Steam–air fluidized bed gasification of distillers grains: Effects of steam to biomass ratio, equivalence ratio and gasification temperature

Ajay Kumar [a,b], Kent Eskridge [c], David D. Jones [b], Milford A. Hanna [a,b,*]

[a] *Industrial Agricultural Products Center, University of Nebraska-Lincoln, Lincoln, NE 68583, USA*
[b] *Department of Biological Systems Engineering, University of Nebraska-Lincoln, 211 L.W. Chase Hall, Lincoln, NE 68583-0730, USA*
[c] *Department of Statistics, University of Nebraska-Lincoln, Lincoln, NE 68588, USA*

ARTICLE INFO

ABSTRACT

In this study, thermochemical biomass gasification was performed on a bench-scale fluidized-bed gasifier with steam and air as fluidizing and oxidizing agents. Distillers grains, a non-fermentable byproduct of ethanol production, were used as the biomass feedstock for the gasification. The goal was to investigate the effects of furnace temperature, steam to biomass ratio and equivalence ratio on gas composition, carbon conversion efficiency and energy conversion efficiency of the product gas. The experiments were conducted using a $3 \times 3 \times 3$ full factorial design with temperatures of 650, 750 and 850 °C, steam to biomass ratios of 0, 7.30 and 14.29 and equivalence ratios of 0.07, 0.15 and 0.29. Gasification temperature was found to be the most influential factor. Increasing the temperature resulted in increases in hydrogen and methane contents, carbon conversion and energy efficiencies. Increasing equivalence ratio decreased the hydrogen content but increased carbon conversion and energy efficiencies. The steam to biomass ratio was optimal in the intermediate levels for maximal carbon conversion and energy efficiencies.

© 2008 Elsevier Ltd. All rights reserved.

1. Introduction

A limited supply of fossil fuels and an ever growing demand for energy sources have intensified the search for alternative renewable energy sources. Energy from biomass has the potential to meet the energy demands of the world given its abundant availability on a sustainable basis (Demirbas, 2001).

Two main pathways of converting biomass to biofuels are biochemical and thermochemical conversion technologies. Biochemical conversion technologies are based on fermentation processes using microorganisms to convert the fermentable portion of the biomass to ethanol or other fuels.

Thermochemical conversion technologies use combinations of heat, oxidizing agents, pressure and catalytic conversions to breakdown the polymers of biomass into smaller molecules which are either gaseous or liquid fuels. Two main categories of thermochemical conversion are gasification and pyrolysis. Pyrolysis is a thermochemical conversion in the absence of oxidizing agents, which results in formation of a liquid fuel called bio-oil, whereas, in gasification, with the help of an oxidizing environment, the products are smaller molecules (e.g., CO, H_2, CH_4, CO_2, N_2 and lighter hydrocarbons) which are gaseous fuels. The product gas can be used either directly in the gas turbine or gas burner or subsequently converted to liquid fuels, hydrogen-rich gases or other chemicals such as methanol and urea (McKendry, 2002).

The processes of gasification and pyrolysis are old, but the commercialization of the process has been limited (Stiegel and Maxwell, 2001). More research is needed to capture the full potential of the process by making it economically viable. The difficulties in cleaning the produced gas from tar and other contaminants, capturing the heat loss associated with the high temperatures, and lack of extensive information about the gasification operation are the main obstacles facing large-scale commercialization of the gasification process.

Many authors have investigated the effects of the operating conditions on the performance and efficiency of gasification using various types of biomass such as woody biomass in downdraft gasifier (Hanaoka et al., 2005), sugarcane residue in a cyclone gasifier (Gabra et al., 2001), rice hull (Boateng et al., 1992), pine sawdust (Lv et al., 2004), almond shells (Rapagna et al., 2000), almond (Rapagna and Latif, 1997), wheat straw (Ergudenler and Ghaly, 1993), and food waste (Ko et al., 2001) in a fluidized bed gasifiers. However, extensive experimental studies of biomass gasification are scarce in the literature. In this study, a full factorial design, with replications, was used to extensively investigate the effects of the operating conditions on the performance of the gasification process.

Ethanol production from corn in US has increased in recent years to a total production of 32×10^3 m³ (8.5 billion gallons) per year and with an additional capacity of 19×10^3 m³ (5 billion

gallons) per year being developed (RFA, 2008). Distillers grains with solubles (DDGS) is a non-fermentable byproduct of ethanol production. On an average, for 3.78×10^{-3} m^3 (each gallon) of ethanol produced at a dry-grind facility, 9.53 kg (21 pounds) of corn and 51 MJ of combined thermal and electrical energy are used, producing 3.04 kg (6.7 pounds) of DDGS as a byproduct (Shapouri et al., 2003). Currently, most of the DDGS is fed to livestock. With the growth of the ethanol industry, the increasing supply of DDGS may saturate the livestock feed market unless other useful application is found. For this research, we used DDGS as a biomass feedstock to investigate its potential uses for gasification to produce gaseous fuel.

