PERGAMON

Energy Conversion & Management 40 (1999) 447–458

ENERGY CONVERSION & MANAGEMENT

Effect of coal particle size distribution on agglomerate formation in a fluidized bed combustor (FBC)

G. Venkat Reddy[a, 1], S.K. Mahapatra[b]

[a]Department of Chemical Engineering, Regional Engineering College, Warangal, India
[b]Department of Mechanical Engineering, S.L. Institute of Engineering and Technology, Longowal, Sangroor, Punjab, India

Received 2 September 1997

Abstract

An investigation was conducted to find out the reasons for agglomeration in a fluidized bed combustion (FBC) power plant. Two typical coal samples were collected for investigation. The first sample was collected when the plant was operating smoothly, and the second was collected immediately after agglomeration. These two samples were subjected to analysis. It was observed that agglomeration of the bed material in FBC plants takes place when the coal sample contains either too many very fine particles or too many very coarse particles or both in very large proportion. The very fine particles present in the coal have considerable plastic properties (caking and swelling tendency) which cause agglomeration. The coarser particles are denser because of their higher ash content, and their higher density causes an increase in minimum fluidization velocity which, in turn, leads to agglomeration and defluidization. Medium size coal particles are most suitable for combustion in a FBC. © 1998 Elsevier Science Ltd. All rights reserved.

Keywords: Fluidised bed; Agglomeration; Size distribution; Minimum fluidisation velocity; Distributor Plate; Defluidisation; Caking propensity

Nomenclature

a_o	Area of orifice (cm^2)
A_1, A_2	Area of Sections 1 and 2 in model, respectively (cm^2)
A_T	Total area of fluidized bed (cm^2)

[1] E-mail: vreddy@recw.ernet.in

PII: S1096-8904(98)00113-7

d_o	Actual diameter of orifice (cm)
d_{or}	Diameter of the distributor orifice (cm)
d_p	Average particle diameter (cm)
e_{mf}	Bed voidage at minimum fluidization
e_B	Average bubble voidage in Section 1 of model
$\mathbf{g}$	Acceleration due to gravity (cm/sec^2)
h_g	Height of spout or jet (cm)
H	Bed height at minimum fluidization (cm)
N	Total number of orifices
N_1/N	Fraction of active number of orifices
P	Pitch of holes in distributor (cm)
P_{av}	Average pressure in distributor (atm)
P_B	Pressure drop across fluidized bed, subscripts 1 and 2 denote Sections 1 and 2, respectively
P_D	Pressure drop across the distributor, subscripts 1 and 2 denote Sections 1 and 2, respectively
T_b	Absolute bed temperature (K)
U	Total superficial gas velocity (cm/sec)
U_1, U_2	Superficial gas velocity in Sections 1 and 2, respectively (cm/sec)
U_{mf}	Minimum fluidization velocity (cm/sec)

Greek symbols

ρ_g	Density of fluidizing gas (g/cm^3)
ρ_s	Density of solids in bed (g/cm^3)
Φ	Orifice density (N_{ao}/A_T)
μ_v	Viscosity of fluidizing gas (g/(cm/sec))

1. Introduction

The fluidized bed combustion technique is being extensively used in India and elsewhere to generate power from coal washery rejects and low grade coals. This technique makes it possible to utilize high ash coal washery rejects in an environmentally friendly manner. A number of small scale power plants have been installed in India near coal washeries to extract energy from this waste material. Almost all coal washeries in India clean the coal that is meant for steel plants. Therefore, the rejects from these washeries have some degree of caking propensity left in them. These small scale FBC power plants have been running successfully, except occasionally, when they face an agglomeration problem. When this problem occurs due to the formation of very big size agglomerates, defluidization follows, and the operation of the plant comes to a sudden halt. The plant can be restarted only after the agglomerates from the bed are removed, which takes considerable time. The economic implications are quite obvious.

It is a well known fact that the melting of ash takes place inside a boiler when the temperature generated by burning coal exceeds the ash fusion temperature. The molten ash then engulfs the unburned coal particles and forms a solidified mass which is known as clinker.

