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COLD FLOW MODEL STUDY ON INTERCONNECTED FLUIDIZED BED REACTORS FOR MULTI-GENERATION SYSTEMS AND CHEMICAL LOOPING PROCESSES

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ABSTRACT

Interconnected fluidized bed reactors (DCFB) were implemented in multi-generation systems (pyrolysis FB reactor and CFB boiler), chemical looping combustion (CLC) systems (double metal oxides or carbonate oxide reactors) and three reactor chemical looping gasification processes. The presented data focus on the solids circulation rate and on pressure profiles of the DCFB depended on selected operating parameters such as fluidization gas flow rate, loop seal fluidization, and solids inventory. Most work was devoted to studying standpipe and valve operation.

INTRODUCTION

The important feature of the DCFB concept is the inherent stabilization of solids hold-up obtained by the direct hydraulic link between the two CFB systems, i.e. the loop seal connection in the bottom region of the risers. Only the air reactor entrainment is responsible for the global solids circulation between the reactors, while the fuel reactor operation can be optimized. The hydrodynamics of interconnected reactor systems and loop seal (standpipe and non-mechanical valves) are very important in that optimization.

Modern multi-generation technologies can be subdivided into three categories:
- Multi-generation systems based on complete coal gasification, used by Shell Co.;
- Multi-generation systems based on partial coal gasification (combination of coal combustion and gasification);
- Multi-generation systems based on pyrolysis of coal.

Multi-generation systems based on pyrolysis of coal include many of configurations. An example is the shale processing units in the USSR. The resulting technology was implemented in industrial units. Two high-capacity units of this type (UTT – 3000, 139 tons of shale per hour each) were erected at in the Estonian thermal power plants and they with still show good performance. VTI has an agreement on scientific cooperation with the Institute for Thermal Power Engineering, Zhejiang University in this field. In studies carried out by Zhejiang University (1), (2)) data were received from their tests on a 1 MW pilot plant. The first results showed that a pilot unit can produce gas with a heating value of about 12 to 14 MJ/m^3 with a 5 to 6 % tar content at 500 to 600 ^{o}C in the fluidized bed (FB). It was established that solids circulation is a critical parameter for stable plant operation. The investigation was financed by the Russian Ministry of Science and Education as a part of a China – Russia scientific cooperation project.

Another aim of the investigation of the interconnected FB and CFB reactors is to study the hydrodynamics of the CLC (double metal oxides or carbonate oxides reactors) and the three reactor chemical looping gasification processes (3) and (4).

RESULTS

It is important for interconnected units containing CFBs to provide a stable circulation rate between reactors as well as an efficient solids separation. Such a study has been connected in VTI`s experimental unit shown in Fig. 1. This unit consists of a CFB reactor and a transportation reactor (TR) that are interconnected with pneumatic valves and standpipes to a FB reactor. The system consists of three interconnected fluidized bed reactors: The CFB (0.2*0.3 m cross section, 5 m height), the FB reactor (0.4*0.4 m cross section, 1.6 m height) and the transport reactor (0.1 m diameter, 4.8 m height). Standpipes and J or L valves connected the FB reactor and the bottom parts of the CFB and transport reactors with the cyclones of the CFB and transport reactors.

