

Research Article

Adiabatic Fixed-Bed Gasification of Colombian Coffee Husk Using Air-Steam Blends for Partial Oxidation

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The increasing energy consumption, mostly supplied by fossil fuels, has motivated the research and development of alternative fuel technologies to decrease the humanity's dependence on fossil fuels, which leads to pollution of natural sources. Small-scale biomass gasification, using air-steam blends for partial oxidation, is a good alternative since biomass is a neutral carbon feedstock for sustainable energy generation. This research presents results obtained from an experimental study on coffee husk (CH) gasification, using air-steam blends for partial oxidation in a 10 kW fixed-bed gasifier. Parametric studies on equivalence ratio (ER) ($1.53 < ER < 6.11$) and steam-fuel (SF) ratio ($0.23 < SF < 0.89$) were carried out. The results show that increasing both SF and ER results in a syngas rich in CH_4 and H_2 but poor in CO. Also, decreased SF and ER decrease the peak temperature (T_{peak}) at the gasifier combustion zone. The syngas high heating value (HHV) ranged from 3112 kJ/SATPm³ to 5085 kJ/SATPm³ and its maximum value was obtained at SF = 0.87 and ER = 4.09. The dry basis molar concentrations of the species, produced under those operating conditions ($1.53 < ER < 6.11$ and $0.23 < SF < 0.89$), were between 1.12 and 4.1% for CH_4 , between 7.77 and 13.49% for CO, and between 7.54 and 19.07% for H_2 . Other species were in trace amount.

1. Introduction

The actual growth in energy consumption, supplied mostly by fossil fuels that produce greenhouse and pollutant gases (NO_x , SO_x) through combustion processes, requires effective solutions to decrease the humanity's dependence on this type of fuels. Biomass feedstock (neutral carbon fuel), which includes energy crops and municipal, animal, and crops wastes, is a good renewable source for energy conversion via thermal processes such as full combustion and gasification. Combustion and gasification of biomass are processes that do not increase the carbon in the atmosphere since the carbon content in biomass has been previously taken from the ambient through the photosynthesis process. Due to the fact that biomass is considered a low heating value fuel, it is more appropriate for gasification than for direct combustion since combustion biomass can lead to combustion instability. Full combustion and air-biomass gasification have been deeply studied for a long time [1, 2]. However, air-steam,

oxygen-steam gasification, pyrolysis, and steam reforming gasification are new technologies recently studied, in order to improve the quality of syngas and liquid fuel produced. Colombia is the third coffee-producer around the world [3]. During 2015, Colombia produced 14.2 millions of green coffee (1 sack = 60 kg.) sacks (the terminus green coffee or gold coffee is the name for coffee bean after it has been separated from its wrappers through the shelling process) [4]. Coffee husk (CH) is a by-product from coffee seed shelling; for example, 100 kg of coffee seed produces about 86 kg of green coffee and 14 kg of CH [5], which leads to a production of about 138697 tons of CH per year in Colombia. In Colombia, many coffee agroindustries do not have the appropriate CH, waste handling infrastructure, which could result in the pollution of natural sources [6]. Using CH wastes as feedstock for locally placed gasification can reduce both the pollution generated by coffee industries and the dependence of these industries on fossil fuels.

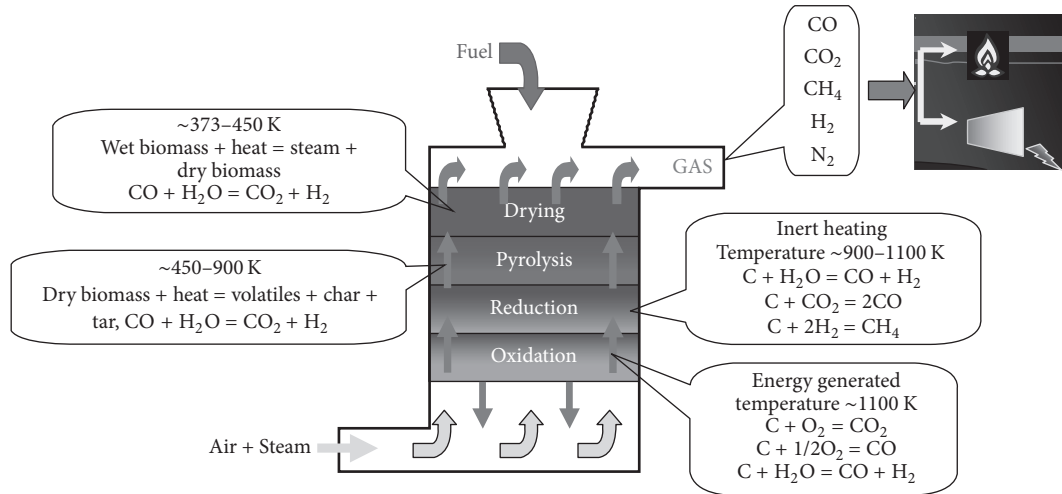


FIGURE 1: Stages and reactions in a fixed-bed gasifier, adopted from [17].

In an updraft fixed-bed gasifier, the biomass undergoes four stages, known as drying, pyrolysis, reduction, and oxidation [7]. Each of these stages occurs in a certain zone, as shown in Figure 1. The biomass enters at the top of the gasifier and goes down to the drying zone ($373\text{ K} < T < 450\text{ K}$) where the moisture content is released [8]. Afterwards, the biomass descends to the pyrolysis zone ($450\text{ K} < T < 900\text{ K}$) to release the Volatile Matter (VM) and to produce char, which descends to the reduction zone where the C atoms reacts with gases (H_2O , CO_2 , and H_2) coming from the combustion zone to produce secondary products such as CO , H_2 , and CH_4 [9]. The remaining char, from the reduction zone, passes to the combustion zone where the C atoms are oxidized with the oxygen from the air supplied to produce CO and CO_2 and the heat required to drive the system. Also, in the combustion zone the C atoms from biomass react with the H_2O supplied to produce CO and H_2 . Due to the fact that in a fixed-bed gasifier the different stages of biomass gasification occur in different zones, those gasifiers are the most appropriate to control some reactions and to produce certain desired species as required [10]. The oxidizer in gasification processes may be air [11], pure oxygen, pure steam [12], or blends of these [13–16].

