Co-Fluidization of Fine Particles and Straw Pellets at Room and Elevated Temperatures

Rafail L. Isemin, Sergey N. Kuzmin, and Dmitry M. Viryasov

Abstract—**The results of the research on fluidization of multi-solid materials in cold and hot models have been introduced. It turned out that the change graphs of the statistical characteristic of pressure fluctuation can be used for evaluate values of fluidization velocity.**

Index Terms—**Multi-solid bed, Pressure fluctuation, Straw pellets combustion, Turbulent fluidization.**

I. Introduction

The technology of dedicated biomass combustion and coal and biomass co-combustion in a fluidized bed is increasingly applied aiming to achieve a better control over a combustion process with a simultaneous decrease in greenhouse gases, ash and sulfur oxides emissions into the atmosphere. Straw is one of the most easily available biomass resources and its energy utilization is also constantly expanding requiring longer transport distances of the material. This stimulates the production of straw pellets, with a bulk density of 650-750 kg/m³ over 100-150 kg/m³ of baled straw.

Fuel combustion can be carried out in a co-fluidized or multi-solid fluidized bed of fine coal ash particles or straw ash and straw char particles loaded with straw pellets. A multi – solid fluidized bed is a circulating fluidized bed in which an entrained bed consisting of fine particles is transiting through a fluidized dense bed of coarse particles, i.e. a bubbling bed in the bottom of the riser [1]-[4].

The value of the minimum fluidization velocity of particles forming the bed is required for the design of fluidized bed furnaces. However, as has been proven [5]-[14] it is not possible to apply the known method of a minimum fluidization velocity evaluation from the pressure drop through a bed versus gas flow rate curve for a confined, multi solid or binary particle mixtures fluidized bed.

The purpose of the present study is to define a method for the experimental evaluation of the minimum fluidization velocity of a bed compounded of fine and coarse particles mixtures at room and elevated temperatures.

II. Experimental

A. Cold Model Unit

Coal ash particles and straw pellet beds were subjected to analysis. The particle size distribution of coal ash were as follows: a mass fraction of particles in the size of: up to 1.0 mm – 20.87%; from 1.0 to 1.2 mm – 61.79%; from 1.2 to 1.5 mm – 5.66%; from 1.5 to 1.7 mm – 2.7%; from 1.7 to 2.0 mm – 1.81%; from 2.0 to 2.5 mm – 1.14%; from 2.5 to 3.0 mm – 0.47 %; from 3.0 to 4.0 mm – 4.07%; more than 4.0 – 1.49%. The moisture content of ash particles was 5.35% on the average; the density of coal ash particles was 1200 kg/m3. Straw pellets had the following characteristics: the granule diameter was 6 mm, the pellet average length to diameter ratio was 0.59; the pellet particle density was 1190 kg/m3.

The fluidization experiments were conducted by means of an apparatus (Fig. 1) with the rectangular cross-section of 194 mm × 485 mm and height of 1500 mm which was rested upon an air distribution grill with an open area of 5%. The air flow rate was measured by a thermo-anemometer Delta-OHM HD 2103-1after the air left the apparatus. Not less than 100 measurements of the air flow rate were taken in each experiment. The pressure drop in the bed was measured by means of a differential micro manometer which allowed taking 1200 measurements of pressure drop within 60 seconds. A digital signal from the micromanometer was transmitted to a personal computer for the subsequent processing.

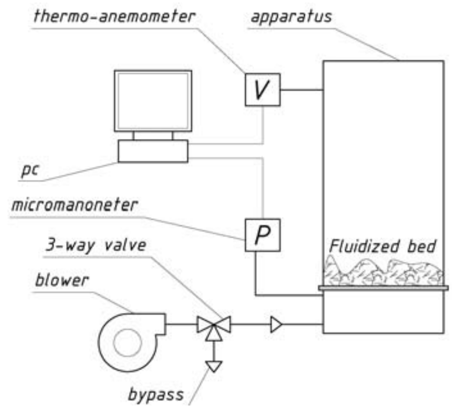

Fig. 1. Schematic diagram of the cold experimental unit

In the course of the pilot experiments it was established that if the content of coal ash particles in a bed was higher than 40 %, the complete segregation of particles by size was evidenced, in this case pellets remained motionless and rested on the air distribution grill. For this reason, the pressure fluctuations were measured in beds containing 100%, 95%, 90%, 85%, 80%, 70%, 65% and 60% of pellets and corresponding amount of coal ash.

