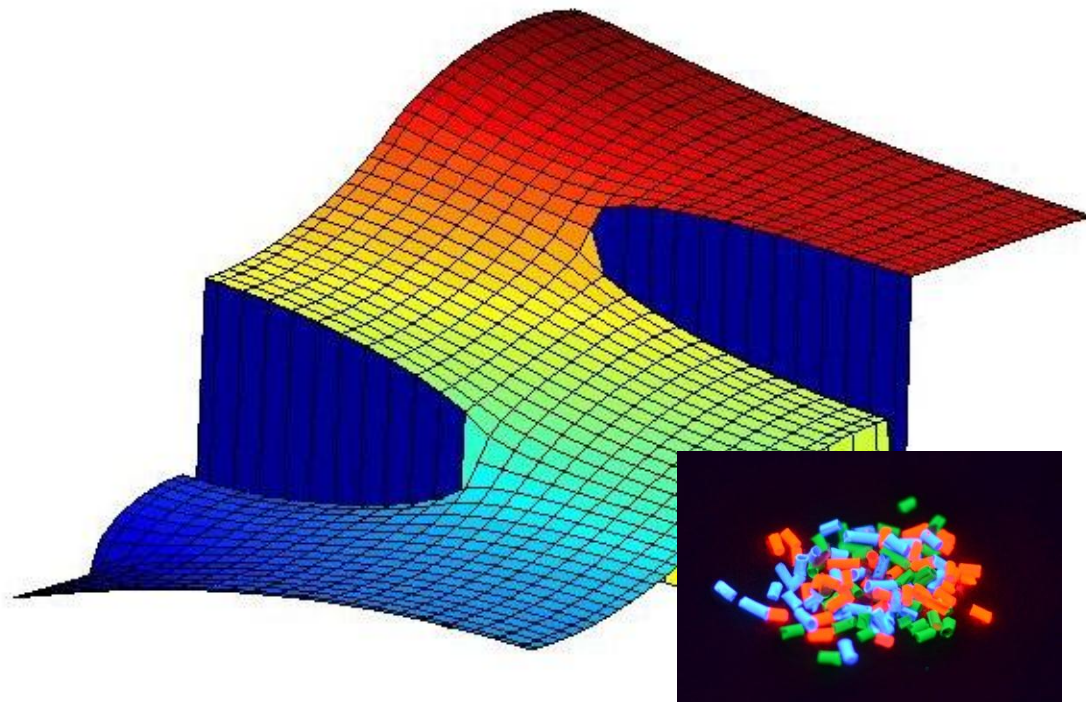




Modelling and optimization of fuel conversion in an indirect bubbling fluidized bed gasifier

Master's Thesis within the Sustainable Energy Systems programme

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CHALMERS UNIVERSITY OF TECHNOLOGY
Göteborg, Sweden 2013
Master Thesis 2013: T2013-389

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Cover:

Temperature field inside gasifier obtained in this work and fuel particles painted with UV activated paint under UV light.

Chalmers Reproservice

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ABSTRACT

The aim of this thesis was to improve the fuel conversion in Chalmers bubbling fluidized bed (BFB) gasifier. A mathematical model has been developed and experiments in a fluid-dynamically scaled reactor have been conducted. To increase fuel conversion the use of horizontal tubes and partitioning walls was studied.

The mathematical model is 2-dimensional and accounts for fluid dynamics, heat transfer and fuel conversion. The model was also used for determining the dispersion coefficient, D , and velocity slip factor, α , from experimental data.

Lateral bed material mixing was evaluated through an indirect method by tracing iron powder in the bed material. The lateral fuel mixing was evaluated through a direct method using digital image analysis. A new indirect method to investigate the fuel lateral fuel dispersion and the convective bed material flow influence was developed and evaluated. The method was easy, quick and robust, thus showing great potential.

The resulting up-scaled values of dispersion coefficients for the bed material were in a range of $2.44 - 4.77 \cdot 10^{-3}$ and $2.15 - 6.17 \cdot 10^{-3} \text{ m}^2/\text{s}$ for fuel particles. The dispersion coefficients at tube banks for bed material and fuel particles were $1.08 - 1.34 \cdot 10^{-3}$ and $1.07 - 1.79 \cdot 10^{-3} \text{ m}^2/\text{s}$ respectively. This shows that tube banks reduced the lateral dispersion of bed material and fuel by around 70% whereas the partitioning walls had little impact on the modelled char conversion. Tube banks also reduced the influence of the convective flow of bed material on fuel, α , with 40% to 100% thereby increasing the residence time of the fuel.

Through simulations of the tuned model for different arrangements of internals, it was found that the residence time of the fuel is the main parameter for improved fuel conversion. The modelled char conversion was increased from 6% for a bed without internals to 22% when two thirds of the beds were covered with a properly placed tube bank.

Key words: Solids mixing, Fuel mixing, Gasification, Biomass, Glicksman, CFD modelling

Contents

ABSTRACT	I
CONTENTS	III
ACKNOWLEDGMENT	V
NOTATIONS	VI
1 INTRODUCTION	1
1.1 Aim	1
1.2 Background	1
1.3 Problem Statement	2
2 METHOD	3
2.1 Overview	3
2.2 Solids mixing - Theory	3
2.2.1 Bed material mixing	3
2.2.2 Fuel mixing	4
2.3 Mathematical model - Transport equation	5
2.4 Solids mixing - Mathematical model	6
2.4.2 2D Unsteady Dispersion	7
2.4.3 2D Unsteady Convection	9
2.5 Fuel conversion	12
2.5.1 Allocation of drying/devolatilisation and char conversion	13
2.6 Energy balance - Mathematical model	14
2.6.1 Steady-state heat transfer through conduction	14
2.6.2 Steady state heat transfer through solids dispersion	16
2.6.3 Steady state heat transfer through solids convection	17
2.6.4 Solver by a Gauss-Seidel iterative method	18
2.7 Impact of internals	19
2.8 Experimental setup	20
2.8.1 Scaling of fluidised beds	20
2.8.2 Scale model setup	21
2.8.3 Redesign of bed material feed system	24
2.8.4 Experimental work	26
3 RESULTS	28
3.1 Dispersion of bed material	28
3.2 Dispersion of fuel	31
3.2.1 Indirect method	31
3.2.2 Direct method	33
3.3 Up-scaling of experimental results	34
3.4 Model tuning	34

3.5	Optimisation	36
3.5.1	Impact of internals	36
3.5.2	Redesign options	38
3.5.3	Best design vs. current design	40
3.5.4	Sensitivity analysis	43
4	CONCLUSION	44
5	FURTHER WORK	45
6	LITERATURE LIST AND REFERENCES	46

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Göteborg February 2013

Joakim Larsson & Jens Olsson

Notations

Roman upper case letters

A	Area, [m ²]
Bi	Biot number, [-]
C	Concentration, [kg/m ²]
$C_{d,or}$	Orifice coefficient, [-]
C_p	Heat capacity, [J/kg·K]
CV	Control volume, [m ²]
D_{bed}	Dispersion coefficient for bed material, [m ² /s]
D_{fuel}	Dispersion coefficient for fuel particles, [m ² /s]
E	Activation energy, [J/mol]
E_f	Local stress tensor fluid, [Pa]
E_p	Local stress tensor particle, [Pa]
F	Drag force between gas and solids, [kg m ⁻² s ⁻²]
Fr	Froude number, [-]
G_s	Solids flux, [kg/s]
H	Height, [m]
H_m	Specific energy, [J/K]
L	Length, [m]
L_p	Particle length, [m]
M	Molar mass, [kg/mol]
N_{or}	Number of nozzles, [no./m ²]
P	Pressure, [Pa]
R	Gas constant, [J K ⁻¹ mol ⁻¹]
Re	Reynolds number, [-]
T	Temperature, [K]
T_{sat}	Saturation temperature, [K]
T_{∞}	Surrounding temperature, [K]
V	Volume, [m ³]
W	Width, [m]
Y	Mass fraction, [-]

Roman lower case letters

d	Diameter, [m]
d_h	Hydraulic diameter, [m]
d_p	Particle diameter, [m]
d_{or}	Orifice diameter, [m]
dz	Bed height, [m]
f	Frequency, [Hz]
g	Universal gravitational constant, [m/s ²]
h	Heat transfer coefficient, [W m ⁻² K ⁻¹]
k	Thermal conductivity, [W m ⁻¹ K ⁻¹]
m	Mass, [kg]
$\dot{m}$	Mass flow, [kg/s]
$\mathbf{n}$	Unit vector, [-]
r	Particle radius, [m]

t	Time, [s]
u_m	Superficial gas velocity, [m/s]
u_{or}	Orifice velocity, [m/s]
$\mathbf{u}$	Velocity field, [m/s]

Greek letters

Δt	Time step, [s]
α	Velocity slip factor, [-]
ϵ	Bed voidage, [-]
ε	Convergence value, [-]
λ	Characteristic length, [m]
μ	Dynamic viscosity, [kg s ⁻¹ m ⁻¹]
ρ	Density, [kg/m ³]
ρ_A	2D density, [kg/m ²]
ρ_f	Density of fluidisation medium, [kg/m ³]
ρ_s	Density of solids, [kg/m ³]
ϕ	General variable

Subscripts/Superscripts

dry	Drying
devol	Devolatilisation
char	Char conversion
eff	Effective
L	Large scale unit
M	Model scale unit
P	Centre node
N	North
E	East
S	South
W	West
n	north
e	east
s	south
w	west

1 Introduction

1.1 Aim

The aim of this thesis is to develop a method to improve fuel conversion in a bubbling fluidized bed (BFB) gasifier through experiments and mathematical modelling. Experimental work was performed in a fluid-dynamically down-scaled model of the Chalmers bubbling fluidized bed biomass gasifier and the mathematical model was built in Matlab.

1.2 Background

In 2007 the European Union set ambitious goals for an increased share of renewable energy in both electricity and transport sectors [1]. Gaseous fuels derived from biomass can be used in both sectors either in gaseous state or as refined liquid fuels obtained from conversion processes such as Fischer-Tropsch synthesis [2].

In a gasifier solid fuel is converted to gaseous state through endothermic reactions. In the case of direct gasification a small amount of oxidiser is used to partially react with the fuel and product gases and release the heat needed to sustain the endothermic reactions. In an indirect gasifier heat to the process is supplied by means of hot bed material from an external combustion reactor, i.e. no combustion takes place in the gasifier. The bed material is circulated through the gasification reactor and then back to the combustion reactor, see Figure 1. Inside the gasification reactor the fuel is dried, devolatilized and char is gasified through heat transfer and reactions. The main final products from indirect gasification are hydrogen and carbon monoxide whereas the product from direct gasification process also includes combustion products and nitrogen (if air is used as oxidiser carrier).

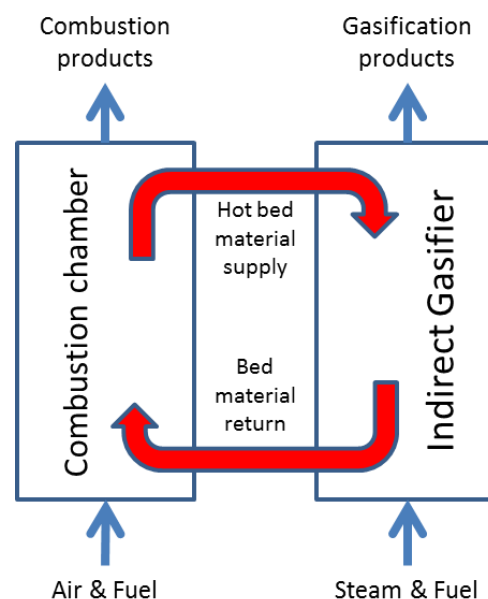


Figure 1 Illustration of an indirect gasifier.

Due to the low extent of commercialization of the gasification process it is less known than the combustion process [3]. Also, most research, with respect to fuel and bed dispersion inside bubbling fluidized beds, has been done without any significant through flow of bed material [4] [5].

Recent measurements suggest that as little as 0-5% of the char is converted in the gasification process while the rest is transported to the combustion chamber with the bed material returning to the combustion chamber.¹

1.3 Problem Statement

The conversion rate for a gasification process is slower than for combustion, in particular for char where combustion reaction is one to two orders of magnitude faster than for gasification [6]. As the processes involved in gasification are endothermic there is also a need for a significant amount of heat, which in indirect gasification is provided by the mass flow of hot bed material. The mass flow yields the need for large gasification chambers. In order to improve the fuel conversion in the gasification chamber, means to increase reaction rate and prolong fuel particle residence time need to be investigated.

In indirect gasifiers the lateral dispersion of bed material and fuel has great importance as it influences the temperature field in the reactor, thereby having a large impact on the rate and efficiency of conversion especially for char gasification as stated above. Currently, research has mainly been done on stationary fluidized bed combustors/gasifiers where lateral dispersion of bed material, [7] - [9], and fuel particles, [4] [5] [10] - [12], has been the focus.

The influence of a significant high mass flow of bed material across the reactor poses a convective problem previously not addressed. In an ideal gasifier the bed and fuel flow would be in a counter current system but this is apparently not attainable in a standard BFB [13]. Instead, the gasification chamber needs to be designed to promote this behaviour by means of altering the bed and fuel feeding points, introducing internals modifying the flow pattern or varying operating conditions.

¹ Anton Larsson, verbally on the 10th of January 2013

2 Method

2.1 Overview

In a BFB furnace inadequate lateral mixing can lead to high concentrations of fuel, creating hot pockets, unburned fuel due to low oxygen concentration and large temperature gradients in the freeboard of the furnace lowering its performance [4]. In BFB gasifiers the mixing of fuel and solids are governed by the same parameters as for BFB furnaces but they have the opposite impact creating cold pockets due to its endothermic reactions. The temperature field is also influenced by a high through flow of bed material. The use of internals such as baffles, tube banks and walls can promote reactions and increase residence time if designed correctly.

In this thesis a mathematical model has been developed and experiments in a fluid-dynamically scaled reactor have been conducted in order to optimize the fuel conversion efficiency of the Chalmers BFB gasifier.

2.2 Solids mixing - Theory

2.2.1 Bed material mixing

In a BFB reactor the fluidisation medium is introduced through nozzles placed in the bottom of the bed creating an upward flow of bubbles. The bubbles create paths through the bed and, depending on operating condition and gas distributor design, coalesce to form larger bubbles. The flow pattern has been shown to form horizontally aligned mixing cells [10], see Figure 2.

