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CO - FLUIDIZATION OF FINE PARTICLES AND STRAW PELLETS AT ROOM AND ELEVATED TEMPERATURES

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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 to evaluate values of fluidization velocity.

INTRODUCTION

The technology of 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. Fuel combustion can be carried out in a co-fluidized bed of fine straw ash and straw char particles loaded with straw pellets.

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 (1)-(10) 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.

EXPERIMENTAL

Cold model unit

Straw ash particles and straw pellet beds were subjected to analysis. The particle size distribution of 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/m^3. 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/m^3.

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. 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.

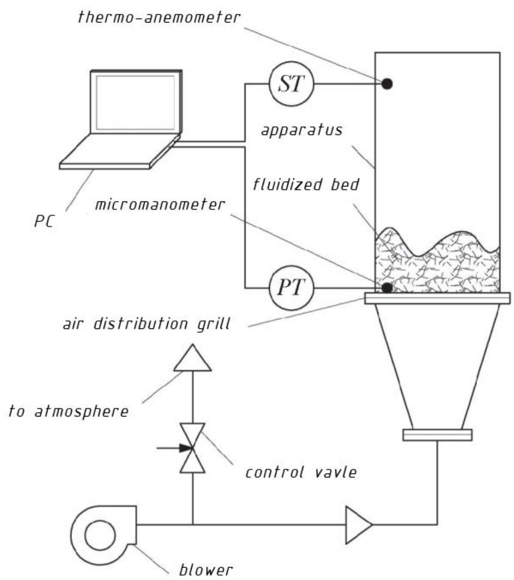


Fig. 1. Schematic diagram of the cold experimental unit

In the course of the pilot experiments it was established that if the content of 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.

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).

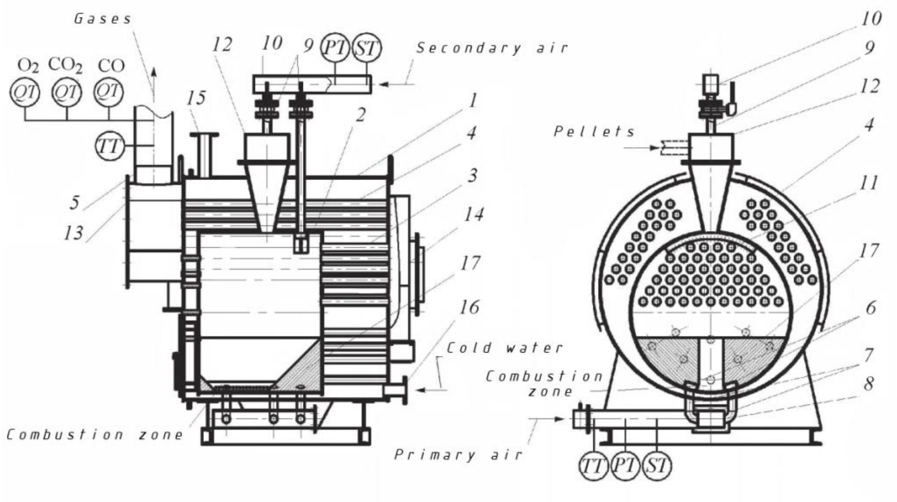
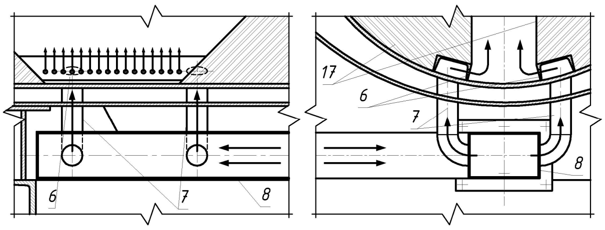


Fig. 2. Experimental boiler (hot model) diagram
1 – cylindrical body; 2 – furnace; 3 – short heating tubes; 4 – long heating tubes; 5 – exhaust gas tube;6 – air distributor; 7 – air pipes; 8 – air collector; 9 – secondary air pipes; 10 – secondary air collector;11 – secondary air distributor; 12 – feeder brunch; 13 – front gases box; 14 – back gases box;15 – outlet water brunch; 16 – inlet water brunch; 17 – ceramic blocks.



a) b)

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

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:

$$\Delta P_m = \Sigma(\Delta P_i / N);$$

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

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

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.