During biomass gasification, many parallel and competitive chemical reactions take place, making it a chemically complex process [19]. However, the global biomass gasification process can be reduced to a set of global reactions as shown in Table 1. Reaction enthalpies (ΔH_r^θ) lower than zero and higher than zero imply exothermic and endothermic reactions, respectively. Reactions (R_2), (R_3), and (R_5) are heterogeneous reactions occurring in the oxidation zone, whereas reactions (R_4) and (R_6) are typically heterogeneous reactions that occur in the reduction zone. Homogeneous reactions usually happen in the upper reactor zones. The species leaving the reactor depend on operating parameters such as temperature, pressure, species concentration, biomass particle size, and residence time [20]. The time for developing

TABLE 1: Chemical reactions in gasification process, adapted from [9, 21, 22].

Chemical reaction	ΔH_r^θ [kJ/mol]
R_1 : Solid Biomass \rightarrow VM + Char + Tar	
<i>Char combustion</i>	
R_2 : $\text{C}_2 + (1/2)\text{O}_2 \rightarrow \text{C} + \text{CO}$	-111
R_3 : $\text{C} + \text{O}_2 \rightarrow \text{CO}_2$	-394
<i>Char gasification</i>	
R_4 : $\text{C} + \text{CO}_2 \rightarrow 2\text{CO}$	173
R_5 : $\text{C} + \text{H}_2\text{O} \rightarrow \text{CO} + \text{H}_2$	131
R_6 : $\text{C} + 2\text{H}_2 \rightarrow \text{CH}_4$	-75
<i>Homogeneous reactions</i>	
R_7 : $\text{CO} + (1/2)\text{O}_2 \rightarrow \text{CO}_2$	-283
R_8 : $\text{H}_2 + (1/2)\text{O}_2 \rightarrow \text{H}_2\text{O}$	-242
R_9 : $\text{CH}_4 + 2\text{O}_2 \rightarrow \text{CO}_2 + 2\text{H}_2\text{O}$	-803
R_{10} : $\text{CO} + \text{H}_2\text{O} \rightarrow \text{CO}_2 + \text{H}_2$	-41
R_{11} : $\text{CH}_4 + \text{H}_2\text{O} \rightarrow \text{CO} + 3\text{H}_2$	206

heterogeneous reactions is much longer than those for homogeneous reactions.

Air-blown gasification processes of wood and coal have been used during a long time to produce a syngas basically composed of CH_4 , CO_2 , CO , N_2 , H_2 , and H_2O . Due to the necessity of improving syngas quality (gases rich in H_2), other gasification processes, which use air-steam, oxygen-steam, and pure steam as oxidizing source in different gasifiers (fluidized-bed, cross draft, etc.), have been studied [23, 24]. Mostly, gasification using air or pure oxygen as oxidizing source produces a syngas with high CO and CH_4 concentrations, whereas gasification with steam or mixtures of air or oxygen with steam results in a syngas rich in H_2 [22, 25]. Typically, gasification using only steam (steam reforming) is

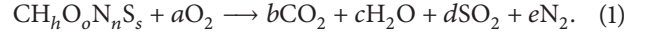
an endothermic process that produces a mixture of H_2 , CO , and CO_2 [26].

In 2008, Gao et al. [12] performed a study on the hydrogen production using pine sawdust in a counter-current fixed-bed gasifier. The parameters investigated were temperature, the steam-biomass (S/B) ratio, and equivalence ratio (ER (In [12] $ER = Air_{actual}/Air_{stoichiometric}$)). In this research, the tar generated in the gasification process was cracked using a porous ceramic reformer. Thus, hydrogen concentration was higher than that without using a porous ceramic reforming. In 2009, Gordillo and Annamalai [9] performed a study on dairy biomass adiabatic gasification carried out in a fixed-bed gasifier and using air-steam blends as oxidizing source. The effects of ER and SF on syngas composition (H_2 , CO , CO_2 , N_2 , CH_4 , and C_2H_6), temperature profiles, syngas HHV, and energy conversion were discussed. In 2008, Plis and Wilk [13] discussed the gas yield in a 50 kW/hr fixed-bed gasifier using air as oxidizer and wood pellets as fuel biomass. Parametric studies on the experimental performance, ($\lambda = Air_{actual}/Air_{stoichiometric}$), and the biomass MC (moisture content) were developed. The syngas produced in this study was blended with coal to run a cocombustor. In 2009, Vélez et al. [14] carried out an experimental study on fluidized-bed cogasification of coal with biomass using a mixture of air-steam as oxidizing source. Blends of coal with 6%–15% of coffee husk (CH), rice husk (RH), and sawdust contents were used as feedstock. The parameters studied were R_{ac} ($R_{ac} = kg_{air}/kg_{fuel}$) and R_{vc} ($R_{vc} = kg_{steam}/kg_{fuel}$) ranging from 2 kg/kg to 3 kg/kg and from 0.1 kg/kg to 0.8 kg/kg, respectively. The effects of these parameters on syngas HHV were discussed. In 2010, Wilson et al. [15] gasified coffee husk using oxygen-nitrogen/steam blends at different gasification temperatures (900°C, 800°C, and 700°C), concluding that increasing temperature resulted in a linear increase in the CO content in syngas. At lower O_2 concentrations, the above effect is more pronounced. When gasification temperature was increased from 700°C to 900°C, the CO produced, at 2% of O_2 , was enhanced by 6 and 2.5 times in a pure N_2 environment. In 2011, Gordillo and Rodriguez [27] published a modeling study on coffee gasification using air-steam as oxidizing source. The Chemical Equilibrium with Applications program (CEA), developed by NASA, was used to estimate the effect of equivalence ratio (ER) and steam to fuel (SF) ratio on equilibrium temperature and gas composition of ~150 species. The results showed that increased ER and (S:F) ratios produce mixtures that are rich in H_2 and CO_2 but poor in CO.