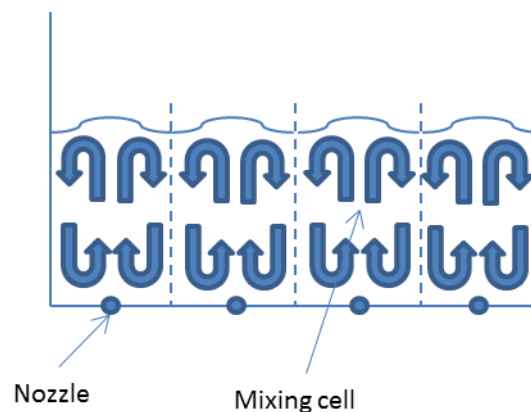


Figure 2 Nozzle placement and mixing cell alignment

Each bubble path creates a flux of solids in its vicinity which is governed by emulsion drift sinking, bubble wake lifting and bubble eruption. These parameters can be described by one variable, the lateral dispersion coefficient, D , treating the macroscopically horizontal solids movement inside the reactor as a random walk [9]. The dispersion coefficient is based on an analogy between Brownian movement theory and lateral mixing. The bed material and/or fuel particles are assumed to randomly move in the same way as particles diffuse in a gas, although at different length scales. However, this analogy is only valid for reactors where sufficiently many mixing cells are present in the bed so a continuum can be assumed as [5]. This

diffusive motion is described in mathematical terms by Fick's law (1) where difference in concentration governs the net flow.

$$j_i = -D_i \frac{\partial \rho_i}{\partial x} \quad (1)$$

There have been numerous attempts to theoretically predict the lateral dispersion coefficient, as listed by [4]. Most commonly indirect methods as manual sampling and concentration measurement over bed have been used to experimentally determine the dispersion coefficient. However, as most of experiments were conducted in small scale units under cold conditions without applying any scaling laws for conservation of the fluid dynamical behaviour the relevance of these results are limited. Thus, dispersion coefficients found in these works should not be used outside the experimental ranges from which they were derived.

2.2.2 Fuel mixing

Fuel mixing can be described in a similar way as for bed material mixing described in Section 2.2.1. However, the fuel particles physical properties differ strongly from the bed material in terms of size, shape and density. Fuel is lighter, bigger and the shape is non-spherical and varies in size. In addition, fuel undergoes strong gas releases, i.e. drying and devolatilisation, which lift the fuel particles to the bed surface much faster than for inert particles with similar properties [12] [14] [15].

Most previous research determining the fuel lateral dispersion has been based in indirect estimations where (2) is fitted to experimental data [4] [12] [16]. Also, direct methods have been developed, [10] [11], which allow the use of the Einstein's expression for particle dispersion [17]:

$$D_x = \frac{\lambda_x^2}{2\Delta t} \quad (2)$$

where λ_x is the characteristic length as per Figure 3.

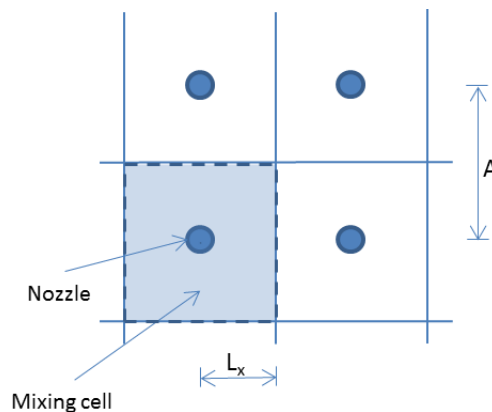


Figure 3 Nozzle placements and mixing cells where A is the nozzle spacing and L_x is the characteristic length.

2.3 Mathematical model - Transport equation

A mathematical model of the gasification chamber was implemented through the construction of a computational fluid dynamics, CFD, model. The model simulates the gasification chamber in terms of mass transport (bed material and fuel mixing), gas flow, chemical reactions/processes and heat transfer. The numerical method used in the model is the finite volume method, a special finite difference formulation [18].

The governing equation of the model is the *transport equation* of a general property ϕ expressed in equation (3).

$$\frac{\partial(\rho\phi)}{\partial t} + \text{div}(\rho\phi\mathbf{u}) = \text{div}(\Gamma \text{ grad } \phi) + S_\phi \quad (3)$$

The physical meaning of the terms in the transport equation is explained in Table 1 below.

Table 1

Rate of increase of ϕ inside the control volume	+	Net rate of decrease of ϕ due to convection across the control volume boundaries	=	Net rate of increase of ϕ due to diffusion across the control volume boundaries	+	Net rate of creation of ϕ inside the control volume
--	---	--	---	---	---	---

The numerical algorithms to solve the transport equation (3) in Matlab consist of the following steps [18]:

1. Integration of the transport equation over all the (finite) control volumes of the domain as well as over time.
2. Discretization – conversion of the resulting integral equations into a system of algebraic equations.
3. Solution of the algebraic equations.

The resulting statements from the control volume integration express the exact conservation of relevant properties for each finite size cell [18].

2.4 Solids mixing - Mathematical model

In the case of solids mixing with the application of the transport equation the general property is represented by the mass fraction Y_k of species k , where $\rho Y_k = C$ and coefficient Γ , is set to ρD . Equation (3) with substitutions gives:

$$\frac{\partial(C)}{\partial t} + \text{div}(C\mathbf{u}) = \text{div}(D \text{ grad } C) + S_\phi \quad (4)$$

The first step for solving equation (3) through finite volume method is the integration over a control volume

$$\int_{CV} \frac{\partial(C)}{\partial t} dV + \int_{CV} \text{div}(C\mathbf{u}) dV = \int_{CV} \text{div}(D \text{ grad } C) dV + \int_{CV} S_\phi dV \quad (5)$$

To solve the convection and dispersion (second and third) terms of the integrated equation (5), the Gauss's divergence theorem is implemented

$$\int_{CV} \text{div}(\mathbf{a}) dV = \int_A \mathbf{n} \cdot \mathbf{a} dA \quad (6)$$

The divergence theorem (6) states that the volume integral of the divergence over the region inside the surface is equal to the outward flux of a vector field through a closed surface in the direction normal to the surface. Rewriting equation (5) with the divergence theorem (6) yields:

$$\int_{CV} \frac{\partial(C)}{\partial t} dV + \int_A \mathbf{n} \cdot (C\mathbf{u}) dA = \int_A \mathbf{n} \cdot (D \text{ grad } C) dA + \int_{CV} S_\phi dV \quad (7)$$

Since the system that was analysed is time-dependent, integration with respect to time had to be implemented:

$$\begin{aligned} \int_{\Delta t} \frac{\partial}{\partial t} \left(\int_{CV} C dV \right) dt + \int_{\Delta t} \int_A \mathbf{n} \cdot (C\mathbf{u}) dA dt \\ = \int_{\Delta t} \int_A \mathbf{n} \cdot (D \text{ grad } C) dA dt + \int_{\Delta t} \int_{CV} S_\phi dV dt \end{aligned} \quad (8)$$

Equation (8) is the general integrated form of the transport equation (3) and will be the base from which the numerical algorithm will be built.

2.4.1.1 Mesh generation

In order to solve the transport equation (3) the system domain is divided into discrete control volumes according to step 2 above. This is carried out through dividing the geometry into quadratic discrete control volumes, i.e. cells. The accuracy of the simulations improves with a larger number of cells at the cost of computational time. The generated cells construct a so-called mesh.

At the centre of each cell a nodal point is located, see Figure 4. A general node is designated as node P and the adjacent nodes as W, E, S and N representing the western, eastern, southern and northern nodes. The faces of the control volumes are located between adjacent nodes and are the control volume boundaries. These are denoted as w, e, s and n representing the western, eastern, southern and northern boundaries.

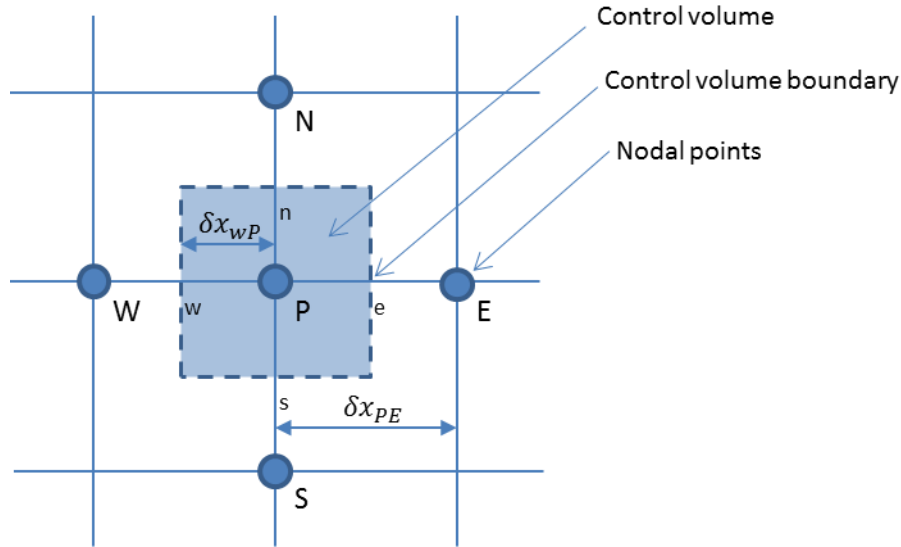


Figure 4 Mesh construction and notations.

2.4.2 2D Unsteady Dispersion

Dispersion of bed material and fuel corresponds to the random movement of material through emulsion drift sinking, bubble wake lifting and bubble eruption scattering.

In the dispersion model the second term of the transport equation (4) which concerns the convective transport is removed. A 2D approach is implemented for the solids and fuel mixing which gives the concentration C in $[\text{kg}/\text{m}^2]$ and $[\text{particles}/\text{m}^2]$ respectively. Equation (8) rewritten including above parameters over time from t to $t + \Delta t$ gives

$$\int_t^{t+\Delta t} \frac{\partial}{\partial t} \left(\int_{CV} C dV \right) dt = \int_t^{t+\Delta t} \int_A (D \text{ grad } C) dA dt + \int_t^{t+\Delta t} \int_{CV} S_\phi dV dt \quad (9)$$

The discretization of equation (9) converts the integral equations into a system of algebraic equations

$$\begin{aligned}
& (C_P^{t+\Delta t} - C_P^t)\Delta V \\
&= \int_t^{t+\Delta t} \left[\left(DA \frac{\partial C}{\partial x} \right)_e - \left(DA \frac{\partial C}{\partial x} \right)_w + \left(DA \frac{\partial C}{\partial y} \right)_n \right. \\
&\quad \left. - \left(DA \frac{\partial C}{\partial y} \right)_s \right] dt + \bar{S}\Delta V\Delta t
\end{aligned} \tag{10}$$

Where $\bar{S}$ is the average source term comprising of an independent term S_u and a dependent term S_p accordingly: $\bar{S}\Delta V = S_u + S_p C_p$ [18].

The superscripts t and $t + \Delta t$ indicate at which iteration time step the values are evaluated, henceforth the superscripts ' o ' will represent the values at time t and values at time $t + \Delta t$ will have no superscripts. A central differencing scheme is implemented where the changes of concentration between nodes are approximated to be linear, e.g.

$$\left(\frac{\partial C}{\partial x} \right)_e = \frac{C_E - C_P}{\delta x_{EP}} \tag{11}$$

An explicit scheme is used to evaluate the concentrations C where that values for the concentration comes from the previous time step t superscript ' o '. Integration of equation (10) with the assumption above becomes:

$$\begin{aligned}
(C_P - C_P^0)\Delta V = & \left[D_e A_e \frac{C_E^0 - C_P^0}{\delta x_{EP}} - D_w A_w \frac{C_P^0 - C_W^0}{\delta x_{PW}} + D_n A_n \frac{C_N^0 - C_P^0}{\delta x_{NP}} \right. \\
& \left. - D_s A_s \frac{C_P^0 - C_S^0}{\delta x_{PS}} \right] \Delta t + \bar{S}\Delta V\Delta t
\end{aligned} \tag{12}$$

Where the boundary face areas A , volume ΔV and distances are defined as follows

$$\begin{cases} A_e = A_w = \delta x_{NP} = \delta x_{PS} = \Delta y \\ A_n = A_s = \delta x_{EP} = \delta x_{PW} = \Delta x \\ \Delta V = \Delta x \Delta y \\ \frac{(C_P - C_P^0)}{\Delta t} = \frac{\partial C}{\partial t} \end{cases} \tag{13}$$

Combining equations (13) and (12) gives

$$\begin{aligned}
& \frac{\partial C}{\partial t} \\
&= \frac{\left[D_e \Delta y \frac{C_E^0 - C_P^0}{\Delta x} - D_w \Delta y \frac{C_P^0 - C_W^0}{\Delta x} + D_n \Delta x \frac{C_N^0 - C_P^0}{\Delta y} - D_s \Delta x \frac{C_P^0 - C_S^0}{\Delta y} \right]}{\Delta x \Delta y} \\
&+ \bar{S}
\end{aligned} \tag{14}$$

Equation (14) is the end product of the discretization of the transport equation (4) which expresses the change of concentration over time, $\frac{\partial C}{\partial t}$, for each discrete control volume or cell

The third step is to solve the numerical algorithms by an iterative method. Equation (14) is solved through the built in Matlab solver, ode45, based on an explicit Runge-Kutta formula Dormand-Prince pair [19].

2.4.2.1 Boundary and initial conditions

To solve the discretised transport equation (14) the boundary and initial conditions need to be defined to correctly interpret the real geometry, restrictions and different run modes.

Wall cells in the mathematical model have a zero-mass flux boundary condition forcing all wall cells in the mathematical model have the same concentration C as its adjacent cell in the bed.

The initial conditions of the model concern the concentration C at the start of the simulation corresponding to time zero. Thus, if the mathematical model is to compute a simulation of a batch of tracer material that is released into the real chamber and spreads across it, a non-zero concentration C at the corresponding location is implemented.

2.4.3 2D Unsteady Convection

Transport of solids and fuel through convection corresponds to the bulk motion of a medium. An analogy describing convection is the ink drop in a river that moves with the bulk motion of water downstream.

The physical attribute of the convection is the appearance of bed material movement and velocity field that occurs when bed material is pressed into the chamber from the loop seal and flows towards the exit by means of convection.

