A novel biomass air gasification process for producing tar-free higher heating value fuel gas

Yan Cao, Yang Wang, John T. Riley, Wei-Ping Pan

Abstract

Biomass is a promising sustainable energy source. A tar-free fuel gas can be obtained in a properly designed biomass gasification process. In the current study, a tar-free biomass gasification process by air was proposed. This concept was demonstrated on a lab-scale fluidized bed using sawdust under autothermic conditions. This lab-scale model gasifier combined two individual regions of pyrolysis, gasification, and combustion of biomass in one reactor, in which the primary air stream and the biomass feedstock were introduced into the gasifier from the bottom and the top of the gasifier respectively to prevent the biomass pyrolysis product from burning out. The biomass was initially pyrolyzed and the produced char was partially gasified in the upper reduction region of the reactor, and further, char residue was combusted at the bottom region of the reactor in an oxidization atmosphere. An assisting fuel gas and second air were injected into the upper region of the reactor to maintain elevated temperature. The tar in the flue gas entered the upper region of the reactor and was decomposed under the elevated temperature and certain residence time. This study indicated that under the optimum operating conditions, a fuel gas could be produced with a production rate of about 3.0 Nm³/kg biomass and heating value of about 5000 kJ/Nm³. The concentration of hydrogen, carbon monoxide and methane in the fuel gas produced were 9.27%, 9.25% and 4.21%, respectively. The tar formation could be efficiently controlled below 10 mg/Nm³. The system carbon conversion and cold gasification efficiency reached above 87.1% and 56.9%, respectively. In addition, the investigation of energy balance for the scale-up of the proposed biomass gasification process showed that the heat loss could be recovered by approximately 23% of total energy input. Thus, partial fuel gas that was produced could be re-circulated and used to meet need of energy input to maintain the elevated temperature at the upper region of reactor for tar decomposition. It was predicted the heating value of product fuel gas would be 8000 kJ/Nm³ if the system was scaled up.

Keywords: Biomass; Gasification; Tar free

1. Introduction

Managing renewable energy such as biomass in an effective way is a challenge for the energy industry [1]. Transplant of commercial coal or oil-based gasification process to biomass utilization is an optimal choice [2–4]. The biomass gasification technologies to produce hydrogen-rich fuel gas present highly interesting possibilities for biomass utilization as the sustainable energy [3]. Biomass used as energy source can reduce CO₂ greenhouse effect and SO₂ and NOₓ atmospheric pollution [2] due to its characterizations of neutral carbon and less sulfur and nitrogen contents. Moreover, utilization of the gasification process can produce fuel gas with only 1/3 of the volume compared to that of the combustion process, which can decrease the burden of the downstream Air Pollutants Control Devices (APCD) [1,4].

The extensive studies on biomass gasification process have been conducted to investigate how to utilize this potential energy source with lower energy density, higher volatile content and higher tar production [5–12]. The gasification agents include air, oxygen-rich air, oxygen, steam and a mixture of air and steam. The formation of tar is one of the major problems to deal with during biomass gasification process [13]. Generally, it is reported that the average tar production in an air–steam gasification process was approximately 20 g/Nm³ [1,2]. Tar is a complex mixture of condensable hydrocarbons, which includes single ring to 5-ring aromatic compounds along with other oxygen-containing hydrocarbons and complex PAH. In the EU/IEA/US-DOE
meeting, it was agreed by a number of experts to define tar as all organic contaminants with a molecular weight larger than benzene \[14\]. Tar will result in: 1) the shut-down of meeting, it was agreed by a number of experts to define tar as mentioned 10 mg/Nm\(^3\) limitation for engine \[15\], and Milne [Milne] minimum allowable limit for tar is highly dependent on the lower heating value of the fuel gas produced. However, the minimum allowable limit for tar is highly dependent on the kind of process and the end user application. Bui et al. mentioned 10 mg/Nm\(^3\) limitation for engine \[15\], and Milne and Evans tabulated the tar tolerance limit for various end use devices, which was suggested by different researchers \[16\].

Control technologies of tar production can broadly be divided into two approaches; treatments inside the gasifier (primary methods) and hot gas cleaning after the gasifier (secondary methods). Although secondary methods are proven to be effective, treatments inside the gasifier are gaining much attention due to economic benefits \[17\]. In primary methods, the operating parameters such as temperature, gasifying agent, equivalence ratio, residence time and catalytic additives play important roles in the formation and decomposition of tar. Primary methods are not yet fully understood and have not to be implemented commercially. However, the utilization of some catalysts in the gasifier \[17\] and the concepts of two-stage gasification are of prime importance \[18\].

