriable material, which is responsible for the entrainment and therefore for the initiation of a sufficiently high external solids recirculation. Many modern CFBCs have all or most of their heat transfer surfaces located in the membrane wall of the combustion chamber. This is not the case in the presently considered waste combustor, where in order to reduce the risk of corrosion, the heat transfer surfaces are located outside the combustion chamber in the external fluidized bed heat exchanger. This arrangement requires a sufficient external solids circulation in order to avoid overheating of the combustion chamber.

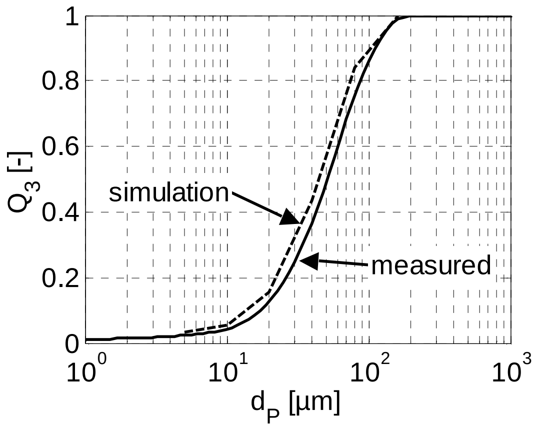

Figure 8: Comparison of measured and simulated cyclone overflow mass distribution for a recirculation with 840 kg/h over screen A. (R840a)

The PSD in the bottom bed is mainly influenced by the ash characteristics of the fuel. It can be influenced by adding a separate bed material e.g. sand. But it can also be influenced by recirculation of a sieve fraction of the bottom ash offtake. Table 2 lists the situations investigated: We start with feeding fuel only (R0). R0s is the case where 420 kg/h sand are additionally fed. R420A is the case where the same amount of recirculated fines obtained as the underlow of sieve A is fed. R840As is the case where both sand and twice as much of the underflow of sieve A is supplied. This is the actual operating case in Neumünster. R840Bs is the same as the previous one with the exception that sieve A has been exchanged against the finer one B. R1260A and R3360A are similar to R420A, but with increased mass fluxes of the recirculated fine ash.

name	ash feed rate [kg/h]	sand feed rate [kg/h]	recirculation rate [kg/h]	sieve
R0	2912	-	-	-
R0s	2912	420	-	-
R420A	2912	-	420	A
R840As	2912	420	840	A
R840Bs	2912	420	840	B
R1260A	2912	-	1260	A
R3360A	2912	-	3360	A

Table 2: Used setups of operating conditions evaluated in the simulation

The resulting cumulative mass distributions of the particle sizes in the bottom bed are plotted in figure 9. For convenience, a second abscissa has been added on top of the figure, which presents the terminal settling velocity of the particles under combustion conditions (850°C, 1 bar). Especially marked is the operating velocity of 4.7 m/s in the combustor, which separates the PSD into an elutriable and a non-elutriable fraction, respectively.

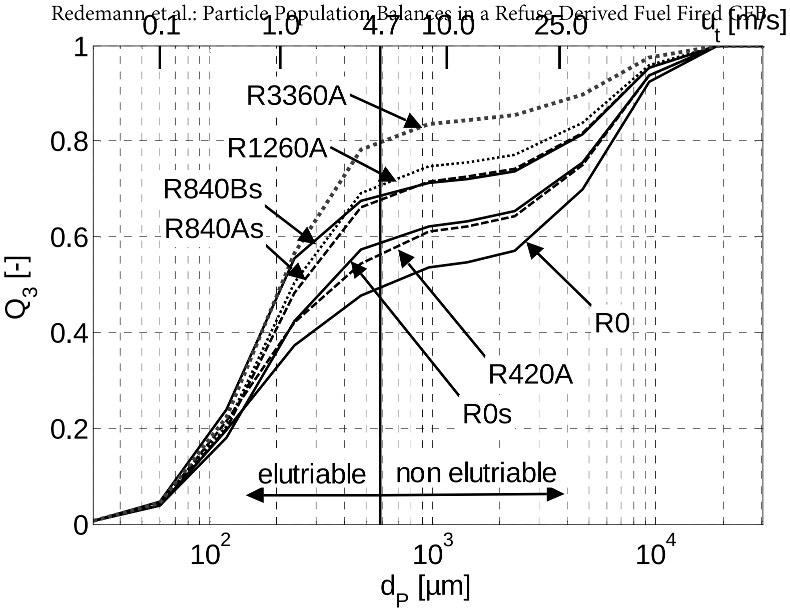

Figure 9: Particle size distributions in the bottom bed of the combustion chamber (terminal velocities calculated for 850°C, 1 bar).

We see that in the base case R0, i.e. without recycling of fines and feeding sand, nearly 50% of the bed mass is not elutriable, which will result in a minimum of external solids recirculation. Adding the comparatively fine sand (R0s) reduces the non-elutriable fraction to 40% and shifts the PSD significantly to the left. The same effect can be achieved with a replacement of the sand by recirculating the same mass flow of fine ash (R420A). The present operating conditions (R840As) are characterized by a non-elutriable mass fraction of 30% and a 50%-value of the PSD of 250 µm. The sand, which is presently added, costs twice. Namely, once as a raw material and secondly for disposal. So there is an incentive to operate without sand addition. The simulation shows that this could be done by simply increasing the ash mass flow to 1260 kg/h (R1260A). Exchanging the sieve A against B (R840Bs) causes some further shift of the bed PSD into the fine direction ($dp_{,50,3}$ = 250 µm) but leaves the non-elutriable fraction unchanged. A further calculation with the drastically increased fines recycle (R3360A) yields a mean particle size of 200 µm and decreases the non-elutriable fraction of 20%, which should provide optimum operating conditions for the heat extraction with a strong solids circulation. It should be noted in this context that measurements in CFB coal combustors in Duisburg and Flensburg (10), which are operated without sand addition, have shown that in these cases the non-elutriable fraction was also between 20% and 30%.

CONCLUSIONS

A simulation model, which is able to calculate particle balances for a CBFC system including cyclone and external fluidized heat exchanger, has been developed. The model was validated with measurements at the refuse-derived-fuel firing incineration plant at Neumünster / Germany. The prediction of the model with regard to mass flow and PSD of the fly ash from the cyclone overflow, the pressure profile and solid volume concentration inside the combustion chamber are in good agreement with the measurements.

The practical use of the model is illustrated with the simulation of different means to influence and adjust the bottom bed PSD in the combustion chamber. It is shown

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that by adding a fine sand or by recycling part or all of the fine fraction of the sieved bottom bed offtake, the bottom bed PSD can be shifted over a wide range in order to adjust a suitable solids circulation rate and to avoid a too large fraction of coarse material in the bottom bed.

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