DESIGN OF A FAST INTERNAL CIRCULATING FLUIDIZED BED GASIFIER WITH A CONICAL BED ANGLE

by

***Jernej MELE*[a], *Andrej SENEGAČNIK*[b*]**

[a] CPPE d.o.o., Celje, Slovenia
[b] Faculty of Mechanical Engineering, University of Ljubljana, Ljubljana, Slovenia

Original scientific paper
https://doi.org/10.2298/TSCI161129171M

The main purpose of a fast internal circulating fluidized bed gasifier is the steam reforming of solid organic matter, like biomass, to a nearly nitrogen-free syngas. The calorific value of this syngas is approximately three times higher than the gas from common air-driven gasifiers. This article deals with a study of the particle dynamics in a 1 MW_t fast internal circulating fluidized bed plant and focuses on the design of the gasification reactor's geometry. Superheated steam is used for the fluidization and gasification in the reactor. The gasification of solid fuels causes an increase in the volume flow of the fluidizing gas and at the same time also a change in the fluidization regime. Approaching a turbulent fluidization regime or even fast fluidization is not desirable. However, with the proper design of reactor, i. e., an appropriately conical bed angle, suitable gasification conditions in the form of a fluidizing regime can be achieved across the entire height of the bed. For the purposes of the experimental research, a semi-industrial unit was set-up. The process was designed and experimentally tested on a lab-scale, cold-flow model and scaled-up to a semi-industrial process. The guidelines for designing the geometry of the gasification reactor were set.

Key words: *biomass-steam gasification, conical fluidized bed, gasifier design, fluidization regime, modal method*

Introduction

While researching a 1 MW_t fast internal circulating fluidized bed (FICFB) gasification process, questions concerning the particle dynamics in the gas-flows arose. The gas-solid flow of a FICFB is almost impossible to model accurately, either analytically or numerically. Building a semi-industrial plant is very expensive and there is no certainty that the semi-industrial apparatus will perform the planned task. Therefore, building a much cheaper, cold-flow model can be economically justified in order to verify the design and to reduce the possibility of an incorrect scale-up procedure to a hot, semi-industrial plant.

The FICFB gasifier being studied and discussed in this article is divided into two major fluidized bed zones: a gasification zone (also called the reactor) and a combustion zone (also called the riser), see fig. 1(a). The solid fuel is fed into the high temperature (~850 °C) tapered bed (gasification zone), which is fluidized with superheated steam (~600 °C). The hot bed material (quartz sand and catalyst) together with some remaining char moves from the gasification reactor to the riser though the chute. In the riser, the remaining carbon burns and heats up the bed material. The

* Corresponding author, e-mail: andrej.senegacnik@fs.uni-lj.si

red-hot bed material is pneumatically transported through the riser to the cyclone, where it is separated and collected in a syphon. The syphon, fully filled with red-hot bed material, acts as a gas barrier between the reactors, *i. e.*, the syngas and the flue gas. Surplus red-hot bed material from the syphon streams back to the gasification reactor and supports the endothermic gasification reactions with heat. A dual fluidized bed system enables the production of a high-grade syngas that is practically nitrogen free. A more detailed description of this process is given in Hofbauer *et al.* [1].

The hydrodynamics of such a system is essential for its design and operation, as discussed in a detailed study by Kaiser *et al.* [2]. Some studies of circulating fluidized bed loop predictions can be found in the literature. Breault and Mathur [3] analysed a high-velocity-loop fluidized bed. Bai *et al.* [4] investigated a dual fluidized bed system with two risers, down-comers and two valves. The authors showed the dominant influence of the solids inventory, the superficial gas velocities and the opening of the solids valves during the operation of the system. Kaiser *et al.* [5] and Loffler *et al.* [6] studied the hydrodynamics of a dual fluidized bed gasifier. Recently Leckner [7] describe the fluidization behaviour at high velocities in wide vessels like commercial solid-fuel converters (12 MW_t), such as boiler furnaces and gasifiers. Bubbling, turbulent, and fast-fluidization regimes are analysed [7]. A brief review by Rudisuli *et al.* [8] presents pathways to the successful scale-up of fluidized bed reactors. The most commonly described approach in the literature to achieve a hydrodynamic and reactive similarity in two fluidized beds is to use sets of dimensionless numbers, which should be kept constant in both scales [9]. Very challenging processes for scaling-up are reactive fluidized beds like gasification. A rather simplistic method of appropriately control the bubble growth in a scaled reactive fluidized bed is suggested in [8] and similar idea is also used in this article. Recently a lot of research was done on gas-solid conical spouted beds and their advantages in case of solid fuel gasification Lopez *et al.* [10] and Saldarriaga *et al.* [11]. Alia *et al.* [12] introduced for the first time radioactive particle tracking as an advanced non-invasive technique to validate scale-up methodology. Resent review of biomass gasification modelling was made by La Viletta *et al.* in [13]. Review of gasification fundamentals with new findings can be found in Mahinpey and Gomez [14].

Main goal of this article is to present the iterative procedure used for determining the cone angle of the fluidized bed biomass gasifier. The gasification of solid fuels is a high temperature process involving heterogeneous solid-gas reactions and phase changing. This type of chemically reactive fluidized bed requires more effort for scaling up from a cold-flow model to the hot pilot scale. First, it is necessary to explore the tapered cold-flow model to get basic experimental data, like pressure drop and fluidizing velocities, to verify the mathematical model of fluidization. Then the mathematical model of gasification is set-up. Scale-up process starts with both models. During the scale-up process, the main idea is to keep superficial velocity near to the state of full fluidization. In a case of inner-bed gasification, the reactor cross-section needs to be increased to keep the superficial velocity inside the limits for the desired *i. e.* bubbling fluidization regime. If some parameters of models are unknown, the best possible estimations need to be made. The results from one model are then entered into the other model. This way, the dimensions, *i. e.* cone angle of the gasification reactor, is then iteratively calculated/corrected until the proper, *i. e.*, a similar, fluidizing regime is achieved over the entire layer of the bed. The proposed scale-up procedure was experimentally verified with measurements carried out during stable operation of 1 MW_t semi-industrial gasifier.

