Comprehensive Design of a Fast Pyrolysis Reactor for Waste Utilization

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Abstract- This study deals with the comprehensive design of a biomass fast pyrolysis bubbling fluidising bed (BFB) reactor. The solid lignocellulosic residue of sugarcane industries, bagasse, was considered as feedstock. A detailed kinetic model was developed in order to simulate the behaviour of the pyrolyser and the kinetic parameters of bagasse pyrolysis were determined by fitting the model to appropriate experimental data. Subsequently, energy balances were applied in order to calculate the necessary heat of pyrolysis. This is 1.4 MJ/kg of bagasse and it is in accordance with relevant literature data. The model was, also, validated with respective experimental data and thereby it can be effectively used to simulate the performance of similar fast pyrolysis systems at various process conditions. Finally, the study concludes with the estimation of the capital and labour costs associated with the reactor and bio-oil storage units. In effect, this study exhaustively incorporates the basic mass, energy and economic calculations of the core unit of a fast pyrolysis plant. The latter will be fully designed in the near future as an expansion of the current model. Overall, the current research suggests and investigates the utilisation of a relatively abundant agricultural residue via fast pyrolysis for the production of bio-oil. The conditions that maximise the liquids yield is 525°C and residence time of 0.5s.

Keywords biomass fast pyrolysis, kinetic modelling, simulation, process economics, bagasse

1. Introduction

In most regions of the world the sharpest growth in biofuel manufacture has been observed during the last two decades, mainly due to recent government policies. The focal drive for these supportive policies are associated with energy security concerns along with the desire to enhance the agricultural economy [1]. Furthermore, the goal of reducing CO_2 emissions in the transport sector is also a significant driver for the establishment of a sustainable biofuel sector, principally in countries belonging to the Organisation for Economic Cooperation and Development (OECD). The most widely implemented measure in this direction is a blending mandate; essentially, it determines the portion of biofuel that has to be blended with conventional fossil fuels [2].

An efficient way to produce biofuels is via pyrolysis. Pyrolysis is a thermochemical process via which char, gas and organic liquids can be produced through the anaerobic decomposition of biomass. It takes places at temperatures lower than gasification (300-800°C) and usually the desired product is termed bio-oil [3]. Based on the operating parameters, pyrolysis can be categorized into three main groups: slow, intermediate and fast pyrolysis. These are different in process temperature, heating rate, solid residence time and biomass particle size. In addition the product distribution (char, liquids and non-condensable gases) depend on pyrolysis type and operating conditions [4] as depicted in Table 1. As mentioned above, in most cases the desired product is bio-oil since production of combustible gases is more efficient via gasification and if char is to be combusted then it is preferable to burn directly the entire biomass [5]. Thus, in this study the fast pyrolysis of bagasse was considered and examined. Fast pyrolysis can have moderately low capital costs and high thermodynamic efficiency compared to other procedures, particularly on a small scale. Production of bio-oil via fast pyrolysis has received more consideration in recent years due to the following potential advantages [6, 7].

- The bio-oil is a renewable fuel suitable, in many cases, for power generation purposes

- The process can exploit lignocellulosic feedstock
- The bio-oil is easy to store and transport
- It is possible to be upgrade the bio-oil into motor fuels
- The bio-oil has higher energy content than gasification producer gas
- The bio-oil can be also gasified similarly to solid biomass

During fast pyrolysis procedure, biomass is rapidly heated to a high temperature (approximately 500-600°C) in the absence of air. Normally on a weight basis, fast pyrolysis yields 55%–70% of liquid hydrocarbon products with 15%–25% of solids (mainly biochar) and 10%–20% of non-condensable gases, based on the characteristics of the feedstock. The production of liquids is typically generated from biomass at moderate temperature, high heating rate and short resident time conditions. The main features of the fast pyrolysis method are high heat transfer and heating rate, very short vapour residence time, fast quenching of vapours, increased bio-oil production and accurate control of reactor temperature. In addition, it has potential to provide several high value chemicals such as benzene and toluene [4]. The core of the pyrolysis procedure is the reactor (pyrolyser), and extensive investigation and development has been dedicated to various reactor systems. Over the past two decades, numerous reactor technologies have been examined and many processes have been tested and explored under different operating conditions. Scott et al. [5] have conducted a comprehensive study on pyrolysis reactor systems and concluded that 1) the bio-oil cooling demand will be minimum if the gas to feed ratio is minimized, 2) the reactor's temperature should remain as low as possible and 3) the pyrolyser should operate effectively on a small scale and at the same time have the potential to scale up easily. The bubbling fluidized bed (BFB) reactor appears to satisfy these criteria best and its construction and operation are now well established. Control of temperature and heat transfer can be accomplished simply and efficiently. High bio-oil yields can be attained via this reactor type. An effective way to control the vapour residence time is through controlling the flow rate of the fluidising agent (normally nitrogen). One restraint of the BFB is that biomass has to be supplied at very small particle size (2-4 mm) in order to sustain the high heat transfer rate and self-cleaning of products from the bed of the reactor.