To solve the bed material and fuel mixing with convection included the entirety of the transport equation (4) needs to be solved. The generation and notations for the mesh is identical as for diffusion. Since the discretisation of the diffusive transport term and source term is identical as in the previous section this will not be written out again. Equation (8) rewritten with above mentioned parameters over time from t to $t + \Delta t$ applied gives:

$$\int_t^{t+\Delta t} \frac{\partial}{\partial t} \left(\int_{CV} C dV \right) dt + \int_t^{t+\Delta t} \int_A \mathbf{u} C dA dt = Diffusion + Source \quad (15)$$

Integration of the convective term and accumulative term (which is also identical as in the previous section) is performed.

$$\begin{aligned}
& (C_P^{t+\Delta t} - C_P^t)\Delta V \\
& + \int_t^{t+\Delta t} [(AuC)_e - (AuC)_w + (AuC)_n - (AuC)_s]dt \quad (16) \\
& = Diffusion + Source
\end{aligned}$$

A central differencing scheme is implemented where the changes of concentration between nodes are approximated to be linear, e.g.

$$C_e = \frac{C_p + C_E}{2} \quad (17)$$

an explicit scheme is used to evaluate the concentrations C where that values for the concentration comes from the previous time step t superscript 'o'. Integration of equation (10) with the assumption above becomes:

$$\begin{aligned}
& (C_P - C_P^o)\Delta V + \left[\frac{A_e \mathbf{u}_e (C_P^o + C_E^o)}{2} - \frac{A_w \mathbf{u}_w (C_P^o + C_W^o)}{2} \right. \\
& \quad \left. + \frac{A_n \mathbf{u}_n (C_P^o + C_N^o)}{2} - \frac{A_s \mathbf{u}_s (C_P^o + C_S^o)}{2} \right] \Delta t \quad (18) \\
& = Diffusion + Source
\end{aligned}$$

The inclusion of the diffusion and source terms from equation (14) and the definitions from equations (13) into (18) and divide by $\Delta V \Delta t$ you get:

$$\begin{aligned}
& \frac{\partial C}{\partial t} \\
& = - \frac{[A_e \mathbf{u}_e (C_P^o + C_E^o) - A_w \mathbf{u}_w (C_P^o + C_W^o) + A_n \mathbf{u}_n (C_P^o + C_N^o) - A_s \mathbf{u}_s (C_P^o + C_S^o)]}{2\Delta x \Delta y} \quad (19) \\
& + \frac{\left[D_e \Delta y \frac{C_E^o - C_P^o}{\Delta x} - D_w \Delta y \frac{C_P^o - C_W^o}{\Delta x} + D_n \Delta x \frac{C_N^o - C_P^o}{\Delta y} - D_s \Delta x \frac{C_P^o - C_S^o}{\Delta y} \right]}{\Delta x \Delta y} + \bar{S}
\end{aligned}$$

Equation (19) is the end product of the discretization of the transport equation (4) for convective and diffusive mass transfer which expresses the change of concentration over time, $\frac{\partial C}{\partial t}$.

2.4.3.1 Velocity field, u

To solve the discretized transport equation (19) the velocity field inside chamber that comprise the system domain must be attained.

The mathematical computation of the velocity field is based on the dispersion model. A mesh is generated with cells and nodal points according to section 2.4.3 above. Boundary conditions for the system are identical as for the diffusion model discussed in section above with the addition for the outlet cells (node points), see Figure 5 below, of the system where the concentrations C_{outlet} are set to zero. The initial

condition for C_{bed} is zero for all cells in the system. A source term S_u of bed material is introduced in the corresponding location of the real bed material inlet.

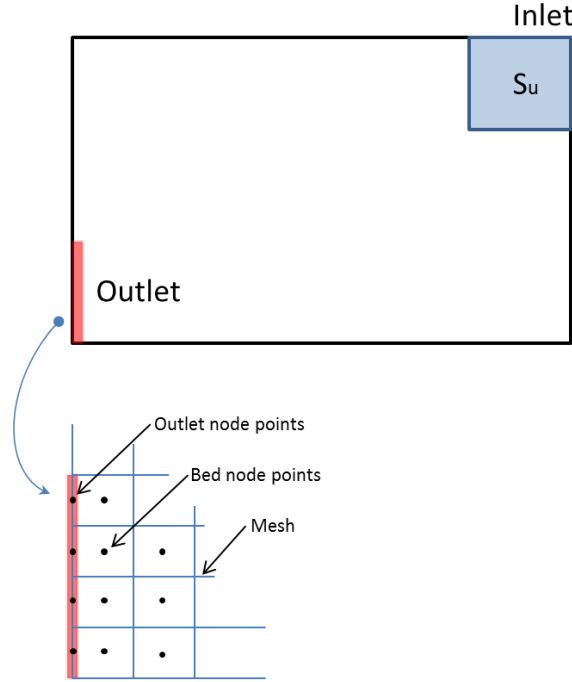


Figure 5 Illustration of chamber and mesh configuration for velocity field computations.

The iterative solver ode45 is employed to compute the concentration field from equation (14). A convergence argument is set to terminate the simulation when the difference between inlet and outlet flow reaches convergence, in this case $\varepsilon_{velocity} = 1e-6$.

$$\varepsilon_{velocity} * S_u < S_u - \sum_{\text{Outlet nodes}} DA \frac{C_{bed \text{ node}} - C_{outlet \text{ node}}}{\partial x} \quad (20)$$

The concentration field at time of termination is the basis for the computation of the velocity field. The velocity through each cell face is computed accordingly, a cell face in the x direction in this case:

$$\begin{aligned} \dot{m} &= \rho \mathbf{u} A = D \frac{\partial C}{\partial x} A \\ \rightarrow \mathbf{u} &= \frac{D \partial C}{\rho \partial x} \end{aligned} \quad (21)$$

Since it is a 2D model the concentration C unit is $[\text{kg}/\text{m}^2]$ and the density is the area density $\rho_A = \rho_V dz$ which depends on the volume density and bed height. The velocity $\mathbf{u}$ unit is $[\text{m}/\text{s}]$. As the velocity field is computed for the bed material flow a velocity slip factor, α , is introduced to describe the influence of bed material flow on the fuel transport. The transport equation with the velocity slip factor, alpha, is presented in equation (22).

$$\frac{\partial(C)}{\partial t} + \alpha \cdot \text{div}(C\mathbf{u}) = \text{div}(D \text{ grad } C) + S_\phi \quad (22)$$

2.5 Fuel conversion

The fuel used in the Chalmers gasifier is biomass, mainly as pellets. The conversion of fuel in a gasifier is a complex chemical process which can be simplified into three main stages: drying, devolatilisation and char conversion.

The first two stages of the process are drying (heating and evaporation of the moisture) and devolatilisation. The process of devolatilisation involves breaking the particle down thermally into a large number of chemical species which further break down by the cracking of long hydrocarbons.

There are two types of approaches for modelling the drying and devolatilisation: thermally large or small particle. In a thermally small particle the temperature gradient through the particle is close to zero whereas for a large particle devolatilisation follows the drying front as the moisture leaves the particle. Thus, small particles have sequential drying and devolatilisation whereas large particles dry and devolatilise simultaneously.

The final step in the fuel conversion process is conversion of the porous char particle that remains after the drying and devolatilisation. The char is oxidized to CO/CO₂ in the presence of O₂, CO₂ or H₂O_(v). In indirect gasification, the C – CO₂ and C – H₂O_(v) reactions are utilised. This process is endothermic and the rate of reactions is closely governed by parameters of the surrounding, mainly temperature and species concentration.

In this thesis the complex process of fuel conversion is simplified into two main processes: simultaneous drying and devolatilisation of a thermally large particle and char gasification through C – H₂O_(v) reaction. Any further reactions or processes are overlooked in this work.

For drying and devolatilisation the following expressions are applied:

$$Bi_t = \frac{h_{c,eff} * r_o}{k_{c,eff}} \quad (23)$$

$$t_{dry/devol} = \frac{Y_m}{1 - Y_m} r_o^2 \frac{\rho_{df}}{k_{c,df}} \frac{H_{m,eff}}{T_\infty - T_{sat}} \frac{2 + Bi_t}{2 * (n + 1) * Bi_t} \quad (24)$$

where n=1 for infinite cylinders [20].

Above expressions assumes that devolatilisation starts instantaneous and is valid for fuel with a moisture content of 10% or higher which corresponds with biomass pellet.

For char the reaction rate for the C – H₂O_(v) reaction governs the conversion and following expression was used [21]

$$\frac{dm_c}{dt} = -\Omega A_p k_{rC} M_c C_{h_2o,\infty}^n \quad (25)$$

$$k_{rC} = 10^{(0.2*10^{-4}E-2)} e^{\left(\frac{E}{RT}\right)} \quad (26)$$

where the activation energy, E, was determined from gasifier operating conditions.

2.5.1 Allocation of drying/devolatilisation and char conversion

The time of drying/devolatilisation, $t_{dry/devol}$, is computed through equation (24) while the time for char conversion, t_{char} , is derived from equation (25). A simulation of the fuel concentrations over time for convection and/or diffusion is performed. The simulation is terminated after the simulation time has reached $t_f = t_{dry/devol} + t_{char}$.

An assumption that mass loss rate in all three fuel conversion processes under a constant temperature is constant was taken. The simulation of the concentration field is divided into time steps where a specific concentration field C_i is computed for each time step. For each time step a fraction $d_{dd_i} = \frac{\Delta t_i}{t_{dry/devol}} C_i$ of the fuel is dried/devolatilized. After time $t = t_{dry/devol}$ the char conversion starts where the fraction of char converted for each time step and cell is $d_{char_i} = \frac{\Delta t_i}{t_{char}} C_i$. All the fractions for each time step of drying/devolatilizing are added and a corresponding profile of the allocation is compiled. An analogous procedure is followed for char conversion.

2.6 Energy balance - Mathematical model

An energy balance simulation is implemented to calculate the temperature profile of the gasification chamber. The energy balance incorporates heat transfer by conduction, solids dispersion and convection and gas convection. It also incorporates the heat of fuel conversion processes (evaporation and heating of the fuel moisture, devolatilisation and char gasification).

2.6.1 Steady-state heat transfer through conduction

The heat transfer simulation uses the transport equation (3) at steady state. The same numerical algorithm steps as for steady state diffusion is used to solve the heat transfer equation.

The convective transport term is eliminated when the heat conduction is evaluated. Setting the property $\phi = T$ temperature and $\Gamma = k$ thermal conductivity of the bed material with the eliminations mention above give the applicable transport equation

$$0 = \text{div}(k \text{ grad } T) \quad (27)$$

Applying the divergence theorem from equation (6) and integrate over the discrete control volumes gives

$$0 = \left[\left(kA \frac{\partial T}{\partial x} \right)_e - \left(kA \frac{\partial T}{\partial x} \right)_w \right] + \left[\left(kA \frac{\partial T}{\partial y} \right)_n - \left(kA \frac{\partial T}{\partial y} \right)_s \right] + \bar{S} \Delta V \quad (28)$$

A central differencing scheme is implemented and where $\bar{S}$ is the average source term comprising of an independent term S_u and a dependent term S_p accordingly $\bar{S} \Delta V = S_u + S_p T_p$.

$$0 = \left[k_e A_e \frac{(T_E - T_P)}{\partial x_{PE}} - k_w A_w \frac{(T_P - T_W)}{\partial x_{WP}} + k_n A_n \frac{(T_N - T_P)}{\partial x_{PN}} - k_s A_s \frac{(T_P - T_S)}{\partial x_{SP}} \right] + S_u + S_p T_p \quad (29)$$

Rearranging equation (29) gives the general equation

$$a_P T_P = a_E T_E + a_W T_W + a_N T_N + a_S T_S + S_u \quad (30)$$

Where a_x is as per Table 2, with the notation of a uniform bed material conductivity k in all directions

Table 2

a_E	a_W	a_N	a_S	a_P
$\frac{kA_e}{\partial x_{PE}}$	$\frac{kA_w}{\partial x_{WP}}$	$\frac{kA_n}{\partial x_{PN}}$	$\frac{kA_s}{\partial x_{SP}}$	$a_E + a_W + a_N + a_S - S_p$

2.6.1.1 Boundary and initial conditions

The walls of the gasification chamber are assumed to be insulated as per Neumanns thermal boundary condition, meaning no heat flux goes through them and that they have the same temperature as the closest bed node point giving it a zero gradients to the wall cell. For example the bed nodes closest to the eastern wall as Node 2 in Figure 6 below have the parameters according to Table 3. The cells at the walls as Node 3 and Node 1 have all parameters a_x from Table 2 equal to zero to force a wall that perfectly insulated.

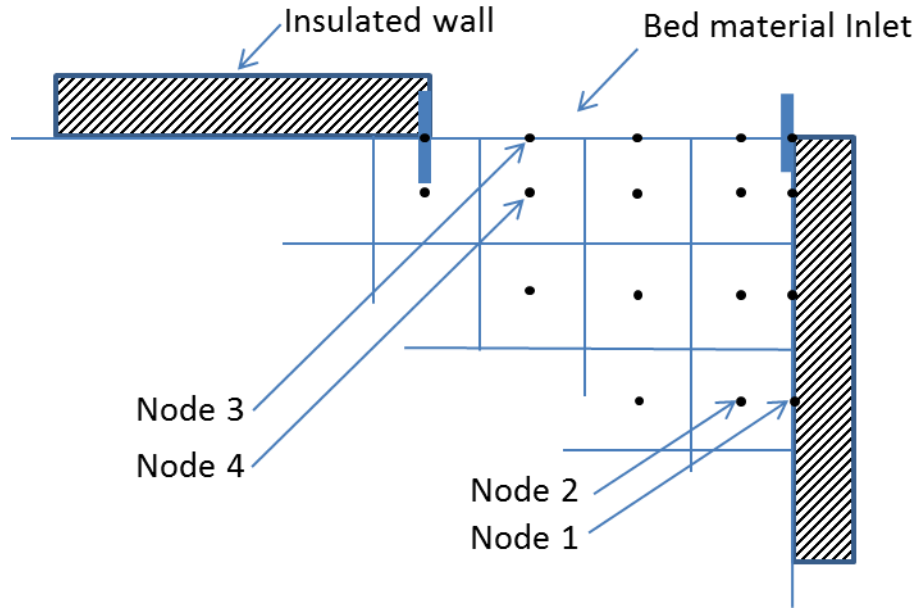


Figure 6 Illustration of mesh configuration for inlet and insulated walls.

Table 3

a_E	a_W	a_N	a_S	a_P	S_P	S_u
0	$\frac{kA_W}{\partial x_{WP}}$	$\frac{kA_n}{\partial x_{PN}}$	$\frac{kA_s}{\partial x_{SP}}$	$a_E + a_W + a_N + a_S - S_P$	0	0

The only boundary conditions that differ from the form of Table 3 are the wall cells that represent the bed material inlet as Node 4 in the figure above. Node 3 is inert at the wall and has all parameters a_x from Table 2 equal to zero. The bed material that enters the chamber has the constant temperature of $T_{bed\ inlet}$ which is represented in the heat conductivity model as a Dirichlet thermal boundary condition with a constant wall temperature of $T_{bed\ inlet}$. The example of an inlet at the north wall as Node 4 gives you with the integration of equation (27) with the above mentioned parameters

$$0 = \left[kA_e \frac{(T_E - T_P)}{\partial x_{PE}} - kA_w \frac{(T_P - T_W)}{\partial x_{WP}} + kA_n \frac{(T_{bed\ inlet} - T_P)}{\partial x_{PN}/2} - kA_s \frac{(T_P - T_S)}{\partial x_{SP}} \right] \quad (31)$$

Rearranged

$$\begin{aligned}
0 &= \left(\frac{kA_e}{\partial x_{PE}} + \frac{kA_w}{\partial x_{WP}} + \frac{kA_s}{\partial x_{SP}} + \frac{kA_n}{\partial x_{PN}/2} \right) T_P \\
&= \left(\frac{kA_e}{\partial x_{PE}} \right) T_E + \left(\frac{kA_w}{\partial x_{WP}} \right) T_W + \left(\frac{kA_s}{\partial x_{SP}} \right) T_S \\
&\quad + \left(\frac{kA_n}{\partial x_{PN}/2} \right) T_{bed \text{ inlet}}
\end{aligned} \tag{32}$$

At the notation for equation (30) parameters as per Table 4 are used.