Experimental equipment

The experiments were performed in a cold-flow reactor and later in a 1 MW_t semi-industrial gasifier that operates in city of Celje, Slovenia.

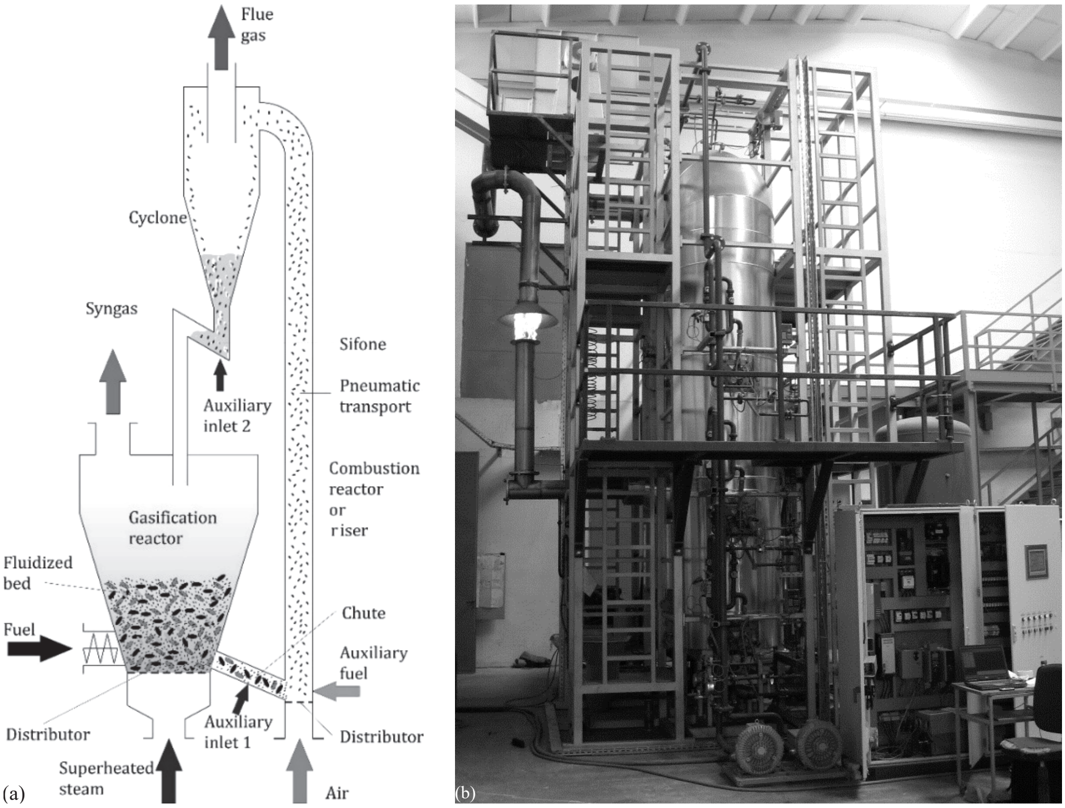


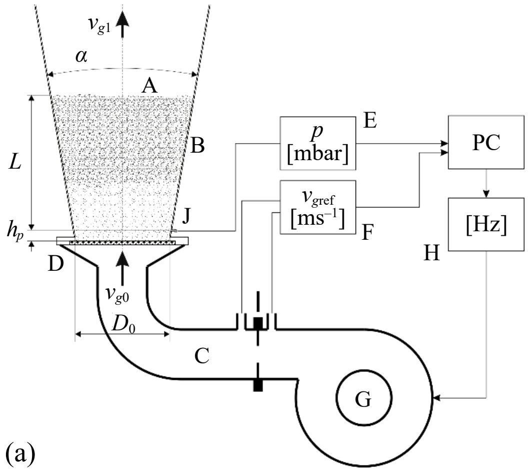
Figure 1. (a) Principle of a FICFB gasifier (b) the FICFB in operation (author's photo)

First, the basic engineering plans were made for the semi-industrial gasifier. To avoid the high costs of probable corrections to the semi-industrial unit, a decision was made to test the process in a small-scale, cold-flow unit. Therefore, the laboratory unit was designed and built in order to simulate the hydrodynamic process of the fluidization with air, under arbitrary conditions. Figure 2(a) shows a laboratory unit.

The semi-industrial plant, figs. 1(b) and 2(b), was designed to take into account the heavy thermal and abrasive conditions that occur in it. The gas chambers are made of heat-resistant stainless steel and protected with refractories, so that the hot gas-solid flow does not affect the steel walls. The main gas-stream inputs are superheated steam and preheated air and the main outlets are syngas from the reactor and hot flue gas from the cyclone. The 1 MW_t FICFB semi-industrial plant was designed on the basis of a study carried out using a cold-flow model.

