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Fluidized Bed Straw Pellets Combustion with Minimal Emissions of Carbon Monoxide

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The type of air entry distribution is essential in the formation of hydrodynamic structures of fluidized bed. The present study is aimed to form a hydrodynamic structure of fluidized bed optimal for burning low-grade solids fuels. In order to determine the uniformity of the fluidized bed in the bed heat exchange rate between the bed and a heat exchange member immersed therein was measured. At a first approximation, heat exchange with the member immersed into the bed imitates heat exchange between the fluidized bed and a fuel particle. It is allowed to determine that an optimal hydrodynamic structure of the bed can be formed by using a lattice supplying the major portion of air near the furnace wall. The above conclusion was confirmed by experiments using a real furnace with a fluidized bed.

1. Introduction

The use of fluidization technique in heat-and-power engineering provides a comprehensive solution for reducing environmental pollution with noxious emissions, reducing size and metal consumption of boiler units, and increasing reliability thereof without excessive fuel quality and performance stability requirements (Borodulya and Vinogradov, 1980; Radovanovic, 1990; Leckner, 2015; Guangxi; 2015; Nowak, 2015).

2. Problem Formulation

However, the hydrodynamic structure of fluidized bed is far from uniform, and said non-uniformity increases further along with an increase in the amount of fluidization. In industrial scale apparatuses, fluidization gas can be supplied via regularly spaced but discrete devices (nozzles, tubes, etc.). In this case, gas bubbles would rise along certain preferred lines, and local vertical solid particle circulation cells would form, hindering complete mixing of particles and gas in the bed (Eights, 1986;; Liu et al., 2009; Tian et al., 2009). With regards to combustion of fine solids fuels and solid fuels with high volatile matter content, the formation of preferred gas bubble ascent lines can lead to burning of fine solids particles and gas flare formation that penetrates the bed and burns chaotically in the freeboard (Eights, 1986).

3. Problem Solution

Particle movement in a fluidized bed can be affected by providing non-uniform entry gas distribution through a distribution lattice. The above method provides strong circulatory particle movement of "Gulfstream movement" type and more effective gas mixing in the bed when supplying greater gas flow through the peripheral area of the gas distribution lattice, as opposed to the central area thereof (Eights, 1986).
Evidently, the approach of providing intensive internal particle circulation in the bed by means of irregular air entry distribution is the simplest and most effective solution. However, results of systematic studies of hydrodynamic structure of fluidized bed with the purpose of discovering optimal air entry distribution in the context of liquid fuel combustion and combustion of fuels with high volatile matter content are not yet available.

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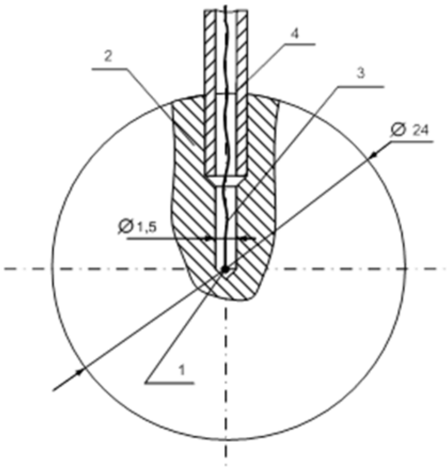

Figure 1. Diagram of sensor for determining heat exchange rate: 1 – thermocouple head with diameter of 1.2 mm, 2 – copper ball, 3 – thermocouple cables, 4 – connecting tube with diameter of 3.0 mm