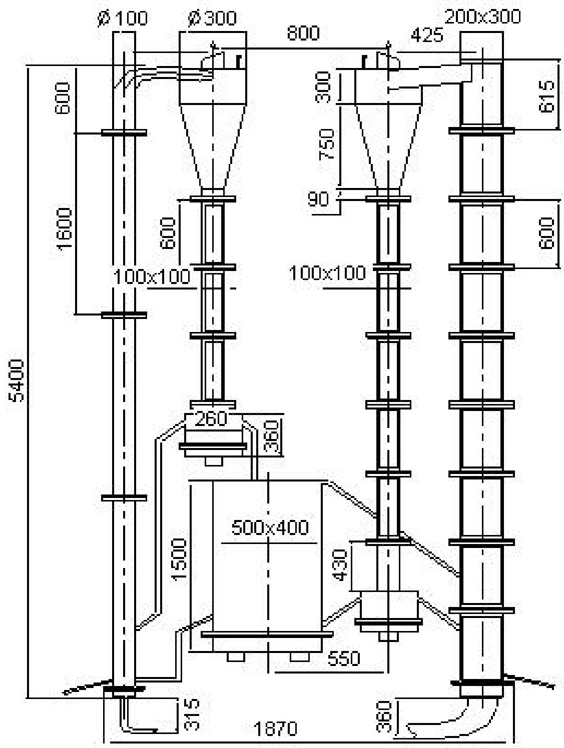


Fig. 1 –Drawing of the VTI cold model of interconnected reactors

The CFB reactor consisted of a vertical column 5.4 m high, made of 9 sections 600 mm high each with a 300*200 mm cross-section. The upper section was connected to the cyclone inlet. The cyclone outlet was removable and was located 50 mm away from the cyclone central axis. The cyclone cone was connected to a standpipe, consisting of 3 sections, each of 100*100 mm cross-section. The lower section of the standpipe was connected to a pneumatic valve. The design of the valve allowed solids to flow in different directions. Part of the flow could be directed back to the bottom part of the CFB reactor, and the other flow to the FB reactor. The transportation reactor consisted of a vertical column 5.4 m high, made of 3 sections of 100 mm diameter and each were 1600 mm high. The reactor and standpipes were equipped with taps to measure pressure along their heights.

The FB reactor consisted of a vertical vessel with a square (400*400 mm) cross-section and 1500 mm in height. There was a nozzle grid at its bottom for distributing the air. The reactor was subdivided into two equal 200*400 mm sections by a partition. Solids were discharged from the FB reactor through pipes equipped with a pneumatic L-valve. The L-valve connected the bottom of the first section of the FB reactor to the bottom part of the transportation reactor. In the first section of the FB reactor, air flowed counter-currently to the flow of solids. In the second section flow ws co-current. In the second section, solids were discharged through the top part of the section, and then fell into the CFB reactor. The VTI aerodynamic test unit has been described in other references (5) and (6). Local mass flows were measured with isokinetic S-type probes, and local heat exchange with gauges while heating. Also mini-turbine flowmeters and differential concentration-measurement gauges were employed.

Based on these tests it was established that:
- pressure profiles and vertical solid concentrations in the CFB and FB reactors are depended on gas velocity and solids loading;
-pressure drops in the unit depended on operating parameters and on design features of the units and return legs.