Table 4

a_E	a_W	a_N	a_S	a_P	S_P	S_u
$\frac{kA_e}{\partial x_{PE}}$	$\frac{kA_w}{\partial x_{WP}}$	0	$\frac{kA_s}{\partial x_{SP}}$	$a_E + a_W + a_N + a_S - S_P$	$-\frac{kA_n}{\partial x_{PN}/2}$	$\left(\frac{kA_n}{\partial x_{PN}/2} \right) T_{bed \text{ inlet}}$

2.6.2 Steady state heat transfer through solids dispersion

The implementation of heat transfer through dispersion of bed material complies with the solution algorithm of conductive heat transfer explained above. Since there are no new source terms or boundary conditions the only changes concerns the pre temperature parameter a_x in equation (30). The employment of an artificial dispersion heat transfer parameter, k' , where for example $a_e = \frac{(k+k')A_e}{\partial x_{PE}}$ is realized accordingly

$$\begin{aligned}
\text{Heat transfer} &= \dot{m}c_p\Delta T = Ddy\frac{\rho_A}{dx}c_p\Delta T = k'\frac{dydz}{dx}\Delta T \\
\rightarrow k' &= D\frac{\rho_A}{dz}c_p \text{ [W m}^{-1}\text{K}^{-1}]
\end{aligned} \tag{33}$$

This gives the following parameters for the bed node points

Table 5

a_E	a_W	a_N	a_S	a_P	S_P	S_u
$\frac{(k+k'_e)A_e}{\partial x_{PE}}$	$\frac{(k+k'_w)A_w}{\partial x_{WP}}$	$\frac{(k+k'_n)A_n}{\partial x_{PN}}$	$\frac{(k+k'_s)A_s}{\partial x_{SP}}$	$a_E + a_W + a_N + a_S - S_P$	0	0

Note that the boundary (wall) and inlet (source) cells get the addition of k' at the a_x parameters also note that k' has directionality dependence because of the dispersion coefficient D while thermal conductivity k has not. The ratio between the bed material heat conductivity k and the artificial dispersion heat transfer coefficient k' show which heat transfer term is dominating. The ration in the simulation were larger than 1/1000 indicating a dispersion dominated heat transfer.

2.6.3 Steady state heat transfer through solids convection

Heat transfer through the convective movement of hot bed material is implemented into the above mentioned solution algorithm, applying the representation of conduction and diffusion into a single variable, $De = \frac{(k+k'_e)A_e}{\partial x_{PE}}$ [W/K].

The contribution of the convection is computed through for example $Fe = \rho_A u_e d y c_p$ [W/K] which is implemented with an upwind differentiating scheme. The upwind scheme for the bed cells is given in Table 6 [18].

Table 6

a_E	a_W	
$\max((De - Fe), De)$	$\max((Dw + Fw), Dw)$	
a_N	a_S	a_P
$\max((Dn - Fn), Dn)$	$\max((Ds + Fs), Ds)$	$a_E + a_W + a_N + a_S + (Fe - Fw) + (Fn - Fs) - S_P$

While the inlet (source) cells get the appendage to equation (36) accordingly

$$\begin{cases} S_u = 2DnT_{bed\ inlet} + \frac{Su_{bed\ tot}}{N_{inlet\ nodes}} c_{p,sand} T_{bed\ inlet} \\ Dn = \text{conduction and diffusion contribution} \\ Su_{bed\ tot} = \text{Total mass of bed material source} \\ N_{inlet\ nodes} = \text{Number of inlet nodes} \end{cases} \quad (34)$$

By combining the complimented parameters of equation (34) and Table 6 into the equation system (35), the Gauss-Seidel iterative solving method can be applied. With this, the temperature field of the gasification chamber is obtained.

2.6.4 Solver by a Gauss-Seidel iterative method

Since the model is a steady state model the general equation (30) can be applied at all nodes and expressed in a linear equation system according to $Ax = b$, with zero elements in A except for the a_x parameters:

[illegible]

Where

$$\left\{ \begin{array}{l} b = S_u - E_{dry} - E_{devol} - E_{char} - E_{heat\ steam} - E_{heat\ moist} - E_{heat\ vol} \\ S_u = \text{Bed material heat source} \\ E_{dry} = \text{Energy requirement to dry fuel} \\ E_{devol} = \text{" to devolatalize fuel} \\ E_{char} = \text{" to convert char} \\ E_{heat\ steam} = \text{" to heat up the steam} \\ E_{heat\ moist} = \text{" to heat up the moisture from the fuel} \\ E_{heat\ vol} = \text{" to heat up volatile from the fuel} \end{array} \right. \quad (36)$$

To solve the equation system (35) a Gauss-Seidel iterative method is implemented according to [18]

$$T_i^{(k)} = \sum_{j=1}^{i-1} \left(\frac{-A_{ij}}{A_{ii}} \right) T_j^{(k)} + \sum_{j=i+1}^n \left(\frac{-A_{ij}}{A_{ii}} \right) T_j^{(k-1)} + \frac{b_i}{A_{ii}} \quad (37)$$

Where the superscript (k) indicates the temperature from the new iteration while ($k - 1$) is the temperatures from the iteration that came before. With a convergence argument that monitors the largest temperature difference for a cell between iterations reading:

$$\varepsilon_{gauss} < abs(\max(T^{(k)} - T^{(k-1)})).$$

A value of $\varepsilon_{gauss} = 1e - 5$ is taken.

2.7 Impact of internals

The main objective in this thesis is to increase fuel conversion by means of controlling the bed temperature gradient, improving the gas solid contact and increase residence time in a BFB reactor, therefore the use of horizontal tubes and a partitioning wall was studied.

Internals have been widely applied in order to improve the overall performance by changing fluidisation behaviour, enhancing heat and mass transfer or improving gas-solid contact. A number of internal structures have been studied such as wire meshes, perforated plate, Louver plate, tubes and different packing's [22]. Most of these internals aim at improving the performance in turbulent, fast fluidised beds and circulating fluidised beds whereas tubes mainly have been used as heat exchangers in BFB reactors.

Rising bubbles split when hitting horizontal tubes, reaching diameters of a similar size as the tube spacing [22]. In BFB's there is a direct link between bubble size, frequency and distribution to mixing and gas-solid contact. A bed with uniform small bubbles will have a higher conversion rate than beds with fewer and large bubbles [23].

The use of partitioning walls can also be applied as a mean to control bed flow pattern and to create sections inside the reactor. Little research has been done on the influence of this in BFB reactors. Vertical baffles above the surface have been shown to reduce the lateral dispersion up to 65% [4]. However, such a baffle would only influence the magnitude of the mixing of particles on and above the bed surface rather than redirecting it.

2.8 Experimental setup

2.8.1 Scaling of fluidised beds

In order to simulate the behaviour in a commercial fluidized bed a down-scaled model operating at ambient temperature has been used. However, the dynamics of a fluidized bed vary in relation to operational conditions, size, fluidization media and bed material. The results obtained in small pilot plants or small experimental models are seldom comparable to those from large commercial units. Therefore a set of scaling relations was established by Glicksman et al. [23] derived from the governing equations of motion for individual particles and for a continuum equation (38) to (41):

Conservation of mass the fluid

$$\frac{\partial \epsilon}{\partial t} + \text{div}(\epsilon \mathbf{u}) = 0 \quad (38)$$

and for the solid phase

$$\frac{\partial}{\partial t}(1 - \epsilon) + \text{div}[(1 - \epsilon)\mathbf{v}] = 0 \quad (39)$$

Equation of motion for the fluid

$$\rho_f \epsilon \left[\frac{\partial \mathbf{u}}{\partial t} + \mathbf{u} \cdot \nabla \mathbf{u} \right] + \rho_f g \epsilon - \nabla E_f + \mathbf{F} = 0 \quad (40)$$

and for solid phase

$$\rho_s (1 - \epsilon) \left[\frac{\partial \mathbf{v}}{\partial t} + \mathbf{v} \cdot \nabla \mathbf{v} \right] + \rho_s g (1 - \epsilon) - \nabla E_p - \mathbf{F} = 0 \quad (41)$$

Complementing the equations above with a set of boundary conditions, a set of dimensionless parameters was found. The parameters are used in this thesis and presented in Table 7.

Table 7 Glicksman's dimensionless scaling relations

Relation	Parameter	Dimensionless number and/or explanation
1	$\frac{u_m^2}{gL}$	Ratio of inertial to gravity forces (Froude number)
2	$\frac{\rho_s}{\rho_f}$	Ratio of particle to fluid inertial forces
3	$\frac{\rho_s u_m d_p}{\mu}$	Ratio of inertia to viscous particle forces (Particle Reynolds number,)
4	$\frac{\rho_f u_m L}{\mu}$	Ratio of inertia to viscous fluid forces (Fluid Reynolds number)
5	$\frac{G_s}{\rho_s u_m}$	Ratio of solids flux to fluid velocity
6	Bed geometry	Length scaling
7	ϕ	Particle sphericity
8	Particle size distribution	Length scaling

With above scaling relations one can, at ambient temperature, model the behaviour of commercial fluidized beds at operational pressure and temperature.

2.8.2 Scale model setup

The experiments were conducted in an existing fluid dynamically down-scaled model of the Chalmers BFB gasifier [7]. Bronze powder ranging from 45-125 μm was used as bed material and air as fluidisation media. With scaling relation 2 from Table 7 and knowing that the bed material density in the gasifier is 2600kg/m³ the corresponding scale model bed material density was calculated where indices M denote the down-scaled model and L the large scale unit.

$$\rho_{s,M} = \rho_{f,M} \frac{\rho_{s,L}}{\rho_{f,L}} = 1.206 \frac{2600}{0.20} \approx 15700 \text{kg/m}^3 \quad (42)$$

However, the bronze powder used due to safety and cost reason has a density of 8800kg/m³ which results in an error of 43% compared with above result. The impact on the results due to this error is unknown but one can assume that lateral solid mixing is overestimated. Heavier particles are more likely to travel shorter distances than lighter particles when ejected by bubble eruptions.

The characteristic length for the scaled model was computed through a ratio of scaling relation 4 and square root of relation 1:

$$\frac{L_M}{L_L} = 0.167 \quad (43)$$

With this, the density of bronze and the length scale are used to determine the down-scaled operating conditions, bed material recirculation rate and down-scaled fuel characteristics.

2.8.2.1 Operating conditions

The variable operating parameters for the experiments were bed height and superficial gas velocity. A reference operating condition used in the Chalmers gasifier, Setting 1 in Table 8, was used as base for the experiments. Increased bed height, higher superficial gas velocity and obstacles were also tested. The up-scaled bed material flow rate was set to 4.72kg/s. See Table 8 for corresponding gasifier operating conditions used in the experiments.

Table 8 Gasifier main operating conditions

Setting	1	2	3	4
Superficial gas velocity, u_m (m/s)	0.16	0.16	0.23	0.23
Bed height (m)	0.34	0.42	0.34	0.42

2.8.2.2 Scaling of operating condition

Using scaling relations 1, Froude number, and values for the superficial gas velocity from Table 8 the scale model superficial gas velocity was found. Calculations for setting 1 are presented below (see Table 9 for settings 2-4).

$$u_M = \sqrt{L_M \frac{u_L^2}{L_L}} = \sqrt{1 \frac{0.158^2}{6}} = 0.064 \text{ m/s}$$

Table 9 Scale model main operating conditions

Setting	1	2	3	4
Superficial gas velocity, u_m (m/s)	0.064	0.064	0.098	0.098
Bed height (mm)	57	70	57	70

One single distributor plate is used throughout the experiments and was determined by the bed height and pressure ratio of 0.33 for setting 1 in Table 8. The bed pressure drop for a 57 mm bed was determined to 3000 Pa, yielding a distributor plate pressure drop of 990 Pa.

In order to calculate the Reynolds number the characteristic length was calculated

$$d_h = \frac{2 \cdot L \cdot W}{L + W} = \frac{2 \cdot 0.30 \cdot 0.13}{0.30 + 0.13} = 0.18 \text{ m}$$

which gives following Reynolds number

$$Re = \frac{d_h u_M \rho_{f,M}}{\mu_{f,M}} = \frac{0.18 \cdot 0.064 \cdot 1.206}{1.8 \cdot 10^{-5}} = 783$$

According to [24] the corresponding orifice coefficient, $C_{d,or}$, for above Reynolds number is 0.66. Using equation (44) from [24] the orifice gas velocity is determined.

$$u_{or} = C_{d,or} \left(\frac{2\Delta p_d}{\rho_f} \right)^{1/2} = 0.66 \left(\frac{2 \cdot 990}{1.206} \right)^{1/2} \approx 27 \text{ m/s} \quad (44)$$

And equation (45) from [24] determined the orifice diameter in mm

$$d_{or} = 1000 \sqrt{\frac{4u_m}{\pi u_{or} N_{or}}} = 1000 \sqrt{\frac{4 \cdot 0.064}{\pi \cdot 27 \cdot 1500}} \approx 1.4 \quad (45)$$

where N_{or} is the number of nozzles per unit area

For settings 2-4 where the orifice diameter was pre-set the pressure ratio was calculated instead. Above calculations was made for all operating conditions from Table 8 and are presented in Table 9

2.8.2.3 Scaling of fuel particles

Fuel particles were scaled down in a similar way as for bed material. A constant density ratio fuel/bed material was kept, i.e. the error in density of 43% from the bed material scaling was carried through. As the wood particle density changes in the gasifier an average density between moist wood pellet and char was used, 1175 kg/m³ and 233 kg/m³ respectively. The standard wood pellet has a diameter of 8 mm and an average length of 16mm according to [25]. Scaling relation 2 from Table 7 and accounting for the density error:

$$\rho_{s,M} = \rho_{f,M} \frac{\rho_{s,L}}{\rho_{f,L}} \cdot (1 - 0.43) = 1.206 \frac{\left(\frac{1175 + 233}{2} \right)}{0.20} \cdot (1 - 0.43) \approx 2400 \text{ kg/m}^3$$

and down-scaled diameter and length respectively

$$d_{p,M} = 0.167 \cdot 8 = 1.3 \text{ mm}$$

$$L_{p,M} = 0.167 \cdot 16 = 2.6 \text{ mm}$$

The obtained density is very similar to that of aluminium, 2700kg/m³. Aluminium thread with a diameter of 1.5 mm was selected as representative of the fuel particle and cut into 3 mm lengths. For the direct method experiments the particles was also painted with UV-activated paint.

2.8.2.4 Scaling of bed material flow rate

Due to the use of bronze powder as bed material and the inherent error of density (see Section 2.8.2) a different method for calculating mass flow rate was implemented. The method is based on down-scaling the solid residence time instead of the mass flow. The methods are derived from Glicksman's dimensionless parameters presented above.

The down-scaled bed material mass flow rate was determined by equation (46) and (47) using a real bed material mass flow rate of 4.72kg/s. The resulting flow rates for the two methods differ drastically as shown below.