The objective of this study was to investigate the effects of steam to biomass ratio (SBR), equivalence ratio (ER) and gasifier temperature on the composition of synthesis gas produced, carbon conversion efficiency and energy conversion efficiency. This was the first replicated full-factorial experiment on gasification of DDGS.

2. Methods

2.1. Biomass sample

DDGS samples (biomass) were obtained from a local ethanol plant (Nebraska Energy LLC, Aurora, NE). The DDGS particles had a geometric mean diameter of 0.85 ± 0.55 mm. The particle size distribution was determined using ASABE standard method of determining and expressing fineness of feed materials by sieving (ASABE Standards). The moisture content of DDGS sample was 13.84% on a dry basis (db). The ultimate analyses showed that the composition of the DDGS was $49.93 \pm 0.21\%$ dry and ash-free basis (daf) carbon, $7.26 \pm 0.07\%$ daf hydrogen, $36.45 \pm 0.28\%$ daf oxygen, $5.31 \pm 0.11\%$ daf nitrogen, $1.04 \pm 0.11\%$ daf sulfur and

$5.84 \pm 0.42\%$ db ash. The heating value of the DDGS was 27.2 MJ/ kg (db) (Wang et al., 2007a).

2.2. Gasification setup

Experiments were performed on a bench-scale, indirectly heated, atmospheric pressure, and fluidized bed gasification system. The maximum throughput of the system was approximately 1–1.5 kg/h depending on the density of biomass and flowrate of fluidizing agents used. Fig. 1 shows the schematic diagram of the gasification system. The gasification system consisted of a feeder, a steam boiler and a superheater, air supply to the feeder and gasifier bed, a tubular furnace, a cylindrical fluidized-bed gasifier, a cyclone separator, a heat exchanger, a condensate collecting flask, an air filter, a dessiccator column, a gas chromatography system and a data acquisition system. The steam and air were used as fluidizing and oxidizing agents. Steam from the boiler (Model: MBA9, Sussman-Automatic Corporation, Long Island City, NY) was supplied to the bottom of the gasifier bed below the distributor plate. A steam superheater prevented the steam from condensing before passing into the gasifier bed. Air was supplied to the gasifier at the bottom of the gasifier bed across the steam port. The volumetric feeder (K-tron America, Inc., Pitman, NJ) was filled with DDGS and a cover was mounted on the top. Air was supplied to the feeder through the port on the feeder cover. Air supplied to the feeder prevented the steam and volatile gases from backflowing towards the feeder. The air supply to the gasifier bed and feeder were controlled and monitored using two flowmeters and two valves separately. The gasifier was heated by a tube furnace (Model: HTF55342C Lindberg/Blue M, Kendro Laboratory Products, Inc., Asheville, NC) with a maximum temperature of 1200 °C and power of 5.4 kW. The furnace temperature was controlled by a temperature controller. A separate heating tape (Model: AWH-051-080D,
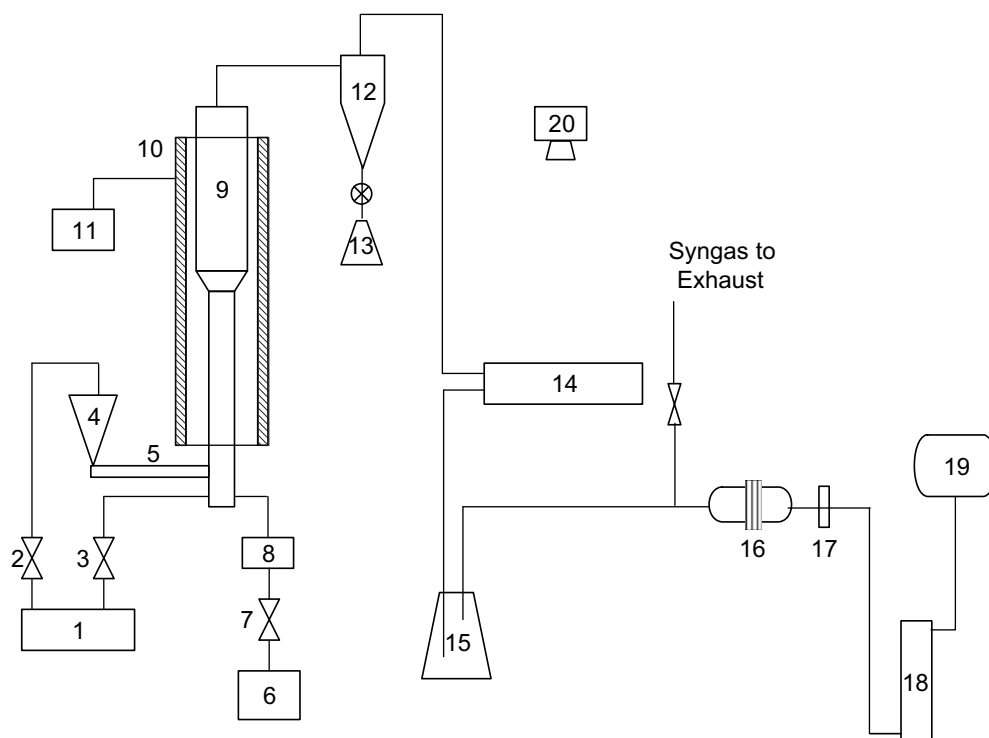

