CARBON CAPTURE ON FLUIDIZED BED COMBUSTION OF BIOMASS AND WASTES FUELS

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Abstrak

Usaha yang dilakukan secara terus menerus dalam rangka mengurangi emisi CO_2 untuk mencegah terjadinya pemanasan global telah mendorong penggunaan energi baru seperti biomasa maupun sampah sebagai bahan bakar alternatif pengganti energi bahan bakar fosil. Akan tetapi, bila bahan bakar biomasa/sampah dimasukkan pada sistem pembakaran fluidized bed, evolusi zat volatil akan terjadi sangat cepat. Ini dikarenakan oleh tingginya laju perpindahan panas oleh material hamparan di dalam ruang bakar sehingga zat volatil hanya berevolusi di sekitar tempat pemasukan bahan bakar. Karena ketidakcukupan oksigen di bagian atas ruang bakar maka pembakaran sempurna sangat sulit terwujud. Formasi hidrokarbon sering terjadi dan diantisipasi akan memunculkan dioksin pada gas buang. Evolusi volatil juga menyebabkan temperatur sangat tinggi di sembarang tempat pada ruang bakar dan kondisi ini akan menyebabkan formasi emisi NOx . Penelitian ini bertujuan untuk menangkap karbon dari evolusi zat volatile yang terjadi pada pembakaran sistem fluidized bed berbahan bakar sampah. Karbon ditangkap dengan menggunakan partikel berpori sehingga terbentuk karbon-deposit. Pada penelitian ini, karbon-deposit dibakar kembali pada suhu pembakaran fluidized bed untuk meningkatkan efisiensi pembakaran. Dispersi horisontal karbon-deposit di atas hamparan material diukur berdasarkan besarnya laju pembakaran, skala horisontal ruang bakar dan dispersi horizontal karbon yang terbakar. Dengan memperhitungkan tingkat oksidasi dan konsentrasi oksigen pada pembakaran karbon-deposit, model perhitungan yang dimodifikasi dari model Kunni-Levenspiel dikembangkan untuk menggambarkan dispersi horizontal karbon-deposit tersebut. Hasil eksperimen dapat dibuktikan sama dengan model perhitungan yang dikembangkan.

Kata kunci*: biomasa dan sampah, fluidized bed, karbon-deposit, dispersi horizontal*

1. Introduction

The increased efforts of the need to reduce CO_2 emission to prevent global warming from combustion systems have led to an interest in biomass and wastes as fuel sources. As a potentially energy renewable resource, biomass and wastes are gaining more attention worldwide. The use of these fuels to provide partial substitution of fossil fuels has an additional importance to be CO_2 neutral. This is particularly the case with regard to energy plants, which are periodically planted and harvested. During their growth, these plants have consumed CO_2 from the atmosphere for photosynthesis which is released again during combustion. An additional benefit from using wastes as fuels is that the volume of waste is reduced, consequently the need for large landfill areas can be minimized (Sandelin *et al.*, 2001).

Unlike fossil fuels, some physical and chemical properties of biomass and wastes fuels complicate their processing and combustion. These properties may include moisture content, low bulk density, low melting point of the ash and high content of volatile matter. In general, moisture content of biomass is higher than coal fuel, however, it can be lower since the crop products like rice, coconuts, groundnuts must be dried before the husks removed. The bulk density of biomass is low that causes a complication of their processing, transportation and combustion (Werther *et al.*, 2000). In addition, biomass is relatively rich in alkali and alkaline earth metal, causing it to melt at relatively low bed temperatures which in the worst case may result in total defluidization and unscheduled shutdown .

Biomass and wastes are characterized to be high volatile matter (VM) content fuels. The amount of VM which contains in biomass and wastes fuels might range from 60% to 80% and more, compared with a typical figure of 20–30% of medium rank coals. The most of the energy is associated with its volatile content rather than with the solid residue i.e., char. Water vapor is one of the first volatile components to be produced at just over 100 °C, when a solid fuel is heated gradually up to, e.g., 900 °C. Subsequently, H2, CO, and CO_2 are produced, together with a wide suite of hydrocarbons, ranging from CH_4 up to tars. Occasionally, soot is generated during devolatilisation; the elements N and S can appear as NH_3, HCN, CH_3CN, H_2S, COS, and CS_2 (Chern and Hayhurst, 2004).