Table 1. Features of different pyrolysis types [4]

Pyrolysis type	Gas	Bio-oil	Char	Heating rate (°C/s)	Residence time (s)
Slow	35	30	35	0.1-2	300-1200
Intermediate	25	50	25	2-10	10-200
Fast	13	75	12	10-200	<2

A promising feedstock for a sustainable biofuel sector is the solid residue of sugarcane industries, known as bagasse. It is the major agro-industrial residue in Brazil and, despite its suitability for fuels and/or chemicals production [8], it is mainly utilised to raise steam and generate electricity; nevertheless there is still some excess [9]. Furthermore, if more efficient combustion boilers are implemented, it has been estimated that less bagasse would be incinerated (approximately 36% reduction) [9]. To this direction, research has focused on exploiting bagasse in biorefining technologies that are capable of producing saleable fuels from sugars, pyrolysis and syngas [10]. In addition, the expected escalation in cane production in order to satisfy the increase of global ethanol demand (from 70 to 130 million m3 between 2009 and 2025), will also give rise to further bagasse availability [11]. The ultimate and proximate analyses of bagasse are illustrated in Table 2 [12].

In view of these, the present study focuses on developing a comprehensive model of a fast pyrolysis BFB reactor for bagasse utilisation. In addition, the model was validated with respective experimental data and thereby it can be effectively used to simulate the performance of similar fast pyrolysis systems at various process conditions.

Table 2. Bagasse composition **[12]**

Proximate analysis	
Parameters	**Mass fraction (%)**
Moisture	10 (wb)
Ash	3.2 (db)
Volatile matter	83.65 (db)
Fixed Carbon	13.15 (db)
Ultimate analysis	
Element	**Dry Weight (%)**
C	45.38
H	5.96
O	45.21
N	0.15
S	0.1

2. Mass Balance on the Pyrolyser

With the aim of predicting the behaviour of a chemical reactor, information on the reaction kinetics, thermodynamics, heat and mass transfer, and flow patterns are commonly necessary. Particularly, in the case of bubbling fluidized bed reactors the flow and the phase interaction phenomena are, in general, the most challenging modelling tasks. Initially, the dominant proposed modelling approach was the adaption of ideal flow models including plug flow, continuous stirred tank or mixed flow, dispersion, and the tank in series [13]. Toomey and Johnstone [14] were the first to propose a two-region model, which has subsequently been modified in several other studies. The chief improvement that these models have offered was that they give the opportunity to the researchers to take into consideration the observed non-homogeneity of dense fluidized beds, identifying the dilute bubble and the dense emulsion phases. The fluidised bed reactor models are 1) the Basic Two-Zone Model, 2) the van Deemter Two-Zone Model, 3) the Davidson-Harrison Two-Zone Model and 4) the Kunii-Levenspiel Three-Zone Model. In this study the Basic Two-Zone Model was used due to its simplicity compared to the Kunii-Levenspiel model and its accuracy compared to the other two models [15]. The overall model assumptions can be summarised as follows [16]:

- Two phases were taken into consideration, the emulsion and the bubble phase. In the emulsion phase the heterogeneous reactions take place while in the bubble phase the homogeneous reactions.

- The flow of gas in the bubble phase is in plug flow.
- The emulsion phase is well mixed.
- No reactions occur in the freeboard of the reactor.
- The bed is under isothermal conditions.
- Char only consists of carbon and is confined to the emulsion phase.

