The effects of air flow rate simulation on the combustion characteristics of luecaena woodchip in a fixed bed

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Abstract. Combustion of woodchip is one of the key areas in the clean energy strategy, but there is still a lack of detailed and systematic theoretical study on the fixed bed burning. The advantage of theoretical study lies in its ability to reveal features of the detailed structure of the burning process inside a solid bed, such as reaction zone thickness, the rate of individual sub-processes, gas emission, and char burning characteristics. These characteristics are hard to measure by conventional experimental techniques. In this study, a mathematical simulation was carried out for the combustion of luecaena woodchip in bench-top fixed bed and the effects of air flow rate have been assessed over wide ranges were investigated. It was found that volatile release as well as char burning intensify with an increase in the air flow until critical point was reached where a further increase in the air flow results in slowing down of the combustion process; a higher air flow also reduces the char fraction burned in the char burning stage and reduces CO emission in the gas exiting at the bed top.

1. Introduction
Global warming and climate change are recent complex issues which are impacting directly on environmental and people around the world. It is changing our economy, health and communities in diverse ways. Emission of CO₂ and other greenhouse gases from fossil fuel are main causes of global warming. Along with this, increasing consumption and limited availability of fossil fuels has urged us to look for the renewable energy sources. Biomass is becoming an important source to produce thermal energy and electricity now a day in the energy industry.

Comparing other fossil fuels, biomass consists of lower carbon, higher oxygen, higher volatile contents and lower specific heating value. So, the combustion of solid biomass is a complex phenomenon due to characteristics. Grate firing of biomass being the most popular technology has the advantage of firing fuels of different moisture content and requires less fuel preparation. A fixed bed is a place inside the grate furnace where thermal transformation of biomass occurs while the flammable volatiles burn inside the combustion chamber. CFD simulations any combustion system is very helpful to develop an efficient combustion system by combining theoretical model and experimental data.
To understand the combustion phenomena inside the fixed bed, in this study, a steady state CFD model was develop for woodchip combustion and a simulation study was accomplished. The CFD model was validated with data from a woodchip fixed bed reactor. The CFD model represents the temperature distribution in the fixed bed. Although this study focused on a bench-scale fixed bed reactor, the fixed bed combustion model is a generalized model. The modelling approach can be used to simulate steady state combustion process. Reactions and thermo-physical properties were evaluated by using empirical models, which suits biomass combustion.

2. Mathematical description of biomass in a fixed bed

Fixed bed combustion models can be primarily categorized into two different approaches according to the directional variations and treatment of the bed particles. Directional variations were considered from 0-D models to 3-D transient models. 1-D transient model is most popular among the fixed bed models and it is able to validate macro scale properties like burning rate [1-2, 4].

Fixed bed solid fuels can be considered either as a continuum or macro scale [3-6] or discrete particles by using Lagrange particle tracking or particle resolved micro scale [7]. Apart from those, physical and chemical properties and reaction models are significant for modelling of fixed bed combustion. Effective thermal conductivity is evaluated by considering combination of thermal conductivities for a quiescent bed with the correction of turbulence created due to mixing. This method is used for both homogeneous and heterogeneous bed models. In some models, radiation effect is also lumped into the effective thermal conductivity [8] while others were used separate radiation models [9].

Drying models can be represented by first order kinetic rate models assume drying rate depend only on the particle temperature and moisture content [8]. Devolatilization of fuels can be represented by a one-step global reaction, very slow to very fast reaction rates are shown a weak influence to the burning rate but it effects on ignition rate and reaction zone thickness [9]. Char combustion rate is controlled by both diffusion and reaction kinetics; hence it is modelled by considering combination of mass transfer coefficient and reaction kinetic rate constant [5, 9]. Volatile combustion reaction can be reduced to single reaction by considering major single volatile gas component and neglecting effects of other volatile reactions. Often minimum of mixing rate and kinetic reaction rate are taken as the volatile combustion reaction rate, although there are some instances only kinetic reaction rate is used [1, 9-10].

In this study, fixed bed combustion was considered as CFD model, a model was developed in MATLAB for fixed bed combustion of leucaena woodchip based on the thermally thin woodchip particles, where intra particle temperature gradients were negligible and the Biot number should be less than 0.1 (Bi < 0.1) [11]. Woodchip considered as a composite of moisture, volatile matter, char, and ash. Combustion process can be divided into 4 sub-processes: drying, pyrolysis, combustion of volatile and combustion of char.