Fluidized bed combustors (FBCs) using biomass and wastes as fuels for power generation are recognized to have benefits such as the recovery of large amounts of energy remaining in fuels, reducing waste volume, etc. One feature of biomass and wastes as fuels is their high VM content. Because of the high heat transfer from the bed material to fuel, VM evolution occurs very rapidly when fuel is fed into the bed. Consequently, local VM evolution takes place only in the vicinity of the fuel feed point. Complete combustion does not readily occur if the mixing of gas in the upper freeboard is insufficient in fluidized bed combustors.

Numerous investigations have been conducted to minimize the harmful of emission and dioxins caused by the high rate of VM evolution. Reducing of bed temperature from the optimum temperature normally used at 1123 K and by minimizing gas superficial into the bed lead to lower a high reaction during combustion. Controlling the waste feed rate and introducing baffles in the freeboard enable to enhance the mixing of air to VM. Another possible measure to suppress the rapid VM evolution is the use of a porous bed material that captures hydrocarbons in the pores as a carbon deposit (Franke *et al.*., 2001, Shimizu *et. al.*, 2001, 2003, Namioka *et.al.* 2003, Ito *et al.*, 2003, Winaya *et al.*, 2006, 2007). The approach in this study is to capture volatile matter by using porous bed material as an adsorbent and to utilize the carbon deposit for regeneration or conversion process as illustrated in Figure 1. The carbon-deposited solid is burned in the bed for carbon removal process under combustion/gasification temperature of bubbling fluidized bed. As the carbon releases, the solid is regenerated in the system.0The horizontal concentration of carbon is observed. The method of horizontal solid dispersion evaluation based on the horizontal measurements of CO_2 which is formed by the reaction of carbon in the tracers with O_2 in the fluidizing gas.

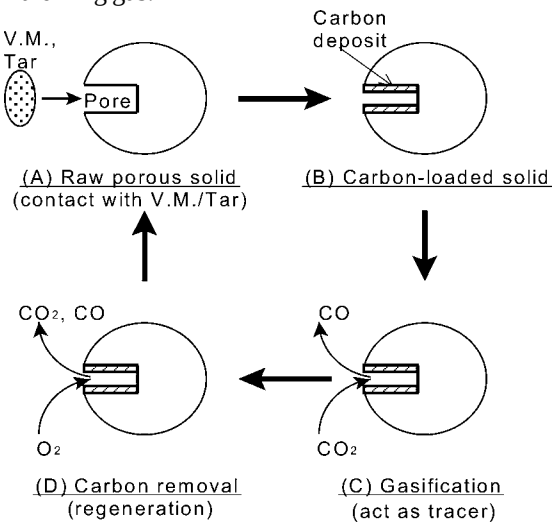

Figure 1. Capacitance effect of porous particle

2. Experimental works

The laboratory-scale two-dimensional bubbling fluidized bed reactor used in this work is presented in the Figure 2. The reactor has a cross section of 0.16 m × 0.04 m and height above the distributor to the top of 0.71 m. The bed material was porous alumina MS1B (particle size, 0.4 mm), whose properties were described elsewhere. The static bed height was 0.10 m. The bed temperatures were operated at 873 K and 943 K. During the combustion process, the gas was mixed between N_2 and O_2 at feed rate of 0.13 m/s. The carbon-deposited solid was continuously fed from the screw feeder at L = 0.12 m from left wall and 0.15 m above the distributor plate. The feed speed of solids was fixed at 180 rpm. The desired concentration of O_2 for combustion was designed at 5%, 10%, 15% and 21% and it was controlled in the flue gas. When the carbon-deposited solids are exposed to O_2 at high temperatures, the carbon forms CO_2 as:

$$C + O_2 \longrightarrow CO_2. \quad (1)$$

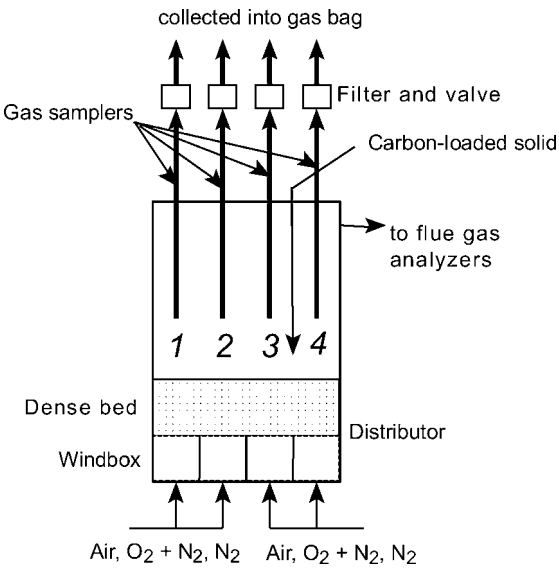

Figure 2. Experimental apparatus of carbon-loaded solids combustion

Assuming that the CO_2 emission rate from the carbon deposit is proportional to the amount of carbon in the solids, the amount of carbon in the bed material can be determined by measuring the CO_2 concentration in the flue gas sampled. Four gas-sampling tubes were installed in the freeboard at distances from the left wall of 0.02 m, 0.06 m, 0.10 m and 0.14 m. The tube inlet height was 0.20 m. The gas samples from the sampling tubes and the flue gas from the top reactor were collected in a gas bag and then analyzed using a gas chromatography.