Petroleum refinery catalysts could be used to effectively eliminate tar in biomass gasification process \[17–19\]. Delgado and Aznar \[20\] explored the effectiveness and life-span of calcined dolomite, magnesite, and calcite for tar decomposition in a fluidized bed biomass gasifier with steam as a gasification agent. Commercial steam reforming catalysts were used to improve biomass gasification, and steam–oxygen mixtures were also included \[21,22\]. Courson et al. \[23\] developed the nickel-based catalysts to study its effects on reforming of a produced gas from biomass. Wu et al. \[24,25\] developed a pilot-scale circulating fluidized gasification (CFG) facility to investigate the optimal operation parameters of wood gasification process. However, it was reported that tar could not be decomposed completely with calcium oxide as a catalyst that resulted in a limitation of the utilization of producing fuel gas in gas turbine. Problems regarding catalyst deactivation, increase of system pressure drop, and carryover of fines were severe in many catalyst utilizations for tar decomposition \[18\].

Reactor design is very crucial for gasification in terms of efficiency, heating value of the product gas, and also for tar formation. Secondary air injection to the gasifier resulted in a significant tar reduction due to an elevated temperature maintained at freeboard region. Pan et al. \[26\] reported 88.7 wt.% of tar reduction by injecting secondary air just above the biomass feeding point in the fluidized bed at temperature of 840–880 °C. Narv et al. \[27\] performed secondary air injection in the freeboard of fluidized bed gasifier and observed a temperature rise of about 70 °C which resulted in a tar reduction from 28 to 16 g/Nm\(^3\). The Asian Institute of Technology (AIT), Thailand, modified a biomass gasifier and resulted in a fuel gas with tar production of about 50 mg/Nm\(^3\), which is about 40 times less than a single-stage reactor under similar operating conditions \[16\]. This concept involves a downdraft gasifier with two levels of air intakes. The produced tar in the biomass pyrolysis process will pass through a high-temperature residue char bed at the bottom and will be decomposed at the elevated temperature. Bhattacharya et al. \[28\] reported a similar gasifier with char produced inside the gasifier itself to act as a filter to further reduce tar production considerably at 19 mg/Nm\(^3\) in fuel gas.

In the current study, a novel tar-free gasification process using air as gasification agent was proposed. The concept was based on a properly designed elevated temperature region in the biomass gasifier where the tar produced in biomass pyrolysis process was decomposed by injection assisting fuel gas and the secondary air. The optimal operation parameters and the factors affecting LHV (low heating value) of fuel gas in a higher cold gas efficiency and carbon conversion efficiency were investigated in a lab-scale testing facility. In addition, the material and the heat balance were given to demonstrate the improvement of LHV of fuel gas in the scale-up of industrial units.

2. Conceptive design of two-region biomass air gasifier

The high volatile content of biomass results in rapid devolatization without much heat required, as shown in Eq. (1). The energy required for the pyrolysis step can be provided through combustion of partial residual fixed carbon in an air atmosphere, as shown in Eqs. (2) and (3) which are both exothermic reactions. The water shift reaction, as shown in Eq. (4), always occurred in any gasification process due to the presence of water vapor. Water vapor and carbon dioxide promotes hydrogen production in biomass gasification process, through Eq. (6) by water vapor and through the combination of Eq. (7) by carbon dioxide and Eq. (4) by water vapor. However, the steam and carbon dioxide gasification processes are both intensive endothermic reactions which will result in temperature drop in the gasifier, and then further promote the formation of tar. Un-reacted steam fed into the system will also result in a low heating value of fuel gas produced due to heat loss in the steam heating process. Thus, air is selected as the only gasification agent in this conceptive design for biomass gasification.

