THE INFLUENCE OF OPERATING CONDITIONS AND FUEL-FEED LOCATION ON FUEL RESIDENCE TIME IN AN INDIRECT GASIFIER

Erik Sette*, David Pallarès, Filip Johnsson

Department of Energy and Environment

Chalmers University of Technology, S412 96 Göteborg, Sweden

**Email: sette@chalmers.se*

ABSTRACT

This work investigates the lateral mixing process of fuel in a cold flow model of a fluidized-bed gasification chamber in an indirect gasification system. The bed has a significant cross-flow of bulk solids and the fuel mixing is investigated with respect to how operational conditions (fluidization velocity and solids circulation rate) and the fuel-feed location influence the fuel residence time in the gasification chamber. Fluidization velocity and fuel feeding location are found to influence fuel residence time the most, whereas solids circulation rate has no significant effect at the conditions tested.

INTRODUCTION

The performance of a chemical reactor is dependent on the residence time of the reacting species. In order to achieve high conversion the reactants need to have enough time to react. In the case of indirect gasification the fuel particles need to have enough time to react with the gasification agent (fluidization medium), otherwise unconverted fuel will leave the gasifier bed and enter the combustor (Thunman and Seeman, 2009) (Pfeifer et al., 2009) reducing the gas yield from the system. Since gasification reactions are endothermic, heat has to be transferred to the gasifier. In indirect gasification this heat is provided from a combustor through circulation of bed material between the combustor and gasifier. Thus for good performance of an indirect gasifier sufficient amount of heat has to be transferred to the reactor (i.e. sufficient circulation of bed material) without compromising the fuel residence time in the reactor.

Several methods for determining the residence time distribution in fluidized beds have been proposed in the literature and a summary is given by Harris et al. (2002). Most of the techniques have in common that they measure the impulse response of a tracer according to the method proposed by Danckwerts (1953). Harris et al. classifies the tracer methods as either disruptive or non-disruptive, where disruptive methods would disturb the field as opposed to non-disruptive methods (Harris et al., 2002). According to Harris a method is disruptive either because the tracer particles have different properties compared to the bulk phase or the sampling of the tracer particles disturbs the flow field. Although the importance of keeping the tracer particle properties similar to the bulk material properties is stressed in literature, few authors consider fluid-dynamic scaling. An exception is the work by Guío-Pérez et al. (2013), who also pointed out the importance of using a tracer which has similar fluid-dynamic properties as the bulk solids when determining the residence time distribution of the bed material. In the present work, a further step is taken by not only scaling the properties of the bulk solids but also those of fuel particles, thereby preserving the density ratio between bulk solids and fuel particles.

This work characterizes the fuel residence time through non-disruptive measurements of the transient outlet concentration of tracer particles at the solids outlet of the cold flow model of the Chalmers gasifier (Figure 1). The influence of fluidization velocity, solids circulation rate in the system and the location of the fuel feeding on the fuel residence time is studied. Furthermore, possibilities are investigated to estimate the fuel residence time by means of simple model formulations.

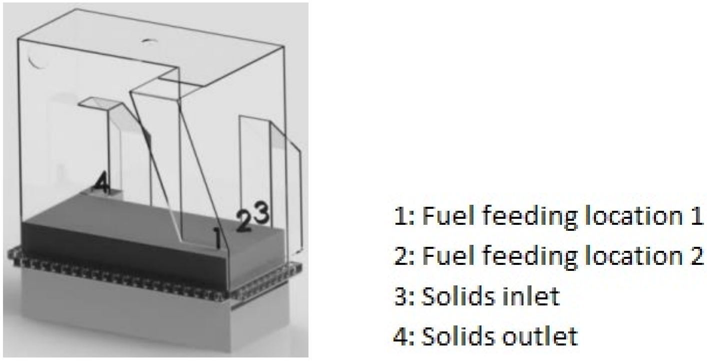


Figure 1: The fluid dynamically downscaled fluidized bed resembling the Chalmers 4 MW indirect gasifier.