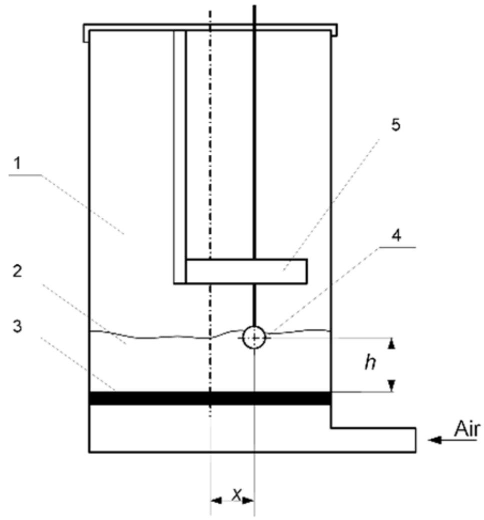

Figure 2: Diagram of sensor arrangement for measuring heat exchange rate in a fluidized bed. 1 – chamber with fluidized bed, 2 – fluidized bed of sand, 3 – gas distribution lattice, 4 – sensor, 5 – sensor positioning system

3.1 Methods and equipment

The hydrodynamic structure of a fluidized bed can be studied using various methods, each having certain advantages and disadvantages.

The results of a fluidized bed hydrodynamic structure study based on measuring heat exchange rate between the bed and a heat exchange member immersed therein (in various areas of the fluidized bed) are of particular practical interest. At a first approximation, heat exchange with the member immersed into the bed imitates heat exchange between the fluidized bed and a fuel particle, which is important for evaluation of heating and combustion velocity of the fuel particle in the bed, as well as for eliminating local overheating of the bed, fuel ash fusion and bed defluidization.

External heat exchanged between the fluidized bed and the surface (copper ball D = 24 mm) was measured using a conventional regular temperature conditions method (Lykov, 1967). A type L thermocouple was welded into the ball center, said thermocouple allowing for continuous temperature measurement of the pre-heated ball after immersing it into the cold fluidized bed. The sensor is shown schematically in Figure 1.

In order to determine heat exchange coefficient distribution in the fluidized bed, a sensor positioning system shown in Figure 2 was used. The sensor positioning system in the fluidized bed chamber allows placing the copper ball both on chamber axis and at a variable distance x from the axis. The height h of the ball position above the gas distribution lattice was adjustable. During the experiments, the sensor was placed at a height of 30 mm and 80 mm above the lattice. Quartz sand with particle size of 0.5…0.8 mm was used as the particulate material (d = 0.65 mm; u_{mf} = 0.26 m/s).

In order to provide various types of air entry distribution, three perforated lattices were used. The first configuration of lattice comprises a regular open area (the lattice forms a "flat" air velocity profile at bed entry), the second configuration comprises a greater open area in the center of lattice (the lattice forms a "convex" entry air velocity profile), and in the third configuration, the greater open area is arranged at the periphery (the lattice forms a "concave" entry air velocity profile).

Prior to measurement, the copper ball was heated with a hot air source to 200-240 °C, then the ball was immersed into target location in the bed, and the reduction in copper ball temperature was determined until the temperature equaled that of the fluidized bed. In order to determine the effect of air velocity in the chamber, all experiments were carried out at three different air filtering velocities u = 0.49; 0.61 and 0.73 m/s.

The air velocityat the entrance to the empty apparatus was measured by using a thermo-anemometer (Delta OHM HD 2103-1. The probe of the anemometer was mounted at a height of 5 mm at the – gas distribution lattice.

The error in determining the heat transfer coefficient did not exceed 10 %.