Fig. 1. Schematic diagram of the fluidized bed gasification system. 1 – air supply; 2 – air supply controller and flowmeter to the feeder; 3 – air supply controller and flowmeter to the gasifier bed; 4 – biomass feeder; 5 – screw feeder; 6 – steam boiler; 7 – steam controller and flowmeter; 8 – steam superheater; 9 – fluidized-bed gasifier; 10 – electrical furnace; 11 – electrical furnace controller; 12 – cyclone separator; 13 – char collector; 14 – heat exchanger; 15 – condensate collector; 16 – syngas filter holder with filter; 17 – syngas flowmeter; 18 – dessiccator; 19 – gas sample bag; 20 – data acquisition system.

HTS/Amptek Company, Stafford, TX, USA), with an open control, was wrapped around the gasifier bed to raise the bed temperature. The gasifier tube was made of 316 stainless steel pipe. The lower part of the gasifier (bed) had an inside diameter of 3.81 cm (1.5") and a length of 700 mm. The upper part of the gasifier (freeboard) had an inside diameter of 6.35 cm (2.5") and a length of 500 mm. The data acquisition system (Model: NI SCXI-1102 with 32-channel thermocouple terminal block) and LabView 5.1 (National Instruments Corporation, Austin, TX, USA) collected the temperatures at many locations throughout the system using T and K type thermocouples. The thermocouples collected temperatures of the steam before entering the bed, 7 zones of the gasifier, the gas before the cyclone separator and the gas before and after heat exchanger.

The particulates from the hot gas were separated by the cyclone separator and collected at the bottom. The hot gas was cooled to about 18 °C by a water-cooled shell and tube type heat exchanger. The condensed steam collected in the flask. The gas then passed through an air filter (size: 90 mm, Toyo Roshi Kaisha, Tokyo, Japan) to remove the fine particulates. The filtered gas then passed through a dessiccator column to remove the moisture in the gas before collecting gas samples in the gas bags.

The collected gas samples were analyzed using a GC (Model: AutoSystem GC, PerkinElmer Inc., Waltham, MA). The GC had two packed columns (a 3.66 m (12 ft) Haysep P 60/80 and a 2.74 m (9 ft) Molecular sieve 13X 45/60, Supelco, Bellefonte, PA), two flow controllers and a thermal conductivity detector. Helium with 8.5% hydrogen was used as a carrier gas. For detecting the concentrations of all the gas components, two methods were used. The first method detected the gas components CO_2, CO, CH_4, N_2, and O_2. The gas samples were again injected and the concentration of H_2 was determined using the second method. The areas for the peaks of CO_2, CO, CH_4, N_2, O_2 and H_2 in the sample gas were obtained using the TotalChrom software (Version 6.3.1, PerkinElmer Inc., Waltham, MA). The concentration of each gas component (% v/v) was determined by comparing the results to a calibration curve for each gas using gas standards.

2.3. Experiment

Each experimental run during this study was conducted in the following three steps.

2.3.1. Start-up
At the start of each experiment, the steam boiler and the gasifier furnace were allowed to reach the set point boiler pressure and gasifier temperature. Air was circulated through the bed to heat the pipes and the cyclone separator. The feeder was filled with DDGS, the cover was mounted and the feeder was connected to the gasifier bed after putting 30 g of sand into the gasifier bed. The steam flowrate was allowed to stabilize to the steam flowrate set point using the bypass valve.