DATA EVALUATION

From the measured concentration of tracer particles, *C*, the average residence time of fuel particles, τ_f, is determined according to Eqs. (1) and (2) (Fogler, 2005).

$$\tau_f = \int_0^{\infty} t \cdot E(t) dt \tag{1}$$

$$E(t) = \frac{C(t)}{\int_0^{\infty} C(t) dt} \tag{2}$$

Solids lateral mixing is commonly quantified using a dispersion coefficient (see Niklasson et al., 2002 and Liu and Chen, 2010) for earlier experimental determination of such in large fluidized bed units) which accounts for the mixing induced by gas bubbles. In dual fluidized bed systems such as the indirect gasification process studied in this work there is also a continuous cross-flow of bulk solids which affects the solids mixing. Thus there are two mechanisms which contribute to lateral mixing of solids: dispersive bubble-induced mixing and a convective mixing mechanism induced by the cross-flow of solids. The convective cross-flow timescale can be expressed as,

$$\tau_{bm} = \frac{m}{\dot{m}} \tag{3}$$

knowing the amount of bed material in the reactor and rate of solids cross-flow. The timescale for dispersive bubble-induced mixing can be determined as,

$$\tau_d = \frac{L^2}{2 \cdot D} \tag{4}$$

where L is the characteristic distance travelled by the fuel and D is the lateral dispersion coefficient. For the latter, this work uses experimentally-determined values by means of the method presented by Sette et al. (2013).

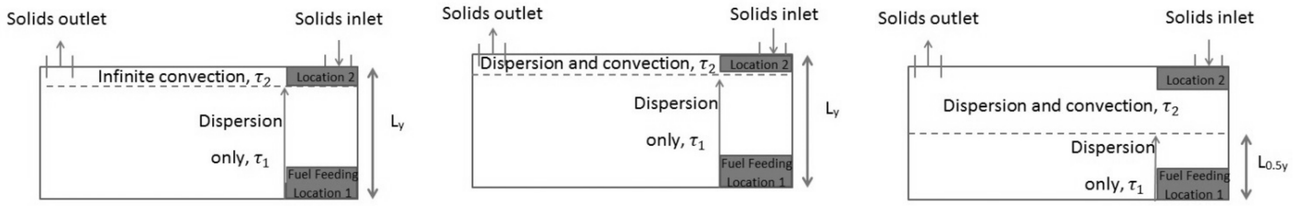
The characteristic timescales of the two mechanisms contributing to solids lateral mixing can be combined to yield the mean fuel residence time in the reactor according to,

$$\tau_f = \frac{1}{\frac{1}{\tau_{bm}} + \frac{1}{\tau_d}} \tag{5}$$

It has been experimentally shown by Olsson et al. (2014) that fuel particles do not fully follow the velocity field induced by the cross-flow of bulk solids but have a certain slip. This is accounted for through the cross-flow impact factor, θ. If fuel particles adopt the velocity of the bed material, θ equals 1, and if the fuel has a certain slip θ becomes lower than 1. With this, the mean fuel residence time (τ_f) accounting for both dispersion and convection can be estimated using Eq. (6).