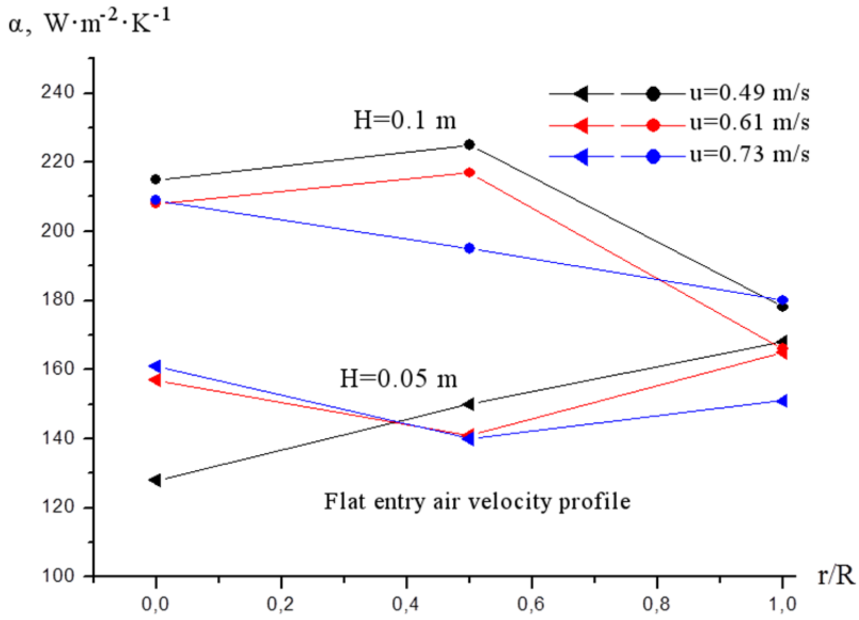

Figure 3.1: Heat exchange coefficient values between the fluidized bed and the heat exchange member plotted against air velocity at uniform entry air distribution (Flat entry air velocity profile).

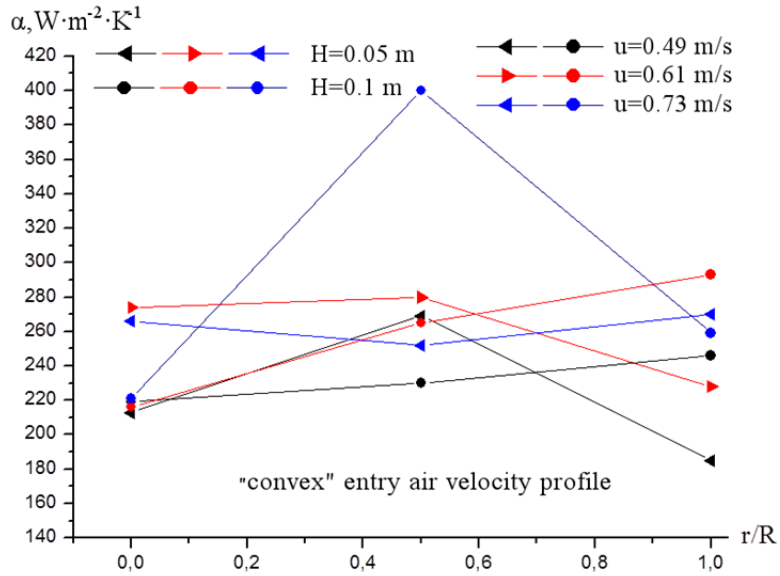

Figure 3.2: Heat exchange coefficient values between the fluidized bed and the heat exchange member plotted against air velocity in a "convex" entry air velocity profile.

3.2 Results and discussion

Figures 3.1 - 3.3 show mean values of heat exchange coefficients between the sensor and the fluidized bed for three types of air distribution lattice and two initial bed height values (0.05 m and 0.1 m). In case of the "flat" entry air velocity profile (Figure 3.1), the hydrodynamic structure of the bed was non-uniform due to the fact that at the initial bed height of 0.05 m and air velocity of 0.49 m/s, heat exchange rate near the wall of the apparatus is 1.31 times higher than that in the center of the apparatus. An increase in air velocity to 0.61 m/s does not lead to an increase in uniformity, as the heat exchange coefficient value at half-radius point of the apparatus is 1.19 times lower than near the apparatus wall. Finally, at air velocity of 0.73 m/s, the non-uniformity is maintained, as the heat exchange rate in the center of the apparatus is 1.16 times higher than in other zones of the fluidized bed.