2.3.2. Run
After the steam temperature and flowrate, and gasifier temperature were stabilized at their setpoints, air was supplied to the feeder and gasifier bed. The steam bypass valve was closed so the steam started flowing into the gasifier bed. The feeder was started and the speed of the feeder was kept constant by varying the feeder setting. The time for each revolution of the gear wheel of the feeder was recorded. After running the system for 3–4 min, the gas began to exit from the syngas bypass valve before the air filter. The experiment was allowed to run for the next 3–4 min to make sure that the system was stable. The gas bypass valve was then partially closed to allow the gas to pass through dessiccator column for approximately 1–2 min for purging before collecting the gas samples. Four gas sam-

ples were collected at the 10–15 s intervals. The gas bypass valve was then opened completely. The gas samples were analyzed using the GC before the start of the next experiment. The gas compositions reported here were the averages of these four measurements. In most cases, the deviations between the gas compositions for the four gas sample measurements were within 5%.

2.3.3. Cleaning
After collecting the gas samples, the feeder was stopped and steam was bypassed. The condensate collected in the flask and char collected at the bottom of cyclone separator were measured. The air filter was removed from the filter holder, weighed and replaced with a new filter. Before starting the next experiment, all char and ash remaining at the bottom of the gasifier was removed. The feeder was disconnected and air supply was increased to burn all the residual combustible particles including char. The remaining particles were then forced out by increased air supply to the gasifier bed and a jet of air using an air supply nozzle. The particles collected at the bottom of cyclone separator during cleaning also were weighed. The system was then ready for the next experiment. The total time of operation from start of one experiment to the start of next experiment was approximately 55 min.

2.4. Experimental design

For this research, DDGS moisture and particle size were kept constant as described above, The DDGS feedrate was kept constant at 0.81 kg/h. Before the experiment, it was verified that the gasification system could consume 0.81 kg/h of DDGS. The experimental design was a $3 \times 3 \times 3$ factorial split-plot with two replications. The gasification temperature was used as whole-plot factor, whereas, steam to biomass ratio and equivalence ratio was used as split-plot factors. The gasification temperature was used as the whole-plot factor since the change of gasification temperature needed more than an hour to reach the equilibrium temperature. Thus, the nine combinations of steam to biomass ratios and equivalence ratios were evaluated under one temperature regime during each experimental run. The three levels of each factors were 0, 7.30 and 14.29 for steam to biomass ratio; 0.07, 0.15 and 0.29 for equivalence ratio; and 650, 750, 850 °C for the gasifier furnace temperature. The steam to biomass ratio was varied by changing the steam flowrate to the gasifier bed. The equivalence ratio was varied by changing the air flow to the gasifier. One-third of the air was supplied to the feeder and the remaining two-thirds were supplied to the gasifier bed. The gasifier temperature was varied by changing the setpoint temperature of the furnace temperature controller.

2.5. Response variable

Data from the experiment were used to determine the following variables.

The ER was calculated as the ratio of air supplied to the air required for the complete stoichiometric combustion of the biomass (DDGS).

Energy conversion efficiency (*Eff*, %) of the gasification was calculated as the ratio of energy in syngas (syngas energy) to the combined energy of biomass (biomass energy) and steam consumed (steam energy) (Eq. (1)). The heat of combustion for each gas was obtained from Perry and Green (1997). Note that the electric energy consumption by the furnace heater is not included in the calculation. Syngas energy was the sum of energy contents of the gas components. The percentage compositions of higher molecular hydrocarbons were assumed to be negligible

$$Eff = \frac{(35.81 x CH_4 + 12.62 x CO + 10.71 x H_2) x F}{B x (1 - Xm/100) x E + (SF - L) x E_s} \quad (1)$$

Carbon conversion efficiency (C_{eff}, %) was the ratio of carbon in the syngas to the carbon in the biomass (Eq. (2))

$$C_{eff} = \frac{(CO_2 + CO + CH_4)x12/22.4xFx60}{Bx(1 - X_m/100)x(1 - X_{ash}/100)xC/100} \quad (2)$$

where CH_4, CO, CO_2 and H_2 were respective concentrations of methane, carbon monoxide, carbon dioxide and hydrogen in the product gas (% v/v), F was the flowrate of syngas (l/min), B was the flowrate of biomass (kg/min), E was the energy content of the biomass (MJ/kg, db), X_m was the moisture content of the biomass (% wb), X_{ash} was the ash content of the biomass (% db), SF was the steam flowrate (kg/min), L was the condensate flowrate (kg/min), E_s was energy of the steam (kJ/kg), and C was the carbon percentage of the biomass (% daf).

The amount of char collected by the cyclone separator was divided by the total amount of biomass fed during the experiment to get the percentage of char.

2.6. Statistical analysis

The statistical analysis of the data was performed using SAS 9.1 (SAS Institute Inc., Cary, NC, USA). Each response variable (energy efficiency, percent gas concentrations and char) was analyzed separately. ANOVA was used to test for factors and their interactions. Polynomial contrasts were used to evaluate linear and quadratic trends. Procedures in Milliken and Johnson (1984) were used in conjunction with the significance of polynomial contrasts to identify the final model. The predictions from the final model were plotted in 3-D using the Excel sheet. The significances of the factors were evaluated at the 0.05 level. R^2 was used to evaluate the fit of the final models.