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.

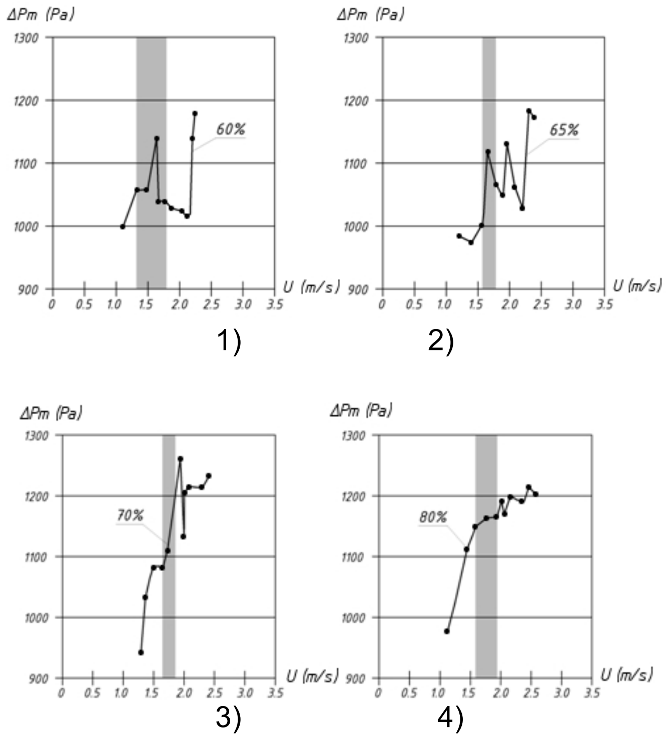


Fig. 4. Dependence of pressure drop in the bed vs. air velocity for mixtures of: 1) 60 and 40%, 2) 65 and 35%, 3) 70 and 30%, 4) 80 and 20% pellets and particles of ash respectively

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 ash and pellet particle mixtures (fig. 4.1 – fig. 4.3).

Fig. 5 shows the dependences $\sigma = f(U)$. The comparison of Fig. 5 with the results of the visual observation allows to draw the conclusion that the dependence diagram $\sigma = f(U)$ can be applied for the experimental evaluation of U_{mf}.

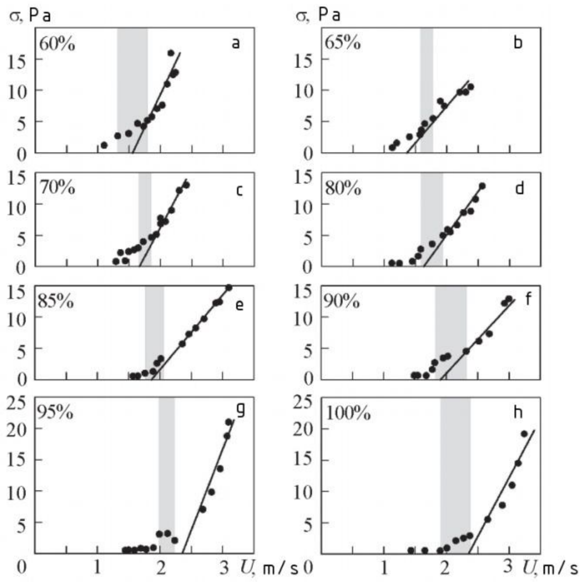


Fig. 5. The dependence of the standard deviation of differential pressure fluctuations vs. of air velocity for mixtures of: a) 60 and 40%, b) 65 and 35%, c) 70 and 30%, d) 80 and 20%, e) 85 and 15 % f) 90 and 10%, g), 95 and 5% pellets and particles of ash respectively, h) for a bed of straw pellets

The suggestion that the dependence diagram $\sigma = f(U)$ can be applied for the experimental evaluation of U_{mf} was initially proposed in (1). However, the cited paper investigated the fluidization of monodisperse particle beds, while the present work demonstrates that this method can also be used to evaluate 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).