\[
\begin{align*}
C_5H_{11}O_p & \rightarrow aCO_2 + bH_2O + cCH_4 + dCO + eH_2 \quad \text{(1)} \\
+ f(C_2 - C_5), & \quad 20.9 \text{kJ/mol} \\
C + O_2 & \rightarrow CO_2, \quad -408.4 \text{kJ/mol} \quad \text{(2)} \\
C + 1/2O_2 & \rightarrow CO, \quad -122.9 \text{kJ/mol} \quad \text{(3)} \\
CO + H_2O & \rightarrow CO_2 + H_2, \quad -42.2 \text{kJ/mol} \quad \text{(4)} \\
H_2 + CO + O_2 & \rightarrow H_2O + CO_2, \quad -524.3 \text{kJ/mol} \quad \text{(5)} \\
C + H_2O & \rightarrow CO + H_2, \quad 118.3 \text{kJ/mol} \quad \text{(6)} \\
C + CO_2 & \rightarrow 2CO, \quad 160.5 \text{kJ/mol} \quad \text{(7)}
\end{align*}
\]
In all reactions above, the value of the reaction enthalpy refers to the temperature of 298.15 K \(^{[29]}\). As indicated by previous studies, 90% of the volatile content in the total weight of biomass will be released instantly if it was heated above 600 °C. If biomass is fed into a low part of the gasifier, the devolatilization compounds such as hydrogen, carbon monoxide, methane, light hydrocarbon and tar will meet primary air to burn out. Loss of these majorities of the combustible contents results in a decrease of the LHV of fuel gas produced. In the conceptive design, the biomass is fed into the gasifier from the top of the reactor so that volatile compounds will be released into an oxygen-lean or reducing environment to prevent from burning out.

The conceptive configuration of the gasification reactor is illustrated in Fig. 1. It is a combination of a sand fluidized bed biomass gasifier at the bottom region and a separated tar decomposer at the freeboard region with the partial re-circulation of fuel gas to maintain an elevated temperature. The secondary air is injected at the amount for the complete combustion of this re-circulating fuel gas. Biomass is fed into reactor from top and it will pass through freeboard regions of the reactor to conduct its drying and pyrolysis processes. The pyrolytic gas with large molecular constituents such as the tar merged with fuel gas upward and passed through freeboard region of the reactor where tar is decomposed under the elevated temperature conditions and a certain residence time. The derived biomass residue char will be partially combusted and partial gasified with injection of the primary dry air in the bottom of reactor by using sand as bed materials. The partial combustion of residue char provides part of heat required for the endothermic biomass pyrolysis and partial gasification process.

Besides tar decomposition, strategies for improvement of LHV of produced fuel gas is considered in the conceptive design. The goal is to produce the self-combusted fuel gas. Generally, the fuel gas with heating value of above 5000 kJ/Nm\(^2\) can meet this requirement. The current effects are control of temperature profile in reactor to promote production of methane and light hydrocarbons which the heating value is higher than those of hydrogen and carbon dioxide. The application of the produced fuel gas is suitable for engine and turbine use. Due to using air as a gasification agent, a large percentage of nitrogen gas is introduced into system which will dilute combustible constituents in the fuel gas. In the future, it may be possible for biomass to be gasified in rich or pure oxygen atmospheres.

3. Experimental setup

3.1. Materials and their characterizations

A mixed wood sawdust from a timber mill was used as the feedstock after it was cut into pieces the average size of 2 to 3 mm. Five mixed wood sawdust samples were taken randomly and tested by proximate and elemental analysis using the standard method described in ASTM-84 (American Society for Testing Materials-84). The average proximate and element analysis of the mixed wood sawdust were shown in Table 1. The volatile content of sawdust is as high as 70 wt.%, and its fixed carbon content is 17.9 wt.%. The oxygen content of sawdust is as high as 35.4 wt.% and thus its heating value was as low as 18810 kJ/kg compared with coal fuel. Air was used as the gasification agent in all tests.

3.2. Test facilities and procedures

Tests were performed in an atmospheric pressure, and directly heated (autothermic, no outside electrical heating system), sand fluidized-bed gasification system with air as gasification agent. The schematic lab-scale configuration was illustrated in Fig. 2. Different from the large-scale facility, a lab-scale facility always meets a larger heat loss. Thus, an additional nature gas was used as an assisting gas to provide heat at the freeboard region and thus maintaining an elevated temperature in this study. Sand was used as a bed material to maintain good heat transfer and fluidization in the reactor.