Biomass pyrolysis comprises several highly complicated reactions that include a significant number of intermediates and end products; thereby the formulation of a precise reaction mechanism pathway and kinetic model is still challenging and consequently the dominant approach for developing kinetic models is to use measurable and simplified kinetics. Hence, this has given the opportunity to researchers, throughout the years to put forward various reaction schemes and kinetic models. By reviewing the biomass pyrolysis mechanisms up to now proposed, the only absolute outcome revealed is that raw biomass is decomposed to non-condensable gases (such as H_2, CO_2 and CO), bio-oil (or tar) and carbon solid (char) [17]. There are three main approaches for pyrolysis reaction mechanism: 1) One step global models, 2) competing reaction models and 3) models with secondary tar cracking. One step global models were mainly developed in the early stages of pyrolysis modelling and propose that biomass degrades to volatiles and coke. The most used model to describe wood pyrolysis is the competing reaction model. It proposes three final products with char yield fluctuating. It is an empirical model and as such it was kept as simple as possible. Finally, most recent models take into consideration the secondary tar cracking. These are actually an expansion of the competing reaction model but feature also tar decomposition to gases and char [17]. One step global models and competing reaction models are inferior compared to models with secondary tar cracking, as they consider a fixed proportion of the char to volatiles yield. In fact, the latter models are more versatile and as so they have been used in the present study to describe the pyrolysis reaction scheme. In particular, the model developed by Di Blasi et al. [18] was used in this research. Figure 1 depicts the reaction scheme of the proposed model.

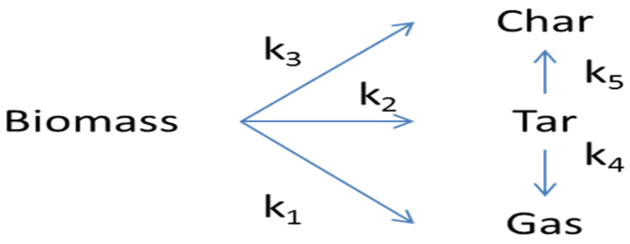


Figure 1. Fast pyrolysis kinetic scheme used in this study [18]

Based on this mechanism, for each component the formation or disappearance rate is given by the following kinetic rate equations:

$$R_B = -(k_1 + k_2 + k_3) * m_B \qquad (1)$$

$$R_g = k_1 * m_B + k_4 * m_t \qquad (2)$$

$$R_t = k_2 * m_B - (k_4 + k_5) * m_t \qquad (3)$$

$$R_c = k_3 * m_B + k_5 * m_t \qquad (4)$$

Where m_B, m_t, m_g and mc are the mass fractions of bagasse, tar, gas and char respectively. The rate constants of the above reactions ($k_{1\text{-}5}$) can be expressed using the Arrhenius equation (temperature dependent). A plug flow term combined with an axial dispersion term was used so as to approximate the gas flows and carbon is present to the emulsion phase. Additionally the mass transfer between the bubble and emulsion phases was taken into account [19, 20, 21]. Finally, an overall mass balance was considered for the bagasse consumption and it was assumed that the exit concentration is zero (full conversion). Table 3 depicts the hydrodynamics associated with the model.

<u>Mass balance for the bubble phase:</u>

$$f_b \frac{\partial m_{ib}}{\partial t} = -f_b U_b \frac{\partial m_{ib}}{\partial z} + f_b D_{ib} \frac{\partial^2 m_{ib}}{\partial z^2} - k_{be} S_{be} (m_{ib} - m_{ie}) + f_b R_{ib} \qquad (5)$$

<u>Mass balance for the emulsion phase:</u>

$$f_e \frac{\partial m_{ie}}{\partial t} = -f_e U_e \frac{\partial m_{ie}}{\partial z} + f_e D_{ie} \frac{\partial^2 m_{ie}}{\partial z^2} + k_{be} S_{be} (m_{ib} - m_{ie}) + f_e R_{ie} \qquad (6)$$

<u>Mass balance for the bagasse decomposition:</u>

$$\frac{\partial m_B}{\partial t} = \frac{W_{B,in} - W_{B,out}}{B} + R_B \qquad (7)$$

<u>Boundary Conditions:</u>

$$At\ z = 0 \ \rightarrow \ m_{ib} - \frac{D_{ib}}{U_b} * \frac{\partial m_{ib}}{\partial z} = C_{io}\ and\ m_{ie} - \frac{D_{ie}}{U_e} * \frac{\partial m_{ie}}{\partial z} = C_{io} \qquad (8)$$

$$At\ z = L \ \rightarrow \ \frac{\partial m_{ib}}{\partial z} = 0\ \ and\ \ \frac{\partial m_{ie}}{\partial z} = 0 \qquad (9)$$

$$At\ z = 0 \ \rightarrow \ \frac{\partial m_B}{\partial z} = 0\ \ and\ at\ z = L \rightarrow \frac{\partial m_B}{\partial z} = 0 \qquad (10)$$