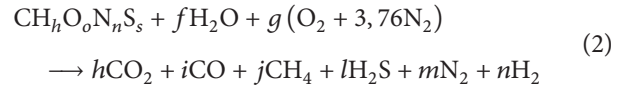
From the literature review, it is apparent that there are no preliminary experimental studies on Colombian coffee husk gasification using air-steam blends in a fixed-bed counter-current gasifier. The current paper presents results obtained from an experimental air-steam fixed gasification study of coffee husk. The effects of operating parameters, such as equivalence ratio (ER) and steam to fuel (SF) ratio, on temperature profile along gasifier axis, syngas composition, syngas HHV, and energy recovery, are discussed. Typically, results obtained from this research were $HHV_{max} = 5085 \text{ kJ/m}^3$ and molar concentration (dry basis) $H_2 = 19.07\%$;

$CO = 13.49\%$; $CO_2 = 21.78\%$; and $CH_4 = 4.1\%$. These results are discussed and analyzed in the current paper.

Full or stoichiometric combustion occurs when it is used only the necessary amount of O_2 for total oxidation of the atoms contained in the fuel (see (1)). This process (stoichiometric combustion) is basically used to produce heat.



On the other hand, partial oxidation or gasification (rich mixtures) uses lower oxygen than that required for theoretical combustion and produces a fuel gas known as syngas. According to the desired syngas composition, the gasification process must be carried out using different oxidizing sources such as pure air, pure oxygen, or pure steam. Also, mixtures of air or oxygen with steam are used. Equation (2) shows an air-steam biomass reaction producing only the most important species. The composition of syngas produced, under air-steam biomass gasification, typically, depends on some operating parameters such as equivalence ratio (ER, see (3)) and steam to fuel (SF, see (4)) ratio.



$$ER = \frac{\text{stoichiometric air moles}}{\text{actual air moles}} = \frac{a}{g} \quad (3)$$

$$SF = \frac{\text{actual steam moles}}{\text{actual fuel moles}} = f. \quad (4)$$

2. Material and Methods

2.1. Experimental Facility. Figure 2 shows a diagram of the experimental facility used in the study. The facility consists of (i) a biomass hopper where biomass is stored; (ii) a 102 cm tall small-scale (10 kW/hr.) updraft fixed-bed gasifier, built of castable alumina refractory tube (inner and outer diameter of 127 mm (5 in) and 25.4 cm (10 in), resp.), surrounded by 5.08 cm fiberglass insulating blanket layer in order to reduce heat lost; (iii) a steam generator built of a 10 cm internal diameter stainless steel tube, surrounded by a heating tape element, with variable power output (350–1000 W) to control vapor flow rate generation from 1.27 g/min to 17.25 g/min; (iv) an ash collector system consisting of a conical vibrating grate and an ash deposit placed under the grate. The conical grate (drilled with several holes) was built of cast iron and attached to a pneumatic vibrator to maintain continuous ash removal and hence quasi-steady-state operating conditions; (v) a cleaning and drying system, composed of filters and condensers to clean and dry, respectively, the syngas samples; condenser was cooled using cold water at 0°C; (vi) a temperature recorder system to record every minute the temperature measured by eight K-type thermocouples placed at 8 different points through the bed; (vii) a gas analyzer system where the syngas, previously cleaned, is analyzed; this gas analyzer analyzes and records mole fraction of CO , CH_4 , H_2 , CO , CO_2 , and C_nH_m ; and (viii) a combustion chamber to burn the syngas leaving both the gasifier and the gas analyzer.