3. Results and Discussion

3.1 Mathematical Modeling

Since the combustion in fluidized bed is rather complicated, the model proposed is based on the carbon material balance gained from carbon feed rate, carbon combustion rate and horizontal dispersion of carbon deposit. The carbon feed rate was determined based on the amount of carbon-deposited solid and percentage of carbon capture fed into the bed. The model of calculations has been developed by modifying K-L model for bubbling fluidized bed is employed to the conditions and characteristics of the experiments (Kunii and Levenspiel,1991). The fluidized bed consisted of two phases, bubble and emulsion phase. Since carbon capture is a first order reaction, thus the change in concentration in the bubble ($C_{b)}$ and that in the emulsion (C_e) along bed height is determined as follows:

$$-\delta u_b^*(dC_b/d_z) = \delta K_{be}(C_b - C_e) + \delta\gamma_b(\rho_s/M)(dX/dt) \quad (2)$$

and,

$$-(1-\delta)u_{mf}(dC_e/d_z) = -\delta K_{be}(C_b - C_e) + (1-\delta)(1-\varepsilon_{mf})(\rho_s/M)(dX/dt) \quad (3)$$

The boundary condition of equations (eq2) and (eq3) is given as:

$$C_b = C_e = C_{in} \text{ at } z = 0 \quad (4)$$

C_{in} is the inlet concentration of O_2 dependent on operation condition such as O_2 feed rate, temperature and velocity. Other parameters in equations (3) and (4) are described below; K_{be} parameter is the gas interchange coefficient between phase given as correlation bellows:

$$K_{be} = 4.5(u_{mf}/d_b) \quad (5)$$

where u_{mf} and d_b are minimum fluidizing velocity and bubble diameter, respectively. Bubble diameter size of 2 cm was adopted. The rise velocity of the bubble gas (u_b*) was given by the bubble rise velocity ($U_{b)}$, the superficial gas velocity (U) and u_{mf}, as follows:

$$u_b^* = u_b + 3u_{mf} \quad (6)$$

$$u_b = U - u_{mf} + 0.711(g.d_b)^{0.5} \quad (7)$$

where g is the gravitational force and d is the bubble fraction which is estimated as:

$$\delta = (U - u_{mf})/(u_b + 2u_{mf}) \quad (8)$$

e_{mf} and g_b are void fractions in the bed at minimum fluidizing velocity condition and volume of solids per unit volume of bubble, respectively. They were assumed to be $_b$ = 0 and e_{mf} = 0.45 respectively. The amount of carbon concentration (r_s/M) for each cell is calculated in kmol/m^3.

The amount of carbon deposit was burned in the bed using a small scale bubbling fluidized bed reactor of 5.3 cm in inner diameter (Nonaka *et al.*, 2005). Effect of oxygen concentration on carbon burn-up rate was investigated. The same porous alumina bed material was packed in the reactor with a static bed height of 0.1 m. Initially, N_2 gas was fed

from the bottom of the bed. After devolatilization, the bed was fluidized by feeding mixing of N_2-O_2 gas to burn carbon deposit and the concentration of CO_2 in the flue gas was measured.

The change in the combustion reaction is proportional to the carbon conversion rate (*1-X*). The conversion rate (*dX/dt*) is calculated as:

$$dX / dt = k(1 - X) \qquad (9)$$

X is the fractional conversion and parameter *k* is reaction rate constant which is determined as:

$$k = k_0 C^n \qquad (10)$$

While *C* is average O_2 concentration, parameter k_0 and *n* were observed from experiments as seen in Figure 3. Comparison for different operating condition of 873 K and 943 K resulted in the different value k_0 and *n*.

burning smaller amount of PE pellet of 0.15 g with higher velocity of 0.21 m/s. The change of the O_2 concentration along the bed height is found small. The concentration difference between C_b and C_e was also found small. Therefore, the concentration that had been obtained from the condition by burning PE pellet at amount of 0.15g and gas velocity of 0.21 m/s was adopted for the model.