Clinker inside a boiler generally occurs when coal contains ash of low fusion temperature. In addition, sintering and agglomeration of particles may also occur in the lower portions. However, this reasoning cannot be extended to agglomerate formation in a fluidized bed combustor because the temperature maintained inside the bed is around 850–900°C only, which is much below the ash fusion temperature.

Agglomeration of particles in fluidized beds has been studied extensively [1–4]. Langston and Stephens [1] studied metallic ore reduction in a self-agglomerating fluidized bed. They reported that agglomeration of particles is directly proportional to the adhesive properties and area of contact and inversely proportional to the particles' relative momentum. Goldberger [2] measured the defluidization velocity at several temperatures, both in a bench scale and pilot plant units. It has been reported that the defluidization velocity is a function of minimum fluidization velocity and bed temperature. Tardeos et al. [5] and Mazzone [6] used a force balance to determine agglomeration break up as a criterion for defluidization. Rehmat and Saxena [7] observed that agglomeration in fluidized bed combustors occurs when high internal temperatures are generated in a reacting coal particle. Moseley and O'Brien [8] developed a mathematical model for agglomeration in a fluidized bed. Their model is in two parts: The first part estimates the defluidization velocity using a two particle collision model, and the second part accounts for the granular energy of the bed material. Manzoori and Aggarwal [9] performed several experiments in a laboratory scale fluid bed system to study the role of inorganic matter in coal in the formation of agglomerates. They observed that the agglomerating propensity of the bed particles increases with the increase of sodium content in coal, and also, it increases with an increase in furnace temperature. According to them, the content and mode of occurrence of sodium in coal is responsible for the ash to melt which then engulfs the surrounding coal particles to form agglomerates in a fluidized bed which results in defluidization.

Dawson and Brown [10] also conducted several tests in a laboratory scale fluid bed combustor to determine the factors and mineralogical reactions that initiate agglomeration and subsequent defluidization. They observed that the loss of fluidization can occur due to agglomeration of bed particles if combustion air falls below 100% theoretical level, even when bed temperatures remain within an acceptable range. It was further observed that the presence of aluminosilicates was responsible for agglomeration and defluidization.

This paper presents the results of a study undertaken to investigate the possible causes for agglomeration in a FBC. The study qualitatively correlates agglomeration to coal properties, e.g. coal particle size distribution, its caking and swelling indices and to operating conditions like bed temperature and, finally, to distributor plate design. The coal properties like caking and swelling indices and particle size distribution are very easy to measure, and in fact, in most of the plants, these tests are performed on a regular basis.

2. Experimental details

To find out the reasons for agglomeration, two typical feed coal samples were collected from the TISCO, Jamadoba, plant. The first one was collected when the plant was running smoothly, and another sample was collected when agglomeration occurred. Both samples were

Table 1. Proximate analysis of coal samples

Composition	Sample 1	Sample 2
Moisture (%)	1.43	2.86
Ash (%)	55.05	59.96
VM(%)	28.23	15.18
FC (%)	15.29	22.00
Distribution modulus	0.699	1.116
Size modulus	0.174	0.378
Average particle diameter	1.91	1.09

subjected to analysis. The particle size distribution, density and ash content of each size fraction were determined, and this data was correlated to minimum fluidization velocity. The proximate analysis, free swelling and caking indices data are presented in Tables 1 and 2. Further experiments were conducted to find out the temperature and the fluidizing conditions under which agglomeration is likely to occur. For this, the feed coal sample (Sample 1, Table 1) was heated in two types of experimental bowls, one with a sufficient number of holes in its bottom for supply of air and another without any holes for restricted air supply. The heating was done in a muffle furnace at different temperatures ranging from 300° to 600°C and for different durations of time. After the heating was over, samples from both bowls were taken out, the amount of loose material in each case was measured and the agglomerates were subjected to a shatter test. The agglomerates were dropped from a height of 6 ft onto a steel plate and the loose material below the size of 2 mm was measured. The less is the loose material, the stronger the agglomeration.