In case of the "flat" entry air velocity profile, the non-uniformity of the fluidized bed hydrodynamic structure is
In case of a "convex" entry air velocity profile, maintained when the initial bed height is increased to 0.1 m, as the heat exchange rate in the center of the apparatus and at half-radius point thereof is 1.1 - 1.35 times higher than in peripheral zones of the bed. Furthermore, at the initial bed height of 0.1 m, heat exchange rate between the bed and the heat exchange member immersed therein decreases with an increase in the velocity of air being passed through the bed (Figure 3.1).

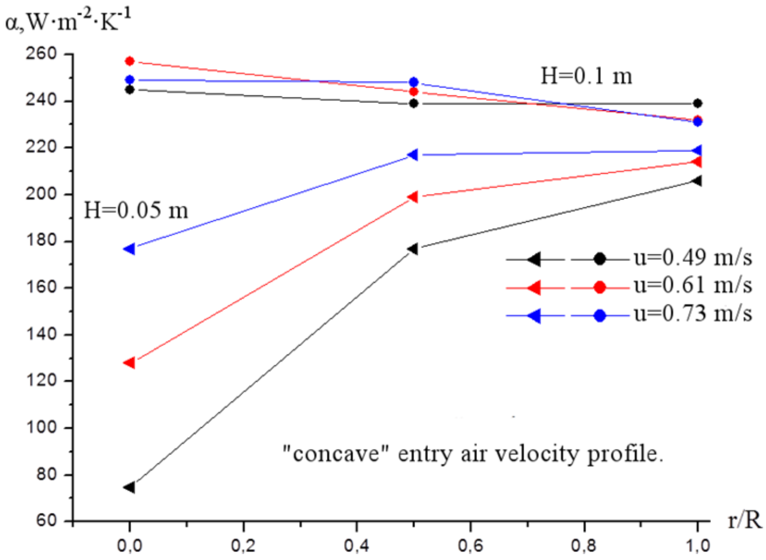

Figure 3.3: Heat exchange coefficient values between the fluidized bed and the heat exchange member plotted against air velocity in a "concave" entry air velocity profile.

In case of a "convex" entry air velocity profile, the rate of heat exchange processes in the bed is significantly higher compared to the "flat" entry air velocity profile (Figure 3.2). At air velocity of 0.49 m/s, the heat exchange rate in the fluidized bed with initial height of 0.05 m is 1.5 - 1.6 times higher than in the "flat" profile at the same air velocity. However, at the initial bed height of 0.05 m, the hydrodynamic structure of the bed remains non-uniform, as at the half-radius point of the apparatus, the heat exchange rate is 1.3 - 1.4 times higher than at the center and near the apparatus wall, respectively. An increase in air velocity to 0.61 m/s leads to an increase in uniformity of the fluidized bed hydrodynamic structure: heat exchange rate values in this case are in the range of 230 to 280 $W \cdot m^{-2} \cdot K^{-1}$. Said increase can be explained by the fact that a strong circulatory movement forms in the bed, spreading over the entire bed volume. However, a further increase in air velocity to 0.73 m/s leads to a decrease in heat exchange rate to 250 - 270 $W \cdot m^{-2} \cdot K^{-1}$. When the initial bed height is increased to 0.1 m in the "convex" entry air velocity profile, said increase does not in fact affect heat exchange rate and particle movement rate. In other words, the "convex" profile forms a uniform structure with approximately equal particle movement rates over the entire bed volume. However, this is only true for relatively low velocities of air being passed through the bed. When air velocity is increased to 0.73 m/s, a sharp spike in heat exchange rate (up to 400 $W \cdot m^{-2} \cdot K^{-1}$) occurs approximately at the central radial point of the apparatus, whereas heat exchange rate in other zones of the bed remains the same.
In case of a "concave" entry air velocity profile, when air velocity increases, the heat exchange rate also increases when compared to the "flat" entry air velocity profile (Figure 3.3). However, the case of a "concave" entry air velocity profile, there is a significant non-uniformity in the hydrodynamic structure of the bed with initial height of 0.05 m, which is completely eliminated upon increasing air velocity to 0.73 m/s. When the initial bed height is increased to 0.1 m, the "concave" entry air velocity profile exhibits heat exchange coefficient values fluctuations in an extremely narrow range of 230 to 260 $W \cdot m^{-2} \cdot K^{-1}$, which indicates uniformity of the bed hydrodynamic structure at any velocity of air being passed through the fluidized bed in the studied air velocity range. Furthermore, when using the "concave" entry air velocity profile, at bed height of 0.1 m, the heat exchange rate is 1.5 - 3.5 higher than that in the bed with initial height of 0.05 m.
It should also be noted that the possibility of maintaining a uniform hydrodynamic structure of the fluidized bed upon an increase in air velocity in the "concave" entry profile suggests that said air velocity profile allows providing a furnace with a wider power adjustment range without decreasing solid fuel combustion efficiency.