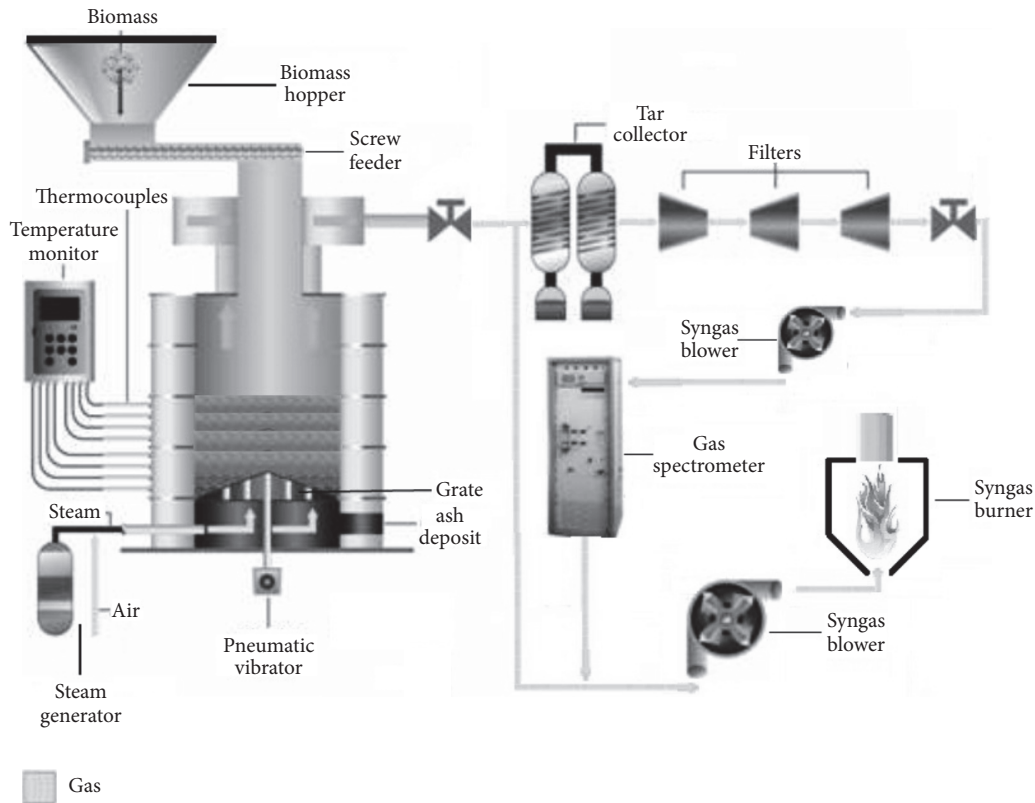


FIGURE 2: Schematic gasification facility, adapted from [18].

TABLE 2: Experimental parameters.

Air pressure and temperature.	0.76 bar, 92°C
Bed height.	180 mm
Fuel biomass.	Coffee husk
Particle size.	Flake where 80% of mass is $d_p < 4$ mm
Biomass flow rate.	0.166 g/s (600 g/h)
Air flow.	17.3 SFCH–69.2 SFCH (0.49 SATP m ³ /h–1.96 SATP m ³ /h)
Steam flow.	1.32 g/min–5.16 g/min (79.2 g/h–309.6 g/h)
ER	1.53–6.11
SF	0.23–0.87

2.2. Operating Conditions and Procedure. The experimental procedure was performed according to the parameters shown in Table 2.

Every experiment started with preheating the gasifier using a propane torch just located under the grate. When the temperature, 2 cm above the grate, was about 550°C (about 25 minutes after having turned on the torch), the torch was turned off and the screw feeder started to supply biomass to the gasifier until the bed height was about 18 cm. Then, the air and steam started to be supplied to the gasifier at flow rates as required to maintain the desired experimental operating parameters (ER and SF) shown in Table 2. At this time, the temperature started to be measured and stored

using the temperature data logger. As biomass was gasified the bed height decreased and ash accumulated on the grate. Then, biomass was added every 10 minutes and in batches as required. Also, ash was continuously removed from the top of the grate to the gasifier plenum through the vibrating grate, whose vibrating frequency was adjusted to maintain a constant 2 cm- thickness ash layer. When the gasification system stabilized, that is, peak temperature, measuring 2 cm above of the grate, started to be constant, without changing its position, the gas samples, taken from the top of the gasifier at a rate of 1 L/min, were cleaned and dried to analyze their composition (CO₂, CO, CH₄, C₂H₆, H₂, O₂, and N₂), at real time and during 10 minutes, by the gas analyzer whose

ranges are CO₂ between 0 and 40%, CO between 0 and 30%, CH₄ between 0 and 10%, H₂ between 0 and 40%, and C₂H₆ between 0 and 5%. For all experiments, the biomass flow rate was maintained constant at 600 g/h while the flows rates of air and steam were adjusted at the values given in Table 2 in order to maintain the desired operating (ER and SF) conditions. For all experiments, the air and steam were supplied to the gasifier at 75 kPa and temperatures of 20°C and 92°C, respectively. The temperatures along the gasifier bed were measured by k-type thermocouples placed at 2, 4, 6, 8, 10, 12, 14, and 16 cm above the cast iron grate and recorded, at real time, every minute by the temperature recorder.

3. Results and Discussion

3.1. Uncertainty Analysis. The uncertainty at measurements of temperature, air flow, fuel flow, and syngas compound were found by the following equations that calculate the total uncertainty and uncertainty propagation, respectively.

$$\sigma_{x_i}^2 = B_{x_i}^2 + P_{x_i}^2, \quad (5)$$

$$\sigma_r^2 = \sum_{i=1}^j \left(\frac{\partial r}{\partial x_i} \right)^2 \sigma_{x_i}^2, \quad (6)$$

where $r = r(x_1, x_2, \dots, x_j)$ is function of x_i measured variables of j , σ_r , σ_{x_i} are uncertainties of r and x_i , respectively, and B_{x_i} and P_{x_i} are systematic uncertainties and random uncertainties of x_i . The temperatures in the gasifier bed were measured by k-type thermocouples with a $\pm 0.75\%$ accuracy in the range of measured temperatures. The uncertainty derived from thermocouple operation was taken to $\pm 0.15\%$. Then, the total uncertainty of temperature is $\pm 0.76\%$ by (5). The air flow was measured by a rotameter with 100 SFCH full scale (2.83 m³/h) and resolution of 5 SFCH (0.14 m³/h) and accuracy of $\pm 4\%$; the total uncertainty of air flow is $\pm 4\%$ by (5). A rotameter with 9.96 cm³/min full scale and 0.066 cm³/min resolution was used to measure the water flow to the gasifier. This rotameter has an accuracy of $\pm 1\%$. Thus, total uncertainty to water flow is $\pm 1\%$. The feeding system (screw feeder) was calibrated previously and its systematic uncertainty was $\pm 3.42\%$. The total uncertainty of scale used to measure the sample weight was $\pm 1\%$, whereas the total uncertainty of the timer used to measure the feed rate was $\pm 3.87\%$. The uncertainties of ER and SF are functions of two variables, which were calculated by using (6), resulting in $\pm 5.57\%$ for ER and $\pm 4.12\%$ for the SF ratio. The total uncertainty to every syngas compound is function of $\pm 1\%$ accuracy only. Table 3 summarizes total uncertainties.

