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Determination of Aspect Ratio and Excess Air Coefficient to Shorten Start-up Time at Small Size Fluidized Bed Boiler

Karani KURTULUŞ¹

¹Süleyman Demirel Üniversitesi, Mühendislik ve Doğa Bilimleri Fakültesi, Maki-
ne Mühendisliği Bölümü, Isparta

e-posta :karanikurtulus@sdu.edu.tr, ORCID No: 0000-0002-2608-3065

Keywords

Start-up, Sand Heating, Fluidized Bed, Bed Aspect Ratio, Multiphase Flow

Abstract

Energy production from clean and renewable sources has gained prominence in recent years. The use of fluidized bed boilers in industrial applications is increasing due to their high combustion efficiency and low NO_x emission. Fluidized beds are suitable for coal and biomass combustion and other gasification applications. Therefore, these systems are expected to become more widespread. Although they are widely used in industry, fluidized bed combustion systems start-up processes are generally based on experience. Start-up is an essential process in the operation of small size fluidized beds. Shorter start-up times are desired, especially in systems requiring frequent start-ups. This study investigated the effect of the height of static bed material and the coefficient of burner excess air on start-up time. Therefore, a bench-scale experimental setup was established. CFD analysis was verified with temperature data measured by the testing apparatus. Then, CFD analyses were carried out for different bed aspect ratios (0.55, 0.65, and 0.90) and different coefficients of burner excess air (1.2, 1.4, and 1.6). The analysis results show start-up time decreases (up to 32%) as the bed aspect ratio decreases. On the other hand, as the excess air coefficient increases, the start-up time increases (up to 44%). In addition, in the start-up process, excess air coefficient 1.4 and then 1.2 were proposed for a shorter start-up time with an aspect ratio of 0.9.

* karanikurtulus@sdu.edu.tr
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Düşük Kapasiteli Akışkan Yataklı Kazanlarda Hızlı Start-Up Yapabilmek İçin Uygun Hava Fazlalık Katsayısının Ve Yatak En/Boy Oranının Belirlenmesi

Anahtar Kelimeler

Öz

Start-up, Kum ısıtma, Akışkan yatak, Yatak en/boy oranı, Çok fazlı akış

Temiz ve yenilenebilir enerji üretimi her geçen yıl önemini arttırmaktadır. Akışkan yataklı kazanlar endüstride kullanımları yüksek yakma verimi ve düşük NOx emisyonlarından dolayı giderek artmaktadır. Bununla birlikte akışkan yataklı kazanlar kömür ve biyokütle yakabildikleri gibi gazlaştırma prosesleri için de uygundur. Bundan dolayı bu sistemlerin ileride daha da yaygınlaşacağı beklenmektedir. Piyasada yaygın olarak kullanılmalarına rağmen bu sistemlerde kalkış (start-up) prosesi daha çok tecrübeye bağlıdır. Küçük kapasiteli akışkan yataklı sistemler için start-up prosesi oldukça önemlidir. Özellikle sık start-up yapan sistemler için kısa sürede hızlı start-up istenmektedir. Bu çalışmada statik yatak yüksekliği ve hava fazlalık katsayısının start-up prosesine olan etkisi araştırılmıştır. Bunun için laboratuvar ölçekli test düzeneği kurulmuştur. Deney düzeneğinden ölçülen sıcaklık verileri ile HAD analizlerinden bulunan değerler karşılaştırılarak doğrulama yapılmıştır. Daha sonra HAD analizleri yatak en/boy oranı 0.55, 0.65, 0.90 ve hava fazlalık katsayısı 1.2, 1.4 ve 1.6 için yürütülmüştür. Analizlere göre yatak en/boy oranı azaldıkça start-up süresinin %32'ye kadar düştüğü görülmüştür. Öte yandan hava fazlalık katsayısı arttıkça start-up süresinin %44'e kadar arttığı bulunmuştur. Ek olarak küçük kapasiteli akışkan yatakta en boy oranı 0.9'da en uygun start-up süresi elde edebilmek için başlangıçta hava fazlalık katsayısı 1.4 daha sonra 1.2 olarak uygulanması önerilmiştir.

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1. Introduction

The escalating energy demand throughout the world has led to an increase in the use of coal and biomass resources. The use of coal in energy production has created various environmental problems, such as CO₂, SO₂, and NO_x emissions. Fluidized bed combustion technology is the cleanest coal combustion method available in the world (Yuansheng and Mengshu, 2021). Fluidized bed boilers (FBBs) are widely used in industrial applications due to their environmentally friendly and clean combustion technology, fuel diversity, and high thermal efficiency (Jain, Groulx and Basu., 2010; Jiang, Li, Yao, Zhang, Lu, Huang and Yue., 2022). In practice, lower SO₂ and NO_x emissions than other boilers make FBB attractive (Basu, 2006). However, although they are known worldwide and widely used in practice, the start-up of fluidized bed boilers is unpredictable and could be based on experience.