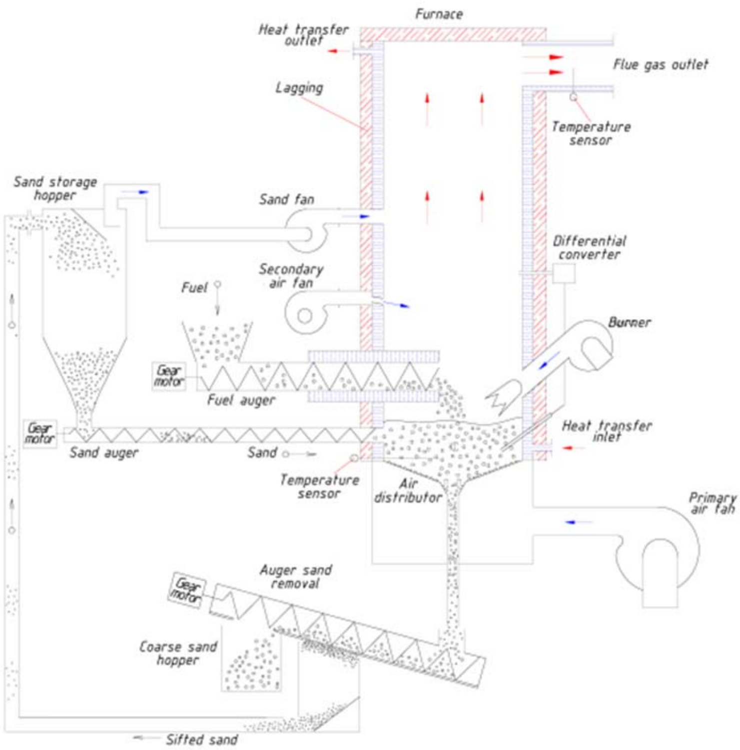

Figure 4. Functional scheme

3.3 Experimental verification of results in practice
In order to verify and substantiate the choice of air distribution lattice forming at the bed entry a "concave" air velocity profile, a prototype furnace was constructed with a fluidized bed with a capacity of 400 kW, the design of which is illustrated in the scheme of Figure 4. The furnace is rested upon the air distribution grill; blasting primary air is forced under this grill. The caps for supplying blast air are uniformly arranged over the entire lattice area. However, due to the fact that air distribution lattice is tilted at 30° towards the center, the resistance of inert material bed or particulate catalyst over the caps arranged on the lattice periphery is lower than that over the central part of the lattice. Therefore, the peripheral part of the lattice experiences greater air flow compared to the central part thereof, thus forming a "concave" entry air velocity profile and providing intensive circulation of bed particulates. The furnace was used to burn straw pellets (diameter – 7 mm, ash content – 6,87 %, low combustion value – 15,52 MJ/kg).