3. Mathematical analysis

To study the effect of minimum fluidization velocity upon the fraction of number of active holes, the approach given by Yue and Kolaczkowski [11] has been adopted. A partially

Table 2. Data on field coal (Sample 1)

Sieve size (mm)	FSI	CI
+ 5.6	—	1
+ 4.0	—	1
+ 3.335	—	1
+ 2.3	—	1
+ 1.7	—	2
+ 0.85	—	2
+ 0.42	1	3
+ 0.125	1.5	8
+ 0.076	2	9
+ 0.0	2	9

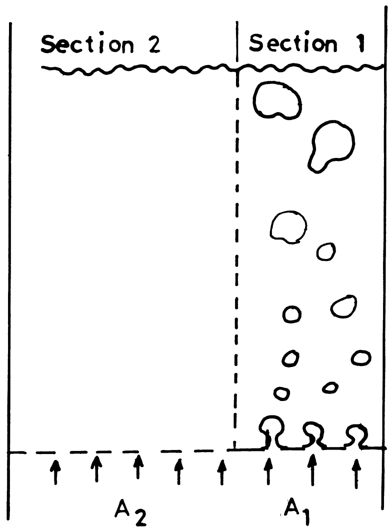

Fig. 1. Model showing the partial operations of a gas distributor.

operated multi-orifice distributor plate is shown in Fig. 1. Normally, gas tends to flow preferentially through that part of the bed which has the higher voidage and, therefore, low bed pressure drop. Thus, two regions of different voidage exist, as shown in Fig. 1. Section 1 of the bed is freely bubbling, and its voidage is greater than e_{mf}. The corresponding superficial gas velocities in the sections are U_1 and U_{mf}.

For the system shown in Fig. 1 to be in equilibrium, the total pressure drop across the distributor and bed in both sections must be the same, i.e.

$$\Delta P_{D1} + \Delta P_{B1} = \Delta P_{D2} + \Delta P_{B2} \tag{1}$$

The pressure drop across a single orifice in the distributor plate is given as

$$\Delta P_o = \frac{\rho_f}{2gC_D^2} U_o^2 \tag{2}$$

Hence, the difference between the distributor pressure drops of the two sections is given by:

$$\Delta P_{D1} - \Delta P_{D2} = \left(\frac{A_1}{N_1 a_o}\right)^2 \frac{\rho_f U_1^2}{2gC_D^2} - \left(\frac{A_2}{N_2 a_o}\right)^2 \frac{\rho_f U_{mf}^2}{2gC_D^2} \tag{3}$$

For a multi-orifice distributor with a regular array of holes, each of the same dimension, we have

$$A_1/N_1 = A_2/N_2 = A_T/N \tag{4}$$

Therefore,

$$\Delta P_{D1} - \Delta P_{D2} = \frac{\rho_f}{2gC_D^2\phi^2}\left(U_1^2 - U_{mf}^2\right) \tag{5}$$

where $\phi = Na_o/A_T$, which may be called orifice density.

The overall increase in bed height can be obtained from a volumetric balance given below;

$$\Delta H = \frac{\bar{e}_B(N_1/N)H_{mf}}{1 - \bar{e}_B N_1/N} \tag{6}$$

The difference between the pressure drops across the two sections because of the presence of bubbles in Section 1 is given by

$$\Delta P_b = \rho_s g(1 - e_{mf})(H_{mf} + \Delta H)\bar{e}_B \tag{7}$$

Using ΔH from Eq. (6), Eq. (7) becomes

$$\Delta P_b = \bar{e}_B \Delta P_{Bmf} + \frac{\bar{e}_B^2 \Delta P_{Bmf} N_1/N}{1 - \bar{e}_B N_1/N} \tag{8}$$

where

$$\Delta P_{Bmf} = \rho_s g(1 - e_{mf})H_{mf} \tag{9}$$

Again the difference between the pressure drops across the two sections because of the spouts formed above the active orifices, according to Fakhimi and Harrison [12] is given by

$$\Delta P_s = (1 - 2/\pi)\frac{h_s \Delta P_{Bmf}}{H_{mf}} \tag{10}$$

Combining the difference with that due to the presence of the bubbles, we obtain:

$$\begin{aligned}\Delta P_{B2} - \Delta P_{B1} &= \Delta P_b + \Delta P_s \\ &= \bar{e}_B \Delta P_{Bmf} + \frac{\bar{e}_B^2 N_1/N}{1 - \bar{e}_B N_1/N}\Delta P_{Bmf} + (1 - 2/\pi)\frac{h_s \Delta P_{Bmf}}{H_{mf}}\end{aligned} \tag{11}$$