Biot number: \[ Bi = \frac{\varepsilon b h d_p}{k} \] (1)

3. Fixed bed combustion experimental facilities

A fixed bed reactor was a vertical cylindrical combustion chamber placed on a weighing scale, figure 1. The height of the chamber was 1.5 m with an inner diameter 0.2 m. It consisted of an interior tube surrounded by a thick layer of insulating material and an external casing. The grate was located at the bottom of the chamber and consisted of a perforated plate made from stainless steel, with approximately 700 holes of 2 mm diameter, representing 7% open area. Thermocouples were used to monitor the temperature of air flow, temperature inside the bed at different height levels and temperature of the flue gases.

There was a gas-sampling probe inside the chamber at 1.25 m above the grate. The main components of the gas measurements of interest were O₂, CO, CO₂. A gas burner was placed at a 45° angle toward the woodchip at 0.45 m above the grate. The gas burner was used to initiate the burning
process of the woodchip sample. The air was fed from the bottom of the fixed-bed reactor through the grate without preheating. Properties of the leucaena woodchip fuels are used in the experiments and mathematical simulation was summarized in table 1. And the air flow rate was spanning a range by 120, 140, 160, 180 and 200 litre/min without preheating (around 30°C).

![Figure 1. Schematic of the experimental fixed bed reactor.](image)

Table 1. Woodchip properties used in experiments and simulation

| Moisture, wt% | Volatile, wt% | Fixed C, wt% | Ash, wt% | LCV, MJ/kg |
|---------------|---------------|--------------|-----------|------------|
| Leucaena      | 2.42          | 82.71        | 14.48     | 0.39       | 15.57      |

4. Fixed bed combustion model

The packed bed combustion model was developed based on following assumption;

1. The fuel bed considered a continuous porous layer consisting of gas and solid phases.
2. No heat transferred through the walls of the packed bed or adiabatic process.
3. Packed bed change with proceeding of combustion with shrinkage of the bed.
4. Gas incompressible flow and pressure drop due to fuel bed resistance to gas flow were included in source terms \( f_x \) and \( f_y \) and could be approximated by the Ergun’s equation [5].
5. The combustion gas was a mixture of species hydrocarbons represented by \( \text{H}_2\text{O}, \text{O}_2, \text{CO}_2, \text{CO}, \text{H}_2, \text{N}_2, \text{CH}_4 \) (light hydrocarbon) and \( \text{C}_x\text{H}_y\text{O}_z \) (tar), respectively [8].
6. The radiative heat transfer inside the bed could be modelled by the effective thermal conductivity.

According to the above assumption, gas and solid phases were considered as inter-penetrating continuous fluid mediums. Therefore, the gas phase was represented by the porosity in a finite volume. Mass, energy and momentum conservation has been separately applied for each phase as shown in following sub-sections.

4.1. Gas phase transport equations

Gas phase is presented in a two-dimensional model including species equation for each gas component.

**Continuity equation:**

\[
\nabla \cdot (\varepsilon_b \rho_g V_g) = r_{\text{dry}} + r_{\text{vol}} + r_{\text{char}}
\]

(2)

**X- momentum equation:**

\[
\nabla \cdot (\varepsilon_b \rho_g V_{g,x}) = -\frac{\partial P}{\partial x} + \nabla \cdot (\varepsilon_b \mu V_{g,x}) - f_x
\]

(3)

**Y-momentum equation:**

\[
\nabla \cdot (\varepsilon_b \rho_g V_{g,y}) = -\frac{\partial P}{\partial y} + \nabla \cdot (\varepsilon_b \mu V_{g,y}) - f_y
\]

(4)
Specie equation: \[ \nabla \cdot \left( \varepsilon_b \rho_g V Y_{g,i} \right) = \nabla \cdot \left( \varepsilon_b D_{g,i} \nabla Y_{g,i} \right) + r_i + \varepsilon_b \sum_{j \neq i} r_{i,j} \] (5)

Energy equation: \[ \nabla \cdot \left( \varepsilon_b \rho_g V h_g \right) = \nabla \cdot \left( \frac{\varepsilon_s \lambda_s}{C_{p,s}} \nabla h_g \right) + h_s A_p \left( t_s - t_g \right) + \left( r_{dry} + r_{vol} + r_{char} \right) h_s + \sum_{j \neq g} \Delta h_j r_j \] (6)

### 4.2. Solid phase transport equations

Solid phase is stationary in a circular tube combustor and then solid phase was modelled in one dimension which are embedded into solid phase mass diffusion coefficient.