3.2 Horizontal concentration profile of carbon

For horizontal dispersion of carbon deposit rate, the bed is divided within cells. The local horizontal dispersion rate (*N*) in each cell can be calculated as follows:

$$N_{i\to i+1} = D_h A\left(\frac{C_{i-1} - C_i}{\Delta x} - \frac{C_i - C_{i-1}}{\Delta x}\right) \text{ with } i = 1,\ldots,J \qquad (11)$$

where D_h is the horizontal dispersion coefficient [m^2/s], *A* is cell area [m^2], ³%x is horizontal length in each

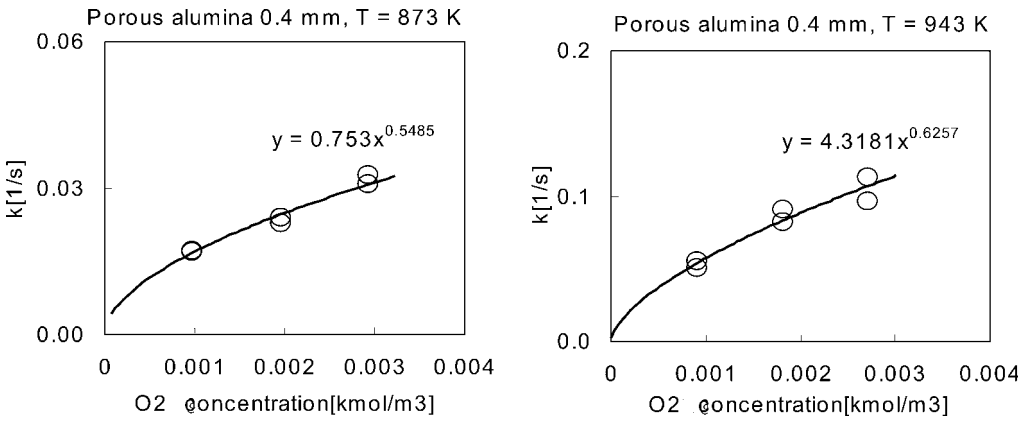

Figure 3. Effect of operating condition and oxygen concentration on k for combustion of carbon deposit over porous alumina at temperature of 873 K and 943 K.

Based on the reaction kinetics for combustion of carbon deposit of Figure 3, and using fluidized bed model, the concentration of O_2 in the bed height is calculated. Figure 4 shows the change in the average O_2 concentration along the bed height. When burning polyethylene (PE) pellet of 0.65g with a superficial gas velocity of 0.13 m/s, the change of the O_2 concentration along the bed height was high and the concentration different between C_b and C_e was also noted high. Comparison was made by

section [m] and *C* is carbon concentration [kmol/m^3-solid]. D_h is 0.0003 m^2/s which is calculated from Borodulya's model (Borodulya *et al.*, 1982) as equation 12.

$$D_h = 0.013(U - u_{mf})H\left(\frac{d_c}{H}\right)^{0.5} Fr^{-0.15} \qquad (12)$$

Figure 5 shows horizontal concentration profile of carbon at upper surface of bed. In this figure, the carbon concentration is normalized by the carbon

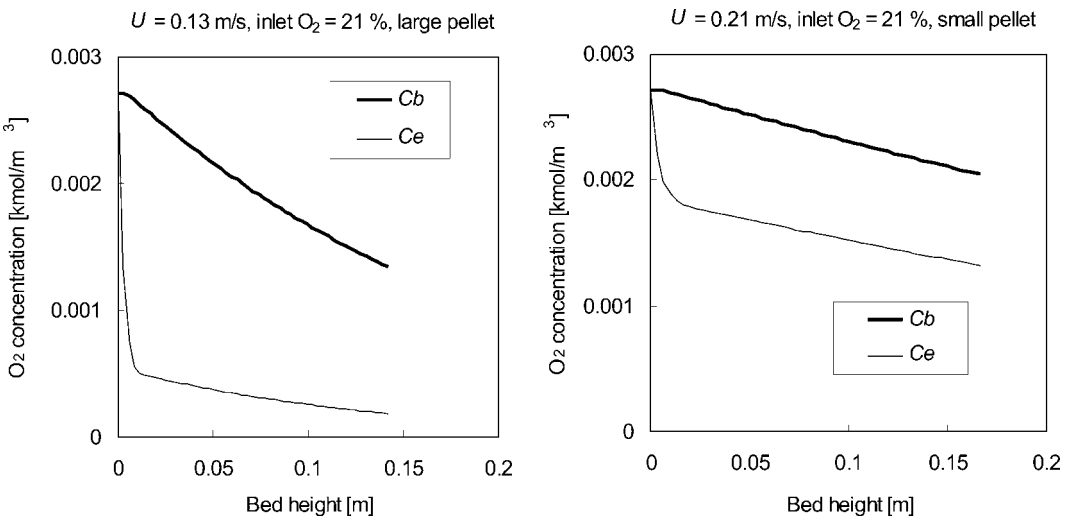

Figure 4. Calculated concentration profile along bed height (1-dimensional model)