Changes of bed behavior were recorded by a video camera Panasonic DVC 30. Then each second of the video recording was broken into 50 video shots that allowed receiving consecutive images of bed behavior in every 0.04 second.

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B. *Hot Model Unit*

To carry out the experiments we used an experimental boiler plant with an experimental hot water boiler of 200 kW capacity (Fig. 2), a fuel storage bunker, fuel supply system, and automation and control equipment. In order to prevent low-melting eutectic formation it is proposed to combust straw pellets in a fluidized bed which is formed by pellets alone and solid products of their combustion (their char particles and ash).

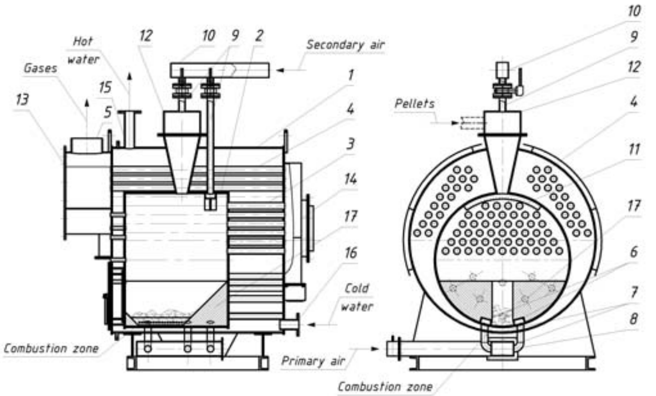
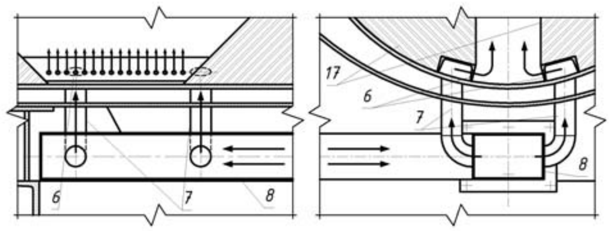


Fig. 2. Experimental boiler (hot model) diagram



a) b)

Fig. 3. Air distribution grill system and primary air supply diagram: a) longitudinal section; b) cross section

The boiler consists of a cylindrical body 1 with a fire tube 2 which is connected by a network of short 3 and long heating tubes 4 to a smoke stack 5 where flue gases are removed from the boiler and released to a smoking pipe (not shown). In the bottom part of a fire tube there is an air distribution grill 6 consisting of two channels punched on the sides (Fig. 3). Primary air is supplied to the grill by a network of pipes 7 from a collector 8 which is connected to a forced-draught fan (not shown in the figure). In the top part of a boiler there is a network of pipes 9 for secondary air supplying which is connected by a collector 10 to a secondary air fan (not shown). In the same top part there is a branch pipe 12 for granule feeding into the boiler furnace which is connected to a granule feeding system (not shown). The exit end of pipes 9 leading to the furnace is covered by a channel 11, blanked at the ends and punched on its flange, faced to the longitudinal axis of a boiler. Turning type smoke boxes 13 and 14 are positioned on the end and back walls of the boiler. In the top part of the boiler cylindrical body 1 there is a branch pipe 15 for the removal of water heated in a boiler. An inlet branch 16 is in the bottom part of a boiler from the side of the back smoke box.

The boiler operates as follows: through an inlet branch 16, the boiler space is filled with water which, heating up, flows out through a branch pipe 15. Biofuel is fed through a branch pipe 12 to a fire tube 2 where it is fluidized by means of blast air supplied under a fuel bed through punched channels 6. Flue gases flow to a smoke stack 5 and a smoking pipe through short 3 and long 4 heating tubes. At the same time water in the boiler body 1 heats up, while moving flue gases rotate by 180 degrees in a back smoke box 14 for more complete flue gas heat recovery. These gases are collected in a front smoke box 13 before they reach a smoke stack 5. Air supplied for combustion of granules is divided into two flows: primary air is supplied through a collector 8 and a network of pipes 7 under channels 6. Secondary air is supplied through a collector 10 and a network of pipes 9 under the channel 11 by the perforations, where it passes through its flange to a fire tube and is used for afterburning of volatile matters. The portion of secondary air flow is also supplied into a branch pipe 12, through which fresh granules are fed into the boiler. This air is also used for afterburning of volatile matters, and also protects a granule feeding system from penetration of flame and hot gases and prevents ignition of granules before they enter the boiler.