Designing the gasification reactor

When designing industrial applications, the gas-flow used to fluidize the bed is typically chosen as a function of the desired fluidization state and the required mass and energy balances [15, 16], and not as a function of the minimum gas velocity necessary to fluidize all the particles. In phase-change processes, large physical changes occur. For example, during the gasification of biomass, a solid fuel is reformed with steam or air into an alternative gaseous fuel. At the bottom of the reactor a gasification agent is vertically entering the reactor and re-



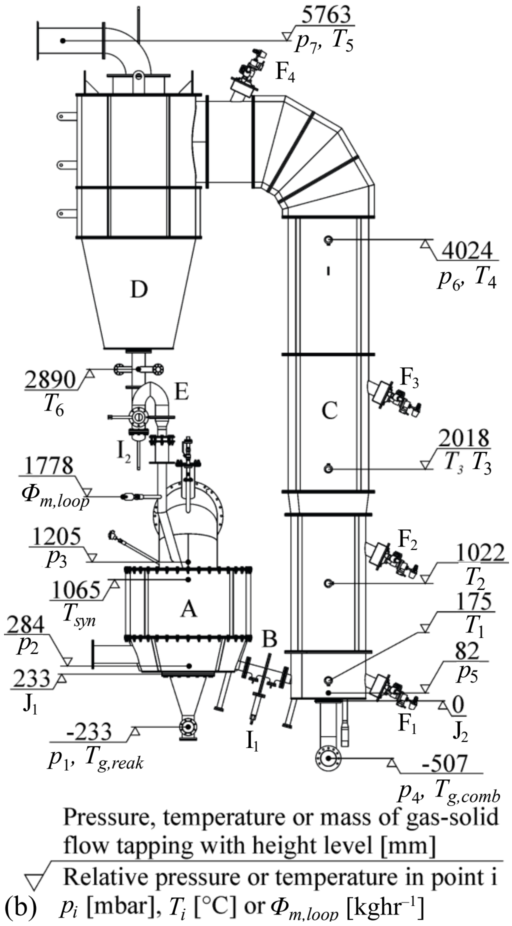


Figure 2. (a) Scheme of laboratory unit
A – conical or non-conical bed of particles, B – housing, C – tube, D – distributor, E – relative static pressure measure, G – driving fan, H – power inverter, I – orifice, J – pressure measure opening, PC – personal computer;
(b) Semi-industrial gasifier scheme; pressure measuring points in semi-industrial unit:
A – conical reactor, B – chute, C – riser, D – cyclone, E – syphon, F – burners, I – auxiliary gas inlet, J – gas distributor

acts with the products of the biomass pyrolysis and exits at the top of the reactor as a syngas. During gasification the volume flow of the fluidizing gas is increased, and also its temperature and composition are modified. This causes the fluidization conditions/regimes to change with the bed height [17]. As shown in this paper, the shape of the reactor must be carefully chosen to achieve suitable process parameters.

For stable, solid-fuel gasification in a fluidized bed, several boundary conditions should be fulfilled:

– For the efficient heat transfer between the gas and solid particles, we need long residence times for the solid-fuel particles inside the bed.
– Establishing the inner circulation of the fuel particles and maintaining the buoyant fuel particles immersed in the bed throughout the whole heterogeneous reaction process [18].
– The fuel pellets should be fed into the fluidized bed and after that the inner bed circulation of the fuel particles should begin [18].
– The fluidization regime at the upper part of the bed should be kept in the lower part of the bubbling regime, to prevent the segregation of wood pellets and char particles on the surface of the bed.
– From other research, it is known that for bed fluidization, more steam is needed than the stoichiometric amount in equation [19].

From other research, it can be predicted that the regime of a fully fluidized conical bed can implement the first three conditions [20, 21]. At the fully fluidized regime, the bed particles in the centre move upwards and form the particle-core upward stream. Particles outside the core move downwards and a proper inner circulation is established [20]. If the superficial velocity is still increasing, the fluidization regime approaches the turbulent fluidization, which is undesirable due to the segregation of light char particles on the top. Simultaneously, the catalytic reaction conditions become worse due to the big-bubbles regime and the voidage increase. The gases and fuel particles need to be in contact with the hot-bed catalyst a much as possible.

Mass and energy balance

For the successful design of the reactor we need to accurately predict the state of fluidization at the bottom and the top of the fluidised bed. To assess the state of fluidization, we need the following data: bed material data, volume flow, density, temperature, and dynamic viscosity of the fluidizing gas. We need to model the mass and energy balance of the reactor and then solve this system in some way. In doing this, we took into account some of our own experiences from a cold model, like the fuel bed material ratio.

More details about the chemical gasification model, the equilibrium reactions and the kinetics can be found in many references, *e. g.*, [22-24]. The common composition of wood biomass (dry, ash free) can be expressed as $CH_{1.44}O_{0.66}$, and with this knowledge the stoichiometric steam demand for gasification can be calculated [19]:

$$m_{\mathrm{H_2O}} = 14.76 \frac{\mathrm{mol_{H_2O}}}{\mathrm{kg_{daf}}} = 0.266 \frac{\mathrm{kg_{H_2O}}}{\mathrm{kg_{daf}}} \tag{1}$$

The volatile matter of wood (dry basis) is ~86 % and the fixed carbon ~13 % [19, 24]. This means that in the case of wood pellets (with ~6 % water) more than 87 % of the fuel mass is reformed. The volume of the products after gasification is much larger than the volume of the gas reactants. This is evident from the chemical model of gasification, *i. e.*, the Boudouard reaction and the water-gas reactions, fig. 3 [24].