Continuity equation: \[ \nabla \cdot \left( (1 - \varepsilon_b) \rho_s \right) = -r_{dry} - r_{vol} - r_{char} \] (7)

Species equation: \[ \nabla \cdot \left( (1 - \varepsilon_b) \rho_s Y_{s,i} \right) = \nabla \cdot \left( (1 - \varepsilon_b) D_{s,i} \nabla Y_{s,i} \right) - r_i \] (8)

Energy equation: \[ \nabla \cdot \left( (1 - \varepsilon_b) \rho_s h_s \right) = \nabla \cdot \left( \frac{(1 - \varepsilon_b) \lambda_s}{C_{p,s}} \nabla h_s \right) + h_s A_p \left( t_g - t_s \right) - \left( r_{dry} + r_{vol} + r_{char} \right) h_s - r_s \Delta h_s - r_{dry} \Delta h_{dry} \] (9)

### 4.3. Gas and solid phase properties modelling

#### 4.3.1. Fixed bed properties

The fixed bed radiation and thermal conductivity effects were lumped into an effective thermal conductivity and this effect was used for the whole bed [12]. Thermal conductivity modelled as a combination of effective thermal conductivity for a quiescent fixed bed along with a correction for fluid flow which this method was used by most of the studies in fixed bed heat transfer models. The solid diffusion coefficient in this model considered as a constant [13]. Bed porosity increased with the advancement of pyrolysis and char combustion reactions in a solid phase. Pressure drop along the bed height depended on air flow resistance of the fixed bed particles, which was modelled by using Ergun’s equation.

#### 4.3.2. Solid phase reactions

Drying and pyrolysis reactions were modelled by considering first-order kinetic rate models. The devolatilization product gas components are H₂, H₂O, CO, CH₄, CO₂ and tar (CₓHᵧOₚ) which are generated in the ratio of 0.00625:0.3125:0.22875:0.05875:0.14375 and 0.25 [14]. Pyrolysis rate selected from [9]. Char combustion reaction rate could not be modelled by the reaction kinetic rate since it depends on the oxygen diffusion into the particle surface. Therefore, the char combustion rate was modelled by considering both mass diffusion rate and first-order kinetic rate.

Drying \[ r_{dry} = 5.6 \times 10^6 \exp\left( -10584/t_i \right) \rho_{Y_{H_g,O_g}} \] (10)

Pyrolysis \[ r_{pr,i} = 7.0 \times 10^6 \exp\left( -9977/t_i \right) \rho_{Y_{s,i}} \] (11)

Char combustion \[ r_{char} = A_p \left( C_{O_2} \left( \frac{1}{k_r} + \frac{1}{k_s} \right) \right); \quad k_r = 290 \exp\left( -10344/t_i \right) \] \[ CO/CO_2 = 33 \exp\left( -4700/t_i \right) \] (12) (13)

#### 4.3.3. Gas phase reactions

Volatile gases combustion reactions assumed that the tar was gaseous species and did not condense, it combusted similar to paraffin.

H₂ combustion \[ R_{H_2} = 51.8t_{g}^{1.5} \exp\left( -3420/t_g \right) C_{H_2}^{1.5} C_{O_2} \] (16)

CH₄ combustion \[ R_{CH_4} = 1.6 \times 10^6 \exp\left( -24157/t_g \right) C_{CH_4}^{0.8} C_{O_2}^{0.8} \] (15)
\[ R_{\text{CO}} = 3.25 \times 10^7 \exp \left( -\frac{15098}{T_g} \right) C_{\text{CO}} C_{\text{H}_2\text{O}}^{0.5} C_{O_2}^{0.5} \]  
\[ R_{\text{C}_4\text{H}_8\text{O}_2} = 1.791 \times 10^{12} \left( \frac{T_g}{t_g} \right)^{0.5} \exp \left( -20131 \right) C_{\text{C}_4\text{H}_8\text{O}_2}^{0.5} C_{O_2} \]  

4.4. Solution of fixed bed combustion model equations

Fixed bed combustion modelling involved co-located grid system and then the Rhie-Chow method was used for pressure-velocity coupling [13]. SIMPLE algorithm used for steady-state conditions and applied to solved the whole system of transport equations which involved energy equations and species equations of both solid and gas phases. Due to the most of the model parameters were temperature dependent and convergence criteria of combustion model solver had been as gas phase temperature residuals of $10^{-6}$.