<table>
<thead>
<tr>
<th>Table 1</th>
<th>The proximate and elemental analysis of wood sawdust used in this study</th>
</tr>
</thead>
<tbody>
<tr>
<td>Proximate analysis (wt.%)</td>
<td>Moisture Volatile matter Ash Fixed carbon Heating value (kJ/kg)</td>
</tr>
<tr>
<td>10.4</td>
<td>70.4</td>
</tr>
<tr>
<td>Elemental analysis (wt.%)</td>
<td>C H O N S</td>
</tr>
<tr>
<td>46.2</td>
<td>5.1</td>
</tr>
</tbody>
</table>
The sand originates from nearby architecture market in Taiyuan City of China for production of a cement mixture. It is used after sieving to an average size of 0.11 mm. The density of sand material is approximately 1470 kg/m³. It is approximately 94.7% silica oxide, 1.7% iron oxide and 2.4% aluminum oxides with the balance of other trace amounts of oxides. The predicted density of the biomass char is 450 kg/m³. The minimum fluidization velocity at atmospheric pressure for sand material was measured to be 0.041 m/s. The primary air in all cases was controlled to achieve a superficial velocity about 2 times of the minimum fluidization velocity required for sand material. At this operating velocity, the fluidization condition for biomass char was about 6–8 times of its minimum fluidization velocity. The reactor was made of 1Cr18Ni9Ti stainless steel pipe and covered with an insulation layer outside. The effective height of the reactor was 1200 mm with an OD of 100 mm. A rotating disc feeding system continuously fed biomass into the bed through a pipe which was connected at the top of the reactor. The gasification agent was provided by an air compressor and divided into three intakes, which were controlled separately by gas flow meters. A primary air was fed into the reactor through a cone-shaped distributor having 100 holes (ID=1 mm, the opening ratio of 1.0%). The cone-shaped distributor was covered with fine metal mesh to prevent the sand material from loss at the gas distributor. A secondary air (I) was injected into the upper region of the reactor and a secondary air (II) was injected in the middle region of the reactor only in the starting period. On two sides of the freeboard region, four nozzles, with two on each side of the reactor, were installed for assisting nature gas injection. Three thermocouples were used to monitor temperature profile along the height of the reactor. A cyclone and a metal filter in series were used as dusting units for fuel gas cleaning which were heat-insulated for prevention of tar condensation. A water cooler and an ice trap in series were used as cooling units for fuel gas cooling and tar capture. The fuel gas exited the freeboard of the reactor and entered the following de-dusting and cooling units after the system was brought to a steady state. A bypass vent was used to exhaust fuel gas produced in the period of start-up.

Initially, the fluidized bed reactor was loaded with sand at a ratio of depth to radius of about 1:1. The heat-assisting fuel gas and the secondary air were injected into the middle and upper regions of the reactor to heat the reactor. Meanwhile, the primary air was injected from the bottom distributor to maintain the sand material to be fluidized. When desired temperatures were achieved simultaneously at the top (in the range of 650–850 °C), the middle, and the bottom of the reactor (in the range of 750–950 °C), the rotating disc feeder was turned on to the desired rotation speed to feed biomass material into the reactor. Then, the assisting fuel gas and the secondary air were adjusted to maintain desired temperatures at the upper region, while bottom temperature was controlled by adjusting the feedstock/primary air ratio. The fuel gas produced was bypassed the cooling system in system tuning period. In general, it took 30 min to make the system stable with respect to the reactor temperatures (the temperatures at the upper and bottom regions of gasifier were kept stable). At a steady state, all parameters were kept constant for about 60 min for gas sampling and analysis.
3.3. Fuel gas sampling and measurement procedures

After a steady state was achieved, the flue gas stream was switched to downstream de-dusting and cooling system. After passing through water cooler and an ice trap, the dry, clean, and cool fuel gas was sampled by a gas sampling bag and taken every 10 min. The permanent gas compositions and light hydrocarbon constituents (mainly C2–C5) of fuel gas were analyzed by a gas chromatograph (model GC-2010, SHIMADZU, Japan) with TCD and FID detectors. A total of three columns were used to separate H2, CO, O2, N2, and CH4 from CO2 and C2–C5 hydrocarbons through a Porapak R column by using argon as a carrier gas. The former stream was further separated through a Porapak Q column and the latter through a 5A molecular sieve column. Standard gas mixtures of H2, CO, O2, N2, CO2 and CH4 were used for 3-point quantitative calibration of GC 2010. Generally, about 5 to 7 gas samples were taken for each test condition. The tar product, if any, was trapped in the water cooler and ice trap and rinsed with dichloromethane to remove tar from the water cooler and the ice trap. The rinsing solution was transferred to a ceramic tray and heated in an oven at 50 °C for about 12 h to evaporate any water and rising solution. The collected tar product was weighed and recorded.

4. Results and discussion

The goal of the study was to demonstrate the concept of a biomass air gasification process to produce fuel gas having a higher heating value with less tar formation. The experiments were conducted to investigate the impacts of parameter variation such as temperature profiles in the reactor and the air/biomass ratio on tar formation, gas composition and energy transformation including carbon conversion and cold gas efficiency.