Where subscripts b and e refer to bubble and emulsion phase respectively, m_{io} the initial mass concentrations (1 for bagasse and 0 for char, tar and gases), U the superficial velocity of gas (m/s), f the volume fraction (m^3/m^3), D the effective gas diffusivity (m^2/s), R the reaction rate ($kmol/m^3s$) in either bubble or emulsion zone, k_{be} (m/s) the mass transfer coefficient between the emulsion and the bubble phase, z the axial distance (m), B the weight of the bagasse in the reactor (kg), Win is the rate of bagasse inlet to the reactor (kg/s), Wout (equal to zero) is the rate of bagasse exiting from reactor (kg/s) and L the reactor length. Furthermore, it should be noted that f_b and f_e are related as $1 = f_b + f_e$. The overall bed void fraction at minimum fluidization, εmf, was assumed to be equal to 0.4 and inlet nitrogen (fluidising agent) velocity, U_0, equal to 1.5 m/s [16]. The term S_{be} refers to the interfacial area between the bubble and the emulsion phase and is calculated per unit volume of the continuous emulsion phase. The surface area of each bubble is πd_b^2 (spherical bubbles were assumed) and the number of bubbles is given by the total volume occupied by the bubbles divided by the volume of each bubble ($\pi d_b^3/6$) [22].

$$S_{be} = \pi d_b^2 \left(\frac{V_b}{\pi d_b^3/6}\right) \frac{1}{V_e} \tag{11}$$

$$f_b = \frac{V_b}{V_e + V_b} \Leftrightarrow \frac{V_b}{V_e} = \frac{f_b}{1-f_b} \tag{12}$$

$$S_{be} = \frac{6}{d_b} \frac{f_b}{1-f_b} \tag{13}$$

Table 3. Model hydrodynamic parameters

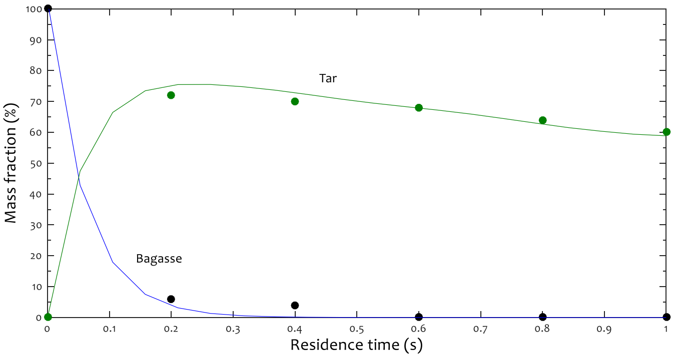
Name	Expression
Bubble diameter	$d_b = d_{bM} - (d_{bM} - d_{b0}) \times \exp(-\frac{0.3x}{D})$
Maximum bubble diameter	$d_{bM} = 1.64 \times [A \times (U_0 - U_{mf})]^{0.4}$
Initial bubble diameter	$d_{b0} = 0.00376(U_0 - U_{mf})^2$
Minimum fluidization velocity	$U_{mf} = \sqrt{\frac{d_p(1-\varepsilon_{mf})(\rho_s - \rho_g)}{\rho_g}}$
Gas velocity in bubble phase	$U_b = U_0 - U_{mf} + 0.711(g d_b)^{0.5}$
Gas velocity in emulsion phase	$U_e = U_{mf}/(1-f_b)$
Bubble fraction	$f_b = (U_0 - U_{mf})/U_b$
Mass transfer coefficient	$k_{be} = 0.11 U_b$
Gas diffusivity in bubble zone	$D_b = 0.001$
Gas diffusivity in emulsion zone	$D_e = 100$