$$\tau_f = \frac{1}{\frac{\theta}{\tau_{bm}} + \frac{1}{\tau_d}} \tag{6}$$

With this approach, two assumptions are made: i) the velocity field created by the solids cross-flow is homogenous over the entire cross section of the bed and ii) the dispersive flow is isotropic. While the latter is supported by the experimental work by Olsson et al. (2012) in the Chalmers gasifier, the first assumption is more questionable. It is reasonable to assume that the velocity field induced by a significant cross-flow of solids is larger in magnitude in the region near the circulating solids inlet and lower in magnitude in locations far apart from the solids inlet (such as the fuel feeding location, see “1” in Figure 1). Since the extent of this effect is not known, three simple model cases are considered in this work, as illustrated in Figure 2. The three model cases differ with respect to the assumption on cross sectional distribution of the region with presence of solids cross-flow. Fuel is assumed to mix only by dispersion from the fuel feeding location to the cross-flow region. The first case (Figure 2a) assumes that the cross-flow is concentrated to an infinitely narrow area along the wall where the bulk solids enter and thus, the magnitude of the velocity field induced by the cross-flow is tending to infinity, i.e. the fuel mixes by dispersion the full width of the unit and is thereafter transported instantly to the solids outlet by convection. In the second case (Figure 2b) the solids cross-flow also occurs along the inlet wall but now the solids cross-flow is given a finite velocity, which is related to the timescale given by Eq. (3) and thus both convection and dispersion are considered in the cross-flow region. In the last case (Figure 2c) the region with presence of cross-flow is assumed to occupy half of the bed cross section, i.e. the fuel mixes exclusively by dispersion over half the width of the bed cross section before reaching the region with presence of dispersion and convection.



a. b. c.

Figure 2: Schematic drawings of the cross section of the assumptions made in the three cases evaluated.

The residence times for the three cases considered are calculated using Eqs. (3), (4) and (6) and are thereafter compared with measured residence times.

EXPERIMENTAL SETUP

The experiments are carried out in a fluid-dynamically down-scaled fluidized bed resembling the Chalmers 4 MW indirect gasifier shown in Figure 1. The fluid-dynamic scaling proposed by Glicksman et al. (1994) is used to achieve dynamic similarity between the full scale and down-scaled units. Table 1 gives the scaling parameters for the downscaled and the full scale unit. In the cold flow model two locations for the fuel feeding are tested (see Figure 1): “1”, which corresponds to the original location in the Chalmers, and “2”, close to the bulk solids inlet (“3” in Figure 1). The scale model has a 1/6 length-scale of the full scale gasifier and is operated with bronze particles (75 μm) as inert bed solids and cylindrical aluminum particles (~1000 μm o.d.) representing pelletized biomass particles, which in this work are used as tracer particles for studying the fuel lateral mixing. As can be seen in

Table 1 there is a small difference in the density ratio between the bed material and fuel particle for the large scale gasifier and the downscaled unit. The tracer used has a somewhat too low density and could therefore exhibit a slightly more flotsam behavior than the real fuel particles. The system is equipped with a screw feeder which transfers the bed material from the solids outlet back to the solids inlet via a loop seal. Since the (aluminum) tracer particles are much larger in size compared to the bulk solids (bronze particles) they are separated at the solids outlet by a sieve. The residence time of the tracer particles is characterized by batch experiments in which tracer particles are sampled at the solids outlet duct of the fluidized bed (“4” in Figure 1). In the beginning of an experiment the unit is fluidized and the screw feeder is operated for some minutes to make sure a steady solids circulation is obtained. When a steady flow of bulk solids is established a batch of 1,000 tracer particles is inserted at the corresponding feeding location. Samples are collected at the solids outlet. The sampling time varies between 10 and 20 seconds per sample depending on the fluidization velocity: the higher the fluidization velocity the shorter the sampling time required to accurately resolve the transient change in concentration at the outlet.

Table 1: The scaling parameters applied.

Parameter	Units	Large scale, hot conditions	Downscaled, ambient conditions
Density of bed material (silica sand, bronze)	kg/m3	2600	8900
Density of fuel particles (wood pellet, aluminum)	kg/m3	950	2700
Density ratio (bed material / fuel)	-	2.7	3.3
Temperature	°C	800	20
Length scale factor	m	L	L/6
Time scale factor	s	t	$t \cdot \sqrt{L/6}$

The fuel residence time is investigated at three different fluidization velocities. In addition, the influence on feeding location and cross-flow circulation rate is analyzed for the lowest fluidization velocity. A more thorough investigation including several fluidization velocities is planned in the future. The cross-flow impact factor, θ, is evaluated using Eqs. (3), (4) and (6) with parameters for evaluating the different timescales given in Table 2.