4.1. Temperature profiles verse tar production

Figs. 3, 4 and 5 represented efforts on the reduction of tar production by variation of temperatures at the bottom and top of the reactor. The effect of temperature variation at bottom of the reactor on tar production was shown in Fig. 3. It was observed that increasing temperature at the bottom from 651 °C to 839 °C could reduce tar production.
from 1227 mg/Nm³ to 338 mg/Nm³ while the temperature at the top of reactor was kept almost constant at 750 °C. Although a majority drop of tar production was found by increasing temperature at the bottom of reactor, tar content in the fuel gas was still too high to be accepted by general application. However, tar content in fuel gas dropped down very rapidly as the temperature at the top of the reactor increased even if the temperature at the bottom of the reactor was maintained at a lower level of 650 °C, as shown in Fig. 4. As the temperature at the top of the reactor reached 854 °C, tar content in the fuel gas produced was just 21.24 mg/Nm³. Further increasing the temperature at the top of the reactor resulted in a much lower tar content in fuel gas. Tar contents of 15.98 mg/Nm³ and 12.34 mg/Nm³ were observed at temperatures of 898 °C and 934 °C, respectively. Thus, in following tests, temperatures at the top of reactor were always maintained above 850 °C due to the purpose of the current study. This means the tar production control was successful at level of ignorance in the demonstration configuration of biomass air gasification process. If the temperature at the top of the reactor was maintained above 850 °C, increasing the temperature at the bottom of the reactor did not apparently improve the reduction of the tar production as shown in Fig. 5. The impact of temperature at the top of the reactor was much more remarkable than that at the bottom of the reactor.

In this test, temperatures at the top and bottom of the reactor could be controlled and maintained separately by adjusting assisting fuel gas and biomass feed rate, respectively. The residence time of pyrolytical product in the reactor may be changed due to adjustment of flow rate of biomass and

![Fig. 5. Tar production as functions of temperatures at the top and the bottom of reactor.](image)

<table>
<thead>
<tr>
<th>Table 2</th>
<th>The test conditions and results of the proposed biomass gasification by air</th>
</tr>
</thead>
<tbody>
<tr>
<td>No. 10</td>
<td>No. 15</td>
</tr>
<tr>
<td>Temperature (°C)</td>
<td></td>
</tr>
<tr>
<td>Upper region</td>
<td>860</td>
</tr>
<tr>
<td>Bottom region</td>
<td>750</td>
</tr>
<tr>
<td>Pressure (bar)</td>
<td>1</td>
</tr>
<tr>
<td>Air/wood (Nm³/kg)</td>
<td>2.325</td>
</tr>
<tr>
<td>Gas production per kilogram of wood (Nm³/kg)</td>
<td>2.989</td>
</tr>
<tr>
<td>Assisting gas/wood (Nm³/kg)</td>
<td>0.475</td>
</tr>
<tr>
<td>Carbon conversion of system (%)</td>
<td>87.1</td>
</tr>
<tr>
<td>Cold gas efficiency of system (%)</td>
<td>56.9</td>
</tr>
<tr>
<td>Heating value of gas produced (kJ/Nm³)</td>
<td>4978.4</td>
</tr>
<tr>
<td>Heating value ratio of assisting gas/gas produced</td>
<td>0.494</td>
</tr>
<tr>
<td>Gas compositions (%)</td>
<td></td>
</tr>
<tr>
<td>H₂</td>
<td>9.27</td>
</tr>
<tr>
<td>N₂</td>
<td>62.47</td>
</tr>
<tr>
<td>CO</td>
<td>9.25</td>
</tr>
<tr>
<td>CH₄</td>
<td>4.21</td>
</tr>
<tr>
<td>CO₂</td>
<td>13.28</td>
</tr>
<tr>
<td>C₂ + C₃</td>
<td>1.524</td>
</tr>
<tr>
<td>C₄ + C₅</td>
<td>0.0116</td>
</tr>
<tr>
<td>Total combustible fuel gases (%)</td>
<td>22.73</td>
</tr>
<tr>
<td>Apparent heat loss carried by gas produced (%)</td>
<td>4.5</td>
</tr>
<tr>
<td>Apparent heat loss of steam (%)</td>
<td>7.8</td>
</tr>
<tr>
<td>Chemical energy of unburnt carbon (%)</td>
<td>9.1</td>
</tr>
<tr>
<td>Heat loss of system (%)</td>
<td>22.0</td>
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</table>
assisting fuel gas and thus coupled that of air. A shorter residence time of pyrolytical product always resulted from increasing feeding rate of biomass and assisting fuel gas to maintain an elevated temperature in the proposed reactor. Test results strongly support the kinetic control mechanisms in the tar decomposition process.