3.2. Fuel Characterization. CH samples were obtained from the Colombian coffee agroindustry and were characterized by ultimate and proximate analyses. Table 4 shows results from these analyses along with the empirical formula that was derived using atom balance on compounds. Also, mesh size particle of samples was determined and results are shown in Table 5.

TABLE 3: Uncertainty of gasification parameters.

Variable	Uncertainty %
Temperature	$\pm 0.76\%$
Air flow	$\pm 4.00\%$
Steam flow	$\pm 1.00\%$
Feed rate	$\pm 3.87\%$
ER	$\pm 5.57\%$
SF	$\pm 4.12\%$
MS compounds	$\pm 1.00\%$

TABLE 4: Proximate and ultimate analysis of CH.

Moisture %	10,10
Volatiles %	79,86
Ash %	1,20
FC %	8,84
C %	44,52
H %	6,03
N %	0,78
O %	48,38
S %	0,29
HHV (kJ/kg)	18740
DRY HHV (kJ/kg)	20845
DAF HHV (kJ/kg)	21127
Empirical formula	CH _{1,63} N _{0,015} O _{0,82} S _{0,0024}

TABLE 5: Mesh particle size characterization of CH.

Particle size [mm]	Mass percentage of the samples
[6.7; ∞)	0.02
[5.6; 6.7)	0.08
[4.0; 5.6)	2.33
[2.8; 4.0)	19.33
[2.0; 2.8)	20.60
[1.7; 2.0)	9.94
[1.0; 1.7)	20.54
[0.5; 1.0)	14.01
[0.0; 0.5)	13,15

3.3. Temperature Profiles. Figures 3–8 show results on temperature profile along the gasifier axis for different ER and SF ratios. Although experimentation was carried out at $1.53 < ER < 6.11$, temperature profiles are shown only for ER of 1.53, 3.07, 4.09, and 6.11. Temperature data is given at quasi-steady-state conditions. Full steady state conditions were impossible to achieve due to batch biomass supply. The tendency of the curves shows that increasing ER tends to decrease temperatures due to less air supplied to the gasifier. This leads to decreased char oxidation rate, because char oxidation is diffusion controlled; hence its rate depends on the oxygen available in the combustion zone. Lower char oxidation rates result in lower temperatures, since char oxidation (heterogeneous reactions 3 and 4) is an exothermic process that releases the heat required for the global gasification. The temperatures peaks, shown in the curves trend from

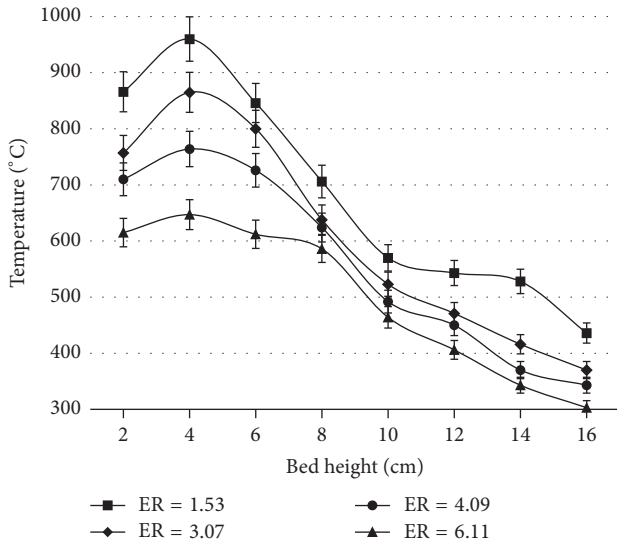


FIGURE 3: Temperature profile along with gasifier axis at SF = 0.

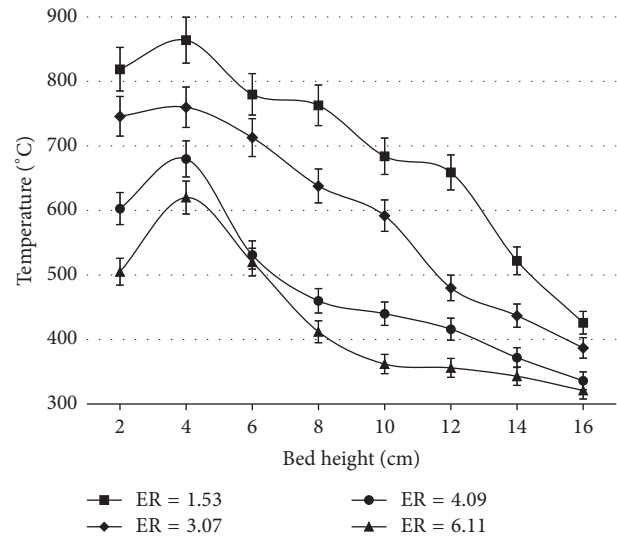


FIGURE 5: Temperature profile along with gasifier axis at SF = 0.38.