There are various problems related to the running of fluidized bed boilers. One of these problems is heating the boiler to reach solid fuel ignition temperature. In practice, there are four methods to heat fluidized bed boilers to the start-up temperature. These methods are over-bed burner, pre-mixed gas, in-bed fired oil burner, and in-duct burner beneath the bed (Moodie and Vickers, 1985; Xin, Changsui, Yufeng and Xiaoping, 1998). The method with the shortest start-up time and the lowest oil or gas consumption is an in-duct burner beneath the bed (Vakkilainen, 2016). In the start-up process, the fuel is fed after the solid fuel ignition temperature is reached. A large amount of heat energy is required to achieve the solid fuel ignition temperature or gasification temperature (Pitsukha et al., 2017; Carlo et al., 2022). For this reason, the start-up time needs to be as short as possible (Ma, Guo, Chang, Yue and Zhang., 2015). In FBBs, the start-up time is related to parameters such as the amount of fuel used in the ignition, the area and contact distance of the flame to the bed material, and the hot exhaust gas flow rate (Basu and Ghosh, 2007). Xin et al. (1998) mentioned the importance of problems experienced during start-up in fluidized bed boilers. They established a fluidized bed boiler experiment setup to simulate the start-up process of the fluidized bed boiler. Then, they compared the experimental data with the simulation data. The study revealed operational parameters such as fluidized bed temperature when the coal is fed, coal feed rate, fluidized bed heat transfer surface area, static bed height, and ignition air flow rate. Jain et al. (2010) investigated the effect of bed material amount on bed heating time in a bubbling fluidized bed boiler with an over-bed burning start-up. They stated that the start-up time increases as the bed material decreases due to the increased distance between the burner flame and the material. There are few studies on FBB start-up parameters in the literature (Ma et al., 2015; Pitsukha et al., 2017; Jain et al., 2010; Xin et al., 1998). A significant research gap exists in the literature, particularly in in-duct burner start-up parameters beneath the bed. This is a crucial aspect that requires further exploration, and its potential benefits could revolutionize the efficiency of small-capacity fluidized bed boilers and gasifiers. Conversely, existing studies have primarily focused on the operation parameters of large-scale

industrial bubbling fluidized bed (BFB) boilers with capacities exceeding 500 MWe. However, most BFBs used in practical applications are of small capacity, typically less than 10 MW_{th}. In the context of high-capacity fluidized beds, the H/D_{bed} (where H denotes the static bed height and D_{bed} denotes the bed diameter) ratio has been shown to be small (Basu, 2015). Consequently, a comparison of start-up parameters with small-capacity combustion systems may prove to be an inadequate assessment tool.

Bubbling fluidized bed technology has proven highly effective for small-capacity industrial systems, particularly for the incineration and gasification of waste materials (Han et al., 2022; Miccio, Raganati, Ammendola, Okasha and Miccio., 2021; Batuecas, Serrano, Horyat and Abelha., 2023). Furthermore, the evolution of technology and environmental constraints suggest that biomass and waste combustion technologies will offer significant advantages over coal. Small-capacity BFBs and fluidized bed gasifiers will continue to expand (Ammendola and Scala, 2012; Koornneef, Junginger and Faaij., 2007; Nascimento, Gonzalez, Lora, Ratner, Palacio and Reinaldo., 2021). Fluidized bed technology has been successfully applied to processes such as coal and biomass gasification, pyrolysis, fluidized catalytic cracking, acrylonitrile, polyethylene, oxychlorination and polycrystalline silicon. This list is likely to continue with current efforts in chemical looping, polymer pyrolysis, methane pyrolysis and propane dehydrogenation, among others (Chew, LaMarche and Cocco., 2022). The present study focuses on small-capacity fluidized bed biomass systems with a thermal output of less than 10 MW_{th}, which are frequently employed in decentralized heat and cogeneration (CHP) systems, small-scale biomass gasifiers, and local waste/biomass incineration applications. Despite the emphasis on the suitability of systems within this capacity range for flexible and intermittent operation in recent years, studies focusing on start-up time and transient behavior appear to be limited (Erbaş, O. 2019; Chen et al., 2024).

Small-capacity fluidized bed boilers and fluidized bed gasifiers start-up frequently due to their operating conditions. As the amount of bed material increases in the start-up process, the amount of heat that must be supplied to the system to reach the ignition temperature also increases. In small-capacity BFBs, the amount of bed material in the combustion chamber during operation is generally also applied in the start-up process. However, start-up with the amount of bed material used for the operation should be discussed. Moreover, the amount of air applied to the start-up also influences the process time. In fluidized bed combustion systems, the start-up process, particularly in operations involving frequent start-stop cycles, results in additional energy consumption due to the reheating of the bed material and reactor mass. As indicated by extant literature, the amount of auxiliary fuel utilized during the initiation phase can account for between 5–15% of the total energy input per cycle, contingent upon the prevailing operating conditions (Basu, 2015; Banarjee, Shahnam, Rogers and Hughes., 2023). Consequently, the reduction of start-up time is regarded as a pivotal domain for enhancement with respect to minimizing total energy consumption, particularly in

small-scale fluidized bed systems that are managed flexibly and intermittently. In addition, there is no study on the effect of burner excess air coefficient (EAC) and static bed height parameters on the start-up time of small size BFB boilers with under bed duct burners. Until a few decades ago, fluidized bed combustion systems were utilized in large-capacity applications. As large-capacity systems do not necessitate frequent start-ups and shut-downs, the parameters influencing start-up time have not been the focus of significant research. However, in the contemporary context, there has been a notable advancement in clean combustion technologies, particularly in the domains of biomass combustion and gasification systems. It is evident that these systems possess a relatively limited capacity and consequently necessitate frequent start-ups. Consequently, research on start-up time and parameters has been limited in recent literature. The present study investigates the impact of adjusting the burner air-fuel ratio on the start-up process, without the necessity for additional systems. Adjustments may be affected by modifying the quantity of combustion air supplied to the burner. In fluidized bed combustion/gasification systems, flow rate adjustment is possible by communicating the flow meter, which is already present, with the automation system to drive an inverter-controlled blower. In circumstances where inverter control is not an option, adjustment can be facilitated by modifying the damper settings.