Overall the system consists of 5 partial differential equations and 1 ordinary differential equation. This set of equations was solved by using the Matlab nested function ‘pdepe’, able to solve parabolic and elliptic differential equations in one space variable and time. The system was solved using a second order discretization with 50 equidistant points. A grid sensitivity analysis showed that the results are precisely the same using more than 50 discretization points. The ordinary differential equations (ODEs) resulting from discretization in space are integrated (a Runge-Kutta integrator is used) to obtain approximate solutions at times specified by the user. Kinetic values for bagasse pyrolysis have not yet been reported in the literature. Thus, in order to determine the values of the Arrhenius based kinetic parameters ($k_i = A_i \exp(-E_i/T)$), fitting of the pyrolysis kinetic model to experimental data for bagasse [23] was performed (limits were adopted from Di Blasi and Branca [24] and refer to wood pyrolysis) by minimizing the sum of squared errors between experimental and computed values. For this purpose an optimization routine based on genetic algorithms followed by a gradient based method (SQP – fmincon nested function in Matlab) was developed in the Matlab environment. Genetic algorithm, a stochastic one, was used to estimate a family of solutions near the (potentially) global optimum and Successive Quadratic Programming (fmincon function) was subsequently applied to compute the final optimum [25]. Table 4 illustrates the calculated values of the above kinetic parameters (A_i and E_i) for the bagasse case as well as their limits.

Table 4. Fast pyrolysis kinetic data calculated for bagasse by fitting the reactor model (Equations 5-7) to available experimental data [23] and checking against limits reported for wood pyrolysis [25]

	Lower limit	Upper limit	Computed
E_1	15,836	19,836	**17,652**
E_2	15,501	19,501	**16,515**
E_3	11,435	15,435	**12,640**
E_4	10,000	14,000	**13,051**
E_5	10,000	14,000	**12,331**
A_1	2.38×10^9	6.37×10^9	**5.91×10^9**
A_2	8.77×10^9	12.76×10^9	**12.55×10^9**
A_3	9.26×10^6	2.92×10^7	**1.25×10^7**
A_4	2×10^6	3×10^6	**2.34×10^6**
A_5	750,000	1.25×10^6	**9.82×10^5**

Where Ei (K) is the ratio of activation energy, Ea (J/mol), over the gas constant R (J/mol K) and A (1/s) is the pre-exponential factor for the Arrhenius equation. Figure 2 illustrates the model predictions for gas, tar and char formation versus the experimental data. The temperature that maximises bio-oil yield was found to be equal to 525°C.



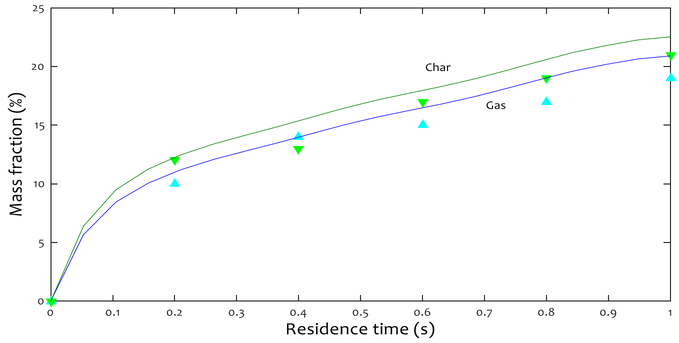
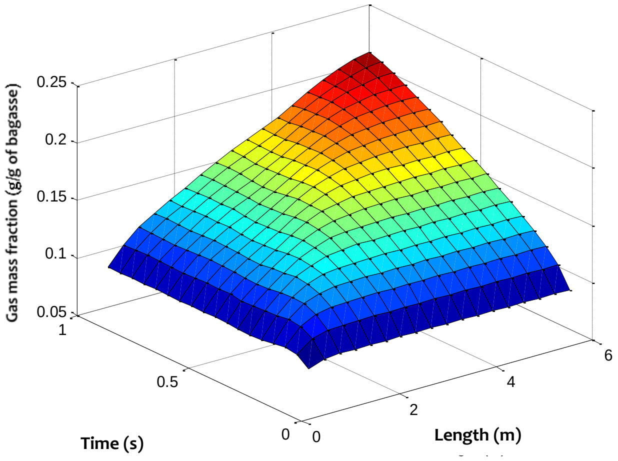
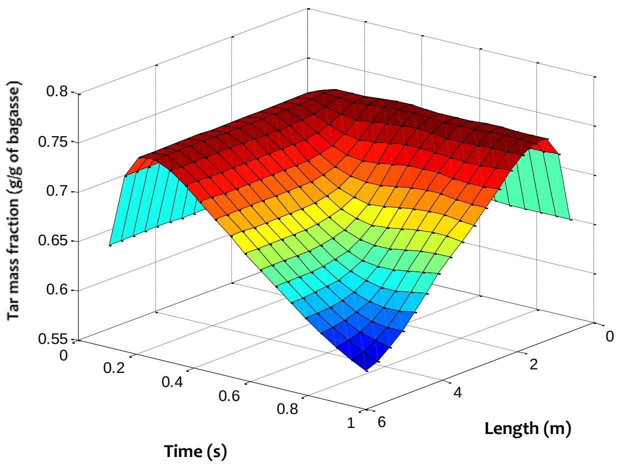