Gasification products provoke changes in the fluidizer density, ρ_g, and the viscosity, η_g. This further impacts on the superficial velocities, v_{mf}, and, v_t, and, consequently, also affects the lower limit of the superficial gas velocity over the zone height. With an increasing gas volume flow in non-conical reactors, on account of the inner gasification, the superficial gas ve-

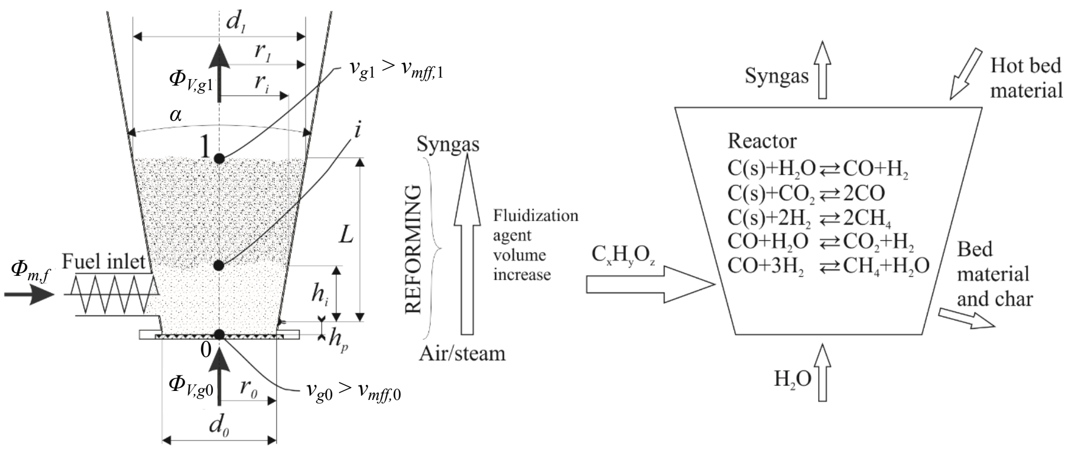


Figure 3. Reactor – dimensions, flows, chemical reactions

locity would rapidly increase along the reactor height. The most important point when defining the gas composition and the physical conditions in the reactor is that the following parameters must be determined: gas density, $\rho_{g,i}$, and gas viscosity, $\eta_{g,i}$, at the bottom (point 0) and at the top (point 1) of the bed.

The bed material

The main characteristics of a bed material are its heat capacity and the ease with which it can fluidize the fuel – only biomass cannot be fluidized in the bubbling state. The most common, simple and effective method is to use dry quartz sand. But the advantage of chemical processes in fluidized beds is the use of a solid catalyst (*e. g.*, olivine). It increases the rate of chemical reactions and lowers the residual time of the fuel in the reactor. The heat capacity of the bed material and its regeneration: circulation rate must be high enough so that the endothermic gasification reactions do not force the temperature below the operating range. From operational experience, the volume of the bed material should be at least three-times higher than the volume of the solid fuel and the char. It is important to ensure that the fuel and char are completely covered and as much as possible in contact with the bed material or the catalyst. We have defined a stationary bed height L = 500 mm. This bed has particles with density, ρ_p, particle sphericity, $\varnothing_S$, particle size distribution, *psd*, and bed voidage at the minimum fluidization velocity, ε_{mf}. At a mass-flow of biomass equal to 170 kg/h the circulation rate of the bed material between the reactor and riser at approximately 562 kg/h maintains 780 °C in the reactor. In experiments, quartz sand was used as the bed material, middle B Geldart's particles.

Fluidizing velocities

The FICFB system is a multiple fluidized bed system, with a bubbling fluidized bed in the reactor, chute and syphon and pneumatic transport in the riser. To achieve circulation of the bed material, the correct regimes must be reached at the same time. For small particles and low Reynolds numbers, the viscous energy losses predominate. Running the gasification in the range of the minimum fluidization velocity lowers the consumption of the gasification agent in the reactor, as well as lowering the consumption of energy and moisture in the raw syngas.

In the mathematical model the bed particles are ranked by size. For each particle size rank the proper equations should be used. In this particular case, we use the equation for calculating the minimum velocity of a fully fluidized bed, v_{mff}, proposed by Kaewklum and Kuprianov [21]:

$$A\left(\frac{r_0}{r_1}\right)^2 v_{mff} + B\left(\frac{r_0}{r_1}\right)^4 v_{mff}^2 - (1-\varepsilon_{mf})(\rho_s - \rho_g)g = 0 \tag{2}$$

where

$$A = 150\frac{(1-\varepsilon_{mf})^2}{\varepsilon_{mf}^3}\frac{\eta_g}{(\varnothing_S d_p)^2} \tag{3}$$

$$B = 1.75\left(\frac{1-\varepsilon_{mf}}{\varepsilon_{mf}{}^3}\right)\frac{\rho_g}{\varnothing_S d_p} \tag{4}$$

In order to solve eq. (6), a conical bed angle, α, and an inlet radius, r_0, (diameter $d_0 = 2r_0$) of the reactor should be presumed to start the iteration:

$$d_0 = \sqrt{\frac{4\Phi_{V,o}}{v_{g,0}\pi}} \tag{5}$$

Pressure drops

An estimation of the pressure drop across the fluidized bed is also very important for dimensioning the actuators (for example, a blower or steam generator) and measuring the components. This estimation gains further importance when designing industrial processes. The right selection of these elements can lower the energy consumption and raise the process efficiency. In order to estimate the pressure drop across the conical beds, a modified Ergun's equation was used [16].

$$\Delta p = A(L-h_b)\frac{r_0^2}{r_1 r_b}v_{g0} + B(L-h_b)\frac{r_0^4(r_b^2+r_b r_1+r_1^2)}{3r_1^3 r_b^3}v_{g0}^2 + (1-\varepsilon)(\rho_s-\rho_g)gh_b +$$
$$+\frac{1}{2}v_{g0}^2\left[\left(\frac{1}{\varepsilon}\right)^2\left(\frac{r_0}{r_1}\right)^4 - \left(\frac{1}{\varepsilon_{fb}}\right)^2\right]\rho_g \tag{6}$$

For industrial use, a further simplification was proposed in [21].