The bubbles voidage e_B usually has a value of 0.1 to 0.2. Therefore, the second term on the right hand side of Eq. (11) is relatively small and can be neglected. Equating eqs. (5) and (11),

$$\frac{\rho_f}{2gC_D^2\phi^2}\left(U_1^2 - U_{mf}^2\right) = \bar{e}_B \Delta P_{Bmf} + (1 - 2/\pi)\frac{h_s \Delta P_{Bmf}}{H_{mf}} \tag{12}$$

Re-arranging the above equation, we can get

$$U_1 = \left[\frac{\Delta P_{Bmf}\{\bar{e}_B + (1 - 2/\pi)h_s/H_{mf}\}}{\rho_f/2gC_D^2\phi^2} + U_{mf}^2\right]^{1/2} \tag{13}$$

However,

$$U = \frac{N_1}{N}(U_1 - U_{mf}) + U_{mf} \tag{14}$$

or

$$\frac{N_1}{N} = (U - U_{mf})/(U_1 - U_{mf}) \tag{15}$$

Putting the value of U_1 from Eq. (13) in Eq. (15), we can finally get,

$$\frac{N_1}{N} = \frac{U - U_{mf}}{\left[\Delta P_{Bmf}\{\bar{e}_B + (1 - 2/\pi)h_s/H_{mf}\} \times 2gC_D^2\phi^2/\rho_f + U_{mf}^2\right]^{1/2} - U_{mf}} \tag{16}$$

Eq. (16) is the relation between U_{mf} required and the number of active nozzles in the distributor plate. It shows that, as the U_{mf} increases, the number of active holes in the distributor plate decreases. When we charge coarser particles to the bed the U_{mf} required increases, consequently, some of the nozzles in the distributor plate become inactive. Around the inactive nozzle, due to the inadequate supply of air, agglomeration and defluidization may take place.

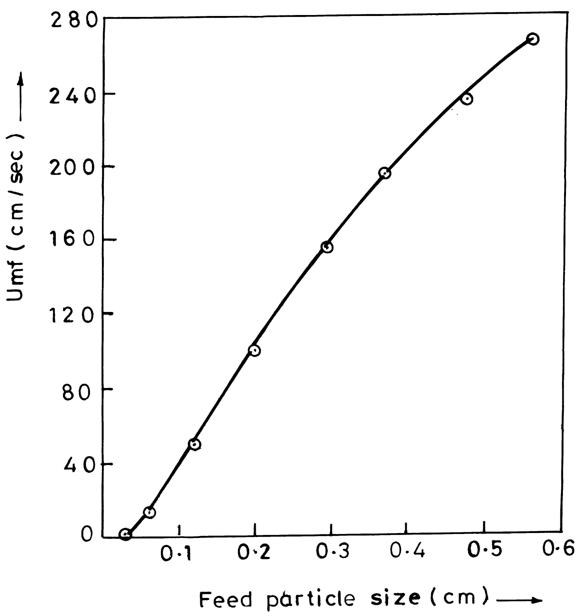

Fig. 2. Effect of feed particle size on minimum fluidization velocity.

The design of the gas distributor affects the physical and chemical performance of a fluidized bed [13–16]. One of the most important requirements of a gas distributor is that it must ensure uniform gas distribution throughout the bed. If gas distribution is not uniform due to some reason, some inactive nozzles in the distributor plate may cause agglomeration and defluidization.