Figure 2. Pyrolysis Model output for gas, tar and char and residual bagasse versus experimental data [23] for the fast pyrolysis of bagasse

In addition, Figure 3 presents the product distribution as a function of reactor length at a specific time point and the time. It can be observed that the tar yield reaches a maximum after 0.2-0.4 seconds and then is decomposed to char and gases as described in Figure 1. As reported in the literature [26], tar formation requires very low vapour residence time in order to minimise secondary reactions. However, pyrolysis reactors with residence time below 1.5 s are difficult to control and operate. Hence, a residence time of 0.5 s can satisfy an efficient reactor design and bio-oil yields of almost 70% of the total biomass input. The computed reactor geometry is length of 6m and diameter of 1m.





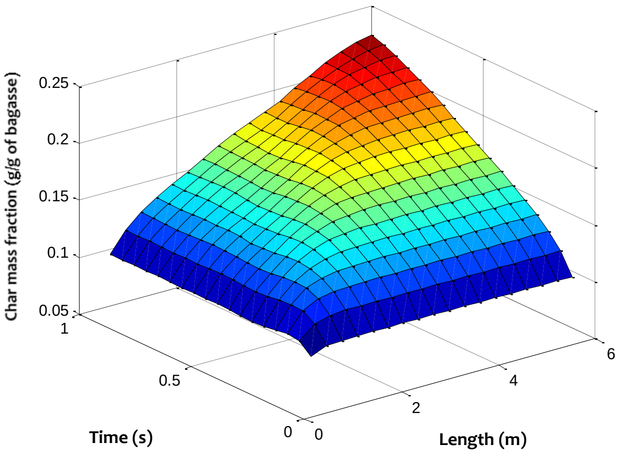


Figure 3. 3D representation of fast pyrolysis products against reactor length and time for the system investigated in the present study

3. Energy Balance on the Pyrolyser

As discussed previously, the pyrolysis reaction mechanism consists of five reactions; the biomass decomposition reactions (reactions 1-3 as depicted in Figure 1) which are highly endothermic (ΔH1-3 = 420 kJ/kg) and the tar cracking reactions (reactions 4-5) which are exothermic (ΔH4-5 = -40 kJ/kg) [27], hence the overall process is endothermic. Therefore, heat balance calculations are essential in order to define the required heat input. This amount of energy, QP, can be calculated from the equation below [26, 27]:

$$Q_P = Q_{W,S} + Q_{W,L} + Q_{V,S} + Q_{Feed,S} + Q_E \qquad (14)$$

Where $Q_{W,S}$ is the sensible specific heat required to heat the water contained in the feed to 100°C, $Q_{W,L}$ is the latent specific heat required to evaporate the water containing in the feed, $Q_{V,S}$ is the sensible specific heat required to heat the water vapour to the pyrolysis temperature, $Q_{Feed,S}$ is the sensible specific heat required to heat the feed to the pyrolysis temperature and Q_E the specific heat required for the pyrolysis reactions. These heat flows can be calculated as follows [26]:

$$Q_{W,S} = wC_{pw}(100 - T_F) \qquad (15)$$

$$Q_{W,L} = wE_w^{100} \qquad (16)$$

$$Q_{V,S} = wC_{pv}(T_P - 100) \qquad (17)$$

$$Q_{Feed,S} = (1 - w)C_{p,Feed}(T_P - T_F) \qquad (18)$$

$$C_{p,Feed} = FC_{p,F} + VC_{p,V} + AC_{p,ash} \qquad (19)$$

$$C_{p,F} = -0.208 + 3.807 \times 10^{-3}T - 1.7558 \times 10^{-3}T^2$$
(20)