$$\Delta p_{mf} = (1-\varepsilon_{mf})(\rho_p-\rho_g)g\,L_{mf} \tag{7}$$

Reactor dimensions – conical-bed angle, α, and gas-inlet diameter, d_0

From the reactor-design point of view, it is important to define the reactor dimensions – especially its cross-section. At the reactor's inlet and outlet the operational superficial gas velocity for the gasification, $v_{lim,g}$, and combustion, $v_{lim,t}$, zones must be greater than the minimum required for the desired fluidization state. We can define the following:

$$v_{mff} < v_{g,0} \text{ and } v_{mff} < v_{g,1} < v_t \text{ for gasification zone, subscript 0-inlet, 1-outlet,} \tag{8}$$

Calculating the conical bed angle, α, is the last factor that defines the shape of the conical bed reactor. By calculating the gas volume flows and the superficial fluidization velocities at points 0 and 1 (bottom and top of the fluidized bed) the diameters of the reactor can be determined for point 1, which follows a trigonometry eq. (9):

As is clear in fig. 3, the top radius of the fluidized part of the bed, r_1, is:

$$r_1 = r_0 + \left[L\tan\left(\frac{\alpha}{2}\right)\right] \tag{9}$$

At the end the actual gas velocity has to be as near as possible to the calculated value:

$$v_{mff} \approx v_{g,0} \text{ and } v_{mff} \approx v_{g,1} \text{ for the gasification zone, subscript 0-inlet, 1-outlet,} \tag{10}$$

If the difference between v_{mff} and $v_{g,i}$ is more than 20% of v_{mff}, the conical bed angle, α, and gas inlet diameter, d_0, have to be evaluated again. The flow chart of iteration loop is presented in fig. 4.

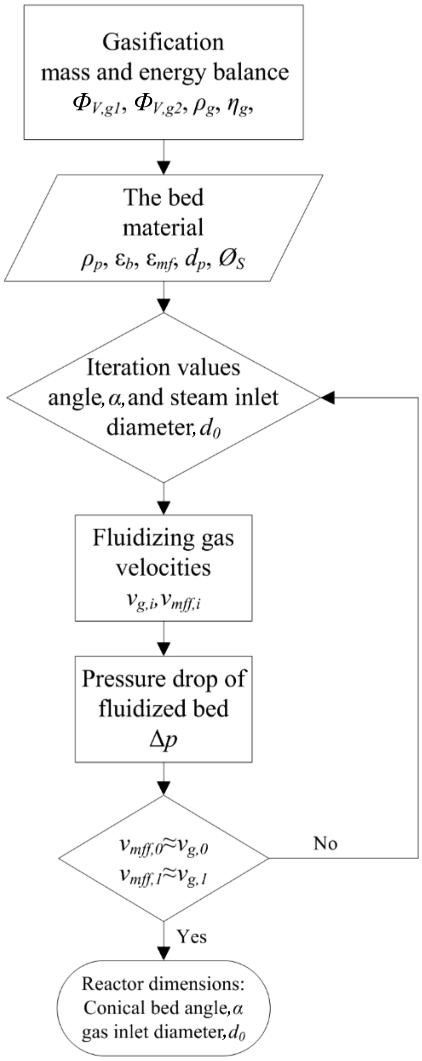


Figure 4. Flowchart of calculation

Experimental work

The continuous gasification process was achieved during stable particle circulation. The experimental units with the locations of the measuring points and height levels are presented in fig. 2(b). In the reactor – A a bubbling fluidized bed was established. The angle of the chute – B is designed in such a way that there is no solid flow if the particles are not fluidized. All the particles transported to the riser – C are pneumatically lifted to the cyclone – D and simultaneously heated up. With long residence times of the particles in the riser, the particles collect as much heat as possible. Red-hot particles are separated from the flue gas in the cyclone and finally gathered in the syphon – E. The auxiliary inlet – I_2 acts to fluidize the gathered hot particles and transport them to the reactor. Fluidization in the reactor is controlled using a statistical method, *i. e.*, the pressure drop across the bed *vs.* the superficial gas velocity, Mele *et al.* [25]. For simulating the particle dynamics in the reactor, the 1:1 cold-flow reactor was built and tested, see fig. 5. The calculated results were compared to the experimental, and the mathematical model was evaluated. The technical and process data for both reactors are gathered in tab. 1.

Monitoring the pressures and temperatures, as well as the gas and solid particle flows, is the most interesting activity during the experimental work. The measuring equipment was an Endress+Hauser PDM75 with a measuring range of 0-1000 Pa ±1 Pa for the gas-flow and a Siemens Sitrans 250 delta bar with a measuring range of 0-100 mbar ±0.5 mbar for the fluidized bed pressure-drop measurements. In the case of the semi-industrial FICFB gasifier, the temperatures were measured using NiCr-Ni thermocouples (type K) with a measuring range of 0-1250 °C ±9.4 °C. The semi-industrial unit has a solid-flow sensor SWR SolidFlow FSM installed for measuring the mass particle flow using the latest microwave technology. The measuring field is generated through the special coupling of the microwave together with the metal pipe. The microwave coupled into the pipe is reflected by the solid particles and received by the sensor. The received signals are evaluated with regard to their frequency and amplitude. The sensor works like a virtual particle counter that counts the streaming particles per time unit. The frequency-selective analysis ensures that only the streaming particles are measured. The calibration of the sensor takes place in the installed state by entering the reference quantity. All the gas-flows are measured with orifices. The locations of the measuring places are presented in fig. 2. The gas analysis in the semi-industrial plant was carried out with a custom-made infrared gas analyser. The analyser measured the main gas components – H, O, CO, CO_2, and CH_4. The heating value of the product gas was calculated from the obtained composition.