2. Material and Method

The numerical approaches used to calculate two-phase flows are quite complex. As a result, the researchers have developed their unique exact or analytical solution as well as numerical approaches (Abbas, Nadeem, Saleem, Issakhov, Abdel-Sattar and Aly., 2021). Analytical solutions can be used for the determination and simplification of the boundary conditions of the problem. This study, analytical model, numerical model and experimental method were used to determine the effect of static bed height and burner air excess coefficient on start-up time. The present study adhered to the standards of research and publication ethics.

2.1 Analytical model used in bed design

Static bed height, minimum fluidization velocity, bed height at minimum fluidization, and operating velocity depend on the capacity of the fluidized bed boiler, coal type and coal size feeding distribution. The amount of bed material required in the combustion chamber during operation can be analytically calculated. This bed material is usually also used in the start-up process. Therefore, this study estimated that static bed height was applied in practice. These parameters are determined by analytical calculations.

The pressure drop in the fixed bed consisting of particles of the same size can be found by the Ergun equation (Ergun, 1952). The equation consists of the viscous and energy losses sum as stated in Eq 1. Where H is the height of the static bed (m), u is the fluid superficial velocity (m/s), and ρ_g and μ_g are the density (kg/m^3) and dynamic viscosity (kg/ms) of the gas, respectively. In the study, the proper-

ties of air at average bed temperature ($T_{ave} = 250^{\circ}\text{C}$) were used as the gas fluid. The ϵ in the Ergun equation is the void fraction of the bed and is obtained by dividing the total void volume by the total bed volume. For randomly packed beds of spherical particles, the static bed void fraction typically lies in the range of 0.36–0.40, as widely reported in classical packed-bed and fluid flow literature (Ergun, 1952; Kunii and Levenspiel, 1991). When the void fraction was restricted to the range typical of randomly packed spherical particles, the predicted pressure drop per unit height varied between –15% and +25% relative to the mean value. is the particle sphericity ratio of bed solids, and a sphericity ratio of 0.8 was applied, assuming that all sand particles were equal.

$$\frac{\Delta P}{H} = 150 \frac{(1-\epsilon)^2}{\epsilon^3} \frac{\mu_g U_g}{(\Phi_s \bar{d}_p)^2} + 1.75 \frac{(1-\epsilon)}{\epsilon^3} \frac{\rho_g U_g^2}{(\Phi_s \bar{d}_p)} \tag{1}$$

where $\bar{d}_p$ is the mean particle diameter (m), and

$$\bar{d}_p = \frac{1}{\sum(\frac{x_i}{d_{pi}})} \tag{2}$$

and , is the mass fraction of particles in the bed (Basu, 2006).-

The average diameter and the type of particles used as bed material affect fluidization. Geldart Group B-type sand particles fluidize easily, and bubbling occurs as soon as they exceed the minimum fluidization velocity (Basu and Fraser, 1991). Therefore, sand particles in Geldart group B were used in the study. Sand sieve analysis was conducted for silica sand; the average particle diameter was 414 μm (Table 1).

Table 1. Particle size distribution of sand sieve analysis

Particle size (mm)	Material retained (g)	Percentage retain (%)	Cumulative percentage retain (%)
1.8	0.00	0.00	0.00
0.85	4.99	0.55	0.55
0.6	322.51	35.22	35.77
0.425	131.58	14.37	50.14
0.3	364.55	39.82	89.96
0.212	54.30	5.93	95.89
0.15	27.48	3.00	98.89
0.106	7.65	0.84	99.72
Pan	2.53		
Fineness modulus of sand (mm)			0.414463

The pressure drop in the fluidized bed can be expressed as,

$$\left[\begin{matrix} \text{Buoyancy} \\ \text{of gas} \end{matrix} \right] = \left[\begin{matrix} \text{Total weight of} \\ \text{particles} \end{matrix} \right]$$

this can be expressed mathematically as follows,

$$\Delta PA_b = A_b H_{mf} (1 - \varepsilon_{mf}) (\rho_p - \rho_g) g \quad (3)$$

where ρ_p is the particle density, 2500 kg/m³ is taken for Geldard B sand (Basu and Fraser, 1991), and A_b is the fluidized bed cross-sectional area (m²). The speed at which fluidization begins in the fluidized bed is called the minimum fluidization velocity (U_{mf}). H and ε_{mf} are the bed height and the void fraction of the fluidized bed at the minimum fluidization velocity, respectively.

The minimum fluidization velocity can be calculated as follows,

$$\frac{1.75}{\Phi_s \varepsilon_{mf}^3} Re_{mf}^2 + \frac{150(1 - \varepsilon_{mf})}{\Phi_s^2 \varepsilon_{mf}^3} Re_{mf} = Ar \quad (4)$$

where Re_{mf} and Ar are the Reynolds number and Archimedes number at the minimum fluidization velocity, respectively. The calculation of Re_{mf} and Ar is performed following Equations (5) and (6):

$$Re_{mf} = \frac{\bar{d}_p U_{mf} \rho_g}{\mu_g} \quad (5)$$

$$Ar = \frac{\bar{d}_p^3 (\rho_p - \rho_g) \rho_g g}{\mu_g^2} \quad (6)$$

Using the approaches in Equation 7 proposed by Wen and Yu (1966), the minimum fluidization velocity can also be found from the Reynolds number.

$$\frac{1}{\Phi_s \varepsilon_{mf}^3} \cong 14, \quad \frac{1 - \varepsilon_{mf}}{\Phi_s^2 \varepsilon_{mf}^3} \cong 11 \quad (7)$$

can be expressed as follows.

$$Re_{mf} = (33.7^2 + 0.0408 Ar)^{(1/2)} - 33.7 \quad (8)$$

The static bed height is related to the height and the void fraction at minimum fluidization velocity (Oka, 2003) (Eq. 9).

$$\frac{H_{mf}}{H} = \frac{1 - \varepsilon}{1 - \varepsilon_{mf}} \quad (9)$$

The expanded bed height H_b is found from the following equation proposed by Kuni and Levenspiel (1991).

$$H_b = 1.38 H_{mf} \quad (10)$$

The parameters of bench scale fluidized bed boiler design are shown in Table 2.