concentration measured from the flue gas. The experimental results are compared with 2-dimension model. For model calculations of bed temperature of 943 K, PE feed rate was designed at 0.005 g/min with carbon capture efficiency of 30%. For bed temperature of 873 K, PE feed rate was designed at 0.1 g/min with carbon capture efficiency of 30%. The experiments conducted at lower bed temperature of 873 K have more effective to capture carbon. In addition, the lower bed operation gave a better performance on horizontal dispersion. As can been seen in this Figure, there is satisfactory agreement with the numerical model.

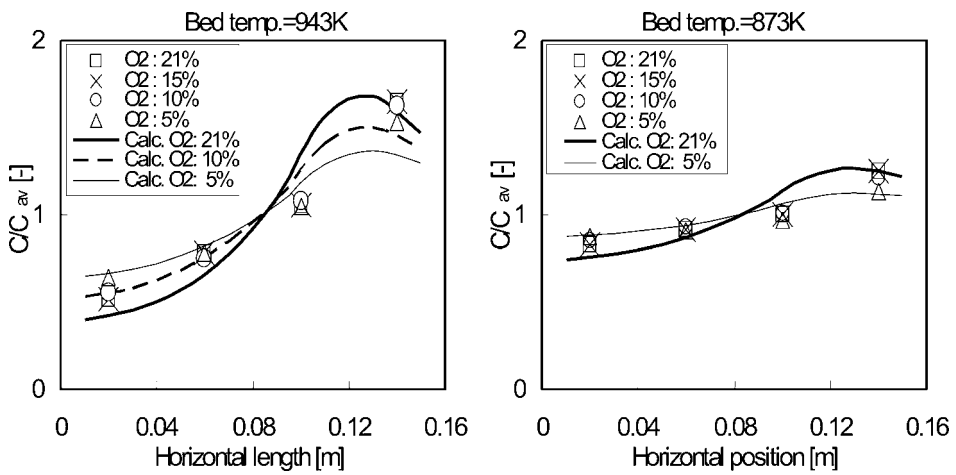

Figure 5. Horizontal concentration profile of carbon at upper surface of bed. (Comparison of experiment results and 2-dimensional model)

4. Conclusions

The fluidized bed combustion to improve conversion process of biomass-wastes fuels was conducted using carbon-deposited solids. Carbon was captured by contacting porous bed material alumina MS1-B during combustion process. In order to avoid undesirable temperature and to increase the conversion efficiency of the system, a good soling mixing in the bed is required. The evaluation was made by measuring horizontal dispersion of solids mixing in the freeboard and a good result was observed. The model of calculations has been developed by modifying K-L model and combustion rate. Horizontal dispersion value was calculated well from Borodulya's correlation. The experimental work has a good agreement with the model.

Nomenclature

A cell area, m^2
C carbon concentration, $kmol/m^3$
C_b carbon concentration in the bubble phase, $kmol/m^3$
C_e carbon concentration in the emulsion phase, $kmol/m^3$
C_{in} carbon concentration in the inlet, $kmol/m^3$
D_h horizontal dispersion coefficient, m^2/s
d_b bubble diameter, m
d_c vessel diameter, m
F_r Froude number $(U-U_{mf})^2/gH_{mf}$, -
g gravitational acceleration, m/s^2
H bed height, m
H_{mf} minimum fluidization bed height, m
K_{be} the gas interchange coefficient between phase, 1/s
k the reaction rate constant, m^3 gas/m^3 solid
k_0 the initial reaction rate constant, m^3 gas/m^3 solid
M the initial carbon mass, kg/kmol
N horizontal dispersion rate
n parameter in eq.8
J number of cell, -
t time, s
U superficial gas velocity, m/s
u_b bubble rising gas velocity, m/s
u_b^* bubble rising velocity, m/s
u_{mf} minimum fluidizing velocity, m/s
X the fractional conversion rate, -
x width of cell, m

Greek symbols

α parameter in eq.9, value= 0.77
δ bubble fraction in dense bed, -
ε_{mf} vioid fraction under minimum fluidization condition,
γ_b volume of solids per unit volume of bubble, -
ρ_s solid density, kg/m^3

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