Measurements were taken during boiler operation at workload equal to 50%, 75% and 100% of the nominal. The workload was controlled by the variation in fuel flow rate with a corresponding change in air supply for combustion.

Measurements of pressure drop in the bed were taken by a differential micro manometer. The obtained range of random values of the pressure drop in a bed was exposed to a statistical analysis. In addition, the mean value of the pressure drop in the bed during the observation period:

$$\Delta P_m = \Sigma(\Delta P_i / N) \tag{1}$$

And the root-mean-square deviation of the pressure fluctuation:

$$\sigma = \sqrt{\Sigma(\Delta P_i - \Delta P_m)^2 / N}\ . \tag{2}$$

Samples of materials were taken out of the bed every 40 and 90 minutes for particle size distribution determination; then, the combustible content of each fraction was determined.

III. Results and Discussion

Results obtained on the cold model (at room temperature).

Fig. 4 shows the change of the pressure drop compared to the air flow rate in the bed, for different bed mixtures. The shaded line on the diagrams indicates the range of the air flow rate values at which the process of bed fluidization U_{mfi} begins and the bed becomes completely fluidized U_{mf}, based on visual observations. The value of the minimum fluidization velocity decreases with the increase of the coal particle fraction in the mixture: the minimum fluidization velocity is 2.4 m/s for a 100% pellet bed and 1.75 m/s – for a bed with 40% of coal ash particles.

As Fig. 4 shows, it is impossible to apply the dependence diagram $\Delta P_m = f(U)$ for the evaluation of the minimum fluidization velocity for coal and pellet particle mixtures. When pellet content in the mixture is in the range from 80 to

100 %, the pressure drop in the bed constantly increases at the increase of the air flow rate, and the portion of the curve parallel to an X-axis, which usually corresponds to a condition of fluidization, isn't observed. If the pellet content in the mixture is lower, some peaks on the influence curve $\Delta P_m = f(U)$ were observed, which obviously correspond to the transition of separate fractions into a fluidization condition, but it is still impossible to evaluate from these curves the value of the air flow rate which would correspond to the complete fluidization of a bed.

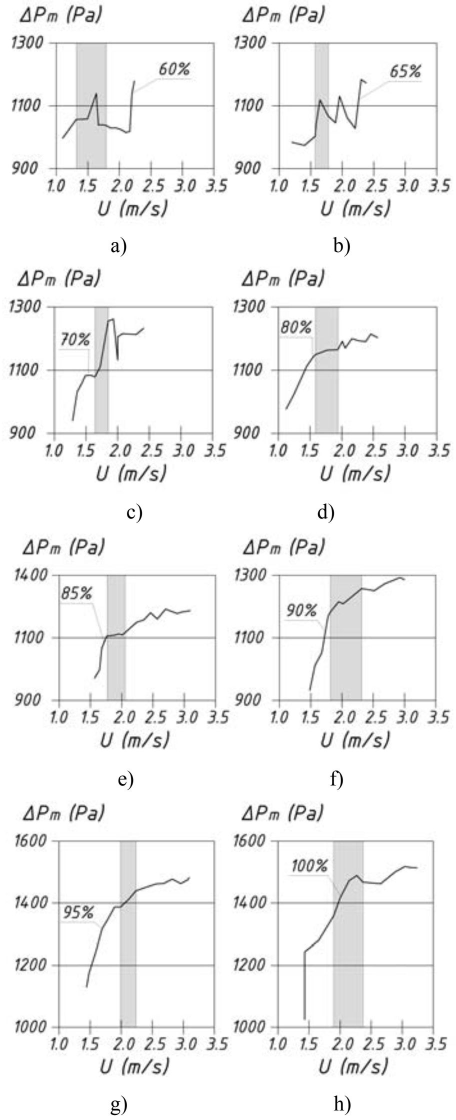


Fig. 4. Dependence of pressure drop in the bed vs. air velocity for mixtures of: a1) 60 and 40%, b) 65 and 35%, c) 70 and 30%, d) 80 and 20%, e) 85 and 15 % f) 90 and 10%, g), 95 and 5% pellets and particles of coal ash respectively, h) for a bed of straw pellets