The design parameters are determined by analytical calculations. The properties of the air at the average bed temperature are used ($T_{ave} = 250^{\circ}\text{C}$). In addition, the minimum bed temperature required for fuel feeding in the start-up process is assumed to be 480°C (De Diego et al., 2013; Stultz and Kitto, 1992). The sensitivity of the minimum fluidization velocity to temperature was assessed by evaluating the dispersion of the analytical values at 200 and 300°C with respect to the reference condition at 250°C . The measured values were found to be 0.09628, 0.08998, and 0.08496 m/s at 200, 250, and 300°C , respectively. Subsequent to an uncertainty analysis using the sample standard deviation ($n-1$), a standard deviation of 0.00800 m/s was obtained, corresponding to a relative standard deviation of approximately 8.9% with respect to the reference value at 250°C . The observed level of variation is considerably higher than that recorded for bed height measurements (0.30%). This finding suggests that the minimum fluidization velocity is significantly more sensitive to temperature changes.

Table 2. Design parameters of experiment start-up system setup

Parameter	Value
Maximum boiler capacity	93.04 kW (for Turkish lignite, LHV = 9215 kJ/kg)
Bed diameter	0.4 m
Minimum fluidization velocity	0.1024 m/s ($T_{ave} = 250^{\circ}\text{C}$)
Static bed height (H)	0.3603 m
Expanded bed height (Δ)	0.4689 m

It is known that the H/D_{bed} ratio decreases as the bed diameter increases. It is predicted that, given the diameters of the small-capacity beds (less than 10 MW_{th}) examined in this study, the aspect ratio will vary between 0.5 and 1.0. Consequently, aspect ratios of 0.55, 0.65, and 0.90 were selected.

2.2 Numerical method and experimental setup

In the start-up process, computational fluid dynamics (CFD) analyses were carried out to determine the effect of hot exhaust gases on the sand heating time according to different excess air coefficients. MP-PIC (multiphase particle-in-cell) method was used to model a multiphase flow of a gas-solid mixture. The mass and momentum relations used in the MP-PIC method are given in Eq. 11 and 12, respectively (Batchelor, 1988; Karimipour and Pugsley, 2012; Gidaspow, 1994; Gürel and İpek, 2019; Williams, 1985). In addition, the SIMPLE algorithm and LES (Large Eddy Simulation) turbulence model were applied in CFD analyses.

$$\frac{\partial(\varepsilon\rho_g)}{\partial t} + \nabla(\varepsilon\rho_g U_g) = 0 \quad (11)$$

$$\frac{\partial(\varepsilon\rho_g U_g)}{\partial t} + \nabla(\varepsilon\rho_g U_g U_g) = -\varepsilon\nabla\rho_g + \varepsilon\mu_g \nabla^2 U_g - F + \varepsilon\rho_g g \quad (12)$$

where g is the gravity (m/s^2), U_g is the gas velocity (m/s) and F is the rate of momentum exchange per unit of volume between the fluid and particles. The Liouville equation (Eq. 13) expresses the time-dependent particle distribution function,

$$\frac{\partial f}{\partial t} + \nabla_x(fU_p) + \nabla_{U_p}(fA) = 0 \quad (13)$$

where x , U_p , and A represent particle position, particle velocity (m/s) and particle acceleration (m/s^2), respectively. The particle acceleration formula is as follows,

$$A = \frac{1}{\rho_p} D(U_g - U_p) - \frac{1}{\rho_p} \nabla_x P + g - \frac{1}{\rho_p \epsilon_p} \nabla_x \tau \quad (14)$$

D represents drag function ($\text{kg/m}^3\text{s}$), P represents pressure (kPa), ϵ_p represents particle volume fraction, and τ represents shear stress between particles (N/m^2).

In drag models, the force on the particle is obtained using the relation given in Eq. 15,

$$F_p = m_p D(U_g - U_p) \quad (15)$$

where m_p is the particle mass (kg). The study used Wen-Yu as a drag model (Atlis, Rohmat, Saptoadi and Sentanuhady., 2022; Wen and Yu, 1966). Usually, the gas-solid momentum transfer is carried out by incorporating a semi-empirical drag coefficient correlation obtained from the experimental study, typically represented by Wu, Liu, Ma and Chen. (2018). In the Wen-Yu model, the drag function is related to the drag coefficient (Eq. 16). D is found with the relations given in Eq. 18, depending on the Reynolds number (Eq. 17).

$$D = C_d \frac{3}{4} \frac{\rho_g}{\rho_p} \frac{(U_g - U_p)}{\bar{d}_p} \epsilon^{-2.65} \quad (16)$$

$$Re_p = \frac{\rho_g (U_g - U_p) \bar{d}_p}{\mu_g} \quad (17)$$

$$C_d = \begin{cases} \frac{24}{Re_p} \epsilon^{-2.65} & Re_p < 0.5 \\ \frac{24}{Re_p} \epsilon^{-2.65} (1 + 0.15 Re_p^{0.687}) & 0.5 \leq Re_p \leq 1000 \\ 0.44 \epsilon^{-2.65} & Re_p > 1000 \end{cases} \quad (18)$$

The fluid to particle heat transfer coefficient ($\text{W/m}^2\text{K}$) in the fluidized bed is expressed as follows,

$$h_p = \left((c_0 Re_p^{n_1} Pr^{0.33} + c_1) \frac{k_g}{\bar{d}_p} + c_2 \right) \quad (19)$$

where β , γ , δ , and ϵ are model constants and are 0.37, 0.1, 0, and 0.6, respectively (Bar-racuda, 2015; Fan and Zhu, 1998). The correlation used is also in agreement with the experimental studies of Levenspiel and Kunii (1991). The Prandtl number (Pr) is expressed as follows,

$$Pr = \frac{\mu_g c_{p,g}}{k_g} \tag{20}$$

here $c_{p,g}$ is the specific heat of gas fluid (J/kgK) and k_g is the conductivity (W/mK).