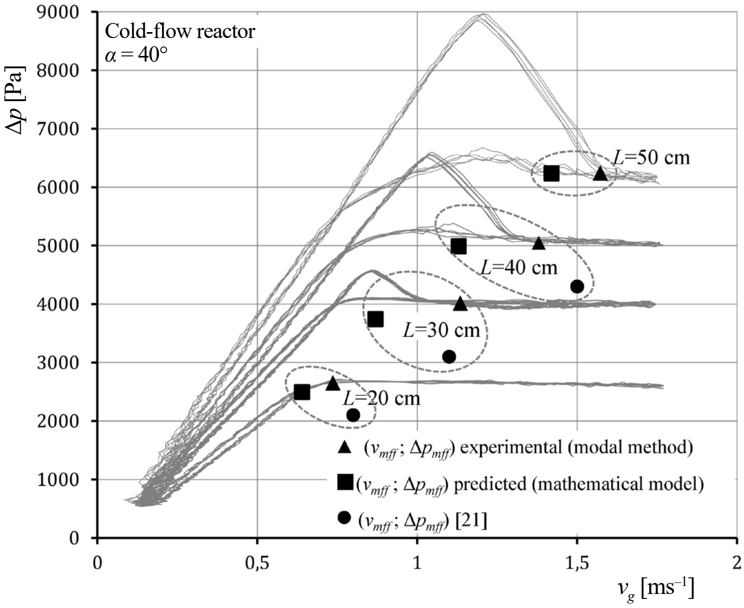


Figure 5. Comparison of the prediction and the experiment (v_{mff}, Δp_{mff}) for the average 400 μm sand particles fluidized in the 40° conical bed at different (static) bed heights

Table 1. Technical and process data for the cold-flow model and the semi-industrial reactor – A

Parameter	Symbol [Unit]	Cold-flow model	Semi-industrial unit	
Reactor diameter at gas entrance	$d_{\text{reak},1}$ [mm]	300	300	
Conical bed angle	α [°]	40	40	
Gas distributor type		Sandwiching nets	Bubble cap	
Bed material		Quartz sand	Quartz sand and char	
Particle size	d_p [μm]	300–500	300–500	
Geldard's classification		middle B	middle B	
Particle density	ρ_p [kgm^{-3}]	2650	2650	
Bulk density	$\rho_{p,b}$ [kgm^{-3}]	1575	1550	
Bed voidage	ε_b	0.6	0.59	
Stationary bed height	L [mm]	500	500	
Fluidization medium		Air	Steam*	Syngas**
Temperature	$T_{\text{g,reak}}/T_{\text{syn}}$ [°C]	40	670	780
Density	ρ_g [kgm^{-3}]	1.124	0.22	0.192
Dynamic viscosity	η_g [Pas]	$1.8 \cdot 10^{-5}$	$3.5 \cdot 10^{-5}$	$4.2 \cdot 10^{-5}$
Volume flow of fluidization agent 1 MW	$\Phi_{V,gi}$ [m^{3}h^{-1}]	399.5	516.5	1563.0
Volume flow of fluidization agent 0.7 MW			318.2	1104.2
Mass-flow of solid fuel (0.7 MW)	$\Phi_{m,f}$[kgh^{-1}]	None	142	
Mass-flow of solid fuel (1 MW)	$\Phi_{m,f}$[kgh^{-1}]		170	

* Gas inlet at the bottom of the reactor; ** Gas outlet above the fluidized bed in the reactor

Results and discussion

The vital relative pressures, temperatures and gas-flows in both units were monitored. Table 2 shows a comparison between the predicted (*i. e.*, the mathematical model) and the

Table 2. Predicted and experimental results for the cold-flow and semi-industrial reactor in a stable loop

		Cold flow		Hot flow	
Gas velocity (hot flow 0.7 MW)	v_{g0}/v_{g1} [ms^{-1}]	1.57	NA	1.25	1.56
Gas velocity (hot flow 1 MW)				2.03	2.2
Gas velocity (modal method)	v_{mff} [ms^{-1}]	1.57		1.25	
Gas velocity (predicted)	v_{g0}/v_{g1} [ms^{-1}]	1.42	1.42	1.46	1.23
Pressure drop (experimental)	Δp [mbar]	62.4		58.6	
Pressure drop (predicted)	Δp [mbar]	62.3		62.4	

experimental values of the minimum gas velocities for the fully fluidized bed and the pressure drop in the bed in the cold-flow and semi-industrial reactor. The measured values are graphically presented in figs. 5 and 6 as (v_{mff}, Δp_{mff}). The comparison was made between the results evaluated using the Modal method [25] and the experimental results of Kaewklum and Kuprianov [21] and by actual experimental values in both units. The main purpose of this experiment was to evaluate and to check our mathematical model in the cold-flow reactor. On the basis of the cold-flow model, a semi-industrial system was designed, built and tested.

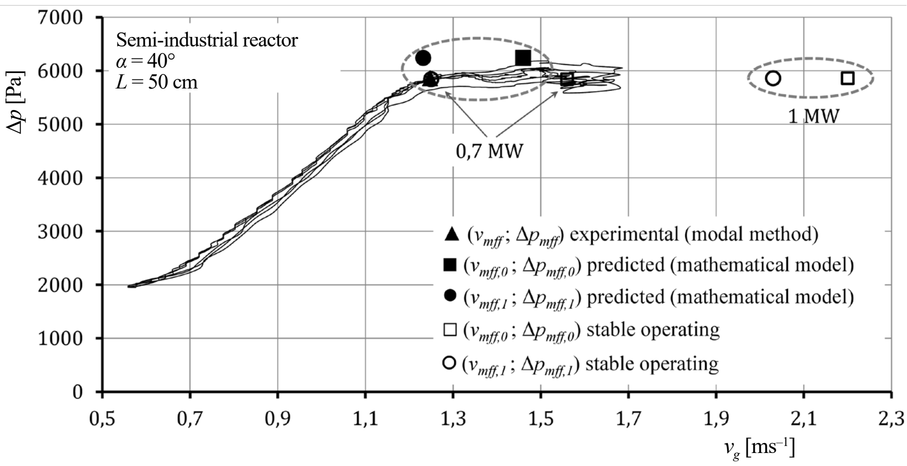


Figure 6. Comparison of the prediction and the experiment (v_{mff}, Δp_{mff}) for the average 400 μm sand particles fluidized in the 40° conical bed reactor at L = 500 mm