4.2. Fuel gas composition, yield and low heating value (LHV)

Two parameters were used to control the operation of the proposed biomass gasifier, the air/biomass ratio and the assisting fuel gas/biomass ratio. The definitions of these two parameters were shown in Eqs. (8) and (9).

\[
\text{air/biomass ratio} = \frac{\text{total air addition into the system (Nm}^3\text{)}}{\text{total biomass fed in the system (kg)}} \quad (8)
\]

\[
\text{assisting fuel gas/biomass ratio} = \frac{\text{total assisting fuel gas addition into the system (Nm}^3\text{)}}{\text{total biomass fed in the system (kg)}} . \quad (9)
\]

In the view of the fuel gas composition as shown in Table 2, at the air/biomass ratio of 2.325 Nm$^3$/kg and an assisting fuel gas/biomass ratio of 0.475 Nm$^3$/kg (test No. 16), the total combustible fuel gas ($\text{H}_2+\text{CO}+\text{CH}_4$) compositions were 22.73% with $\text{H}_2$, CO and $\text{CH}_4$ contents being 9.27%, 9.25% and 4.21%, respectively. The fuel gas production was 2.989 Nm$^3$/kg. As the air/biomass ratio increased to 3.155 Nm$^3$/kg and an assisting fuel gas/biomass ratio of 0.607 Nm$^3$/kg (test No. 8), $\text{H}_2$, CO and $\text{CH}_4$ compositions in fuel gas were 7.73%, 5.52% and 2.45%, respectively, with fuel gas production of 3.588 Nm$^3$/kg. As the air/biomass ratio was 2.070 Nm$^3$/kg and an assisting fuel gas/biomass ratio of 0.423 Nm$^3$/kg (test No. 15), $\text{H}_2$, CO and $\text{CH}_4$ compositions were 11.22%, 9.12% and 4.09%, respectively, with fuel gas production of 2.651 Nm$^3$/kg.

The light hydrocarbon compositions in the fuel gas after the water trap and ice trap are shown in Tables 2 and 3. The accumulative concentrations of light hydrocarbon contents in the fuel gas are shown in Table 2. The nine most abundant gas constituents quantified by GC are shown in Table 3. As the air/biomass ratio was maintained above 2 Nm$^3$/kg but below 2.5 Nm$^3$/kg, and temperatures at the top and bottom of the reactor were maintained at a range of 860–870 $^\circ$C and 730–760 $^\circ$C, respectively, and the total contents of light hydrocarbon species of C$_2$ plus C$_3$ were varied in a range between 1.1415% and 1.5240% and that of C$_4$ plus C$_5$ was varied in a range of 0.0069% and 0.0414%. Due to higher LHVs of methane and light hydrocarbons, the LHV of fuel gas produced was expected to be as high as about 5000 kJ/Nm$^3$. The variation of LHV of fuel gas produced is shown in Fig. 6. The corresponding gas production is shown in Fig. 7. A higher air/biomass ratio coupled with a higher assisting fuel gas/biomass ratio could result in higher top and bottom temperatures in the reactor, as indicated in tests No. 4 and No. 8. However, the quality of the fuel gas deteriorated. In tests No. 4 and No. 8, the total combustible fuel gas composition decreased to about 14.75% and 15.70%, respectively. The LHV of fuel gas produced decreased to about 3000 kJ/Nm$^3$. Thus, maintaining an optimum air/biomass ratio was critical in the designed biomass process.

### Table 3

<table>
<thead>
<tr>
<th>Test no.</th>
<th>C$_2$H$_2$ (vol.%)</th>
<th>C$_2$H$_4$ (vol.%)</th>
<th>C$_2$H$_6$ (vol.%)</th>
<th>C$_3$H$_3$ (vol.%)</th>
<th>C$_3$H$_6$ (vol.%)</th>
<th>C$_3$H$_8$ (vol.%)</th>
<th>C$_4$H$_8$ (vol.%)</th>
<th>C$_4$H$_10$ (vol.%)</th>
<th>C$_5$H$_12$ (vol.%)</th>
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</thead>
<tbody>
<tr>
<td>4</td>
<td>0.3200</td>
<td>0.3500</td>
<td>0.0000</td>
<td>0.0000</td>
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<td>0.0070</td>
<td>0.0000</td>
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<td>8</td>
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<td>0.2900</td>
<td>0.0250</td>
<td>0.0130</td>
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<td>0.0050</td>
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<tr>
<td>10</td>
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<tr>
<td>15</td>
<td>0.0000</td>
<td>1.3100</td>
<td>0.0000</td>
<td>0.0000</td>
<td>0.0080</td>
<td>0.0000</td>
<td>0.0000</td>
<td>0.0069</td>
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<tr>
<td>16</td>
<td>0.5300</td>
<td>0.6000</td>
<td>0.0000</td>
<td>0.0000</td>
<td>0.0046</td>
<td>0.0069</td>
<td>0.0184</td>
<td>0.0000</td>
<td>0.0230</td>
</tr>
</tbody>
</table>

Fig. 6. The variation of low heating value of fuel gas produced with the air/biomass ratio.