In the start-up process, numerical analyses were conducted to determine the sand heating time according to operating parameters (EAC and static bed material height). The flow volume of the bed was obtained, and the boundary conditions in Figure 1 were applied to the created 3D numerical model. Mesh independence analysis was performed. According to the mesh independence analysis, it was deemed suitable to generate 100,198 meshes for the model (Table 3).

Table 3. Mesh independent analysis results

Mesh element number	Bed material average temperature@500 s
20,520	169.45
41,900	172.89
65,250	175.36
100,198	177.19
142,985	177.55

Experimental data were used to validate the analyses. An experimental setup was established for validation. Figure 2 shows the schematic drawing of the experimental setup. The experimental setup consists of a blower, start-up oil burner (44.23 kWh, 5 lt/h diesel fuel with 42 MJ/kg LHV), thermal mass flowmeter (0.5% RD + ± 1.5% FS), differential pressure transmitter (± 0.5% FS), K type (± 0.15%) thermocouples, combustion chamber, and seven nozzles for homogeneous gas distribution. Thermocouple T1 was used to measure the windbox temperature, and T2 and T3 were used to measure the average sand temperature of the fluidized bed. The surfaces in contact with flame after the burner location and the windbox were manufactured with H-solid insulation material. The combustion chamber was insulated with rock wool, and an adiabatic wall was assumed in the analysis.

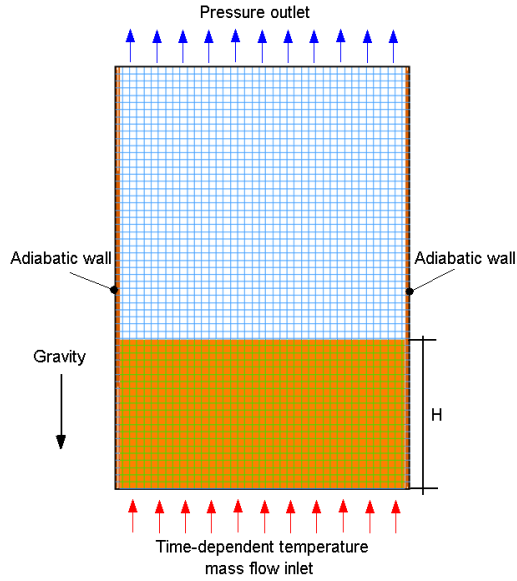


Fig. 1 Boundary conditions of the model

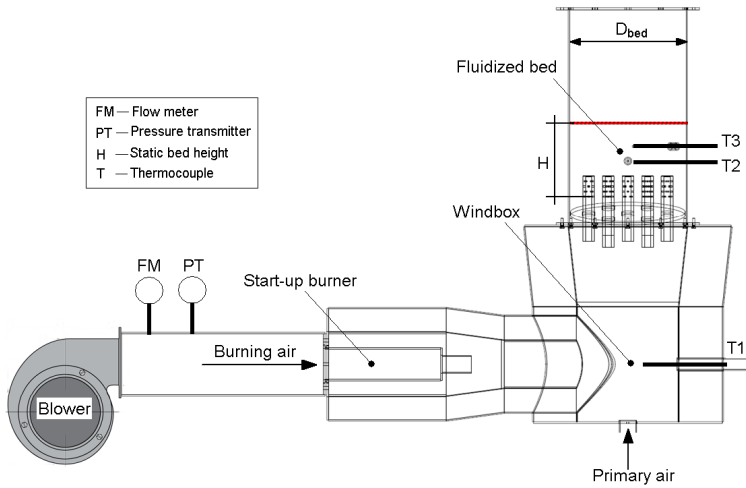


Fig. 2 Experimental setup of fluidized bed

Time-dependent temperature data were used in numerical analysis. The data was obtained by experiment. The time-dependent temperature data of the experiment are given in Figure 3. The equation shown in the figure was obtained with the temperature data measured from the T1 thermocouple located in the windbox. Then, the equation used for time-dependent mass flow inlet boundary

condition in the analyses. The windbox temperature gradually increases because the walls of the fluidized bed boiler are initially cold. Hot exhaust gases produced from the start-up burner heat the cold walls of the boiler and the windbox. It then passes through the nozzles and heats the bed material.