3. Results and discussion

The gasifier was heated by the tubular furnace having single heating zone. As the setpoint temperature of the furnace was changed from the 650 from 850 °C, the temperature along the gasifier stabilized at different values. The equilibrium temperatures from the bed to the top of the gasifier at the 200 mm intervals were 320 552, 667, 784, 703, 484 and 340 °C with the furnace temperature setpoint of 650 °C. The equilibrium temperatures with the setpoint of 750 °C were 319, 610, 736, 847, 799, 557 and 421 °C. The equilibrium temperatures with the setpoint of 850 °C were 319, 613, 854, 974, 908, 612 and 447 °C. It should be noted that when the setpoint temperatures of the furnace were raised to 850 °C, the temperature of the gasifier bed was below 350 °C. The temperature at the center was maximal and reached 970 °C when the setpoint temperature was 850 °C.

The overall ANOVA of the input variables and their interactions showed that the three-way interactions were not significant on any of the response variables (efficiency, percent gas compositions and char). However, the two-way and main effects were significant for most response variables. The resulting significant polynomial contrasts were used to obtain the regression models for predictions of the response variables (Table 1).

3.1. Methane content in the syngas

Plots from the resulting regression equations to predict percentage concentration of methane (Fig. 2a and b) show that the percentage of methane did not change significantly with change in equivalence ratio (ER) but it did change with changing steam to biomass ratio (SBR). Depending on the temperature, the maximum methane was obtained when the SBR was in the range of 7–11. The optimum value of SBR increased with increasing temperature. The methane composition in our study was very high as compared to other authors (Gonzalez et al., 2008; Lv et al., 2004 and Boateng et al., 1992), which may have been due to the much lower temperature of our gasifier bed (below 350 °C) during these experiments. Higher temperature tends to favor steam reforming of methane which decreases the concentration of methane and increases the concentration of hydrogen (Turn et al., 1998 and Lv et al., 2003).

3.2. Hydrogen content in the syngas

The regression equation for the hydrogen content was used to construct 3-D Fig. 3a and b which show that temperature had the most significant effect on the hydrogen content. Many authors (Turn et al., 1998; Lv et al., 2004; Lv et al., 2007; Gupta and Cichonski, 2007 and Gonzalez et al., 2008) also observed that higher temperatures contributed to higher hydrogen content in the syngas. Increased temperature provides the energy for the endothermic reaction of hydrogen production, thereby enhancing the hydrogen content of the syngas (Lv et al., 2003; Gupta and Cichonski, 2007). At low temperature (650 °C), the maximum hydrogen content was approximately 4%, whereas at 850 °C, the hydrogen content reached close to 15%. The relatively low hydrogen content during our experiment, as compared to the studies by Turn et al. (1998) and Lv et al. (2003, 2004) was probably due to the lower gasifier bed temperature (below 350 °C) as compared to their bed temperatures of 700 °C and above. The ER and SBR also affected hydrogen content, but their effects were relatively small compared to effect of temperature. At high temperature of 850 °C, the hydrogen content decreased with increasing ER which may be attributed to the dilution of the syngas by the air and increased partial combustion with increasing ER. The effect of SBR was minimal at all temperature set points, which suggest that the steam did not react with the syngas to increase the hydrogen content at these operating conditions. This observation also can be attributed to the lower gasifier bed temperature which was not favorable for the reforming reactions.

3.3. CO content in the syngas

The content of carbon monoxide in the syngas was less than 7% for all experimental conditions. The three-way interaction between the factors was not significant but most of the two-way effects and all main effects significantly affected the content of carbon monoxide in the syngas. The resulting regression equation is given in the Table 1.

Table 1
Regression equations for the response variables (t is (temperature -650 °C)/100, s is steam to biomass ratio and a is equivalence ratio).