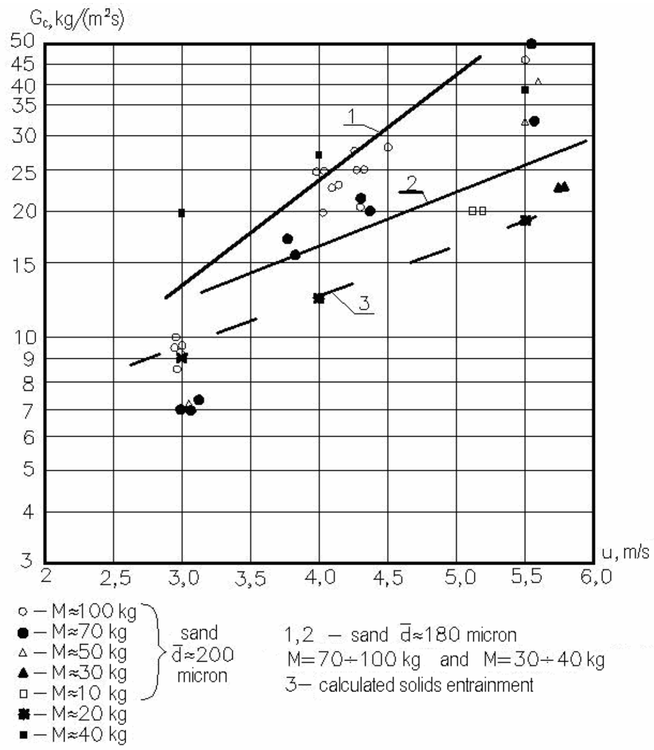


Fig. 2 – Typical solid circulation rate as function of velocity and inventory

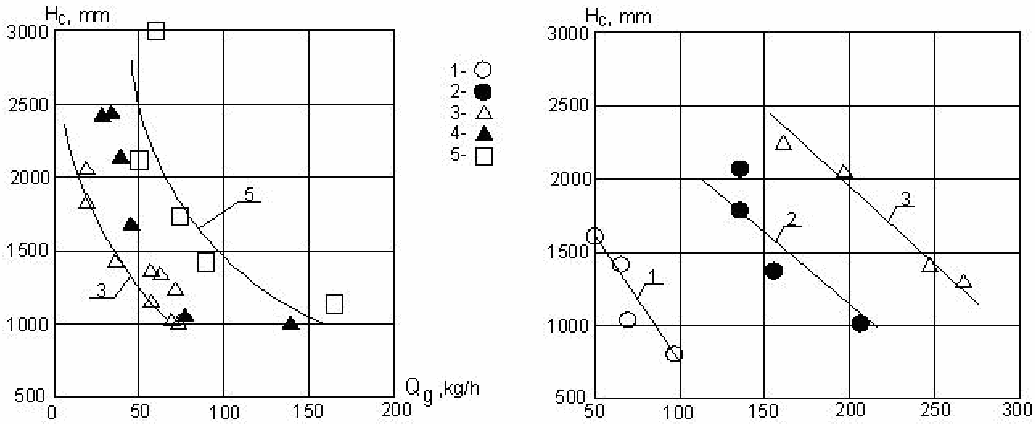
VTI studies carried out on a similar unit showed that particle size determined the value of the dimensionless criteria (Ar, Re, Fr), used for hydrodynamic and heat exchange calculations. It was established that the change in average particle size with respect to unit width or diameter was insignificant. The studies were basically directed to calculate the circulating mass flow rate in the external circuit (Gr), as it is the critical operational parameter for the ireactors. The solid flow rate increased with gas velocity in the CFB unit and with unit inventory. This relationship is shown in Fig. 2. This relationship was obtained at stable unit operation when unit load increases were possible. Stable system performance was made possible with good loop seal operation. A separated solids return system is necessary for reliable operation of the CFB unit. Usually a return system consists of a vertical standpipe, pneumatic valve and a short return leg back to the furnace. It is important to choose the right dimensions (diameters and height) of the standpipes and have them operate in the correct fluidization regime. Observation of the solids motion in the standpipe indicated that in the bottom part was in dense phase flow, but it was in dilute phase flow at the top. During several tests with sand (at a low bed level in the standpipe and a large mass of sand in the unit) a pulsating regime was established with bubble generation and gas overshoot to the upper part of the standpipe. The amplitude of the standpipe solids level oscillations reached 1 m with a frequency of 0.1 Hz. The limiting significant pulsation case could be estimated using Equation (1):

$$\left(\frac{\Delta P}{L}\right)_{mf} \cdot H_s \approx H_v \cdot (1-\varepsilon_v) \cdot \rho_p \cdot g + \Delta P_{ft} + \Delta P_c \qquad (1)$$

At normal conditions the pressure drop in the upper part of the loop seal $H_v \cdot (1-\varepsilon_v) \cdot \rho_p \cdot g$ was equal to or exceeded the pressure drop at the bottom part of the furnace $\Delta P_{fb}=\Delta P_f-\Delta P_{ft}$, so this guaranteed zero furnace gas overshoot into the valve or the standpipe. The limiting bed level in the standpipe was estimated using Equation (2):

$$(H_s)_{\lim} = \frac{\Delta P_f + \Delta P_c}{\left(\frac{\Delta P}{L}\right)_{mf}} \qquad (2)$$

In the tests, the limiting bed level in the standpipe ranged from 0.15 to 0.5 m. This increased with increasing bed mass and increasing particle size. For an industrial CFB boiler this limiting value was about 2-3 m. It was suggested (7) to increase the standpipe height to double its normal height. Fig. 3 shows the dependence of standpipe bed level with constant aeration. Small changes in aeration to the standpipe caused a significant bed level change in the standpipe (Fig. 3a), but the aeration to the valve had little affection on the standpipe bed level (Fig. 3b).