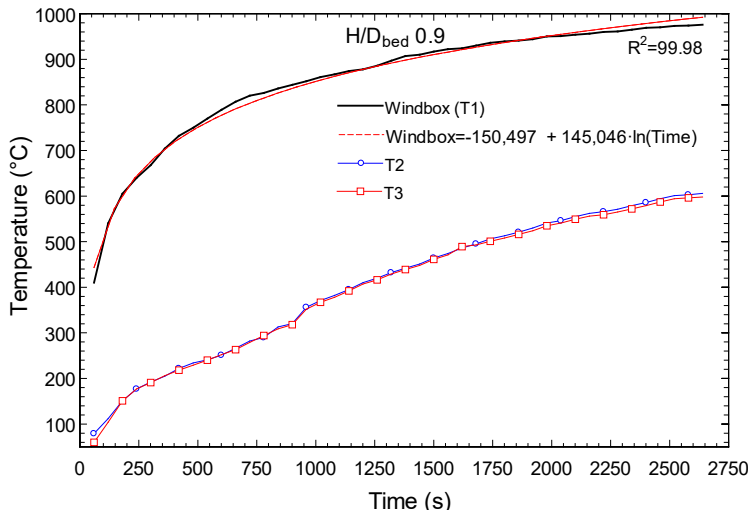


Fig. 3 Experimental time-dependent temperature data

The experiment was carried out in EAC 1.2 conditions. In the analysis, the method given in Figure 4 was used to calculate the time-dependent hot gas temperature values in the windbox with the change of EAC. Here, the windbox was considered a mixing chamber. The first law was applied to the hot exhaust gases (EAC 1.2) and clean air in the mixing chamber (Eq.21). Thus, time-dependent mixture gas temperatures were calculated for EAC 1.4 and 1.6. In Equation 21, h is enthalpy and kJ/kg . The temperature was obtained from the enthalpy of the mixture gas ($\quad$), and the temperatures were entered in the analyses depending on time.

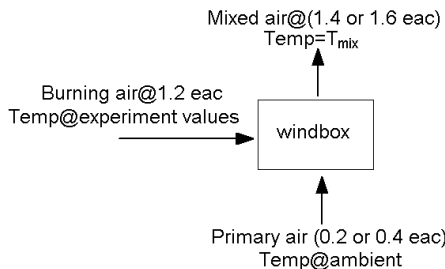


Fig. 4 Windbox mixing chamber

$$\dot{m}_{@1.2\text{ eac}} h_{@T1} + \dot{m}_{\text{primary,air}} h_{@ambient,temp} = \dot{m}_{\text{mix}} h_{\text{mix}} \quad (21)$$

3. Results and Discussion

Numerical analysis was validated with data from the experiment for bed aspect ratio (H/D_{bed}) 0.9 and EAC 1.2. Numerical analyses were then carried out in EAC 1.2, 1.4, and 1.6 for bed aspect ratios 0.55, 0.65, and 0.90. This way, the effect of different excess air coefficients and static bed heights on the start-up time was observed. Figure 5 compares the temperature data from thermocouples T2 and T3 with the average sand temperature data calculated by numerical analysis. The analysis used cross-sectional planes to obtain the average sand temperature from the bed material (Figure 6). The average temperatures of the particles contacting the planes formed at 150 mm intervals (ΔX) from the base of the bed material were taken. The average particle temperatures in the planes were taken for all planes, and the average sand temperature was found by taking the arithmetic averages. It is seen that the analysis and experimental temperature data are in good agreement. However, temperature differences between the experiment and analysis data vary depending on time and reach a maximum of 30°C. Nevertheless, the experimental data are at a lower temperature than the numerical analysis data. The experimental system loses heat through the walls despite its good insulation. The temperature difference was recorded 900 seconds prior. The difference is negligible in the subsequent period. In the simulation, the gas flow is specified to be normal to the boundary, and its composition is described as homogeneous. However, the reality is that this process occurs through the utilization of nozzles. The existence of a temperature difference, especially prior to the occurrence of bubbling, could provide a possible explanation for this phenomenon. Prior to the process of bubbling, the rate of heat transfer is predominantly high in the sand particles located near the nozzles. Following the bubbling process, a more homogeneous distribution was formed, and the difference between the simulation and the experiment narrowed.

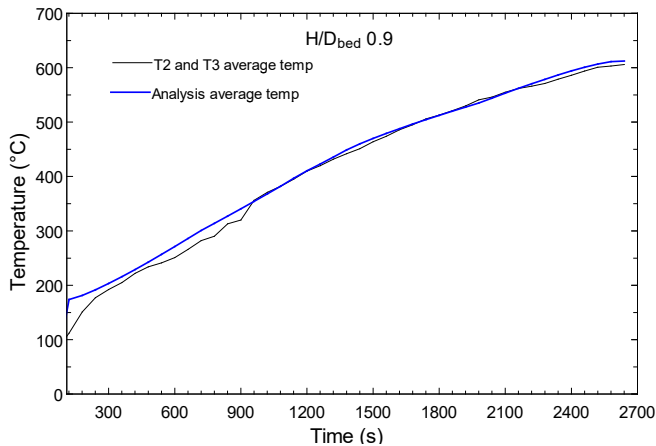


Fig. 5 Numerical analysis validation results

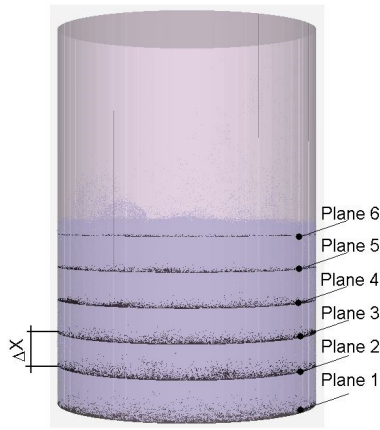


Fig. 6 The method used to obtain average sand temperature in analyses.

The U_g/U_{mf} ratio, a crucial parameter, is a significant indicator of the fluidization effect of hot gases in the fluidized bed. Figure 7 shows the effect of different excess air coefficients on the time-dependent bed material temperature and U_g/U_{mf} ratio for an aspect ratio of 0.9. Up to 1200 s, the U_g/U_{mf} ratio for EAC 1.2 is less than the EAC 1.4 and 1.6. This can be seen in Figure 8. Therefore, slightly bubbling was observed for EAC 1.2 at 1000 seconds. Between 1250 and 1650 seconds, U_g/U_{mf} ratios are very close. Then, the average particle temperature for EAC 1.2 was observed to be the highest. This is likely to be caused by the high mixture of gas temperature. The bed material reached the reference temperature (480°C) in 1390, 1695, and 1913 seconds for EAC 1.2, 1.4, and 1.6, respectively. The results obtained for EAC 1.2 are like the study by Wöss et al. (2009). As the rate of airflow to bed grows, the start-up time increases due to the decreasing mixture temperature in the windbox. It took 22% more time at EAC 1.4 and 44% more time at EAC 1.6 to reach the desired temperature compared to EAC 1.2. It is observed that, for an aspect ratio of 0.9, the use of EAC 1.4 leads to a faster increase in bed material temperature until the 837th second. Based on this behavior, a staged operating strategy using EAC 1.4 up to 837 s followed by EAC 1.2 can be recommended to shorten the start-up duration.