RESULTS AND DISCUSSION

Figure 3 exemplifies a typical tracer transient concentration distribution obtained from the experiments. Included in the graph is the experimental average residence time of the tracer, *t*=291 s (corresponding to 712 s on up-scaled basis), determined using Eqs. (1) and (2). As can be seen in Figure 4, the fuel residence time is reduced considerably as the fuel feeding is moved from *Location 1* to *Location 2*, i.e. closer to the inlet of the circulating solids inlet but still at a similar distance from the reactor outlet. This indicates the existence of the two solids mixing mechanisms included in Eq. (6). When fuel is fed from *Location 2*, the fuel particle transport in the lateral direction will not depend on the dispersive transport (τ_1=0 in Figure 2) to reach the region influenced by the solids cross flow, but is instead directly fed into this cross-flow and, thus yielding a shorter residence time in the bed, i.e. $\tau_f=\tau_2$

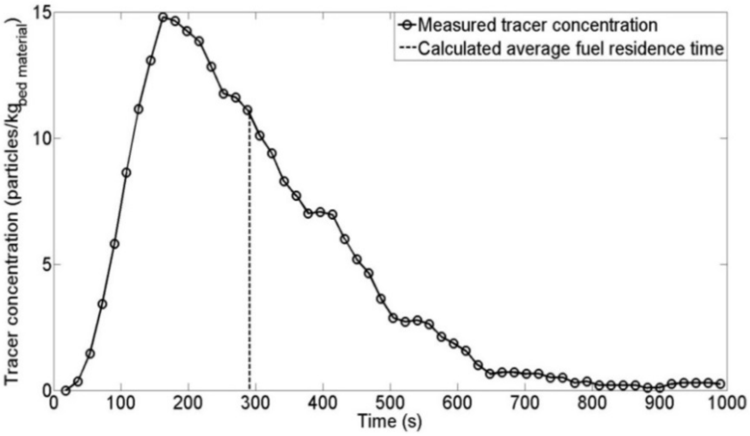


Figure 3: Measured tracer concentration and corresponding average fuel residence time, evaluated using Eqs. (1) and (2) given at (downscaled conditions).

In Figure 4 data points with open round markers show that the fuel residence time decreases as fluidization velocity is increased when the fuel is fed from *Location 1*, i.e. when the fuel particles are required to be transported along a longer path by the dispersive mechanism until they reach the region dominated by the solids cross-flow. This is expected and consistent with literature data (Niklasson et al., 2002, and references therein) showing that lateral fuel dispersion increases with fluidization velocity. Although the influence of fluidization velocity was not investigated for *Location 2* one can expect that the fuel residence time for this location will

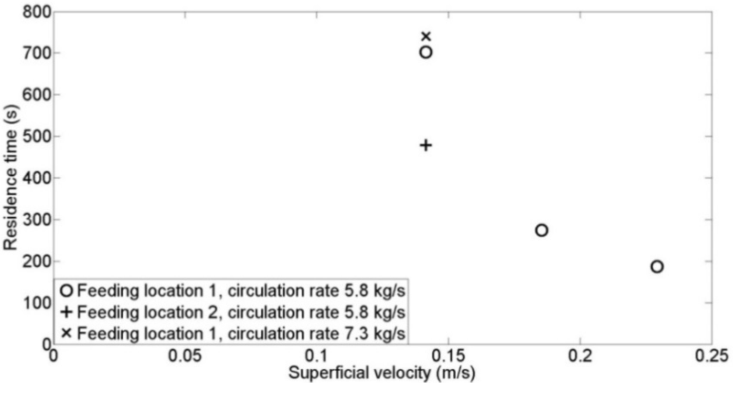


Figure 4: Measured fuel residence times for different fluidization velocities, solids circulation rates and fuel feeding locations (up-scaled values).

be less influenced by fluidization velocity since the importance of dispersive mixing is, as discussed, lower for this case. Figure 4 also shows that an increase in the solids cross-flow (from 5.8 to 7.3 kg/s) does not have any major influence on the residence time of particle fed from *Location 1*. Thus, this further indicates that the dispersive transport path from *Location 1* to the cross-flow region is the main limiting mechanism in the lateral fuel mixing for the configuration and operational conditions investigated.

