THE TRANSIENT PROCESSES LN CIRCULATING FLUIDIZED BED BOILERS (CFBB)

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The transient processes were studied at pilot (*FBB located in boiler-house of Chalmers University of Technology, Gothenburg, Sweden. CFBB nominal capacity is 12 MW. All walls of the furnace (cross-section 1.7x1.7 m2, height 13.5 m) are provided with membrane tube screens. Front and back screens are covered with lining 110 mm thick. Side screen lining is 2 m height from gas distributor. The furnace is followed by cyclone with lining 1,82 m in diameter. Solid particles from cyclone arrive to external heat exchanger (EHE) and then remove to the furnace.

Fluidized bed of sand particles 0,3 mm in average diameter is made by joint action of primary air flow through gas distributor and air flow from EHE Secondary air is blown into the furnace at the height of 2 m above grate. In the experiments on transient processes in CFBB wood chips 8-10 mm in dimension were used as a fuel. The fuel characteristics were following: W^r = 37.2%, C^d = 51.1% H^d=6.3%, S^d = 0.03%, O^d = 41.4%, A^d = 0.8%, V^{daf}=85%.

The boiler is equipped with automatical control system for measuring more than 100 parameters. There are three main regimens of CFBB in experiments, distinguishing by heat capacity: 100, 80 and 60% of nominal capacity (regimes I, II and III respectively). The velocity of combustion products u=3.5-6.5 m/s in all regimes is sufficiently more than particles terminal velocity. It provides intensive particle mixing in furnace as well as circulation of material in a loop: furnace- cyclone-EHE-furnace. Previous investigations of aerodynamics have shown that

- bottom bed about 0.6 m in height H with porosity $\varepsilon \approx 0.6$ is observed,
- the height of splash zone with density 100 kg/ m^3 is as much as 1-1.5 m,
- average density of freeboard is 5-20 kg/m^3, it increases as well as flow rate of circulating material with gas velocity at the bottom u_o.

The furnace exit temperature at regimes 1 and II has 20-30 °C above than bed's temperature, i.e, fuel combustion occurs in bulk bed and freeboard. The temperature of returned material is lower than furnace exit temperature because of looses of bed heat for heating of air conducted to EHE. Moreover, there is a heat transfer through lining to screen surface of cyclone and EHE. The equations between temperatures of bed t, furnace t_f and circulating material t_s depend on intensity of heat transfer in furnace and cyclone, output of material in external loop and equation between amounts of fuel in bed and freeboard.

An additional experiments were made in experimental installation 500 mm in diameter at temperature 700-900 °C in USTU for the purpose of verification of combustion processes. The wood chips 10x20x10 mm in dimensions were arranged to fluidized bed of sand with particles 0,32 mm in diameter. The time of volatilize and char combustion was measured. The time of volatilize combustion was approximated by

$$\tau_{v.c.} = \delta \cdot 10^{-3}\left(\frac{1000}{t}\right)^{0.7137}, c \quad (1)$$

where t- bed temperature, °C; δ - initial diameter of fuel particles, mm to the bed temperature t and heating of material, circulating in external loop with flow rate Gs, kg/(m^2s). The heat of air from EHE with air flow rate Gk, kg/(m^2s), is concidered too.

Overheating of coke particles was taken into consideration at calculation of reaction rate constant based on experimental data [2]. There is a break of k=*fT*4) at 900 °C. This fact is probably concerned with reaction in porous coke particles at low temperature. This break of coke is characteristic of coal for all kinds of lignite [3].

One can distinguish three consecutive processes on combustion of wood: drying of wood particles, devolatilization and volatilize combustion, char combustion. On wood .content V^{daf}=85%, W^r = 37.2% equation between shares of heat is following: heat for moisture vaporization 4%; heat for char combustion 18,6%; heat for volatilize combustion 885,4%.

According to disperse analysis of coke particles heat share of heat combustion in freeboard is less than 8%. On this reason it was considered that coke bums in bottom bed and splash zone, meanwhile volatilize combustion occurs in freeboard too.

Considering that there is no any temperature fluctuation in cross section one-dimensional problem could be approached. Besides it was conceived that intensive internal circulation provides permanent value of freeboard temperature.

The heat and mass balance equation system for particles of coke and fuel in bed with volume V and mass M can be written as

$$M\frac{dz}{z\tau} = B\frac{V_c}{100} - V\,j\,\frac{6\rho_m(1-\varepsilon)}{\delta\,\rho_c}z$$

(2)

$$M\frac{dz_v}{d\tau} = B\frac{V^r}{100} - V\,j_v\,\frac{6\rho_m(1-\varepsilon)}{\delta_v\,\rho_{ch}}z_v$$

(3)

where: B - fuel flow rate, kg/s; ρ_m, ρ_{ch}, ρ_c - density of bed material, char and coke, respectively, kg/m^3; δ and δv - volatilize evaluated coke and fuel particles dimension, m; z and z_v – mass concentration of coke and char particles in the bed, what content volatilize, %; V_c and V^r volatilize and coke content in the fuel as fired, %.