Mass flow based method:

$$\dot{m}_M = \frac{\dot{m}_L * \frac{\rho_{f,M}}{\rho_{f,L}} * L_M^3}{L_L * \sqrt{L_M}} = \frac{4.72 * \frac{1.206}{0.20} * 0.167^3}{1 * \sqrt{0.167}} = 0.322 \text{ kg/s} \quad (46)$$

Residence time based method:

$$\dot{m}_M = \frac{m_M}{\frac{\dot{m}_L * \sqrt{L_M}}{\dot{m}_L}} = \frac{11.9}{\frac{760}{4.72} * \sqrt{0.167}} = 0.181 \text{ kg/s} \quad (47)$$

where m is the mass of bed material inside reactor.

The residence time based bed material mass flow was applied since the inherent density error is avoided.

2.8.3 Redesign of bed material feed system

The down-scaled model was a stationary bubbling bed with possibility to operate with bed material flowing through it. The existing feeding system for the bed material was not able to recirculate material efficiently due over dimensioning (it used a rotary valve to control the flow which was ten times larger than required). Thus, the existing bed material feeding system was redesigned for better operation and easier control.

The redesign options for the feeding system were reduced to three types: a smaller rotary valve, pneumatic conveying through a riser and a flexible screw conveyer. The options were evaluated in terms of cost, operability, powder and operational flexibility, rebuilding needed and time constraints. A summary of the options pros and cons is given in Table 10.

Table 10 Pros. and Cons. for the rebuild option of the bed material feeding system

Technique	Pros.	Cons.
Rotary Valve	Cheap, flexible open design, good flow control, minor rebuild needed, gentle in terms of particle wear, short delivery time	No recirculation, pulsating flow
Pneumatic riser	No electrical components needed, relatively cheap, recirculation, gentle in terms of particle wear, short delivery time	No easy way to verify flow rate, major rebuild needed, limiting test rig flexibility, limiting experiment methods as rig will be completely sealed
Flexible screw	Recirculating, wide and excellent flow control. Open design, flexible in bed material, minor rebuild needed	Relatively expensive, particle wear unknown, long delivery time

The screw conveyer was selected as the best option and installed, see Figure 8. The screw is capable to control a range of 50%-130% of the nominal bed material flow. The screw conveyer was tested with bronze powder and the corresponding flow rate function was found Figure 7.

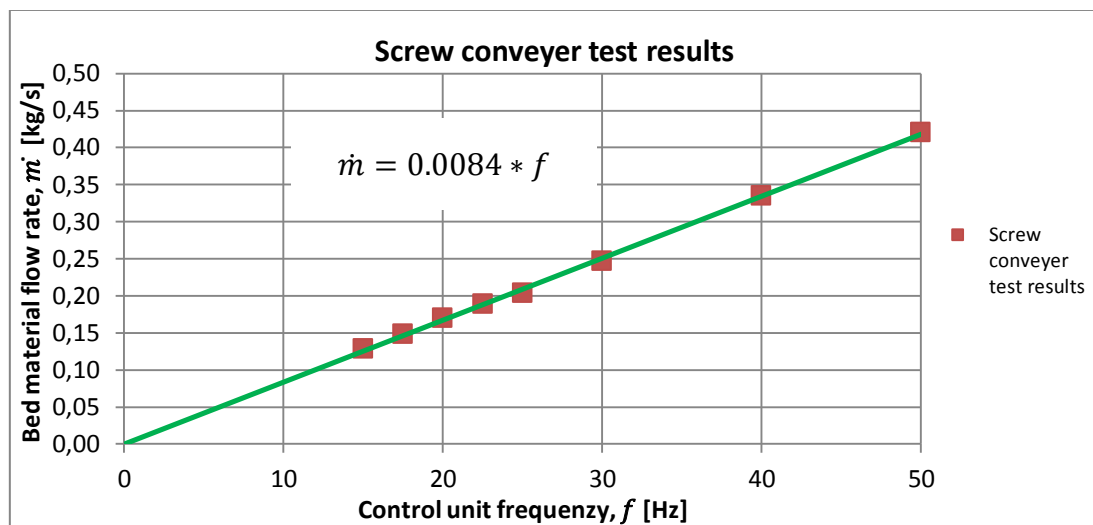


Figure 7 Screw conveyor test results for bronze powder and plotted flow rate function.



Figure 8 Screw conveyor installed and mounted to scale model

2.8.4 Experimental work

2.8.4.1 Bed material dispersion

The method used for determination of the dispersion coefficient for the bed material was an indirect method through concentration measurements of a tracer material in the scaled model [7]. Iron powder with same particle size as the bed material was used as tracer due to its similar density to bronze, 7800kg/m^3 and 8800kg/m^3 respectively. A batch of 80-100g of iron powder was inserted and samples were collected at the outlet over time.

The iron was then separated from each test sample through magnetic separation and the iron concentration was calculated from weighting.

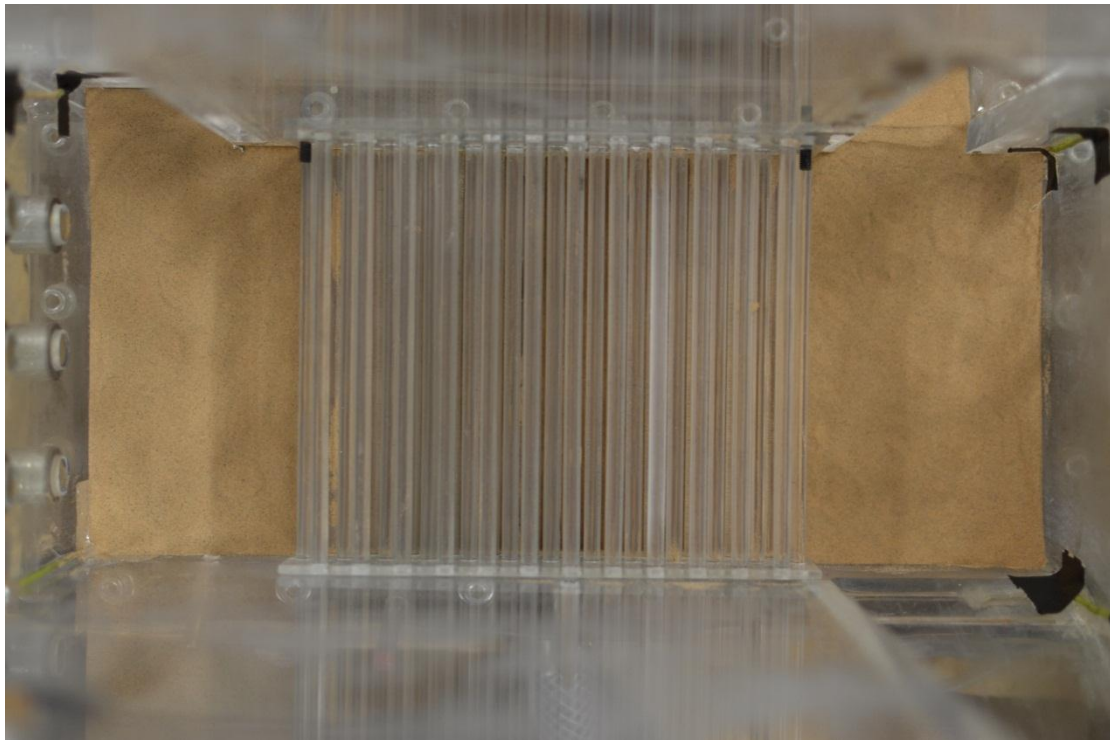


Figure 9 Tube bank covering 50% of the bed

Experiments with a tube bank covering 50%, see Figure 9, of the total area were also conducted for setting 1 and 3. The tube bank was $15 \times 13 \times 6\text{cm}$ (LxWxH), triangular pitch 12.5mm and rod diameter of 5mm. Samples were collected during 10 seconds for 5 minutes for the freely bubbling bed and for 6.5 minutes with tubes. A sample was collected before the experiments and acts as the background concentration value. The experiment results were then matched with the mathematical model to compute the dispersion coefficient, D .

2.8.4.2 Fuel dispersion

The fuel dispersion coefficient was determined through two methods, digital image analysis and concentration measurement of tracer particles in the scaled model, a direct and indirect method respectively.

In the direct method three cylindrical aluminium particles were painted in red, blue and green with UV-activated paint, see Figure 10.



Figure 10 Aluminium fuel particles painted with UV activated paint

A UV light and camera was mounted above the bed and the scale model was covered with a black cloth. The movement of the particles inside the model was recorded using a Nikon D90 Digital camera and a resolution from 640*480 to 1280x720 pixels, 24 frames per second. The movie length was 1 to 2.5 hours per experiment.

The movie was numerically analysed and the dispersion coefficient, D , was determined using Einstein's [17] dispersion model according to equation (2). The characteristic length, L_x , was set to half the nozzle spacing, A , for free bed, see Figure 3, and the tube diameter for tube bank.

Any movement near a wall was discarded as the bounce effect does not count as a clean movement. A statistic measure, interquartile range, was implemented to remove any extreme values from the data.

In the indirect method the bed material was recirculating through the scale model. A batch containing 1000 fuel particles, same as used in the direct method, was inserted in the fuel inlet of the scale model and collected at the outlet during 6-25 s intervals depending on the superficial gas velocity and bed setup. The results from the indirect method was matched with a mathematical dispersion model were the dispersion coefficient, D , and the velocity slip factor, α , was determined.

Both methods were performed for all four settings as per Table 9, with and without tube bank. The method and results from the direct and indirect methods were evaluated in terms of reliability, measurability and applicability.

3 Results

3.1 Dispersion of bed material

The experiments to find the dispersion coefficient for bed material were conducted in accordance with the experimental setup. The lateral solid dispersion coefficient is computed by correlating the experimental and mathematical modelling results applying the least squares method.

The bed material dispersion experiments were conducted using two freely bubbling bed heights of 57 and 70mm and a bed that contained a tube bank. The experiments for three different bed configurations were conducted with a high and a low superficial gas velocity, u_m , of 0.064 and 0.098 m/s. The e dispersion coefficient values found are shown in Table 11.

Table 11 Lateral dispersion coefficient for bed material, m^2/s .

$D_{bed}, [m^2/s]$	$u_m = 0.064m/s$	$u_m = 0.098m/s$
Freely bubbling bed: height 57mm	$2.62 \cdot 10^{-4}$	$2.78 \cdot 10^{-4}$
Freely bubbling bed: height 70mm	$1.65 \cdot 10^{-4}$	$3.25 \cdot 10^{-4}$
Tube banks: height 57mm	$0.73 \cdot 10^{-4}$	$0.91 \cdot 10^{-4}$

Figure 11 shows the experimental results and model simulation from the freely bubbling 57mm bed.

During the low gas velocity experiment a particle separation was noted where some of the iron powder was accumulated at the walls of the model. Due to the low gas velocity the bed material transport through the outlet was low and the initial bed height needed to be above the outlet bricks to be able to collect samples. The loss of bed material from the bed was 8% of initial amount.

No accumulation of iron powder was noted with the high gas velocity. However, more bed material was lost, 15% of initial amount. The loss of bed material was compensated by having a higher initial bed height so the average bed height was according to specification.

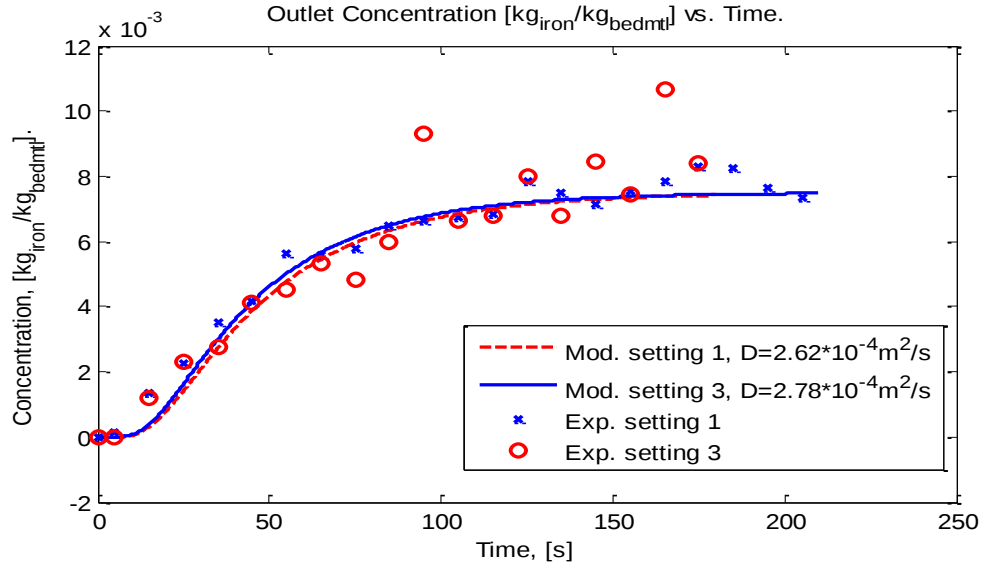


Figure 11 Experimental results and Model fitted to exp. results for setting 1 ($u_m=0.064\text{m/s}$) and 3 ($u_m=0.098\text{m/s}$) with 57mm bed height.

The experiments for the freely bubbling bed with bed height 70mm were conducted using the same two superficial gas velocities, u_m , of 0.064 and 0.098 m/s as used previously.

Figure 12 shows the experimental results and model simulation for the freely bubbling 70mm. During the low gas velocity experiment, the separation and accumulation of iron particles at the walls was noted at a higher extent than for the above-cited run with same velocity and a 57mm bed, yielding a dispersion coefficient 37% lower in comparison which indicated inadequate mixing. This result is contradictory compared to experimental trends in other runs (for the high gas velocity at 57 and 70mm bed, dispersion coefficient increased 17% with bed height). According to theory, as bed height increases the number of bubble paths decrease due to coalescence of bubbles. The surface eruptions of the larger bubbles eject solids further in the lateral direction resulting in higher dispersion coefficients.

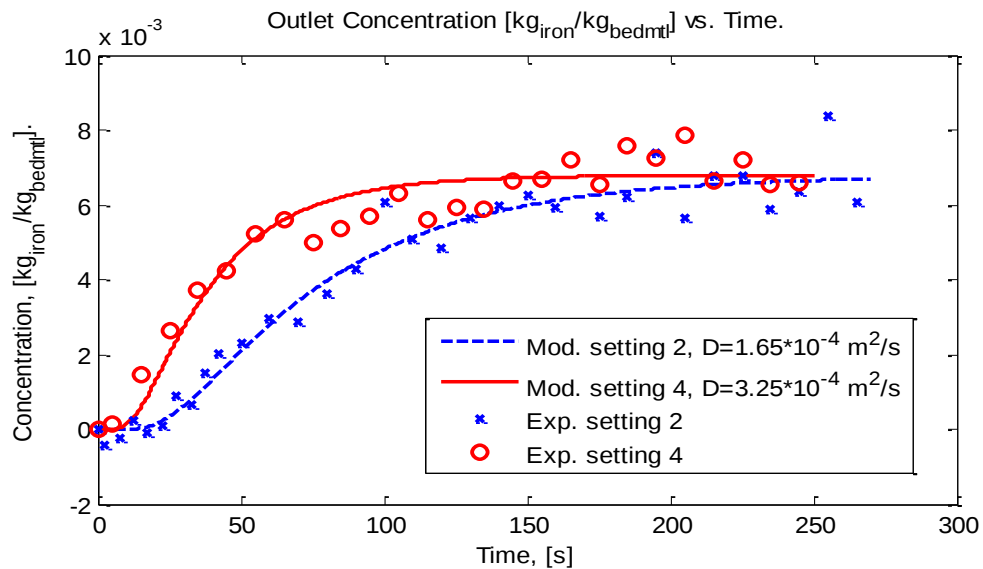


Figure 12 Experimental results and Model fitted to exp. results for setting 2 ($u_m=0.064\text{m/s}$) and 4 ($u_m=0.098\text{m/s}$) with 70mm bed height

Figure 13 shows the experimental results and model simulation from different gas velocities with tube bank installed covering 50% of the bed as per Figure 9. The corresponding bed material dispersion coefficient at tube banks was found according to Table 11 respectively.