The experiments were made by fluidizing the quartz sand with cold air at approximately 40 °C in a cold-flow reactor and with superheated steam at 670 °C in the semi-industrial reactor. The average particle diameter was 400 μm. The stationary bed in both reactors was L = 500 mm. Our primary interest was to control the fluidized bed near the minimum velocity of the fully fluidized bed and consequently to minimize the consumption of superheated steam as a gasification agent. The fully fluidized state was experimentally detected using the Modal method [25].

The fully fluidized state in the cold-flow model was achieved at about 1.57 m/s and with Δp = 62.4 mbar across the bed. The predicted values were 1.42 m/s and 62.3 mbar. The predicted and experimental results were compared to those by Kaewklum and Kuprianov [21] for stationary bed heights of 200, 300, and 400 mm. It is clear that the results are in good agreement, with the difference being less than 20%, fig. 5. This gave us the motivation to continue the design of the semi-industrial reactor.

The main difference between the cold and the hot reactors' operating conditions was a phase change of the solid fuel at high temperatures in the semi-industrial reactor. The volume increase during the process of gasification can be predicted [19, 26, 27]. With the mathematical model, the minimum fluidization velocities at the bottom and at the top of the fluidized bed were calculated. At the bottom of the reactor the predicted superficial velocity of the superheated steam was 1.46 m/s, while at the top of the bed it was 1.23 m/s in order to achieve the fully fluidized state. The gas velocity of the fully fluidized bed determined using the Modal method was 1.25 m/s, which shows a 22% deviation. In this case the predicted pressure drop is 58.6 mbar, which is only 3.8 mbar higher than the experimentally measured value fig. 6.

By observing the volume flows at 0.7 MW_t there was 318.2 m^3/h of steam and 1104.2 m^3/h of syngas, and at 1 MW_t there was 516.5 m^3/h of steam and 1563 m^3/h of syngas at the reactor exit, tab. 1. A large increase of the gas velocity can be expected if we have a non-conical reactor. This shows us that the gasification reactor must have a conical shape with an angle α established by the proposed method.

By optimizing the particle dynamics, we achieved stable operation between 0.7 MW_t and 1 MW_t, and a high calorific syngas composition. The calorific value of the dry syngas was ≈10 MJ/Nm^3. The raw syngas contains ≈30 vol.% of water steam. To avoid short, undesirable defluidizations of the bed material under the minimum fluidization conditions and to produce more hydrogen in the syngas [28] the optimal operational superficial gas velocity in the reactor was $v_{lim,g} = v_{mff} + 20\%$. The superficial gas velocities were 1.25 m/s at the bottom and 1.56 m/s at the top of the fluidized bed when having 0.7 MW_t. At 1 MW_t the superficial velocity of the steam was 2.03 m/s and 2.2 m/s of syngas, fig. 6. Finally, the gasification has stable operation when in excess of the minimum fluidization between 0.7 MW_t and 1 MW_t, and the pressure drop, once fluidization occurs, is no longer governed by Ergun's equation, but instead by support of the bed weight minus the buoyancy.

Conclusions

Fluidization is widely used in process industry. Fluidized beds provide a more homogenous temperature profile and better heat transfer, as well as allowing the possibility to use the catalyst as the bed material, and allow the use of a solid fuel with a wide size distribution of particles. At the outset, we carried out basic engineering procedures for the actual 1 MW_t FICFB gasifier. Due to the high costs that would be incurred if there were errors in our model, the decision was made to test the process in a cold-flow unit. The proper construction and scaling of a cold flow model is still one of the most challenging aspects of fluidization technology, in particular with reactive fluidized beds. Chemical reactions during gasification and combustion cause variations in the temperature, density, composition and dynamic viscosity, all of which affect the particle dynamics, kinetics, mass transfer phenomena, *etc.* and finally the efficiency of the reactor. It is practically impossible to create the similarity by equalizing all the criteria in the scale-up set. However, even if we had been able to equalize all the scale-up parameters, it would still be based on assumptions. In order to evaluate the model, the results obtained from the experiments on the semi-industrial unit and the cold-flow model were compared to the mathematical model's results. In this paper, we propose a new design method for the gasification reactors. Due to gasification, the volume flow within the fluidised bed is increased. To sustain a suitable fluidization regime the use of a conical reactor is proposed. The angle of the reactor is determined iteratively based on the assumption that the fluidization regime should not change too much throughout the whole bed layer.

If the semi-industrial gasifier is operating continuously and producing syngas with the expected chemical composition, and if it has a similar pressure-flow profile as the cold-flow model, then it can be assumed that the scale-up was successfully designed. At this point, the systems are considered to be running in a stable loop.

The scale-up protocol was confirmed with the successful demonstration of the FICFB semi-industrial unit located in city of Celje, Slovenia.