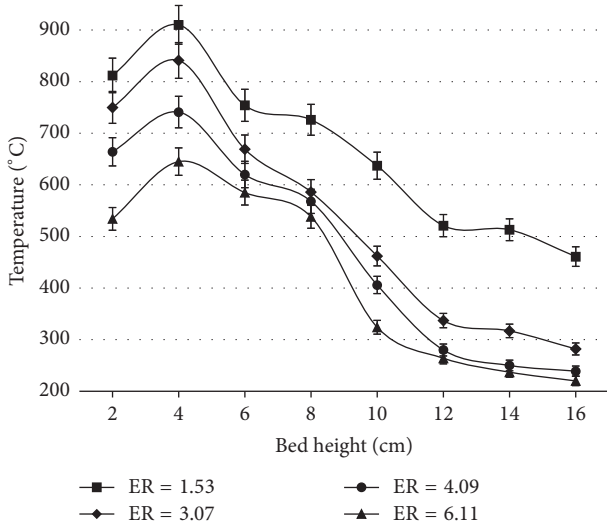


FIGURE 4: Temperature profile along with gasifier axis at SF = 0.23.

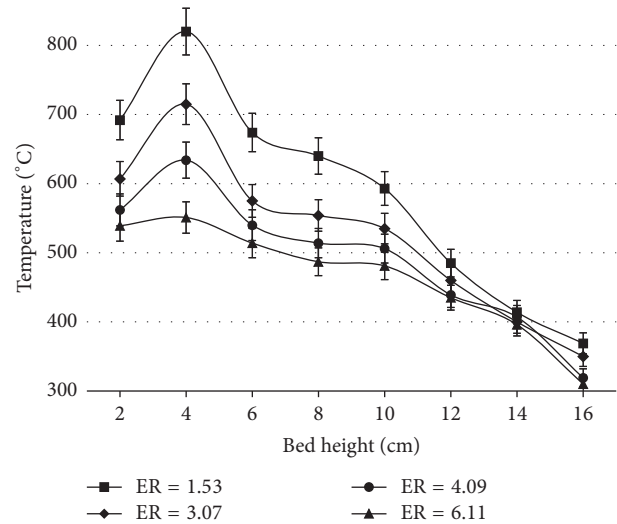


FIGURE 6: Temperature profile along with gasifier axis at SF = 0.59.

Figures 3–8, show the place where the char oxidation occurs. For almost all experimentation cases the char oxidation occurred 4 cm above the grate, indicating that quasi-steady-state was achieved. When fixed gasification is carried out under unsteady operating conditions the maximum temperature in the combustion zone move up through the gasifier axis due to ash accumulation on the grate. In the current experimentation the ash was removed to the gasifier plenum by a vibrating grate.

It is apparent from these results that increased SF ratios also decrease peak temperatures (Figure 8). Increased SF implies more steam supplied to the gasifier; hence the reaction of char in the combustion zone takes place under reach-steam conditions, which implies more of the char reacting with the steam through the endothermic reaction R_5 to produce H_2 and CO. The maximum peak temperature

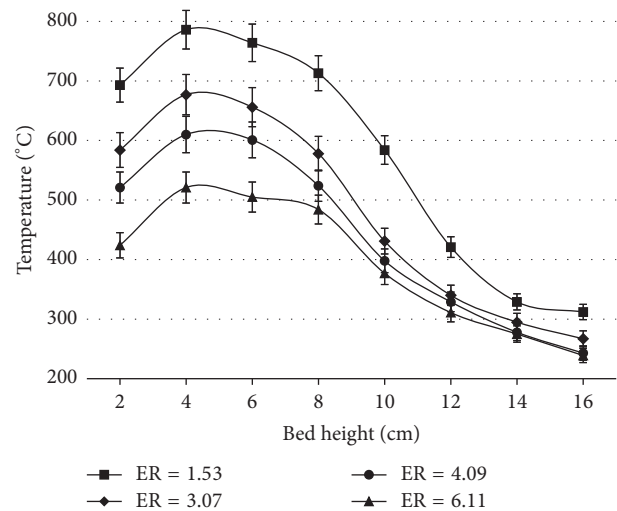


FIGURE 7: Temperature profile along with gasifier axis at SF = 0.87.

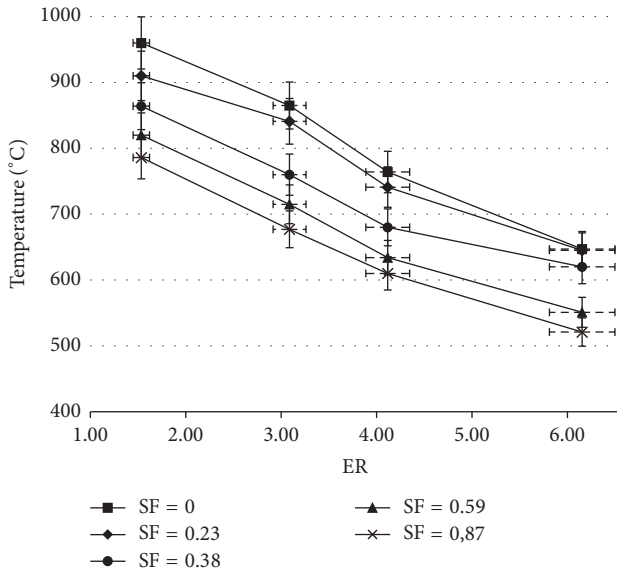


FIGURE 8: Temperature peaks versus ER for several SFs.

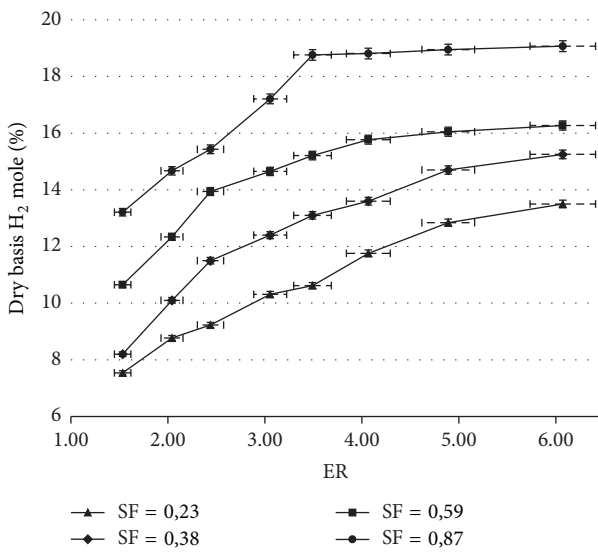


FIGURE 9: H₂ production versus ER for several SFs.