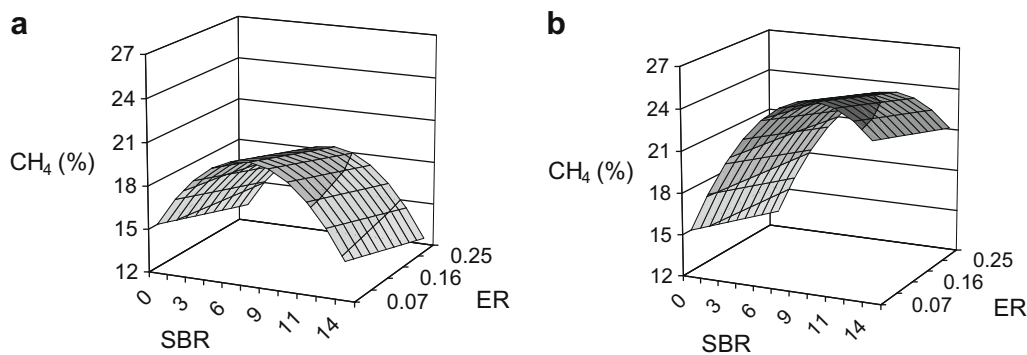| Response (%)	Regression equation	R^2
CH_4	$16.14 - 0.007t + 1.48s - 10.21a - 0.11s^2 + 0.30ts$	0.81
H_2	$3.70 + 0.29t - 0.13s + 18.30a - 22.4ta + 4.13t^2 - 48.85\ a^2 - 11.18t^2a + 77.49ta^2$	0.95
CO	$5.57 + .88t + 0.008s - 14.85a - 0.005s^2 + 28.06a^2 + 0.223ts - 0.01ts^2 + 3.94ta$	0.83
Char	$0.02 + 0.1s + 7.06a$	0.49
C_{eff}	$5.14 + 1.89t + 4.12s + 287.75a - 0.28s^2 + 1.08ts - 5.90sa$	0.91
Efficiency	$4.85 + 4.87t + 3.48s + 187.7a - 0.28s^2 + 1.02ts$	0.89

Fig. 2. Effects of SBR (steam to biomass ratio) and ER (equivalence ratio) on the percentage of methane at gasification temperatures of (a) 650 °C and (b) 850 °C.
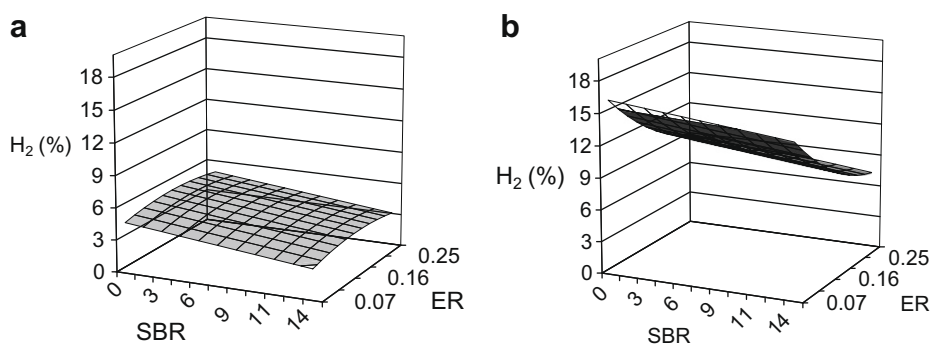
Fig. 3. Effects of SBR (steam to biomass ratio) and ER (equivalence ratio) on the percentage of hydrogen at gasification temperatures of (a) 650 °C and (b) 850 °C.
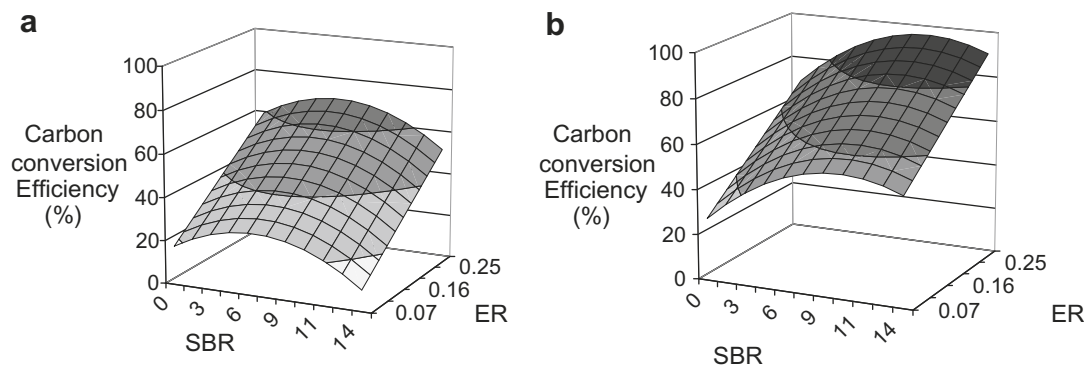
Fig. 4. Effects of SBR (steam to biomass ratio) and ER (equivalence ratio) on the carbon conversion efficiency at gasification temperatures of (a) 650 °C and (b) 850 °C.