Fig. 7. The variation of fuel gas production with the air/biomass ratio.
The total combustible compositions and also individual combustible compositions in the fuel gas reached a maximum at a lower air/biomass ratio, 2.5 Nm$^3$/kg in the current study. Under this condition, the pyrolysis of biomass and decomposition of tar occurred and thus the LHV of fuel gas increased, shown in Table 2. It was also found that the combustion outcome of water and carbon dioxide increased as the air/biomass ratio was greater than 2.5 Nm$^3$/kg. More air input at same biomass feed rate provided more oxygen and a higher probability for fuel gas combustion after fuel gas was produced from the pyrolysis and gasification processes. It seems that the air/biomass ratio at 2.5 Nm$^3$/kg in this study was a transformation point on production and consumption of combustible fuel gas composition. Several processes were included into this transformation, such as biomass pyrolysis, gasification and tar decomposition in one side, and biomass and its pyrolytic product combustion in the other side.

4.3. Carbon conversion and cold gas efficiency

Variation of two parameters of carbon conversion efficiency and cold gas efficiency could be used to investigate the effects of air/biomass ratio on LHV of fuel gas. The definitions of carbon conversion efficiency and cold gasification efficiency were shown in Eqs. (10) and (11), respectively.

\[
\text{carbon conversion efficiency} = \frac{\text{total reacted carbon in the system (kg)}}{\text{total carbon fed in the system (kg)}} \quad (10)
\]

\[
\text{cold gas efficiency} = \frac{\text{LHV of fuel gas (kJ/Nm}^3\text{)} \times \text{fuel gas production (Nm}^3/\text{kg)}}{\text{low heating value of biomass fed in the system (kJ/kg)}}. \quad (11)
\]

The variation of carbon conversion in biomass gasification process versus the air/biomass ratio is shown in Fig. 8. The carbon conversion of this system reached its maximum of 87% at the air/biomass ratio of about 2.5 Nm$^3$/kg. Under conditions where the air/biomass ratio was below about 2.5 Nm$^3$/kg, the carbon conversion of the system increased rapidly with the increase of the air/biomass ratio. However, under conditions where the air/biomass ratio was greater than about 2.5 Nm$^3$/kg, the carbon conversion decreased gradually. The carbon conversion was just 70% when the air/biomass ratio was 4.5 Nm$^3$/kg. After increasing the air input, more fuel gas is produced as shown in Fig. 7. The higher gas velocity in the gasifier would carry the lower density of fly ash with the unburned fixed carbon out of the gasifier so that the unburned carbon did not have enough residence time to further react either with O$_2$ for combustion or with CO$_2$ and H$_2$O for gasification and consequently a decrease in the carbon conversion efficiency was observed. Considering the improvement on carbon conversion efficiency, the circulating fluidized bed technology may be an optimal choice to meet the characterization of biomass air gasification process. The fly ash having higher carbon content could be re-circulated and returned to the gasifier to prolong its residence time. The variation of cold gasification efficiency versus the air/biomass ratio was shown in Fig. 9. As the air/biomass ratio increased, as expected, the cold gas efficiency increased. More air was put into the gasifier to satisfy the needs of the gasification process of the biomass and its product. However, with air being further input, cold gasification efficiency decreased starting at an air/biomass ratio of approximately 2.5 Nm$^3$/kg due to their combustion.

In practice, the chemical energy of biomass can not be transformed completely into that of fuel gas due to unavoidable energy loss by those processes such as cooling of fuel gas and water vapor, system heat loss and chemical energy loss in the unburned carbon of fly ash. As presented in Table 2, under optimum operation conditions with the air/biomass ratio at 2.325 Nm$^3$/kg (test No. 10), the apparent heat loss of fuel gas produced, apparent heat loss of water vapor, the chemical energy of unburned carbon and heat loss of system were 4.5%, 7.8%, 9.1% and 22%, respectively, totally 43.4%. With comparison, in the air/biomass ratio at 2.070 Nm$^3$/kg (test No. 15), the four parameters of heat loss mentioned above were 4.1%, 7.6%, 20.0% and 15.7% respectively, totally 47.4%. At

![Fig. 8. The variation of the carbon conversion with the air/biomass ratio.](image)

![Fig. 9. The variation of the cold gasification efficiency with the air and biomass ratio.](image)
an air/biomass ratio as 3.155 Nm$^3$/kg, the four parameters of heat loss mentioned above were 5.0%, 9.8%, 18.4% and 27.9%, respectively, which was totally 61.1%.