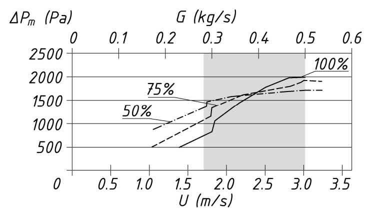


Fig. 6. Pressure drop in a burning bed vs. air flow rate

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

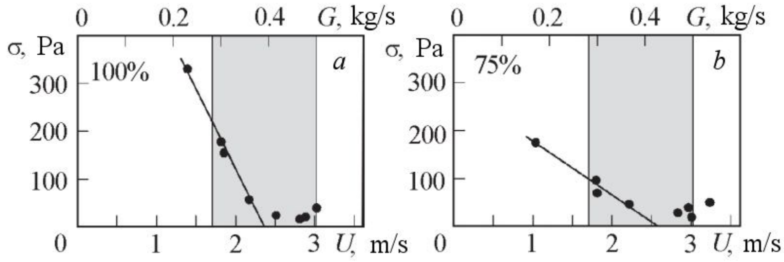
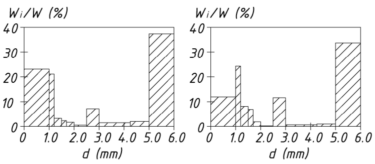


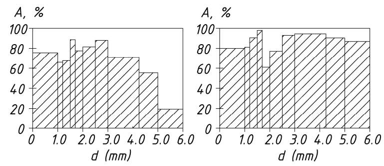
Fig. 7. The dependence of the standard deviation of differential pressure fluctuations vs. of air velocity and air flow rate at 100% (a) and 75% (b) boiler workload of the nominal

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, 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 (12)

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

The minimum fluidization velocity can be determined from the following equation (13)

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

The effective kinematic viscosity of dense phase can be determined from equation (12):

$$\nu_d = 0.000374 \cdot Ar^{0.0764}$$

A value of *Y* equal to 0.8 (12).

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 (14):

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

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 confirms the results of the experiments presented in Fig. 7.

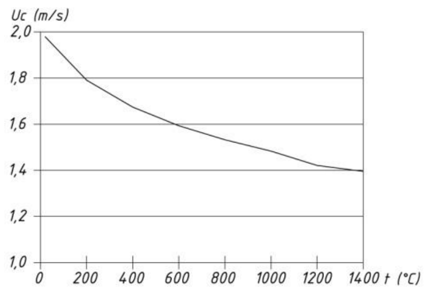


Fig. 10. The dependence of the calculated values U_c on the temperature of the bed, consisting of straw pellets, char and ash particles

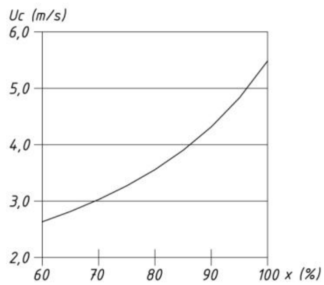


Fig. 11. The dependence of the calculated values U_c on the fraction of pellets in the mixture with particles of coal ash at room temperature

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 is 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.

CONCLUSION

Based on these results the following algorithm was developed to determine the values of gas velocities when the multi-solid bed transfer to fluidized state or to turbulent fluidization state:

1) measure the differential pressure fluctuations in the bed for several values of the gas velocity *U*.

2) determine the values of the standard deviation of the differential pressure pulse for several values of the gas velocity *U*

3) determine the values of the derivative $[\Delta(\sigma_i - \sigma_{i-1})/(U_i - U_{i-1})]$ by numerical differentiation.

4) select two values of the gas velocity for which $(d\sigma / dU) = const$.

5) plot the value of σ corresponding to these gas velocities on the dependence diagram $\sigma = f(U)$; draw through these two points a straight line to the intersection of the line with the abscissa.

6) value of the gas velocity *U*, corresponding to this intersection is complete fluidization velocity (at room temperature) or turbulent fluidization velocity (at elevated temperatures).

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KEY WORDS
Multi-solid bed, pressure fluctuation, straw pellets combustion, turbulent fluidization.