Comparing the results of the tube bank experiment with the experiments without tubes it is clear that the tubes influence the bulk solid movement drastically, 67% decrease of dispersion coefficient between freely bubbling bed and tube banks at the high gas velocity, thus increasing the bed material residence time.

As the bubble growth is limited by the tube bank the maximum bubble size different bed heights are assumed equal [22]. With this assumption made the bed material dispersion coefficient at tubes is set equal for both 57 and 70mm bed height. The tube bank covers the entire height of the bed in both cases.

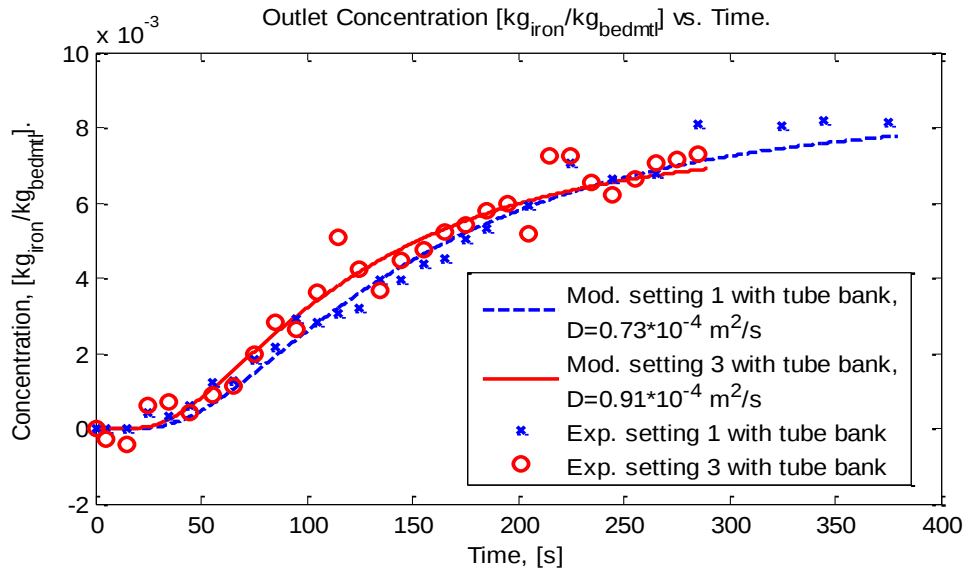


Figure 13 Experimental results and Model fitted to exp. results for setting 1 ($u_m=0.064\text{m/s}$) and 3 ($u_m=0.098\text{m/s}$) with 57mm bed height and tube bank installed.

3.2 Dispersion of fuel

3.2.1 Indirect method

The indirect method for fuel dispersion was conducted in a 70mm-high bed.

In addition to the fuel dispersion coefficient a velocity slip factor, α , was also obtained in this experiment. The fuel dispersion coefficient was found to influence the initial slope of the outlet concentration curve and α influence the tail of the curves seen in Figure 14. The experiment configurations and the corresponding results are displayed in Table 12, Table 13 and Table 14 below.

Table 12 Lateral dispersion coefficient Fuel, m^2/s .

$D_{fuel}, [m^2/s]$	$u_m = 0.064m/s$	$u_m = 0.098m/s$
Freely bubbling bed: height 70mm	$1.45 \cdot 10^{-4}$	$4.20 \cdot 10^{-4}$
Tube banks: height 70mm	$0.72 \cdot 10^{-4}$	$1.22 \cdot 10^{-4}$

Table 13 Velocity slip factor, α .

α	$u_m = 0.064m/s$	$u_m = 0.098m/s$
Freely bubbling bed: height 70mm	0.6	0.9
Tube banks: height 70mm	0.0	0.25

Table 14 Wall accumulation, %, of total number of particles.

Wall accumulation, %	$u_m = 0.064m/s$	$u_m = 0.098m/s$
Freely bubbling bed: height 70mm	35%	5%
Tube banks: height 70mm	70%	25%

During the experiment with the low gas velocity it was found that a large portion of the fuel particles accumulated at the walls close to the fuel inlet. This accumulation was reduced but not eliminated with higher superficial gas velocity as seen in Table 14.

As seen in Table 12, Table 13 and Table 14 all mixing parameters increase with the gas velocity where the fuel dispersion coefficient increased 190%. With the higher gas velocity the velocity slip factor increase meaning that the fuel follow the bed material flow at an higher extent than for the low fluidisation velocity.

The most significant result for the tube banks is the very large accumulation factor where 70% of the fuel particles are accumulated at the walls where they get caught, see Table 14. The combination of the high accumulation factor and a non-existing velocity slip factor give the conclusion that the low gas velocity and tube banks would be and invalid operation condition, no mixing, for the gasifier, see Figure 15

Figure 14 and Figure 15 show the comparison between a freely bubbling bed and a bed with tube banks for the low and high superficial gas velocity where the outlet concentration is displayed. The tube bank influences the fuel in the same way as noted in the solid dispersion experiments, influencing the bulk solid movement drastically. The velocity slip factor found also indicated that the fuel particles are almost unaffected by the velocity field created by the flow of bed material inside the unit, thus increasing the fuel particle residence time. The dispersion factor decreases 71% between freely bubbling bed and tubes at the high gas velocity, see Figure 15.

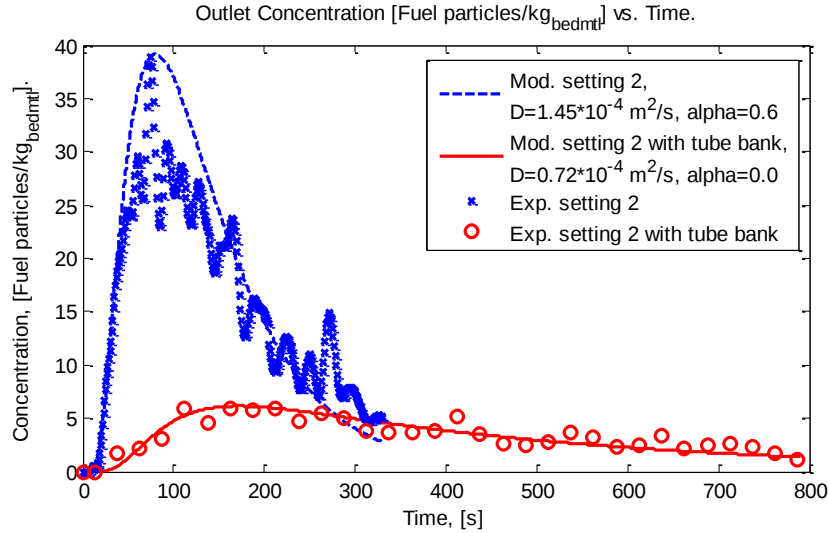


Figure 14 Experimental results and Model fitted to exp. results for setting 2 ($u_m=0.064\text{m/s}$) with 70mm bed height with and without tube bank

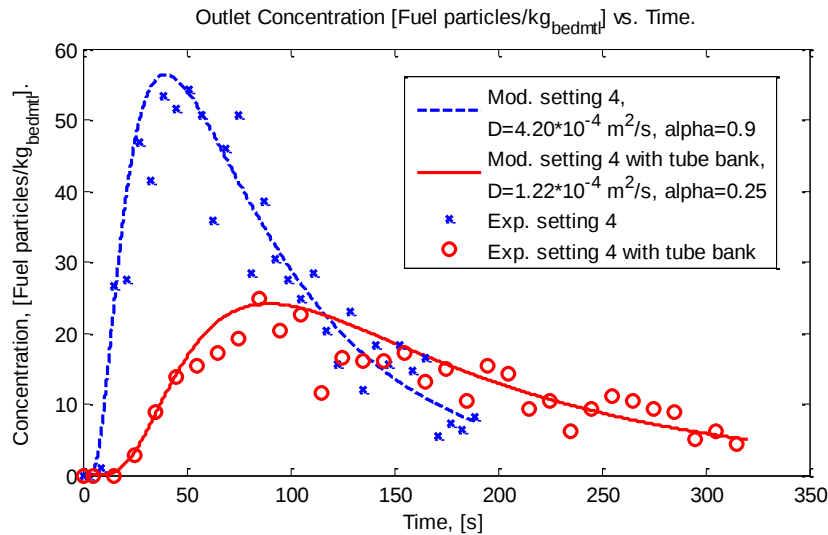


Figure 15 Experimental results and Model fitted to exp. results for setting 4 ($u_m=0.098\text{m/s}$) with 70mm bed height with and without tube bank

The method was easy, quick, and robust and gives valuable data on velocity field impact on fuel particle residence time.

3.2.2 Direct method

The experiments to find the dispersion coefficient for fuel were conducted in accordance with the experimental setup for the direct method. The lateral fuel dispersion coefficient is found by analysing the video recording in the mathematical model for fuel dispersion.

Table 15 Lateral fuel dispersion coefficient, recording time and time at surface

$D_{fuel}, [m^2/s]$	$u_m = 0.064m/s$	$u_m = 0.098m/s$
Freely bubbling bed: height 57mm	$1.77 \cdot 10^{-4}$	$2.64 \cdot 10^{-4}$
Recording time, h:m:s	02:35:36	02:17:52
Particle surface time, %	4.34	1.77
Freely bubbling bed: height 70mm	$4.30 \cdot 10^{-4}$	$6.35 \cdot 10^{-4}$
Recording time, h:m:s	01:18:36	01:21:16
Particle surface time, %	2.73	2.34
Tube banks: height 57mm	$3.89 \cdot 10^{-4}$	$2.49 \cdot 10^{-4}$
Recording time, h:m:s	01:17:59	01:14:59
Particle surface time, %	8.61	5.38

As seen in Table 15 the fuel dispersion coefficient for a freely bubbling bed at 57mm increases 49% with a 53% higher gas velocity. The surface time of the particle decreases significantly with a higher gas velocity indicating favourable mixing conditions.

Comparing freely bubbling beds at 70mm and 57mm the fuel dispersion coefficients increase 143% and 141% with bed height for the low gas velocity and high gas velocity respectively. This is in good correlation with findings of the bed dispersion for the high gas velocity.

For the low gas velocity the fuel dispersion coefficient for the tube bank was found to be higher than for the bed without tube bank. This is in contradiction to the results for the bed material dispersion and the indirect fuel dispersion method where the dispersion in the tube bank was much lower than that of the free bed.

During the analysis it was found that the recording length had more impact on the accuracy of the results than the use of High Definition (HD) recordings. Also, the time for analysing 1hour of HD recording was equal to that of 2 hours of the lower resolution. Therefore the later of the experiments were recorded in the lower resolution but for a longer period of time.

As noted in the indirect method the tube bank influences the fuel in the same way as noted in the solid dispersion experiments, whereas for the direct method the fuel dispersion was found to be higher or equal to that of the free bed.

As the results obtained from the indirect method was more reliable and also gave the velocity slip factor this method was used in the model tuning and optimisation. However, the direct method performed fairly for freely bubbling bed with no significant net flow of solids across the unit.

3.3 Up-scaling of experimental results

The results obtained in the indirect methods were chosen to represent the fuel particle dispersion and as basis for evaluating the fuel conversion. The dispersion coefficients were scaled by using equation (48). Scaling of setting 1 is presented below (and other values in Table 16).

$$D_L = \frac{D_M}{\frac{u_M}{u_L} \cdot \frac{L_M}{L_L}} = \frac{0.000262}{\frac{0.064}{0.158} \cdot \frac{1}{6}} = 0.00388 \text{ m}^2/\text{s} \quad (48)$$

Table 16 Up-scaled bed material and fuel particle lateral dispersion coefficients in m²/s

Setting	1	2	3	4
Bed material dispersion coefficient, m ² /s	0.00388	0.00244	0.00408	0.00477
Bed material dispersion coefficient tube bank, m ² /s	0.00108	0.00134	0.00108	0.00134
Fuel particle dispersion coefficient, m ² /s		0.00215		0.00617
Fuel particle dispersion coefficient tube bank, m ² /s		0.00107		0.00179

The findings for bed material dispersion can be compared with the results found by [7], 0.028 m²/s, where same scale model as in this thesis was used under other operating conditions. For fuel dispersion [5] obtained in the large scale gasifier values in the order of 0.35·10⁻³ to 1·10⁻³ m²/s. However, these values were found in cold conditions and no scaling laws were utilised.

3.4 Model tuning

The mathematical conversion and heat balance model was tuned using the up-scaled dispersion coefficients. As the operating condition used for the tuning was different to those conditions studied an extrapolation to find the corresponding dispersion coefficients was conducted.

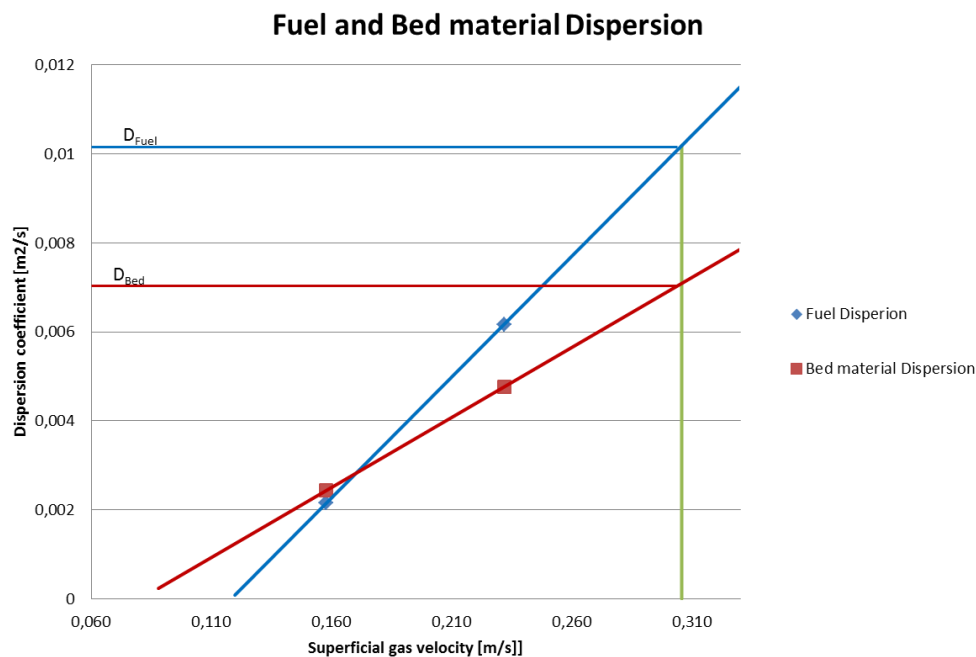


Figure 16 Extrapolation for fuel and bed material dispersion coefficient, setting 2 ($u_m=0.16\text{m/s}$) and 4 ($u_m=0.23\text{m/s}$) and corresponding values for superficial gas velocity of 0.306 m/s.