After analyzing the experimental results, an attempt to estimate the fuel residence time by simple expressions and assumptions has been made. First, the cross-flow impact factor, θ, is estimated from experiments and thereafter Eq. (6) is used to estimate the fuel residence time for each of the 3 cases shown in Figure 2. Finally, the residence time calculated for each the 3 cases are compared with the experimental value in order to assess the validity the 3 cases considered.

Table 2: Values used in the determination of the cross-flow impact factor, θ, with Eqs. (3), (4) and (6).

Fluidization velocity (m/s)	Solids cross-flow, $\dot{m}$, (kg/s)	Mass of bulk solids (kg)	Lateral dispersion coefficient, D, (m^2/s)	Dispersive distance	Cross-flow impact factor, θ
0.14	5.8	750	0.0012	L_y	0.09
0.14	5.8	375	0.0012	$0.5L_y$	0.04

The cross-flow impact factor, θ, is estimated using the measured residence time obtained in the test in which fuel feeding *Location 2* was used in combination with Eqs. (3), (4) and (5). The data in Table 2 is used in evaluating the timescales given by Eqs. (3) and (4). With this, estimated values for the cross-flow impact factor of 0.09 and 0.04 are obtained under the assumptions for model cases b and c. Note that the assumptions in these two cases lead to different convective time scales, thereby yielding different values of the cross-flow impact factor. Note also that case a assumes infinitely fast convective transport, so no cross-flow impact factor is calculated for it.

With the cross-flow impact factors experimentally evaluated, the ability of the three different model cases to provide reasonable estimations of the fuel residence time is evaluated. The estimations for the experimental cases at a fluidization velocity of 0.14 m/s and fuel feeding from *Location 1* are given in Table 3 together with the experimental values. The residence times estimated from model case 2 and 3 show the best match to experimental values. This shows that it is important to account for both the bubble induced dispersion and the convective cross-flow of solids when estimating the residence time of fuel particles in fluidized beds with cross flow of solids. However, all of the three model assumptions considered give relatively large errors in the estimation of the fuel residence time. Thus, simple expressions cannot be used to provide good estimations of the fuel residence time in wide beds with significant solids cross-flow and more advanced modeling is required which can resolve the details of the velocity field of the solids cross flow over the bed, both horizontally and at different depths in the bed.

CONCLUSION

An experimental method to characterize the fuel residence time distribution in a fluidized bed with significant cross-flow flow of solids has been developed. Experimental results show that, in the ranges tested, fluidization velocity and fuel feeding location have a strong impact on the fuel residence time, while solids circulation has not. The experiments confirm that the lateral fuel mixing is influenced by two mechanisms: bubble-induced dispersion and a convective mixing due to the cross-flow of bulk solids through the bed. When fuel is fed from the wall opposite to the solids inlet, *Location 1*, the mixing by dispersive means until reaching the cross-flow region is the limiting mechanism governing the fuel residence time in the bed. This limiting mixing mechanism becomes faster with increased fluidization velocity.

Simple analytical expressions could not provide good estimations of the fuel residence time for different assumption setup, indicating that more advanced (discretized) modeling is required which can resolve the velocity field induced by solids cross-flow.

Table 3: Calculated residence times for the three different flow systems and the difference to the measured values.

Fuel Feeding location: 1, Fluidization velocity: 0.14 m/s, solids cross-flow rate: 5.8kg/s			
Case	Residence time calculated, τ_c	Residence time measured, τ_m	Relative error $(\tau_m - \tau_c)/ \tau_m$
1	267	702	0.62
2	961	702	-0.37
3	533	702	0.24
Feeding location: 1, Fluidization velocity: 0.14 m/s, solids cross-flow rate: 7.3kg/s			
1	267	740	0.64
2	885	740	-0.20
3	467	740	0.36

ACKNOWLEDGEMENT

This work has been supported by Akademiska Hus, Göteborg Energy AB, Metso Power AB and Chalmers Energy Initiative.

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