4. Results and discussion

Fig. 2 shows the minimum fluidization velocity as a function of coal particle size. The minimum fluidization velocity required to initiate fluidization of bed particles increases as the particle size increases. Our laboratory analysis showed that, in feed coal, as the particle size increases, the ash content in the coal particle increases which, in turn, causes an increase in density of the particles. In other words, coarser particles are denser. Therefore, as the proportion of coarser particles in the coal feed to the fluidized bed increases, the minimum fluidization velocity required also increases. Fig. 3 shows a relationship between the minimum fluidization velocity and the fraction of active holes in the distributor plate. This relationship is derived from the mathematical analysis and is given in Eq. (16). It is clear from Fig. 3 that, as the proportion of coarser particles in the feed coal increases, the number of inactive nozzles in the distributor plate increases. Near the inactive nozzles, the air supply will be insufficient for fluidization and also for complete combustion. However, the temperature near these nozzles will be high enough to make coal particles soft and sticky. The soft and sticky particles form into agglomerates and grow in size and ultimately lead to defluidization of the entire bed. Our laboratory experiments confirmed this proposition. Fig. 4 and 5 show the percentage of loose material as a function of temperature with the time as the parameter for bowls without and with holes, respectively. The conditions prevailing in the bowl without holes are similar to the conditions around an inactive nozzle, where there is an insufficient supply of air for fluidization

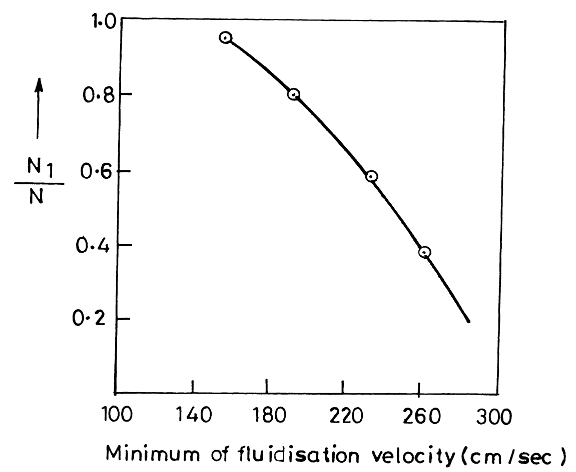

Fig. 3. Effect of minimum fluidization velocity on fraction of active number of holes.

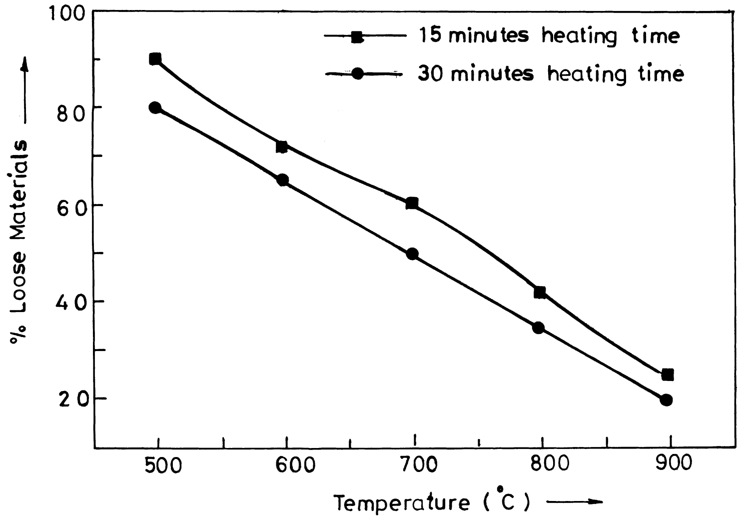

Fig. 4. Effect of temperature on the strength of agglomerates (steel bowl without holes).

and combustion. However, the temperature is high enough to make coal particles soft and sticky and consequently agglomerate. Fig. 4 shows that the higher the temperature, the stronger are the agglomerates formed. At any given temperature, the bigger and stronger agglomerates form near inactive nozzles in the fluidized bed if there is an insufficient supply of air. As shown by our mathematical analysis, some of the nozzles in the distributor plate become inactive, as the minimum fluidization velocity required is increased. The minimum fluidization velocity increases as the particle size increases, as shown in Fig. 2. To put it in another way, agglomeration of coal particles takes place near some inactive air nozzles when