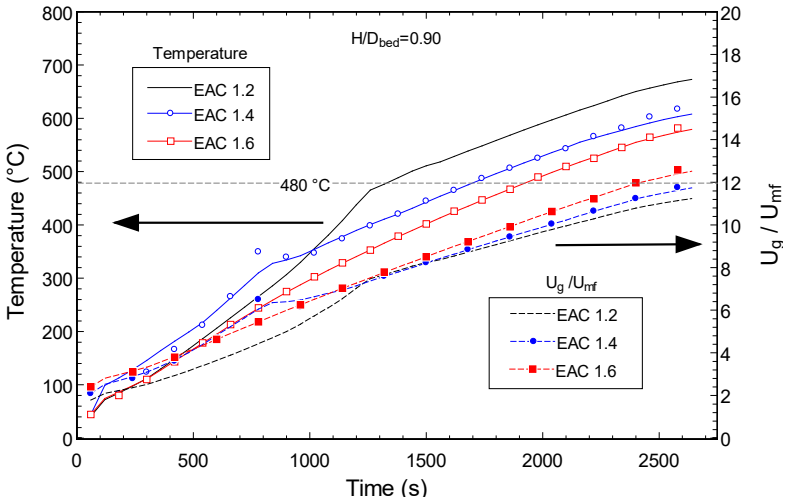


Fig. 7 Variation of bed material temperature and U_g/U_{mf} ratio with time at EAC for $H/D_{bed} = 0.90$

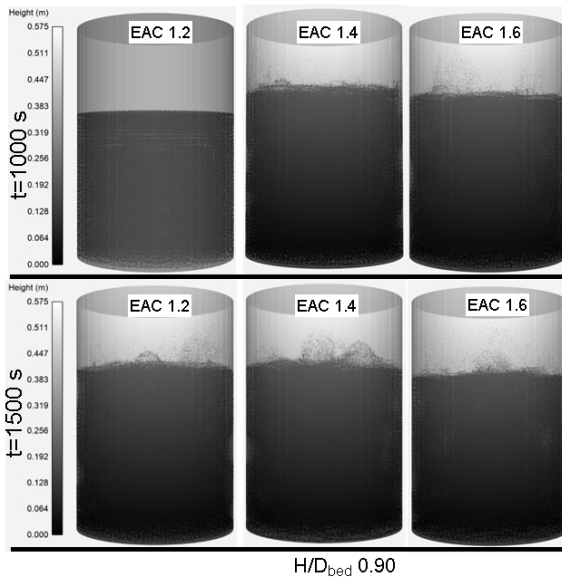


Fig. 8 Variations of expanded bed height with time at different EAC for $H/D_{bed} = 0.90$

Figure 9 shows the effect of different excess air coefficients on the time-dependent bed material temperature and U_g/U_{mf} ratio for an aspect ratio of 0.65. The increase in the air excess coefficient caused a rise in the U_g/U_{mf} ratio. However, it was interesting to note that the bed material temperature decreases contrary to the U_g/U_{mf} ratio, highlighting the complex dynamics in this process. This is

caused by the decrease in the temperature of the hot gas mixed in the windbox with the increase in the air excess coefficient. In EAC 1.6, a bigger bubble size had been observed since the beginning of the process (Figure 10). The bed material reached the reference temperature in 1168, 1407, and 1627 seconds for EAC 1.2, 1.4, and 1.6, respectively. It took 20.4% more time at 1.4 and 39% more time at 1.6 to reach the desired temperature compared to EAC 1.2. Compared to the aspect ratio of 0.9, the time difference between EACs was reduced.

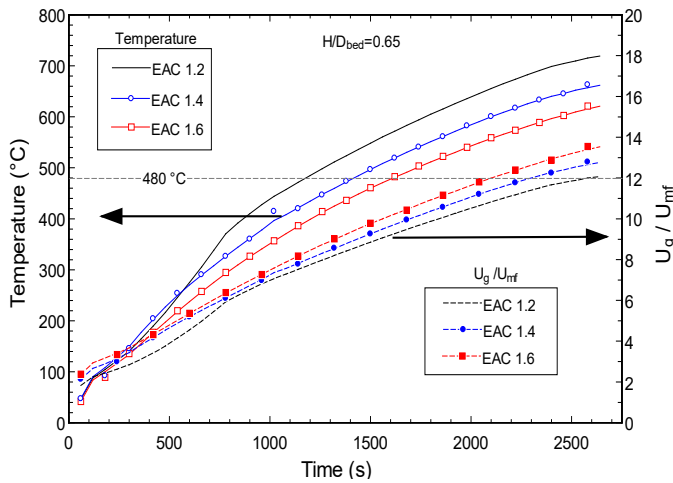


Fig. 9 Variation of bed material temperature and U_g/U_{mf} ratio with time at different excess air coefficients for H/D_{bed} 0.65