The operating conditions for the model tuning can be found in Table 17.

Table 17 Operating conditions for the tuning of the model

Operating condition for tuning	
Superficial gas velocity, m/s	0.306
Bed material mass flow rate, tonnes/h	23
Fuel mass flow rate, kg/h (wet basis)	400
Bed height, m	0.42
Char conversion, %	0-5%
Fuel dispersion coefficient, m ² /s	0.0102
Bed material dispersion coefficient, m ² /s	0.0071

As stated in section 1.2 the char conversion was 0-5% for the Chalmers gasifier. The model was tuned by using the operating conditions in Table 17 and adjusting the char activation energy, E , to achieve a 5% char conversion in the simulated model.

3.5 Optimisation

An optimization was conducted where the conversion of char was examined for setting 4 with a fuel flow rate of 285kg/h (wet basis) and a bed material flow rate of 4.72kg/s.

3.5.1 Impact of internals

3.5.1.1 Walls

The layout of the gasifier was unchanged and a simulation of the impact on char conversion for various positions of a partitioning wall and tube sections were performed. The first analysis concerned the inclusion of a wall placed as in Figure 17 and the resulting char conversion values are presented in Table 18.

Table 18 Partitioning wall placement and corresponding char conversion

Bed configuration	Char conversion, %
1: Freely bubbling bed	6.15
2: Wall NW	6.04
3: Wall NE	5.35
4: Wall SW	6.16
5: Wall SE	6.59
6: Wall West	6.18
7: Wall East	4.86

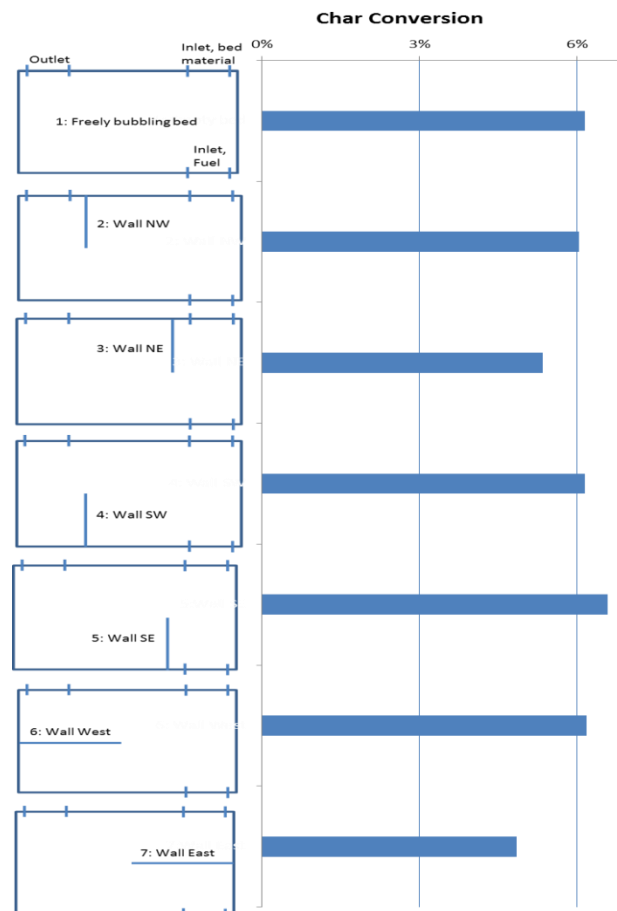


Figure 17 Partitioning wall placements for char conversion simulations

As seen in the table above the use of a partitioning wall had almost no impact on the char conversion no matter wall position. The only wall location that had any significant improvement, 7% increase, was the SE wall that creates a wake behind the wall where the char is held up due to low influence of the flowing bed material, i.e. dominated by dispersion. The east wall, no. 7, creates a temperature sink at the fuel inlet that delays the char conversion that is heavily dependent on high temperature surroundings.

3.5.1.2 Tube banks

The impact of tube banks on the char conversion is analysed where tube placements according to Figure 18 render the char conversions according Table 19.

Table 19 Tube bank placement and corresponding char conversion

Bed configuration	Char conversion, %
8: N to S Tube bank 1/3	6.16
9: N to S Tube bank 2/3	6.62
10: W to E Tube bank 1/3	7.54
11: W to E Tube bank 2/3	8.645
12: N wall Tube bank 2/3	21.61
13: S wall Tube bank 2/3	6.88
14: Tube bank covering 100%	19.65

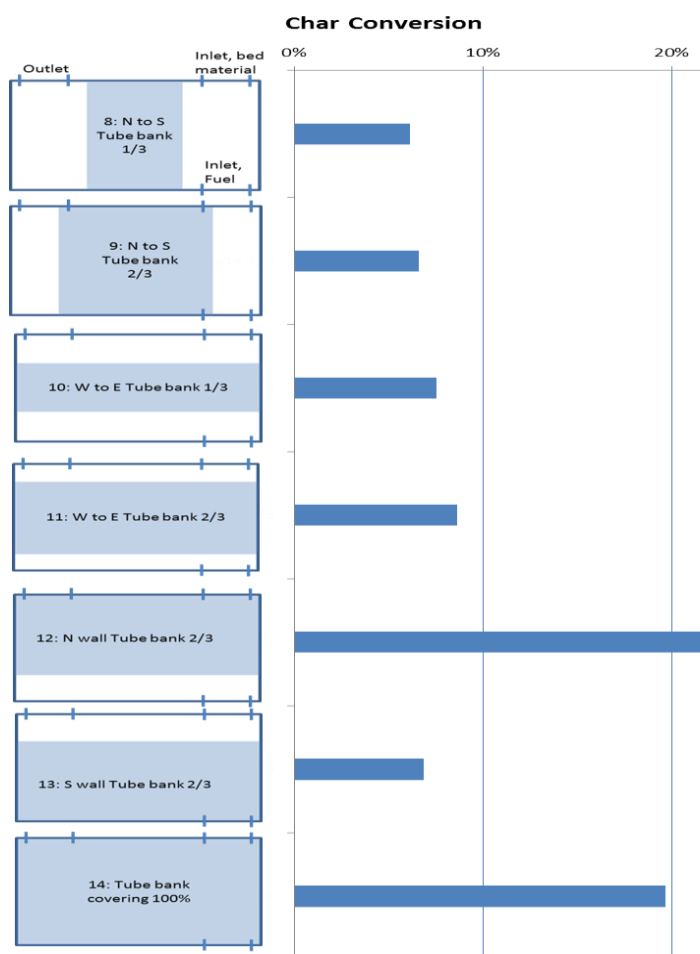


Figure 18 Tube bank placement for char conversion simulations

Compared to that of walls, the impact of the tube banks had a drastic effect on the char conversion. The tube bank configurations that had the largest effect were those that are adjacent to the bed material inlet, i.e. configurations 12 and 14.

Configuration 12 had the highest char conversion at 21.61% which is 250% higher than a freely bubbling bed. The slightly higher conversion rate than the tube bank covering the entire bed can be related to the fact that the fuel can easier travel horizontally away from the fuel inlet thus alleviate the created temperature sink.

3.5.2 Redesign options

Based on the findings in section 3.5.1 simulations where the impact of a combination of various partition walls and tube sections was performed. Also, relocation of the fuel inlet was also evaluated. The setups can be found in Figure 19 and the results are presented in Table 20.

Table 20 Redesign options corresponding char conversion for partitioning walls and tube banks

Bed configuration	Char conversion, %
15: N wall Tube bank 2/3, Wall SE	20.25
16: Tube bank covering 100%, Wall SE	19.89
17: N wall Tube bank 2/3, Wall West, Inlet fuel SW	14.30
18: Tube bank covering 100%, West Wall, Inlet Fuel SW	13.39
19: Tube bank covering 100%, Wall SE, Wall NW	19.95

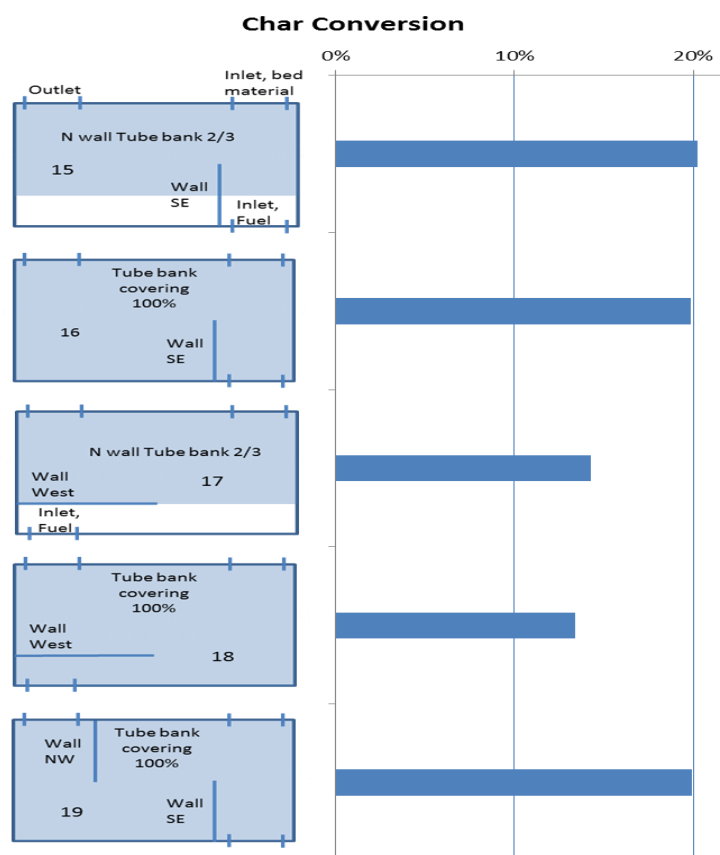


Figure 19 Redesign options corresponding char conversion for partitioning walls and tube banks

While configuration 12 was the best option from section 3.5.1 the implementation of walls did not improve the char conversion since they created temperature sinks that delay the conversion. While the implementation of walls increased the conversion for the entire tube banks configuration it did not reach same level as configuration 12.

The redesign option of moving the fuel inlet to the SW was not successful since it created a temperature pocket more than 100 °C lower than the bed material at the inlet impeding the conversion.

The resulting best design was configuration 12 found in section 3.5.1. This option does not imply any changes to be made on current layout of the gasifier.

3.5.3 Best design vs. current design

The comparison of the devolatilisation allocation between the current and the optimised design (configuration 12) can be seen in Figure 20 and Figure 21 below. Note that the devolatilisation for the freely bubbling bed covers a larger extension than with the tube banks, for which fuel yields higher concentrations at the fuel inlet.

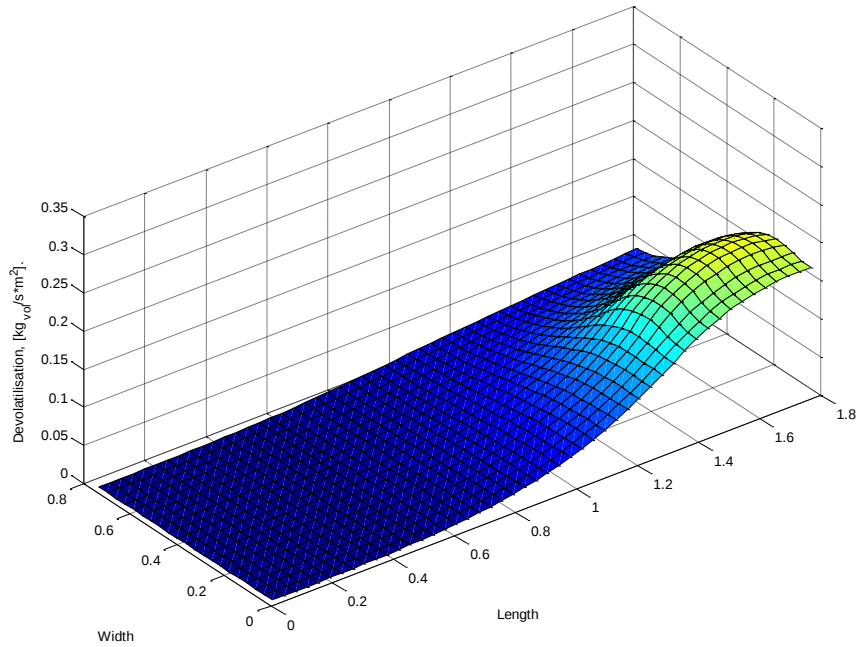


Figure 20 Allocation of devolatilisation for configuration 1: Freely bubbling bed.

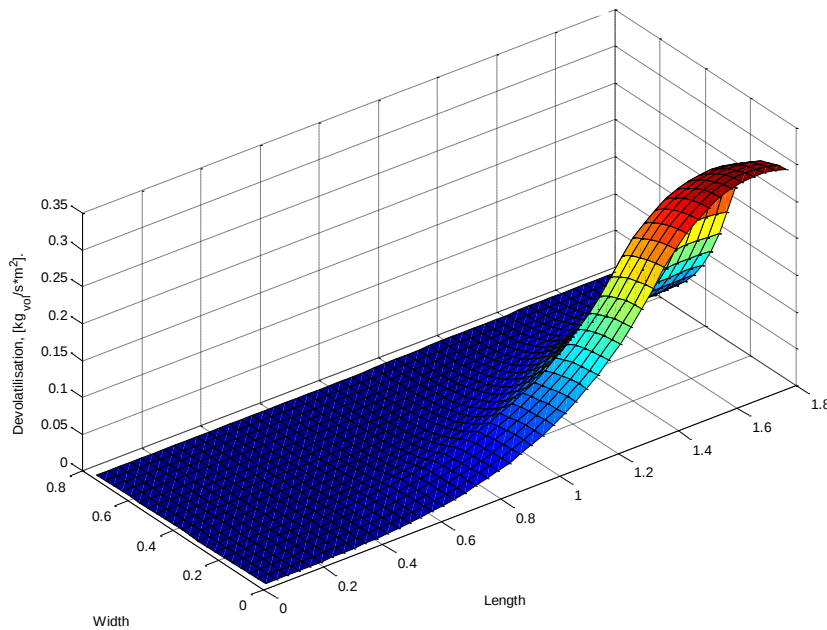


Figure 21 Allocation of devolatilisation for configuration 12: N wall Tube bank 2/3.

The allocation of char conversion in the current and the optimised design is shown in Figure 22 and Figure 23 (note the different scales at the z-axis). The optimised configuration reduces the dispersion and convection in the tube bank allowing the fuel to relocate towards the high temperature bed material inlet where the conversion becomes very high. At the freely bubbling bed the fuel flows directly towards the outlet resulting in a low char conversion.