Fig. 5 shows the dependences $\sigma = f(U)$. For mixtures with a pellet content of 80-100%, the dependence diagram $\sigma = f(U)$ can be divided into three parts: a part, where $\sigma \approx 0$, a part where a weak increase of σ is observed with the increase of U, and a part where a fast increase of σ is observed with the increase of U. For mixtures with pellets content lower than 80%, the first portion is practically absent on the diagram, but the 2-nd and 3-rd are clearly distinct. The comparison of Fig. 4 with the results of the visual observation allows to draw the conclusion that the value of the air flow rate corresponding to the transition of the dependence $\sigma = f(U)$ from the first portion on the curve to the second is the velocity for which fluidization just begins. The air flow rate corresponding to the transition of the dependence $\sigma = f(U)$ from the second part of the curve to the third one is the velocity for which the bed becomes completely fluidized. Hence, the dependence diagram $\sigma = f(U)$ can be applied for the experimental evaluation of U_{mf}.

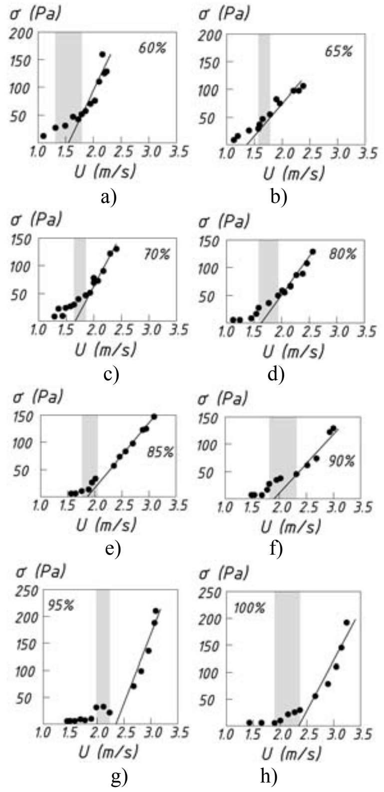


Fig. 5. The dependence of the standard deviation of differential pressure fluctuations vs. of air velocity for mixtures of: a) 60 and 40%, b) 65 and 35%, c) 70 and 30%, d) 80 and 20%, e) 85 and 15 % f) 90 and 10%, g), 95 and 5% pellets and particles of coal ash respectively, h) for a bed of straw pellets

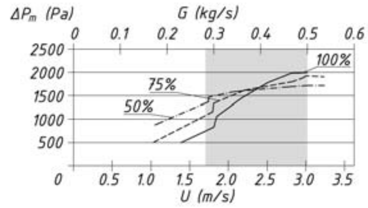


Fig. 6. Pressure drop in a burning bed vs. air flow rate

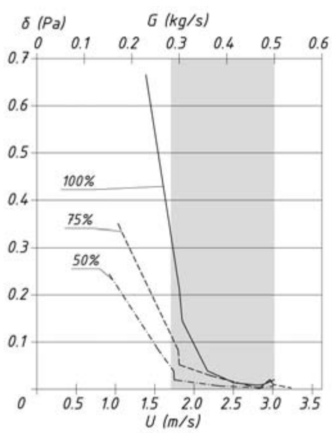


Fig. 7. Dimensionless amplitude of pressure fluctuations vs. primary air flow rate at 50%, 75% and 100% boiler workload of the nominal

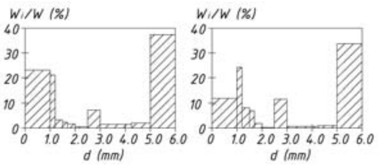
For suggestion that the dependence diagram $\sigma = f(U)$ can be applied for the experimental evaluation of U_{mf} was initially proposed in [15]. However, the cited paper investigated the fluidization of monodisperse particle beds, while the present work demonstrates that this method can also be used to estimate the minimum fluidization velocity for a mixture of fine and coarse particles.

Results obtained on the hot model (at temperature around 1200°C).

The fluidization of burning straw pellets and solid products of their combustion starts with an air flow rate equal to 0.35-0.60 kg/s (this zone of bed transition from fixed condition to fluidized is shown by hatching in Fig. 6, Fig. 7 and Fig. 8).

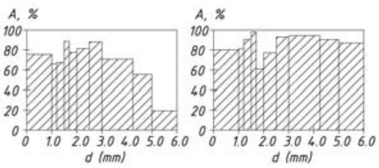
The transition of a bed of burning pellets into a fluidized condition was characterized by sharp drop of values of dimensionless amplitude of pressure fluctuations δ (Fig. 7). Hence, the dependence diagram $\delta = f(U)$ can be applied for the experimental evaluation of U_{mf}.