$$C_{p,V} = 0.728 + 3.3.391 \times 10^{-3}T \qquad (21)$$

$$C_{p,ash} = 80 - 2.4 \times 10^{-2}T - 35.47 \times 10^{-4}T^2 \qquad (22)$$

$$Q_E = (1-w)Q_R \qquad (23)$$

Where w is the water mass fraction in feed, C_{pw} the water specific heat capacity at 1 bar, C_{pv} the water vapour specific heat capacity at 1 bar, T_P is the pyrolysis temperature, T_F is the reference temperature, E_w^{100} the enthalpy of water vaporization at 100°C and 1 bar, Q_R the sum of enthalpies of the pyrolysis reactions, $C_{p,Feed}$ the feed specific heat capacity and F, V and A are the mass fractions of fixed carbon, volatile matter and ash of the bagasse. At this point, it should be noted that bagasse was presumed to have passed through a dryer (an exit moisture content of 10% was assumed) before entering the pyrolysis reactor. Yang et al. [28] have previously estimated the heat of pyrolysis for different types of biomass feedstocks as depicted in Table 5. The computed value for the heat of pyrolysis is in good agreement with published data.

Table 5. Heat of pyrolysis for different raw materials [28] compared with that calculated for bagasse in the present study

Biomass type	**Heat of pyrolysis (MJ/kg)**
Pine	1.6
Oat hulls	1.25
Oak	1.46
Corn stover	1.35
Bagasse (this study)	1.4

4. Effect of Temperature and Model Validation

The effect of temperature on the pyrolysis product yield was tested in the range 300-700°C. As stated by Koufopanos et al. [29], primary reactions for wood pyrolysis commence at approximately 210°C, whereas the secondary slightly exothermic reactions [30], in which active pyrolysis occurs, commence at about 300°C. Hence at low temperatures, char formation is favoured while increased temperatures favour tar formation, which reaches a maximum at 525°C. Gas yield increases almost linearly with temperature, though at a low rate. Figure 4 provides a visual representation of the pyrolysis temperature effect. Also, the proposed model was validated by comparing the product yields (exit mass concentrations) with respective experimental data [23] over a range of temperatures (different from the temperature at which fitting was applied). As depicted in Figure 4, the proposed pyrolysis model predicts relatively well the exit concentrations of pyrolysis products.

Hence at this point it is possible to define the yields of the pyrolysis products as a function of temperature via regression analysis. These are:

$$y_{Char} = -2\times10^{-6}T^3 + 0.0043T^2 - 2.462T + 461.8 \qquad (24)$$

$$y_{bio-oil} = 2\times10^{-6}T^3 + 0.004T^2 + 2.335T - 381.4 \qquad (25)$$

$$y_{gas} = 0.0311T - 4.804 \qquad (26)$$

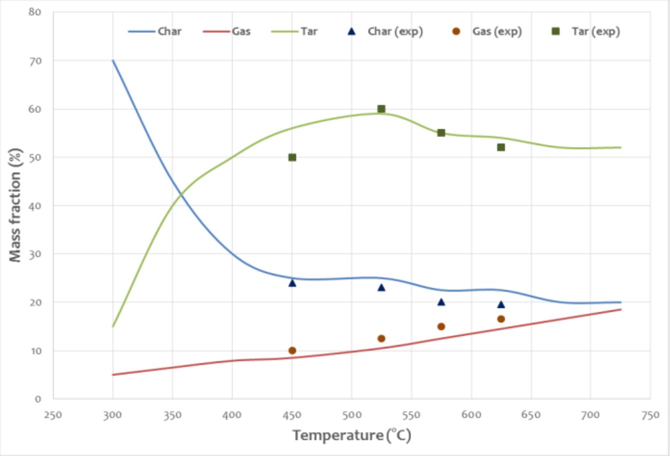


Figure 4. Effect of temperature on the yield of pyrolysis products for the system investigated in the present study