References

[1] Hofbauer, H., *et al.*, Six Years Experience with the FICFB-Gasification Process, *Proseedings*, 12th European Conference and Technology Exhibition on Biomass for Energy, Industry and Climate Protection, Amsterdam, The Netherlands, 2002
[2] Kaiser, S., *et al.*, Simulation of a Highly Efficient Dual Fluidized Bed Gasification Process, *Proceedings*, 3rd European Congress on Chemical Engineering, Nuremberg, Germany, 2001
[3] Breault, R. W., Mathur. V. K., High Velocity Fluidized Bed Hydrodynamic Modelling, 1. Fundamental Studies of Pressure Drop, *Industrial and Engineering Chemistry Research*, *28* (1989), 6, pp. 684-695
[4] Bai, D., *et al.*, Analysis of the Overall Pressure Balance around a High-Density Circulating Fluidised Bed, *Industrial and Engineering Chemistry Research*, *36* (1997), 9, pp. 3898-3907
[5] Kaiser, S., *et al.*, Hydrodynamics of a Dual Fluidized Bed Gasifier – Part II: Simulation of Solid Circulation Rate, Pressure Loop and Stability, *Chemical Engineering Science*, *58* (2003), 18, pp. 4215-4223
[6] Loffler, G., *et al.*, Hydrodynamics of a Dual Fluidized – Bed Gasifier – Part I: Simulation of a Riser with Gas Injection and Diffuser, *Chemical Engineering Science*, *58* (2003), 18, pp. 4197-4213
[7] Leckner, B., Regimes of Large-Scale Fluidized Beds for Solid Fuel Conversion, *Powder Technology, 308* (2017), Feb., pp. 362-367
[8] Rudisuli, M., *et al.*, Scale-up of Bubbling Fluidized Bed Reactors – A Rreview, *Powder Technology, 217* (2012), Feb., pp. 21-38
[9] Glicksman, L. R., Scaling Relationships for Fluidized Beds, *Chemical Engineering Science, 39* (1982), 9, pp. 1373-1384
[10] Lopez., G., *et al.*, Assessment of a Conical Spouted with an Enhanced Fountain Bed for Biomass Gasification, *Fuel, 203* (2017), Sept., pp. 825-831
[11] Saldarriaga, J. F., *et al.*, Correlations for Calculating Peak and Spouting Pressure Drops in Conical Spouted Beds of Biomass, *Journal of the Taiwan Institute of Chemical Engineers, 80* (2017), Nov., pp. 678-685
[12] Alia, N., *et al.*, An Advanced Evaluation of the Mechanistic Scale-Up Methodology of Gas–Solid Spouted Beds Using Radioactive Particle Tracking, *Particuology, 34* (2017), Oct., pp. 48-60
[13] La Villettaa, M., *et al.*, Modelling Approaches to Biomass Gasification: A Review with Emphasis on the Stoichiometric Method, *Renewable and Sustainable Energy Reviews, 74* (2017), July, pp. 71-88
[14] Mahinpey, N., Gomez, A., Review of Gasification Fundamentals and New Findings: Reactors, Feedstock, and Kinetic Studies, *Chemical Engineering Science, 148* (2016), July, pp. 14-31
[15] Gupta, S. K., *et al.*, Prediction of Minimum Fluidization Velocity for Fine Tailing Materials, *Powder Technology, 196* (2009), 3, pp. 263-271
[16] Kunii, D., Levenspiel. O., *Fluidization Engineering*, John Wiley and Sons Inc., New York, USA, 1969
[17] Girimonte, R., Formisani, B., The Minimum Bubbling Velocity of Fluidized Beds Operating at High Temperature, *Powder Technology, 189* (2009), 1, pp. 74-81
[18] Foscolo, P. U., *et al.*, Design and Cold Model Testing of a Biomass Gasifier Consisting of Two Interconnected Fluidized Beds, *Powder Technology*, *173* (2007), 3, pp. 179-188
[19] Kern, S., *et al*., Gasification of Wood in a Dual Fluidized Bed Gasifier: Influence of Fuel Feeding on Process Performance, *Chemical Engineering Science*, *90* (2013), Mar., pp. 284-298
[20] Peng, Y., Fan, L., T., Hydrodynamic Characteristic of Fluidization in Liquid – Solid Tapered Beds, *Chemical Engineering Science, 52* (1997), 14, pp. 2277-2290
[21] Kaewklum, R., Kuprianov, V., Theoretical and Experimental Study on Hydrodynamic Characteristic of Fluidization in Air-Sand Conical Beds, *Chemical Engineering Science*, *63* (2008), 6, pp. 1471-1479
[22] Jankes, G., G., *et al.*, Biomass Gasification with CHP Production, *Thermal Science*, *16* (2012), 1, pp. 115-130
[23] Gomez-Barea, A., Leckner, B., Modelling of Biomass Gasification in Fluidized Bed, *Progress in Energy and Combustion Science*, *36* (2010), 4, pp. 444-509

[24] Kaltschmitt, M., *et al*., *Energy from Biomass* (Energie aus Biomasse – in Germany), Springer, Berlin, Heidelberg, Germany, 2009
[25] Mele, J., *et al.*, A Method to Detect and Control Fully Fluidized Conical Beds with a Wide Size Distribution of Particles in the Vicinity of the Minimum Fluidization Velocity, *Thermal science*, *19* (2015), 1, pp. 267-276
[26] Nguyen, T., D., B., *et al.*, Three-Stage Steady-State Model for Biomass Gasification in a Dual Circulating Fluidized-Bed, *Energy Conversion and Management*, *54* (2012), 1, pp. 100-112
[27] Miao, Q., *et al.*, Modelling Biomass Gasification in Circulating Fluidized Beds, *Renewable Energy*, *50* (2013), Feb., pp. 655-661
[28] Goransson, K., *et al.*, An Experimental Study on Catalytic Bed Materials in a Biomass Dual Fluidised Bed Gasifier, *Renewable Energy, 81* (2015), Sept., pp. 251-261

Paper submitted: November 29, 2016
Paper revised: March 28, 2018
Paper accepted: May 13, 2018