One can observe insignificant changes in the particle size distribution during the course of the experiment: 90 minutes after the experiment begun, the PSD of the bed was the same as after 40 minutes (Fig. 8).



a) b)

Fig. 8. Fractional composition of a bed material: a) after 40 minutes of start-up; b) after 90 minutes of start-up



a) b)

Fig. 9. Ash content in a bed material: a) after 40 minutes of start-up, b) after 90 minutes of start-up

Moreover, as it appears from Fig. 9, of the longer experiment time the sharper the growth of the ash content for the large fraction (by 1.5-4.0 times). Therefore, even if small agglomerates are formed in a bed, the combustion process of char does not stop and this allows the assumption that fuel loss due to mechanical incompleteness of combustion is not significant.

Thus, the abrupt change in the numerical values of statistical characteristics of the pulsations of the pressure drop in the bed at a high temperature as well as at room temperature can be used as a criterion for evaluating the minimum fluidization velocity. But at a high bed temperature, one can see a sharp decline, not growth of the numerical values of these statistical characteristics.

To explain this fact the velocity of turbulent fluidization start for the mixture of pellets and fine particles of coal ash and the composition of the bed formed by burning straw pellets at a temperature of about 1200°C (Fig. 8b) is calculated.

The calculation is carried out with the equation proposed in [16]:

$$U_c = U_{mf} + 1.21 \cdot Ar^{0.04} (g \cdot \nu_d)^{1/3} / (Y - 0.3 \cdot Ar^{0.04}) \quad (3)$$

The minimum fluidization velocity can be determined from the following equation [17]:

$$\mathrm{Re}_{mf} = Ar / (1400 + 5.22 \cdot Ar^{0.5}) \quad (4)$$

The effective kinematic viscosity can be determined from equation [16]:

$$\nu_d = 0.000374 \cdot Ar^{0.0764} \quad (5)$$

A value of Y equal to 0.8 [16] was considerated.

Values of U_{ci} were determined for each i fraction of the particles constituting the bed. The value of U_c for the multi - solid bed consisting of i = 2, 3, ... fractions, determined from equation [18]:

$$U_c = ((W_1/W)U_{c1} + (W_2/W)U_{c2} + ...)^{-1} \quad (6)$$

where U_1, U_2, ... – the velocity of transition to a turbulent state of the first, second, etc. fraction of the bed, W_1, W_2, … – the weight of the first, second and so on fraction of the bed and W is the total weight of the bed.

The calculation results are presented in Fig. 10. As can be seen, the calculated value U_c decreases with increasing temperature. At a temperature of about 1200°C the calculated values of U_c are 1.4-1.5 m/s. Consequently, when the gas velocity is greater than 1.4-1.5 m/s turbulent fluidization should be observed (for the fractional composition of the bed, shown in Fig. 8b). This is confirms the results of the experiments presented in Fig. 7.

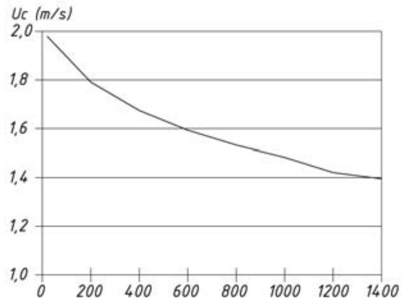


Fig. 10. The dependence of the calculated values U_c on the temperature of the bed, consisting of straw pellets, char and ash particles

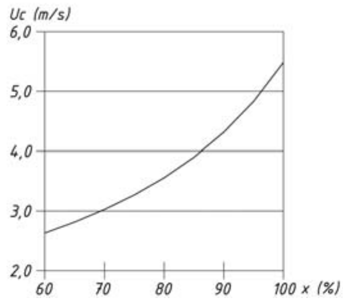


Fig. 11. The dependence of the calculated values U_c on the fraction of pellets in the mixture with particles of coal ash at room temperature

On the other hand, at room temperature, the calculated value of U_c increases if the share pellets in the mixture increases (Fig. 11). U_c for this bed are above the maximum air velocity at which the study was conducted (Fig. 5). That is, at room temperature, the bed of particles of coal ash and pellets (in the investigated range of change the air velocity) should be in a state of bubble fluidization.

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