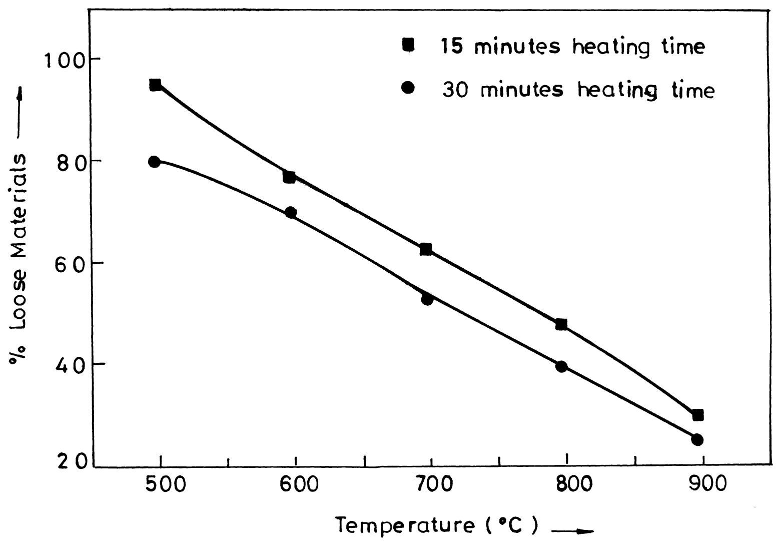

Fig. 5. Effect of temperature on the strength of agglomerates (steel bowl with holes).

the feed coal contains too many coarser particles. Fig. 5 suggests that agglomerate formation is less, when there is sufficient air supply.

Table 2 presents the free swelling index and caking index for different size fractions of feed coal. Both these indices increase as the particle size decreases. Free swelling and caking indices, both together, provide a measure of the binding propensity or sticky nature of the particles. The larger these indices for coal particles, the more sticky the coal becomes when it is heated. In other words, if the feed coal to the FBC unit contains excessive fines, then there are good chances of agglomerate formation because the finer particles become sticky at elevated temperatures. Factors that help agglomerate formation in a FBC are: (i) temperature, (ii) inadequate air supply, (iii) excessive surface area of particles, (iv) caking propensity and (v) presence of binding agents. As mentioned earlier, when there are too many coarse particles in the feed coal to the FBC, due to the increased U_{mf} requirement, some nozzles in the distributor plate become inactive, and consequently, there will be an inadequate air supply around these nozzles. Because of the insufficient air supply around some nozzles and because of sufficiently high temperature prevailing, the fine coal particles become soft and jelly-like and stick to each other, ultimately resulting in agglomerate formation. This agglomerate formation is further helped by the excessive surface area of fine particles. Though water is present initially in the coal particles, it evaporates as soon as the coal enters the bed. Therefore, at a temperature of 800–900°C prevailing in the fluidized bed furnace, the chances of water acting as a binding agent for coal particles is nil. Therefore, all coal particles in the present case become sticky because of their caking nature. If proper conditions exist, these coal particles stick to each other and then grow bigger into agglomerates.

The chances of two particles hitting each other and then sticking together depends upon the surface area of the particles. Obviously, the smaller the particles, the more will be the surface area for a given mass. Therefore, as the particle size distribution shifts towards a smaller size, the tendency to agglomerate will increase.

Therefore, we can say that, in fluidized bed combustion units, agglomeration, which leads to ultimate defluidization and shutdown, occurs when there are either too many coarse particles or too many fine particles in the feed coal. Both these particle fractions cause agglomeration

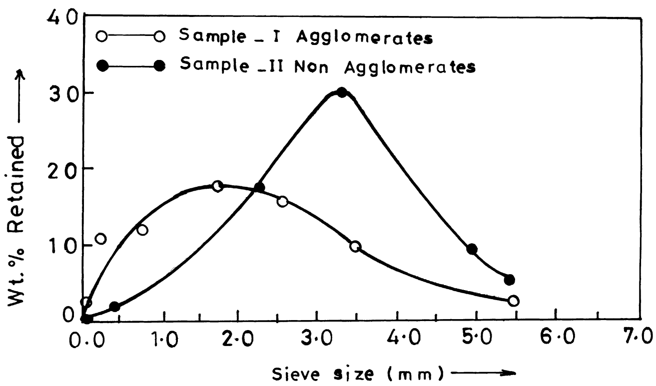

Fig. 6. Size distribution of feed coal samples.

for quite different reasons. Finer particles cause agglomeration because of their stickiness, whereas coarser particles cause agglomeration because of their large U_{mf} requirements.