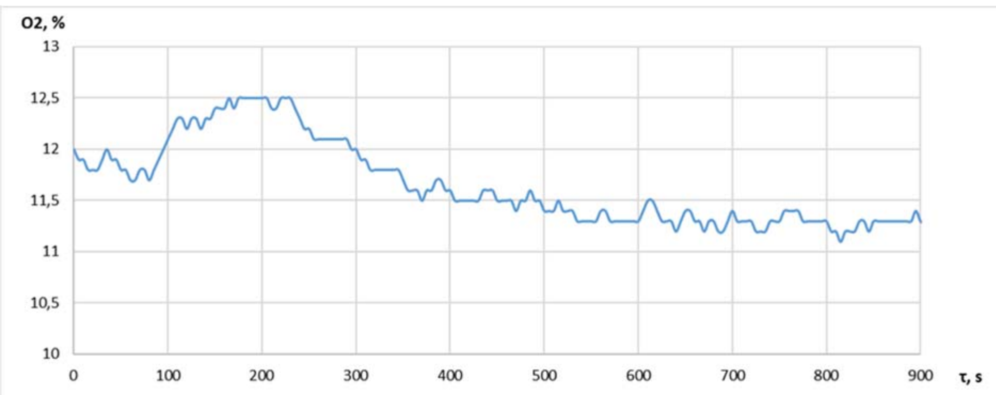

Figure 5. Oxygen content change in furnace gas at furnace exit during combustion of straw pellets.

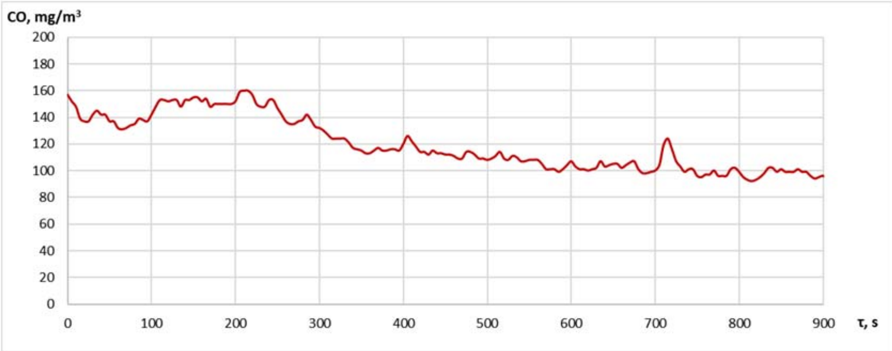

Figure 6. Carbon monoxide content change in furnace gas at furnace exit during combustion of straw pellets.

The furnace was tested in nominal operation. During testing, a «VarioPlus» gas analyzer was continuously used to measure oxygen and carbon monoxide content in furnace gas at furnace exit. The burning of the straw pellets was happening at the gas temperature of 605-625 °C at furnace exit. As seen in the Figure 5, in 300-400 seconds after warming the furnace and initiating straw pellet supply, fuel combustion is stabilized and maintained with furnace gas oxygen concentration changing in the range of 11 – 11.5 %. As can be seen from figure 6, the burning of straw pellets in the experimental furnace with fluidized bed is accompanied by an extremely low emission of CO in the atmosphere (100 – 160 mg/m3 when the concentration of oxygen in flue gases equals 6 % or 52 – 84 mg/m3 when the concentration of oxygen in the flue gas equals 11.4 %). The average concentration of NO_x is 750 mg/m^3 or 394 mg/m^3 when the concentration of oxygen in combustion gases equals 11,4 %. The content of sulfur oxides during combustion of straw pellets was almost equal to zero.

4. Conclusions

Conducted studies of changes in the values of local heat transfer coefficients in a fluidized bed allowed to choose the type of air entry distribution as the optimum for combustion solid fuel with a high volatile matter content. Selected type of air entry distribution supplying the major portion of air near the furnace wall. It provides intensive circulation of solid particles and gas in the bed. This intense circulation provides almost complete combustion of the volatile matter in the bed volume and low emissions of carbon monoxide in the atmosphere by burning biofuels.

Acknowledgments

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