(960°C) was reached for air-gasification (SF = 0) due to no H₂O in the combustion zone.

Also, results from temperature profile show that temperatures at 2 cm above the grate are lower than these at the combustion zone (peak temperature), indicating that at this point (2 cm above the grate) there is ash accumulation. On the other hand, the temperature, above the combustion zone (4 cm up to the grate), decreases since the reactions occurring at these zones (reduction, pyrolysis, and drying) are endothermic.

3.4. Gas Composition. In this section (Figures 9–11), the results on syngas composition are presented for only relevant species in a dry tar free basis. Species in trace amount are not

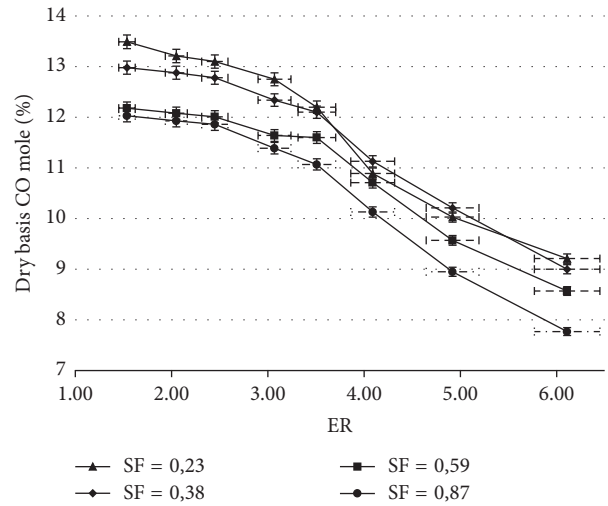


FIGURE 10: CO concentration at several SFs versus ER.

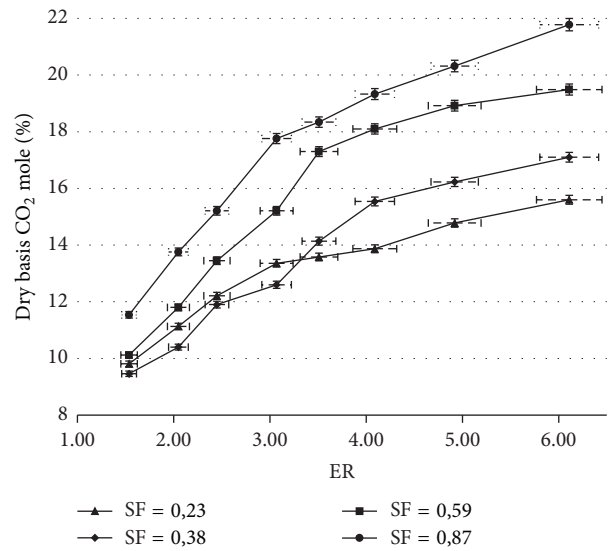


FIGURE 11: CO₂ concentration at several SFs versus ER.

shown. Figure 9 illustrates the H₂ content in syngas as a function of ER for several SFs. It is apparent that H₂ production increases when both ER and SF are increased. Increasing SF, at constant ER, leads to an H₂O-rich combustion environment, so those reactions (R₅ and R₁₀), which produce H₂, are favored. That is, steam reforming reaction (reaction R₅), which produces H₂ and CO through the reaction of C atoms with H₂O, takes place in the combustion and reduction zones (4 cm above the grate) where all water vapor supplied with the air is available. Thus, the char reaction occurs in an H₂O-rich mixture, which increases the reaction rate of char, resulting in high productions of H₂ and CO. However, Figure 10 shows a decreased production of CO while SF increases, which suggest that much of the CO produced by the steam reforming reaction (R₅) in the combustion and reduction zones is consumed by the shift reaction (R₁₀) in the

upper zones (pyrolysis and drying) to produce more H_2 and CO_2 .

In general, results show that H_2 increases and CO decreases with increased ER. At constant SF, increasing ER implies less O_2 supplied to the gasifier, resulting in H_2O -rich mixtures that are favorable for the reaction of char with H_2O (reforming reaction) to produce H_2 and CO . Also, higher ER leads to lower combustion temperatures (peak temperatures), which are more appropriate for the production of CO_2 than CO , via char oxidation (reaction R_3). Oxidation of char to produce CO and CO_2 via reactions R_2 and R_3 is competitive, and their reaction rates are very dependent on the temperature; low temperatures favor the production of CO_2 through reaction R_3 [9]. Less production of CO implies more O_2 free to react with remaining char and hence more CO_2 (Figure 11).

From Figure 9, it is apparent that the variation of ER is more important on the H_2 production at lower SF ratios than at higher SF: at SF = 0.23, increasing the ER from 1.53 to 6.11 increases the H_2 production by 79%, whereas at SF = 0.87, the same increase (1.53 to 6.11) in the ER only rises the H_2 production by 44.3%. On the other hand, at ER = 1.53, increasing SF from 0.23 to 0.87 increases the H_2 concentration by 75.2%, but, at ER = 6.11, the same increase in SF (0.23 to 0.87) only increases the H_2 production by 41.2%. This suggests that the influence of SF on the H_2 production is more important at lower ER than at higher ER.