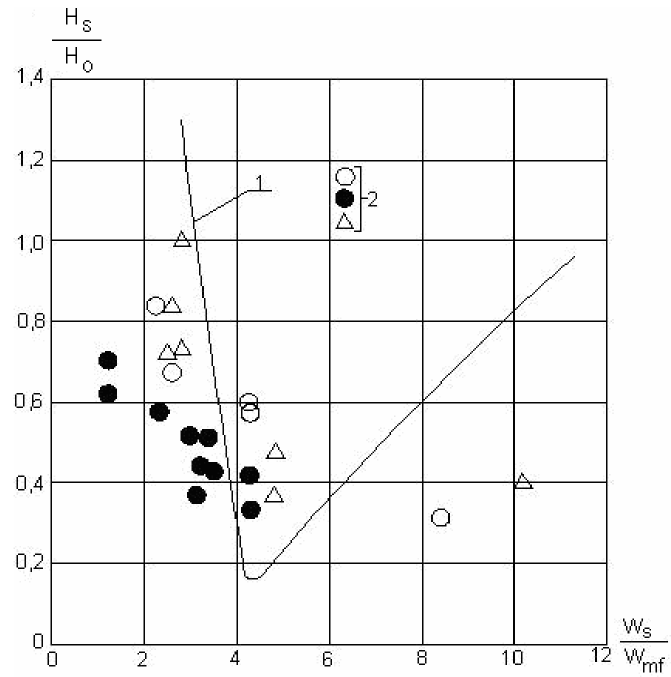


1 – full cross-section bed material reduced speed W_{ms}=0,013 m/s; 2 – W_{ms}=0,02 m/s; 3 – W_{ms}=0,05 m/s; 4 – W_{ms}=0,1 m/s; 5 – W_{ms}=0,15 m/s

Fig. 3 – Standpipe bed level dependence of aeration to the standpipe (a) and to the valve (b)

Thus, parameters shown in Fig. 3 are critical for return system regulation. The best operational mode is with continuous light aeration to the standpipe and regulating the aeration to the valve.

A master curve of standpipe bed height (H_s/H_0) relative to aeration velocity (W_{sv}/W_{mf}) is shown in Fig. 4. Solids inventory in the unit (riser + return system), gas flow rate in the riser and in the solids upflow section of the loop seal were constant, as aeration to the standpipe was varied. A similar curve was given by (8), which says that the minimum bed height level was obtained when the air velocity is about 5 W_{mf}. If the velocity was less than 4 W_{mf} there was an increase in the height of the solids level in the standpipe. If the air velocity was greater than 6 W_{mf}, then the steandpipe height increased with increases in gas velocity. At low velocities – there was a dense bubbling downflow mode and at high velocities air bubbles moved upwards to the cyclone. Our data show that the cyclone efficiency decreased when that occurred. From the data in Fig. 4 it is clear that it was necessary to maintain the bed level in the standpipe so that it did not exceed 0.4 of its height with simultaneous limitation of aeration to the standpipe at la value of 3-6 W_{mf}. Within the recommended regime, the solids flow rate reached 1800 t/h·m^2 (there was no opportunity to reach higher values because of cyclone overload). It is recommended a velocity of 0.1 m/s be used for standpipe diameter calculations.



1 –Knowlton data (8), 2 – VTI experimental data
Fig. 4 – Relative standpipe bed material height level dependence of relative air velocity in the standpipe

In the normal operating mode of the ash return system, a dense flow occurs in the bottom part of the standpipe. In this case, the Ergan Equation pressure gradient will be equal to:

$$\frac{\Delta P}{L} = \frac{150 \cdot \mu_g}{(\varphi \cdot \bar{d})^2} \cdot \left(\frac{1-\varepsilon}{\varepsilon}\right)^2 \cdot V_R + \frac{1{,}75 \cdot \rho_g}{\varphi \cdot \bar{d}} \cdot \left(\frac{1-\varepsilon}{\varepsilon}\right) \cdot V_R \cdot / V_R / \qquad (3),$$

A linear dependence of voidage on relative velocity was suggested by Knowlton (8):

$$\varepsilon = \varepsilon_v + (\varepsilon_{mf} - \varepsilon_v) \cdot \frac{V_R \cdot \varepsilon_{mf}}{W_{mf}} \qquad (4)$$

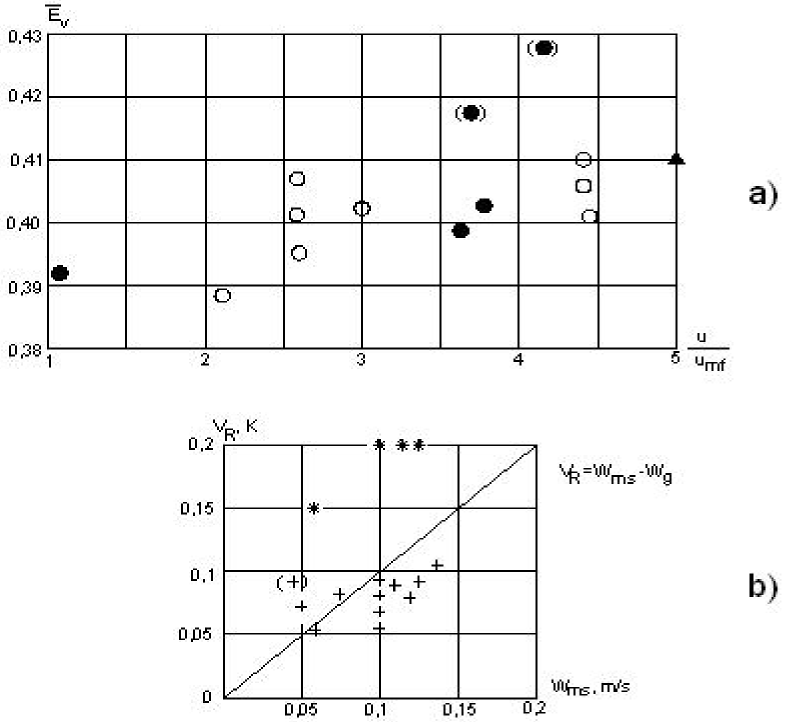
Relative bed velocity is equal to:

$$V_R = W_p - W_g = \frac{G_p}{\rho_p \cdot (1-\varepsilon) \cdot F_s} - \frac{Q_g}{\rho_g \cdot \varepsilon \cdot F_s} \qquad (5)$$

From Equations (3), (4) and (5) voidage as a function of pressure gradient $\frac{\Delta P}{L}$ in the standpipe can be calculated as:

$$\left(\frac{1-\varepsilon}{\varepsilon}\right)^2 \cdot (\varepsilon - \varepsilon_v) = \frac{\frac{\Delta P}{L}}{\left(\frac{\Delta P}{L}\right)_{mf}} \cdot \left(\frac{1-\varepsilon_{mf}}{\varepsilon_{mf}}\right)^2 \cdot (\varepsilon_{mf} - \varepsilon_v) \qquad (6)$$

The standpipe voidage dependence on relative gas velocity (W_g/W_{mf}) is represented in Fig. 5a. Standpipe voidage increased with increasing of the aeration, but was still less than the voidage a minimum fluidization. A comparison of the relative particle velocity from Equation (5) and form the observed solids velocity (Fig. 5b) shows that the calculated relative velocity was lower than the observed velocity in most cases. This indicates that gas flowed down with the bed material. In the pulsation regime, the relative velocity was much higher than the observed velocity, indicating that the gas was flowing up the standpipe.



1-particle velocity in standpipe W_{ms}=0,06 m/s; 2 – W_{ms}=0,1 m/s, 3 - W_{ms}=0,13 m/s
Fig. 5 – Relative air velocity dependence of standpipe voidage (a); Particle velocity in the standpipe dependence of its relative velocity (b)

Thus, calculations confirmed the experimental data from the unit and indicate that this method can be used to design industrial units.

CONCLUSIONS

The studies on solids circulation and solids return systems of interconnected reactors found:

- pressure profile and vertical solids concentrations in interconnected FB and CFB reactors were typical for reactors of this type, and depended on gas velocity and bed mass;
- the axial pressure profile depends on the fluidization velocity and the design features of unit`s structure and return system;
- the recirculating solids flow rate between interconnected FB and CFB reactors depends on gas velocity and solids inventory in the CFB reactor, on high-efficiency cyclones and sufficient capacity of the return system for a certain particle size;
- the solids return system (standpipe and pneumatic valve) must have a reserve capacity and be able to regulate aeration to the standpipe and to the upflow part of the valve. Otherwise, the recirculation solids flow rate is determined by the operational mode of the valve.

ACKNOWLEDGEMENT

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NOTATION

G_c – solid circulation rate
H_o – standpipe height
H_s – bed height in the standpipe
H_v – height of the upper part of the valve
M – unit inventory
ΔP_{ft} – pressure differential at the top of the furnace
ΔP_c – cyclone pressure drop
ΔP_{fb} – pressure differential at the bottom part of the furnace
ΔP_f – pressure differential in the furnace
u – gas velocity in the riser
ρ_p – particle density
ε_v – vibrated bed porosity (≈ 0.37)
ε_u – voidage in the upper part of the valve
W_{mf} – velocity at minimum fluidization
W_{ms} – particle velocity in the standpipe (downflow particle velocity)
φ – shape coefficient ($\varphi \approx 0.73$)
V_R – relative particle velocity
ε – voidage
ε_{mf} – minimum voidage
G_p – solids flow rate
Q_g – flow rate of gas
F_s – cross-section of the standpipe

W_g – relative gas velocity
W_{sv} – standpipe fluidization air velocity
$\left(\frac{\Delta P}{L}\right)_{mf}$ - pressure gradient in the standpipe at minimum fluidization parameters

$\frac{\Delta P}{L}$ - pressure gradient

$(H_S)_{\lim}$ - limiting bed level in the standpipe

$\bar{d}$ - average particle diameter

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