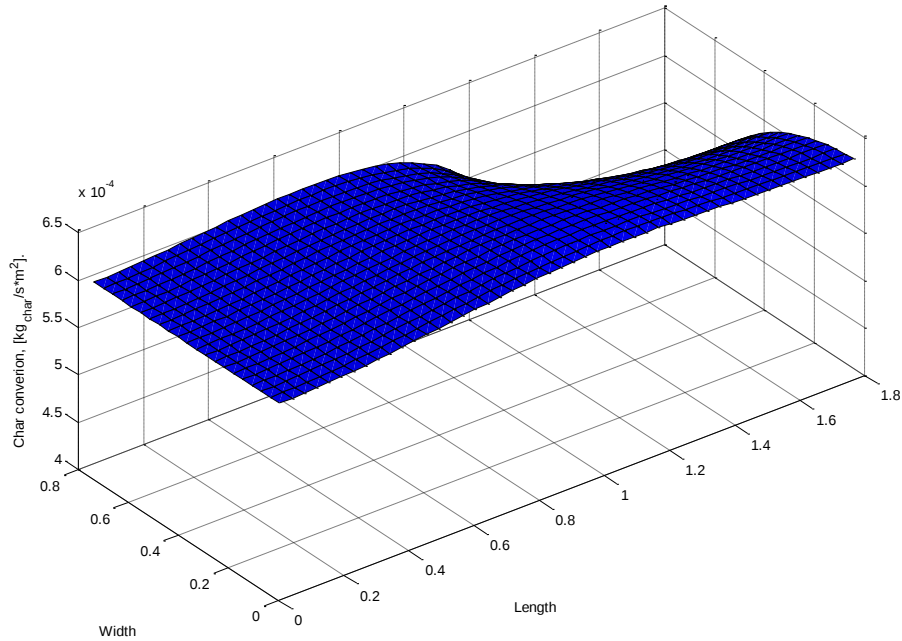


Figure 22 Allocation of char conversion for configuration 1: Freely bubbling bed.

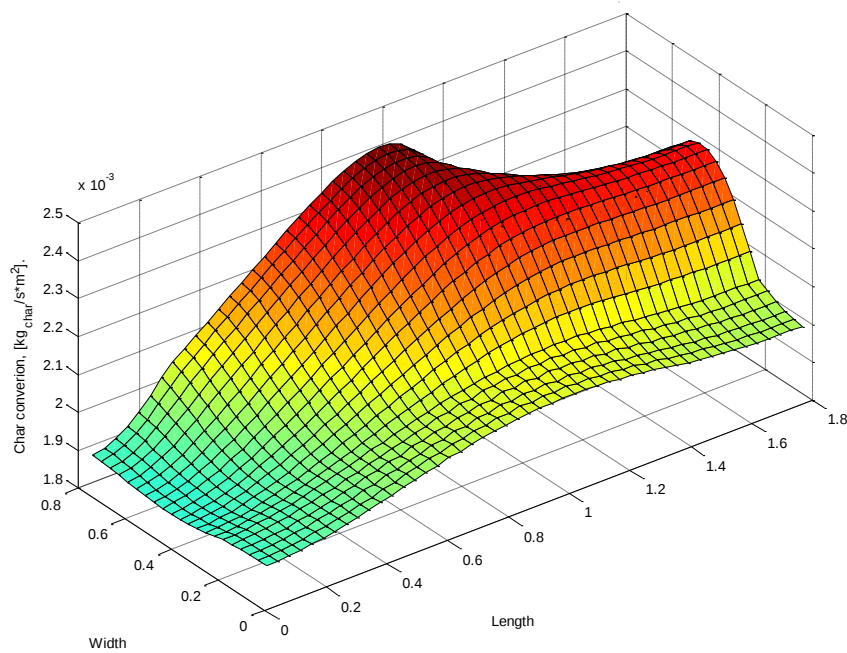


Figure 23 Allocation of char conversion for configuration 12: N wall Tube bank 2/3.

The elevated char conversion for the optimised design compared with current design decreases the temperature in the chamber by more than 15°C, as can be seen at the outlet in Figure 24 and Figure 25. Note in Figure 25 the large temperature gradient where the tube banks are located while the freely bubbling part of the same bed has almost no gradient at all, since the higher dispersion and convection of bed material transfers faster the heat across the bed.

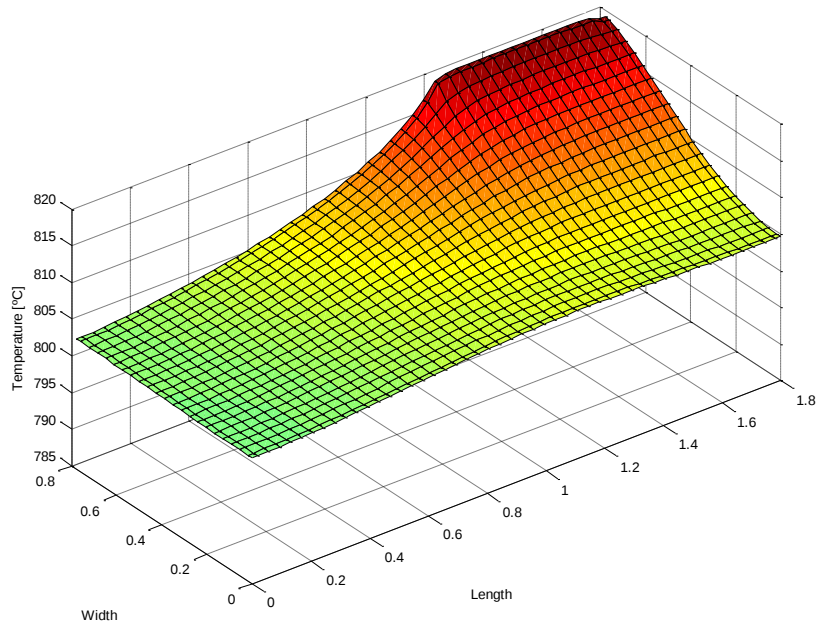


Figure 24 Temperature field for configuration 1: Freely bubbling bed.

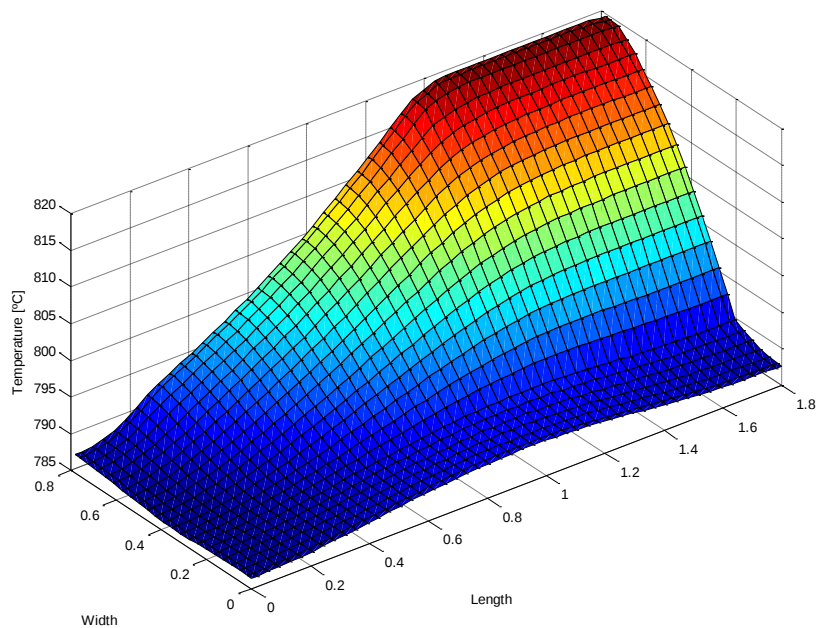


Figure 25 Temperature field for configuration 12: N wall Tube bank 2/3.

3.5.4 Sensitivity analysis

The impact of fuel mass flow was simulated, as seen in Figure 26. The increased fuel mass flow results in large temperature sinks at the fuel inlet delaying the fuel conversion. A five time increase in fuel mass flow results in 35% and 54% decreases for the freely bubbling bed and optimised bed respectively. A decrease of fuel mass flow by half would increase the char conversion to 24.5% for the optimal design.

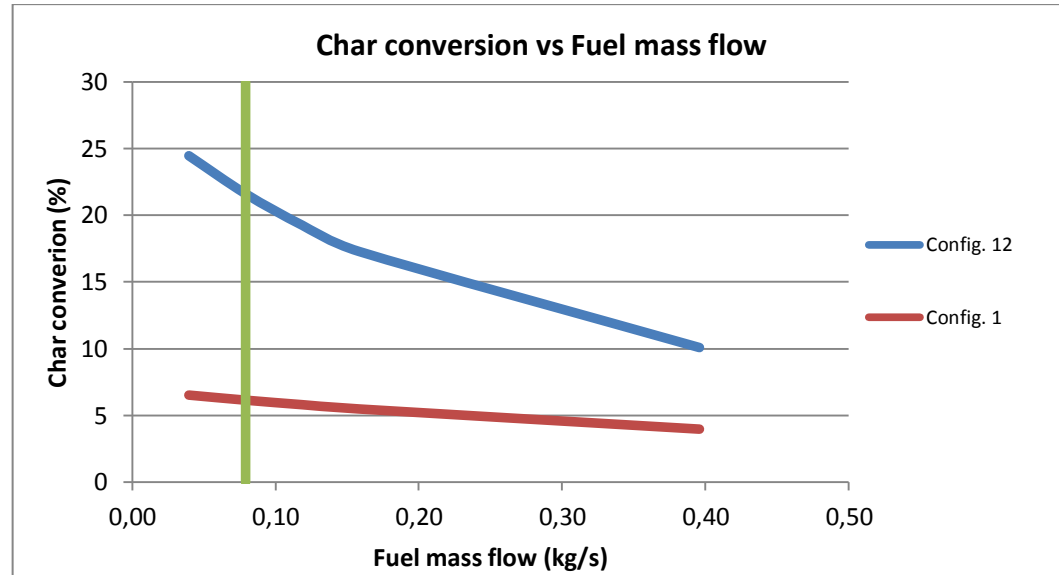


Figure 26 Char conversion for configuration 1 and 12 at different fuel mass flow rates.

The impact of bed material temperature is shown in Figure 27. Increased temperature results in increased char conversion for both designs but the impact is more noticeable for the optimised design. An increased temperature of 40°, from 820°C to 860°C, resulted in a 18% increase in char conversion for the optimum design whereas the same increase in temperature only resulted in a 6% increase for the current design.

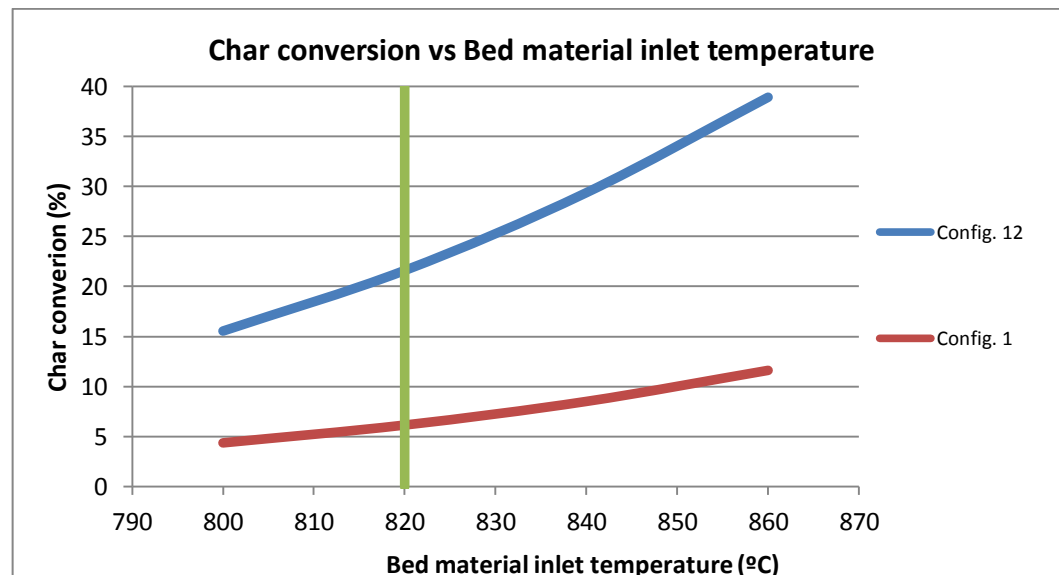


Figure 27 Char conversion for configuration 1 and 12 at different bed material inlet temperatures.

4 Conclusion

The lateral dispersion coefficients for bed material and fuel were investigated in the down-scaled model through experiments. The values obtained from the experiments have been scaled up to the Chalmers gasifier, ranging, with no internals added, within $2.44 - 4.77 \cdot 10^{-3} \text{ m}^2/\text{s}$ for bed material and $2.15 - 6.17 \cdot 10^{-3} \text{ m}^2/\text{s}$ for fuel.

The influence of internals (tube banks and partitioning wall) and relocating the fuel inlet was tested. The dispersion coefficients at tube banks ranged within $1.08 - 1.34 \cdot 10^{-3} \text{ m}^2/\text{s}$ for bed material and $1.07 - 1.79 \cdot 10^{-3} \text{ m}^2/\text{s}$ for fuel.

This shows that tube banks reduced the lateral dispersion of bed material and fuel by around 70% whereas the partitioning wall had little impact. Tube banks also reduced the bed material convective flow influence on fuel, α , with 40% to 100% thereby increasing the residence time of the fuel.

Lateral fuel mixing was evaluated through a direct method using digital image analysis, tracking UV-painted fuel particles. The method performs fairly for bubbling bed units as far as no significant net solids flow across the unit exists.

A new indirect method to investigate the fuel lateral fuel dispersion and the convective bed material flow influence was developed and tested. The method was easy, quick and robust, thus showing great potential.

A mathematical model for fluidized-bed indirect gasifier was constructed. The model is 2-dimensional and accounts for fluid dynamics, heat transfer and fuel conversion.

Through simulation runs of the tuned model for different arrangements of internals, it was found that the residence time of the fuel is the main parameter for improved fuel conversion. The resulting char conversion was increased from 6% for a bed without internals to 22% when two thirds of the bed was covered with a properly placed tube bank. Also, it was found that relocating the fuel inlet did not improve the fuel conversion.

5 Further work

- Investigation of the influence of bed material flow rates and bed heights on fuel dispersion and velocity slip factor.
- Investigation of the lateral solids dispersion in the direction parallel to the tubes and with the convective solids flowing in parallel with the tubes. Currently, values for the perpendicular direction to the tubes are taken.
- Improvement through some kind of automation of the magnetic separation method of tracer material, currently very time-consuming.
- Inclusion of radiative heat transfer in the modelling.
- Use of more advanced models for drying, devolatilisation and char conversion.
- Assess the influence of the choice of inlet boundary conditions in simulations at low flows of bed material.
- Evaluation of the char conversion in the Chalmers BFB gasifier with inserted tube banks.

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