4.5 Boundary conditions

The solid phase temperature of the fixed bed at the bed top modelled as a radiation heat flux incident from the freeboard region. As solid temperature could not be explicitly added, the radiative heat flux at the interface used to calculate the temperature at the interface by the following equation [5].

\[ \frac{T_{\text{top}} - T_{\text{cell}}}{\Delta y/2} = \left( \varepsilon_{\text{rad}} \frac{T_{\text{env}} - T_{\text{top}}}{1 - \varepsilon_{\text{rad}}} \right) \]  

where $T_{\text{cell}}$ was the temperature of the bed top cell, $\Delta y$ was the height of the interface cell of the bed and $T_{\text{env}}$ was the radiation temperature.

5. Results and discussion

5.1. Bed height, solid temperature, and reaction zone thickness versus time

Figure 2a showed the calculated bed height and solid temperature profile versus time for the leucaena woodchip combustion with 160 liter/min of air flow rate. The initial bed height was 170 mm above the grate. It saw that the bed top begins to fall at $t = 120$ s as the local bed temperature rose from room levels to above 533 K. Then, the temperature at the top increases to 823 K while the height of the bed decreased linearly. From around $t = 300$ s, the bed-top temperature began to fall to around 923 K and remains at that level for a considerable time period. During this time, the height of the bed continues to drop linearly. It was also seen that the flame front reached the bottom of the bed (the grate) at around $t = 600$ s. After that, the bed became very hot for a wide period of time and the maximum temperature reached to 1173 K. This last stage the height of the bed undergoes only a slight fall. Further on, the bed residual material cold down as the combustion completes at $t = 1,200$ s.

It was also interesting to note the development of the reaction zone as time goes on. Figure 2a showed that as the bed ignited from the top at around $t = 120-150$ s, the reaction zone thickness increased quickly, with the combustion proceeding, the reaction zone thickness increased and reached a maximum level of 50 mm or 8 times of the original particle diameter when the flame front touches the bed bottom. After that, the reaction zone extended to the whole bed height for a while before reducing towards the bed top as the char burned out.
5.2. Process rates versus time and combustion stages

Figure 2b showed the individual process rates for drying, pyrolysis and char combustion. The first, drying was the initial or ignition stage \( t = 0-120 \) s where only evaporation occurs. The moisture inside the solids was driven out at a temperature of 373 K or 100°C by strong radiation the over-bed ignition source. When all the moisture in the top layer has been evaporated, the local bed temperature then rose quickly to the onset point of devolatilization (taken as 533 K or 260°C) at \( t = 120 \) s and the bed is ignited. Then fall in the drying rate, because the over-bed heat input to the drying front inside the bed decreased as the front moves away from the bed top downward. After that, the moisture evaporation rate coved as the heat supply shifted from over-bed to the newly established flame-front at the bed top.

The primary burning stage started after the bed gets ignited at \( t = 160 \) s and extended to where both the moisture and the volatiles in the whole bed had been completely driven out at \( t = 750 \) s. During this period, drying, pyrolysis and char burning rates maintained a relatively constant level, respectively (the formed char begins to burn at \( t = 400 \) s as the local solid temperature roses a preset onset temperature of 883 K or 610°C). Towards the end of this primary stage, the pyrolysis rate showed a sharp rise before falling to zero. Because, all the moisture has evaporated at this point, causing an upsurge in the bed temperature.

The final stage of combustion follows where only char burning occurs. An initial sharp increase in the char burning rate was showed and this is due to the increased \( O_2 \) availability to the char burning (in the primary stage, most of the \( O_2 \) was consumed by volatile gas burning). The char burning rate fell gradually until the whole combustion process completes at \( t = 1,200 \) s.

5.3. Time-averaged burning rate versus air flow rate

The air flow rate spans a range from 120 to 200 litre/min without preheating (around 30°C). The burning rate increased as the air flow rate increased until a peak point reached, beyond which a further increased in the air flow results in a fall in the burning rate. This is defined as the critical air flow rate and reflects the balance between the burning heat absorbed by the solids and the heat loss to the cooler gas stream from the particles. Thus, at lower air flow rates, more heat is generated than that carried away by the gas flow. At higher flow rates, however, the heat carried away by the gas flow exceeds the heat generated by combustion.
Results of the simulation and the experiments could be compared, an agreement is satisfactory to
good in terms of the general trends. However, the simulated burning rate was higher than the measured
one. This might be caused by the channelling phenomenon at a high air flow rate in an actual bed. The
local bed structure (porosity, particle size distribution and the orientation of particles) could not be
made perfectly uniform in an actual bed. The effect of such structure non-uniformly on the flow
distribution inside the bed became significant at high flow rates and channeling occurred as a
consequence, where some of the combustion air would bypass the solids and flows out of the bed
without reacting, thus reducing the burning efficiency.