If characteristic complex for specific surface of coke and fuel particles will be written as $S = \frac{6\rho_m(1-\varepsilon)}{\delta\,\rho_c}$ and $S_л = \frac{6\rho_m(1-\varepsilon)}{\delta_v\,\rho_{ch}}$ one can get non-stationary equations of heat balances

$$Mc_m\frac{dt}{d\tau} = VS\,j\,z\,Q_c + \beta VS\,j_v z_v Q_v - B\frac{W^r}{100}t - u_0\rho_g c_{fg}F\,t - G_s F_f c_m(t - t_s) + G_{EHE}c_{air}\rho_{air}t_s$$

(4)

where ρ_g,ρ_{air} - gas and air density, respectively, kg/m^3; c_m, c_{fg}, c_{air} - heat capacity of bed material, flue gases and air, respectively, kJ/(kg·K); F, F_f - cross section of bottom and top parts of furnace, m^2; τ-time, s; u_O-flue gas velocity in the top furnace, m/s.

Fuel combustion of char as as well as combustion of volatilize share β in the bottom bed is considered in Eq.(5). The heat of char combustion is symbolized by Qk. That one of volatilize is designated as Qv. The heat is expended in moisture evaporation, heating of combustion products up

The heat released in freeboard after combustion of volatilize share (l-β) goes into heating of secondary air, circulating material and combustion products of primary air on temperature interval from t to t_f, a portion of heat is transmitted to screened walls:

$$V_f\left(c_{fg}\rho_g + c_m\overline{\rho_m}\right)\frac{dt}{d\tau} = (1-\rho)BQ_v V^r / 100 - c_m u_2 F_f t_f - (G_s c_m F_f + u_0\rho_g c_m F)(t_f - t) - k_s F_s(t_f - t_s)$$

(5)

where: $\bar{\rho}_m$ - average material density in freeboard, kg/m^3, u_2 - secondary air velocity per unity of furnace surface, m/s, F_s and t_s - surface area of screen in the furnace, m^2, and temperature of media in it, °C; V_f - - freeboard volume, m^3.

Experimental data were used for estimation of heat transfer coefficient k_s. Heat transfer coefficient *a* g was calculated as a function of velocity at the furnace bottom u_o, m/s: $\alpha = 28{,}6u_0$.

The heat of circulating sand with material flow rate G_s'F_f goes into heating of air for EHE to

temperature tfand heat transfer to the surface of screened walls in cyclone as well as EHE:

$$M_{EHE}c_m\frac{dt}{d\tau} = G_sF_fc_m(t_f - t_s) - G_{EHE}c_{air}\rho_{air}t_s - k_{EHE}F_{EHE}\left(\frac{t_f + t_s}{2} - t_s\right)$$

(6)

where M_{EHE} - sand mass in EHE, kg; k_{EHE} -heat transfer coefficient through lining of cyclone and EHE, W/(m^{2}K).

Average to height of bottom bed oxygen concentration was calculated with consideration for concurrent combustion of coke particles and volatilize Specific flow rate of material Gs, volatilize share p and heat, transmitted by circulating material to cyclone and EHE was not measured So, Gs, β and k_kF_k were calculated on base of fuel flow rate, air flow rate and temperatures, measured in steady state conditions.

According to these calculations Gs rises greatly with increasing of gas velocity in bottom part of the furnace. Dependence Gs = 0.73 $u_o^{0,37}$ resulted in experimental data correlates with that one of [1], where degree was in interval 4.2-4 8. Volatilize share β decreases drastically with rise of velocity. This dependency can be written as $\beta = 51\ m_0^{-4,2}$ The magnitude of k_kF_k was about 5.1 kW/K- at regimes I and II and about 2.3 kW/K in regime Ш

Inertia of thermocouple and transport delay time were considered in comparison of calculation according to equations (2-4) and experimental data obtained in investigation of transient processes after disturbance by fuel flow rate.

Satisfactory agreement between calculations and experiment inspire of extreme simplification can be explained by big mass bed's capacity. Its magnitude is known rather good.

Summarized inertia of temperature in CFBB is determinated by sum of all these factors. Therewith their joint influence is ambiguous [7] The reason is that specific flow of coke and volatilize depends on temperature. Moreover, oxygen concentration at disturbances is decisive factor during the process of coke combustion.

Consequently heat inertia of FB furnace is determinated by time of particles heating in the bed and combustion conditions. The store sand in EHE is the cause of additional heat inertia increasing. The heat inertia can be reduced by rise of circulating material flow rate, especially at Gs<40 kg/(m^2s). The coincidence of calculations and experiment testifies that model is adequate to real transient process**.**

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