5. Economics

In this section, the cost related to the conversion of the prepared bagasse feedstock to a pyrolysis liquid in a fast pyrolysis reactor is calculated. The starting point of the module is the entry of the prepared feedstock into the reactor feeding system. The end point of the module is the storage of the pyrolysis liquid product. A flowrate of 50 t/h on a dry basis ($m_{bagasse,dry}$) was considered as basis for the calculations. The following correlations were used to calculate the capital costs of the reactor and bio-oil storage units [31]:

$$Reactor\ Cost\ (k€) = 40.8 \times (m_{bagasse,dry} \times 1000)^{0.6194} \qquad (27)$$

$$Storage\ Cost\ (k€) = 119 \times (m_{bagasse,dry})^{0.4045} \qquad (28)$$

A scaling factor (exponent) is applied to estimate the cost of a system based on the known cost of the system for a different size, as shown above. This relationship assumes that the equipment or unit operations can be scaled up/down. The estimated costs are M€ 33 and M€ 0.58 for the fast pyrolysis reactor and the storage units respectively. Labour requirements are estimated from Eq. (29) which is derived from the estimates given by Beckman and Graham [32] for the operation of a fast pyrolysis plant and are equal to M€ 0.67.

$$Labour\ Cost\ (k€) = 104 \times (m_{bagasse,dry})^{0.475} \qquad (29)$$

The economic analysis, presented in this section, is the first step of a discounted cash flow analysis of an entire fast pyrolysis plant. This will be a matter of further study by the authors.

6. Learning Effect

The cost of emerging technologies, such as biomass fast pyrolysis, is normally very high in the beginning of the development stage. The cost of the technologies drops as more plants are constructed and productivity increases because of more experience gained by organisations. This effect is known as a learning curve or experience curve or progress curve or learning by doing effect. The rates of learning vary across different organizations. The factors influencing the rate of learning are crucial and particular attention has to be paid to enhance the performance and economics of the plant. A few factors have been identified as follows:

- Organization forgetting
- Employee turnover
- Transfer of knowledge
- Failure to control other factors such as economies of scale

The trend of the learning curve can be described by Eq. (30).

$$y = ax^{-b} \tag{30}$$

Where y is the capital cost of the xth unit, a is the capital cost of the first unit, x is the cumulative number of units and b is a parameter (calculated as b = log (progress ratio)/log 2 and Learning rate = 1 − (Progress ratio) [33]). For a learning rate of 20%, the value of b is 0.33. Figure 5 illustrates the learning effect on the capital cost of a pyrolysis plant (comprising only reactor and storage units). It can be observed that capital costs can be reduced drastically (~50%) for the 12th plant.

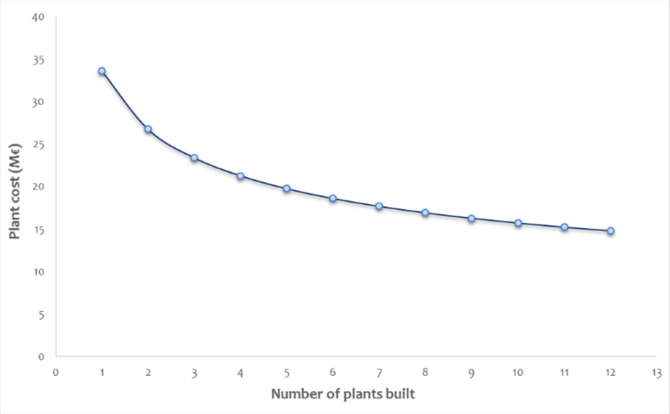


Figure 5. Learning curve effect for a fast pyrolysis unit

7. Conclusions

A detailed design of a fast pyrolysis reactor system for bagasse utilisation, in a sustainable manner, was conducted in the present research. For this purpose, a bubbling fluidised bed reactor was considered. Bubbling fluidised beds have the advantages of a well understood technology that is simple in construction and operation, good temperature control and very efficient heat transfer to biomass particles arising from the high solids density. Equally, exploitation of lignocellulosic biomass derived from forestry or agricultural residues, including bagasse, can positively contribute to the renewable production of biofuels and building block chemicals without competing for land. A detailed kinetic model, involving secondary tar cracking reactions, was developed and the kinetic parameters for bagasse fast pyrolysis were calculated. The model was validated by comparing the exit mass fractions of the pyrolysis products with respective experimental data at a wide range of temperatures. As a result, the proposed model can successfully predict the product distribution at different process conditions and can be used in the future as a tool for process development. In view of a thorough process design, the necessary pyrolysis heat input as well as the capital costs of the reactor and the storage unit were, also, estimated. In the near future, a complete process and economic model of a fast pyrolysis plant (including pretreatment, heat and power generation units as well as bio-oil upgrading units) will be developed using as core the bubbling fluidised bed model proposed in the present study. Overall, the study proposed a reactor operating at 525°C, with residence time of 0.5 s and geometry of 6m length and 1m diameter.

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