3.4. Char content in the syngas

The only significant factors to affect the amount of char were main effects of equivalence ratio and steam to biomass ratio. The resulting regression equation is shown in Table 1. The equation shows that increasing SBR and ER linearly increased the amount of char in the syngas at all levels of temperature. The effect of temperature was not significant. Note that the amount of char analyzed here were char separated by the cyclone separator only. The remaining char in the gasifier bed and very fine char not separated by the cyclone separator could not be measured accurately. Therefore, the data for the char did not represent all the char during the experiment.

3.5. Carbon conversion efficiency (C_{eff})

All factors had significant effects on the carbon conversion efficiency (Fig. 4a and b). With increase in temperature, maximum C_{eff} increased from 62% at 650 °C to 71% at 750 °C, and 82% at 850 °C. C_{eff} increased with increasing ER. These trends were expected as higher temperature and higher ER (equivalent to higher air flow) enhance the oxidation and breakdown of the molecular bonds in the biomass leading to higher amount of conversion of the solid carbon to gaseous molecules. With increasing SBR, the C_{eff} increased and then decreased after reaching a maximal. The optimum values of SBR at the maximum C_{eff} were 2.86 at 650 °C, 4.29 at 750 °C and 7.14 at 850 °C of temperature set points.
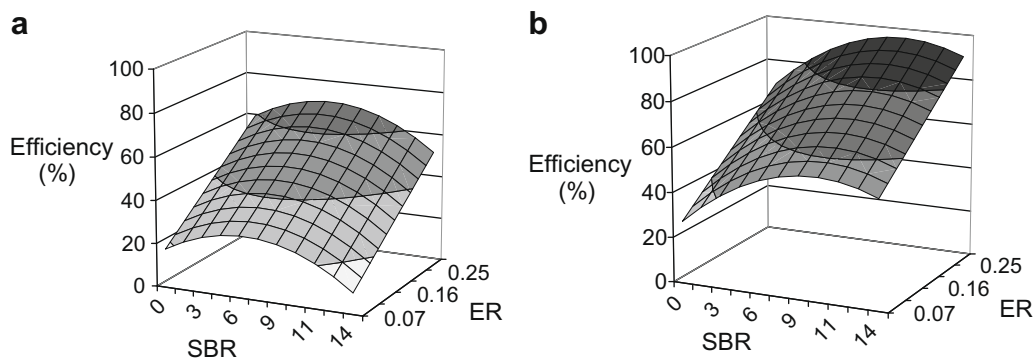
a

b

Fig. 5. Effects of SBR (steam to biomass ratio) and ER (equivalence ratio) on the energy conversion efficiency at gasification temperatures of (a) 650 °C and (b) 850 °C.

The temperature of the steam supplied below the gasifier bed was between 120 and 150 °C. An increase in SBR may have resulted in excess steam which decreased the gasifier bed temperature, which subsequently led to a decrease in the carbon conversion efficiency. Superheating the steam to a temperature comparable to the temperature of the gasifier bed may increase the carbon conversion efficiency of the process.

3.6. Energy efficiency

Fig. 5a and b shows the effects of SBR and ER on the energy efficiency at different gasification set point temperatures. At all temperatures, the plots had similar trends. The efficiency increased with increasing ER. Increase in ER implies more air was supplied to the process which helps to break down and partially oxidize the biomass and subsequent intermediate products such as methane. But supplying excess air leads to further oxidation of the desired product gases such as CO and H_2. Hence, an optimum ER is desired for maximum energy efficiency. The results (Fig. 5) suggest that our experimental operating range of the ER was below the optimum ER for obtaining maximum energy efficiency. With increase in SBR, the energy efficiency increased and then decreased after reaching a maximum. Wang et al. (2007b) reported a similar trend that with an increase supply of steam, the energy content of the produced gas increased and then decreased after reaching a maximum. They attributed that trend to the decrease in gasification temperature due to the excess steam. Effect of temperature was most significant. With increase in temperature, the energy efficiency increased from 10–70% at 650 °C to 22–82% at 750 °C and 28–96% at 850 °C. Boateng et al. (1992) also reported that an increase in temperature increased efficiency. It should be noted that the energy efficiency reached close to 100% because the heat energy supplied to the gasification systems (by the gasifier furnace) was not taken into account. This energy efficiency represents the energy of the syngas as compared to energy of consumed biomass and steam.

4. Conclusion

Extensive statistical analyses of the effects of gasification operating conditions on the syngas composition, char content, carbon conversion efficiency, and energy efficiency were performed. The methane content in the syngas increased with increasing temperature but the effect of equivalence ratio (ER) was minimal. Hydrogen content increased tremendously with increasing temperature, decreased with increasing ER, while the effect of steam to biomass ratio (SBR) was minimal. The carbon monoxide content remained relatively low throughout these experimental conditions. Carbon conversion efficiency increased with increasing temperature and

ER. The methane content, carbon conversion efficiency, and energy efficiency increased and then decreased after reaching maximal with increasing SBR.