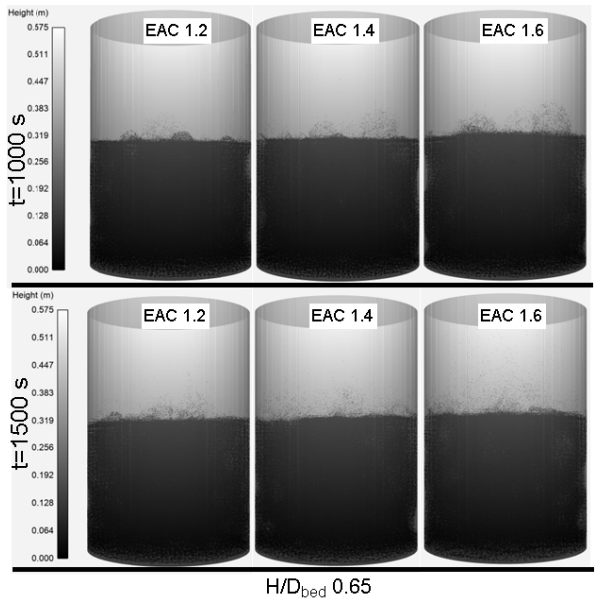


Fig. 10 Variations of expanded bed height with time at different excess air coefficients for H/D_{bed} 0.65

Figure 11 shows the effect of different excess air coefficients on the time-dependent bed material temperature and U_g/U_{mf} ratio for an aspect ratio of 0.55. Similarly, the increase in the excess air coefficient caused a rise in the U_g/U_{mf} ratio. However, the U_g/U_{mf} ratio difference for different EACs is more significant at H/D_{bed} 0.65 and 0.9. The variations of the expanded bed height with time are given in Figure 12. The bed material reached the desired temperature in 1080, 1269, and 1440 seconds for EAC 1.2, 1.4, and 1.6, respectively. It took 17.5% more time at 1.4 and 33% more time at 1.6 to reach the desired temperature compared to EAC 1.2. Compared to the ratios of 0.65 and 0.9, the time difference between EACs decreased.

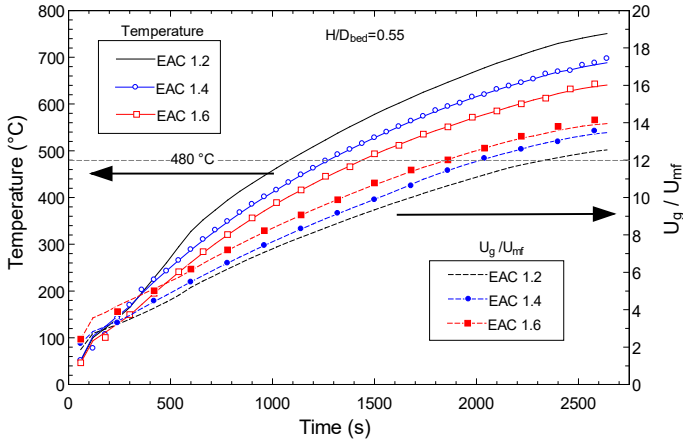


Fig. 11 Variation of bed material temperature and U_g/U_{mf} ratio with time at different excess air coefficients for H/D_{bed} 0.55

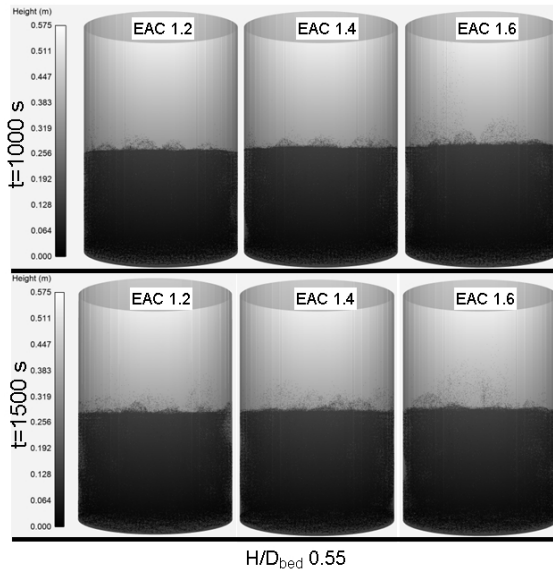


Fig. 12 Variations of expanded bed height with time at different excess air coefficients for H/D_{bed} 0.55

4. Conclusion

The operation of bubbling fluidized bed systems is usually based on experience. There needs to be more literature, especially on the start-up process. The start-up process is energy intensive. Therefore, the process is desired to take a short time. For this purpose, this study investigated the effect of start-up burner excess air coefficient and bed aspect ratio on the start-up process of a bubbling fluidized bed boiler. A bench-scale experimental set was installed. The analytical method was applied to determine the bed aspect ratio. The experiment was carried out at EAC 1.2 for the calculated aspect ratio. Experimental and numerical results were compared and validated. The process was then analyzed using numerical methods for different aspect ratios and EACs.

The results of the analysis show that the start-up time shortens as the aspect ratio decreases (averaging 32%). The results agree with the study of Xin et al. (1998). This is expected as the amount of sand in the bed is reduced. On the other hand, as the EAC increases, the start-up time is prolonged (up to 44%). As the amount of bed material increases, the U_g/U_{mf} ratio is in close ranges for different EACs. Therefore, the importance of the air excess coefficient increases. As the aspect ratio decreases, the time difference between EACs also decreases. In other words, as the amount of bed material increases, the effect of EAC becomes more powerful. Another important finding is that the bed material temperature increases faster in EAC 1.4 until 837th second at an aspect ratio of 0.9. Applying EAC 1.4 until the 837th second and then 1.2 is recommended for a shorter start-up process. Therefore, while the results are valid for similar size units (less than 10 MW_{th}), extrapolation of this data to very large scales should be undertaken with caution.

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