4.4. Process scale-up

The analysis of energy distribution in lab-scale biomass gasification system indicated that totally 43.4% of the chemical energy from the biomass input was lost and could not be transformed into chemical energy of the fuel gas produced. Thus, a nature gas intake was used to maintain an elevated temperature in the current study, and meanwhile to compensate for LHV of fuel gas produced. Keeping in mind that a larger heat loss occurred in the lab-scale reactors such as the chemical energy loss of unburned carbon at 9.1% and heat loss of system at 22% due to the reactor scale and configuration, a modified design in a large-scale gasification facility was conducted. A circulating fluidized bed was proposed as the reactor type. The mass balance and heat distribution under the optimum gasification case is shown in Fig. 10. Due to longer residence time of re-circulating of residue char in the system, the carbon loss would be less than 2% and thus the carbon conversion efficiency could be controlled at level of about 98%. As the gasifier was scaled up, the system heat loss could be controlled at a level of about 5% or less. Assuming that apparent heat loss of fuel gas produced and apparent heat loss of water vapor were just half of that in the lab-scale reactor; the total 14% chemical energy of biomass could not converted into that of fuel gas produced (the apparent heat loss of fuel gas produced, apparent heat loss of water vapor, the chemical energy of unburned carbon and heat loss of system at 3%, 4%, 2% and 5%, respectively). Thus, the cold gas efficiency of about 86% could be achieved in the scaled-up gasifier. Because of the assumption that apparent heat loss of fuel gas and water vapor decreased in half in the scale-up gasifier, the required heat input of the assisting fuel gas and, consequently, air intake will be decreased in half at its amount of that in the small-scale reactor. Therefore, the theoretical total air/biomass ratio should be 1.25 Nm$^3$/kg in the scaled-up gasifier. Reasonably assuming that the primary air/biomass ratio was 0.75 Nm$^3$/kg, the
secondary air/biomass ratio should be 0.5. Actually, the recirculation of fuel gas produced in the scaled-up gasifier would be used as the assisting gas for a self-sustainable purpose. Due to lower LHV of the fuel gas produced with comparison to that of nature gas used in the small reactor, more secondary air was used to maintain a well sustainable flame. Here, the ratio of the secondary air/biomass was selected at 0.5 Nm³/kg. The calculated fuel gas production was found to be about 2.3 Nm³/kg biomass in view of total of air intake and pyrolytical gas produced. Thus, the LHV of fuel gas was about 8000 kJ/kg biomass.

5. Conclusion

In this study the conceptive design of a novel two-region combined reactor for the air gasification of biomass was proposed. This concept was demonstrated in a lab-scale sand fluidized bed for producing tar-free fuel gas with higher heating value. The upper region of the gasifier was maintained at a relatively high temperature to decompose the tar produced from the biomass pyrolysis and gasification process. The biomass was fed into the gasifier from the top to increase its residence time in the gasifier and to prevent its higher heating value constituents from meeting the primary air to be burned out. The residue carbon continued to be gasified and combusted in the bottom region of the reactor. Results of the lab-scale tests showed that the tar formation could be efficiently controlled about 10 mg/Nm³ at optimal conditions in this study. The fuel gas production and its LHV could reach about 3.0 Nm³/kg biomass and about 5000 kJ/Nm³ at the air/biomass ratio and the assisting fuel gas/biomass ratio were maintained at 2.3 Nm³/kg and 0.475 Nm³/kg, respectively. The optimum temperatures at the upper and bottom regions of gasifier were 860 °C and 750 °C, respectively. The carbon conversion efficiency and cold gas efficiency could reach above 87.1% and 56.9%, respectively. The maximum total concentration of combustible fuel gas (H₂ + CO + CH₄) was 22.73% with concentrations of H₂, CO and CH₄ at 9.27%, 9.25% and 4.21%, respectively. The analysis of the energy balance showed that if the gasifier was scaled up, the cold gas efficiency could be improved to be more than 86%, and thus LHV of fuel gas increased to be about 8000 kJ/Nm³.

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References