The size distribution of the feed samples is shown in Fig. 6. The values of the distribution modulus for these two samples are also given in Table 1. The distribution modulus in the Rosin–Rammler [17] equation is a measure of the spread of the sample. The higher the value of distribution modulus of the sample, the narrower will be the particles' size distribution and vice versa. From Fig. 6, the feed coal sample (Sample 1, Table 1) which was responsible for agglomeration in the TISCO, Jamadoba, plant has the smallest distribution modulus or widest distribution, indicating that the increase in coarse and fines contents is responsible for agglomeration. The particles momentum will depend upon the fluidization velocity, particle size and particle density. If there is a wide size distribution of particles, the smaller particles travel faster in the bed than the bigger particles. In other words, smaller particles, because of their higher velocity, over-take the bigger particles during their upward movement. During such over-takes, the smaller particles may collide with bigger particles which may lead to agglomeration. Therefore, we can say, a wide particle size distribution increases the chances of agglomeration in the bed. Intuitively, if the bed is fluidized vigorously, the chances of agglomeration and subsequent defluidization are less because stable agglomerates do not form under such conditions.

5. Conclusion

1. Agglomeration in a FBC occurs when feed coal contains either too many fine particles or too many coarse particles.
2. Coarse particles cause agglomeration due to the increased U_{mf} requirement. Agglomeration and defluidization occur near inactive nozzles.
3. Finer particles in feed coal cause agglomeration because of their caking propensity and very large surface area.
4. Medium size particles (4 mm) are most suitable as feed for a FBC.

Acknowledgements

The facilities provided by the Department of Fuel and Mineral Engineering, Indian School of Mines, Dhanbad, India for carrying out some of the experiments is gratefully acknowledged. The help extended by plant operators and engineers at TISCO, Jamadoba, India, during data collection is highly appreciated.

References

[1] Langston BG, Stephens FM, Jr. Self agglomerating fluidized bed reduction. J. Metals 1960;12:312.
[2] Goldberger WM. Collection of fly ash in a self-agglomerating fluidized-bed coal burner. In: ASME Winter Annual Meeting and Energy Systems Exposition. Houston, Texas, ASME Paper No. 67 WAFU-3, 1967.
[3] Gluckman MJ, Yerushalmi J, Squires AM. Defluidization characteristics of sticky or agglomerating beds. In Fluidization technology. Ed. Keairns D. 1976;2:195.

[4] Siegell JH. Defluidization phenomena in fluidized beds of sticky particles at high temperatures. Ph.D. thesis. City University of New York, 1976.
[5] Tardos G, Mazzone D, Pfeffer R. Destabilization of fluidized beds due to agglomeration. Part I. Theoretical model. Can. J. Chem. Engng 1985;63:377.
[6] Mazzone DN. Agglomeration of particles in fluidized beds. Ph.D. thesis. City University of New York, 1986.
[7] Rehmat A, Saxena SC. Agglomeration of ash in fluidized-bed gasification of coals of steam-oxygen (or air). Ind. Engng Chem. Process Des. Dev. 1980;19:223.
[8] Moseley JL, O'Brien TJ. A model for agglomeration in a fluidized bed. Chem. Engng Sci 1993;48:3043.
[9] Manzoori AR, Aggarwal PK. The role of inorganic matter in coal in the formation of agglomerates in fluid bed combustor. Fuel 1993;72:1069.
[10] Dawson MR, Brown CR. Bed material cohesion and loss of fluidization during fluidized bed combustion of midwestern coal. Fuel 1992;71:586.
[11] Yue PL, Kolaczkowski JA. Trans. I. Chem. Engng 1990;60:164.
[12] Fakhimi S, Harrison D. Chemeca 70 Ichem. Symp. Ser. 33, Session 1 1970;29:.
[13] Geldart D, Kelsey JR. Tripartite Chem. Engng Conf., Montreal 1968;114:.
[14] Whitehead AB, Dent DC. Proc. Int. Symp. Fluidization, 1967. p. 802.
[15] Walker BV. Chem. Engng J. 1975;9:49.
[16] Cooke MJ, Harris W, Highley J, Williams DF. Tripartite Chem. Engng Conf., Montreal 1968;21:.
[17] Rosin P, Rammler E. J. Inst. Fuel 1927;7:29.