5.4. Drying rate versus air flow rate

Figure 4a showed the simulated time-averaged moisture evaporation rate as a function of air flow
rate. It was seen that the evaporation rate of moisture from the solids follows a similar trend to the
overall burning rate as the air flow rates increased. The evaporation rate rises as the air flow increased
and reached a peak value at a critical point, beyond which further increased in the air flow results in a
dropped in the evaporation rate.

It could be explained by the heat input to the evaporation zone. When the air flow is low, an
increased in it causes a rose in the flame front temperature and enhanced the radiation heat transfer
from the flame front to the evaporation zone, hence the evaporation rate increased. But, at high air
flow rates, a decreased more heat is carried away from the evaporation zone by the fresh air at room
temperature. Thus, the net heat input to the evaporation zone actually decreased.

5.5. Pyrolysis rate versus air flow rate

Figure 4b showed the simulated time-averaged pyrolysis rate as a function of both air flow rate.
Similar trends to the situations of the burning rate were found that relate to the change in the air flow
rate. The volatile release rate is a strong function of temperature. The moisture evaporation front is
moved away from the pyrolysis front so that less heat is transferred from the pyrolysis zone downward
to the drying zone, resulting in a higher temperature in the pyrolysis zone which enhances the
pyrolysis rate.
5.6. Char burning rate versus air flow rate
The char burning rate was affected by the air flow rate. Figure 4c showed the char burning rate as a function of air flow. It was seen that the char burning rate rose as the air flow rate increased and reached a peak point at the critical air flow rate. After that, the rate declines as the air flow is further increased. The char burning depends on three factors: the amount of formed char, the O₂ availability and the temperature.

The relationship between its rate and the air flow as observed in figure 4c could be explained in terms of those three factors. In the range of air flow smaller than the critical rate, an increased in the air flow makes more O₂ available to the char burning. It also increased the pyrolysis rate as showed in figure 4b (so more char is produced). When the air flow rate is further increased beyond the critical point, the flame temperature undergoes no further increased and the pyrolysis rate started to decrease, so less char is produced, though more air is available to the char burning.

5.7. Gas emissions from the bed top
Gas compositions exiting from the bed top were important in that they provide information on the extent of further burning reactions needed in the over-bed region to complete the combustion. Figure 5 presented the calculated time-averaged O₂, CO₂ and CO concentrations at the bed top as a function of air flow rate. The O₂ concentration at the bed top increased with the increase in the air flow.

For the CO₂ emission from the bed top initially increased with increasing air flow, reaching a peak value and then gradually falls with a further increase in air flow. The concentration of the combustible gas (CO) exiting from the bed top is inversely proportional to the air flow rate and ranges from 16 to 25%.

![Figure 5](image-url). Calculated gas emissions from the bed top as function of air flow rate.

6. Conclusion and recommendations
Mathematical model simulations, as well as limited experiments, were carried out for the combustion of leucaena wood chips in a bench-top fixed bed and the effects of air flow were assessed and provided detailed information on the bed burning processes which are very difficult, if not possible to obtained by conventional experimental techniques. The major conclusions are:

1. As the combustion proceeds, the reaction zone thickness in the bed increases and reaches a maximum level of 8 times the original particle diameter when the flame front arrives at the bed bottom; the bed becomes very hot afterwards for a short while before the combustion ends.
2. Air flow rate has a significant effect on all the sub-processes: drying, pyrolysis and char burning. Increasing air flow initially increases each of the process rates but causes a decrease in the rates beyond a critical air flow rate.
3. The whole combustion process divided into three successive stages: the ignition stage, the primary combustion stage and the final char-burning stage. Char burned stage is affected by the air flow level. The higher the air flow, the larger the percentage of char burned in the final combustion stage.
4. Emission of the unburned gaseous at the bed top depends on the air flow rate, according to a theoretical calculation. A higher air flow rate produces low emission of CO, but higher O₂ from the bed top. Mathematical model simulations have correctly predicted the trend of change in the bed burning rate and bed temperature with variation in the air flow rate where experimental data are available. A quantitative discrepancy in a certain range of the operating parameters results from the neglect the channelling effect in the bed.

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