References

ASABE Standard ANSI/ASAE S319.3 FEB03. Method of determining and expressing fineness of feed materials by sieving.
Boateng, A.A., Walawender, W.P., Fan, L.T., Chee, C.S., 1992. Fluidized-bed steam gasification of rice hull. Bioresource Technology 40, 235–239.
Demirbas, A., 2001. Biomass resource facilities and biomass conversion processing for fuels and chemicals. Energy Conversion and Management 42, 1357–1378.
Ergudenler, A., Ghaly, A.E., 1993. Agglomeration of alumina sand in a fluidized bed straw gasifier at elevated temperatures. Bioresource Technology 43, 259–268.
Gabra, M., Pettersson, E., Backman, R., Kjellstorm, B., 2001. Evaluation of cyclone gasifier performance for gasification of sugar cane residue – Part 1: gasification of bagasse. Biomass and Bioenergy 21, 351–369.
Gonzalez, J.F., Roman, S., Bragado, D., Calderon, M., 2008. Investigation on the reactions influencing biomass air and air/steam gasification for hydrogen production. Fuel Processing Technology 89 (8), 764–772.
Gupta, A.K., Cichonski, W., 2007. Ultrahigh temperature steam gasification of biomass and solid wastes. Environmental Engineering Science 24 (8), 1179–1189.
Hanaoka, T., Inoue, S., Uno, S., Ogi, T., Minowa, T., 2005. Effect of woody biomass components on air–steam gasification. Biomass and Bioenergy 28, 69–76.
Ko, M.K., Lee, W.-Y., Kim, S.-B., Lee, K.-W., Chun, H.-S., 2001. Gasification of food waste with steam in fluidized bed. Korean Journal of Chemical Engineering 18 (6), 961–964.
Lv, P., Chang, J., Xiong, Z., Huang, H., Wu, C., Chen, Y., 2003. Biomass air–steam gasification in a fluidized bed to produce hydrogen-rich gas. Energy and Fuels 17, 677–682.
Lv, P.M., Xiong, Z.H., Chang, J., Wu, C.Z., Chen, Y., Zhu, J.X., 2004. An experimental study on biomass air–steam gasification in a fluidized bed. Bioresource Technology 95, 95–101.
Lv, P.M., Yuan, Z.H., Wu, C.Z., Ma, L.L., Chen, Y., Tsubaki, N., 2007. Bio-syngas production from biomass catalytic gasification. Energy Conservation and Management 48 (4), 1132–1139.
McKendry, P., 2002. Energy production from biomass (part 2): conversion technologies. Bioresource Technology 83, 47–54.
Milliken, G.A., Johnson, D.E., 1984. Analysis of Messy Data, vol. 1 – Designed Experiments. Chapman and Hall, New York.
Perry, R.H., Green, D.W., 1997. Perry's Chemical Engineers' Handbook. McGraw-Hill Professional.
Rapagna, S., Latif, A., 1997. Steam gasification of almond shells in a fluidised bed reactor: the influence of temperature and particle size on product yield and distribution. Biomass and Bioenergy 12 (4), 281–288.
Rapagna, S., Jand, N., Kiennemann, A., Foscolo, P.U., 2000. Steam – gasification of biomass in a fluidised-bed of olivine particles. Biomass and Bioenergy 19, 187–197.
RFA. 2008. Renewable fuel Association. <http://www.ethanolrfa.org/industry/locations/> (accessed 23.04.08).
SAS. 2003. SAS User's Guide: Statistics. Version 9.1. Cary, N.C.: SAS Institute, Inc.

Shapouri, H., Duffield, J.A., Wang, M., 2003. The energy balance of corn ethanol revisited. Transactions of ASAE 46 (4), 959–968.

Stiegel, G.J., Maxwell, R.C., 2001. Gasification technologies: the path to clean, affordable energy in the 21st century. Fuel Processing Technology 71 (1–3), 79–97.

Turn, S., Kinoshita, C., Zhang, Z., Ishimura, D., Zhou, J., 1998. An experimental investigation of hydrogen production from biomass gasification. International Journal of Hydrogen Energy 23 (8), 641–648.

Wang, L., Kumar, A., Weller, C.L., Jones, D.D., Hanna, M.A., 2007a. Co-production of chemical and energy products from distillers grains using supercritical fluid extraction and thermochemical conversion technologies. ASAE Paper No. 076064. St. Joseph, Mich.: ASABE.

Wang, Y., Yoshikawa, K., Namioka, T., Hashimoto, Y., 2007b. Performance optimization of two-staged gasification system for woody biomass. Fuel Processing Technology 88 (3), 243–250.