For all SF, the slopes of CO -curves are higher at $ER > 3.51$ than those at $ER \leq 3.5$ (Figure 10). In other words, for any increase in ER the decrease in CO is higher at $ER > 3.51$ than at $ER < 3.51$ (i.e., at SF = 0.87, increasing ER from 1.54 to 3.51 decreases CO by about 8% while increasing ER from 3.51 to 6.11 decreases CO by about 30%). The effect of increased ER on the decrease of temperature peak (combustion zone) is higher at $ER > 3.51$ than that at $ER < 3.51$ (Figure 8). Thus, increasing ER decreases more the production of CO , via reaction R_2 , at $ER > 3.51$ than at $ER < 3.51$. Also, it is apparent that R_3 (oxidation of char to produce CO_2) is more important at $ER > 3.51$ than at $ER < 3.51$.

Figure 12 shows CH_4 production as a function of ER for different SF. Increasing both ER and SF increases H_2 production (Figure 9). Thus, the production of CH_4 , via the heterogeneous reaction R_6 , is favored (Figure 12) due to higher H_2 concentration in the reduction zone. The maximum CH_4 was about 4.1%, which indicates that most of the CH_4 produced in the reduction zone, via reaction R_6 , is consumed by the reforming of CH_4 (R_{11}) above of the reduction zone.

3.5. HHV of Syngas and Energy Conversion Efficiency. The syngas higher heating values (HHV), calculated by (7), are presented in Table 6 as a function of ER and SF.

$$HHV_{\text{syngas}} = \sum_{i=1}^j X_i \cdot HHV_i, \quad (7)$$

where X_i and HHV_i are the molar fraction and higher heating value of each gas fuel, contained in syngas, respectively.

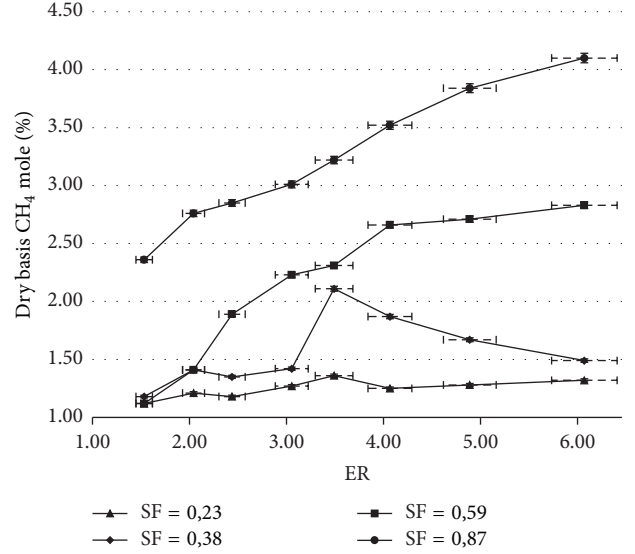


FIGURE 12: CH_4 concentration at several SFs versus ER.

TABLE 6: Dry basis syngas HHV [kJ/SATP m^3] for several SF and ER ratios.

SF	ER							
	1.53	2.05	2.45	3.07	3.51	4.09	4.92	6.11
0.23	3112	3130	3075	3216	3248	3300	3386	3413
0.38	3132	3478	3622	3710	4042	3887	3832	3678
0.59	3445	3764	4147	4318	4391	4464	4376	4312
0.87	4148	4482	4604	4836	5077	5085	5081	5052

HHV_{syngas} is the high heating value (kJ/SATP m^3 of dry syngas).

Table 6 presents results on HHV_{syngas} for different ER and SF. At SF = 0.23 increased ER increases HHV_{syngas} . However, at SF ≥ 0.38 , the HHV_{syngas} presents a maximum value at ER = 3.51 or ER = 4.09. In general, HHV_{syngas} increases with a higher ER, because the syngas has more CH_4 content ($HHV_{CH_4} = 36339$ kJ/SATP m^3 of CH_4). According to the results, the H_2 content in syngas is inversely proportional to the CO content; more H_2 implies less CO . This suggests that the HHV_{syngas} increases with increased ER, mainly, due to more production of CH_4 , since the energy density of CO (11567 kJ/SATP m^3 of CO) is very similar to that of the H_2 (11600 kJ/SATP m^3 of H_2).

Although HHV_{syngas} gives information on syngas energy density, it does not give information on energy recovered in syngas for each biomass unit gasified. The energy recovery (dry tar free basis) was estimated using the following:

$$\eta_{\text{Gas,E}} = \frac{HHV_{\text{syngas}}}{N_{\text{Fuel}} * HHV_{\text{Fuel}} + N_{\text{steam}} * 18 (h_v + 4.18 (365 - 293))}, \quad (8)$$

where HHV_{syngas} is the higher heating value of syngas on a dry basis, N_{Fuel} and N_{steam} are the moles of biomass and steam supplied to the gasifier, respectively, by each normal m^3 of dry

TABLE 7: Dry tar free energy recovery for several ER and SFs.

SF	ER							
	1.53	2.05	2.45	3.07	3.51	4.09	4.92	6.11
0.23	0.44	0.36	0.31	0.26	0.23	0.19	0.17	0.14
0.38	0.44	0.38	0.34	0.29	0.28	0.24	0.2	0.15
0.59	0.47	0.41	0.40	0.35	0.32	0.29	0.24	0.19
0.87	0.62	0.53	0.48	0.43	0.41	0.36	0.30	0.24

syngas produced, HHV_{Fuel} corresponds to the gross heating value of DAF fuel (kJ/m^3), and h_v is the vaporization enthalpy of water (kJ/kg_K). Since it was impossible to measure flows rates of both tar and water in the syngas, the syngas flow rate, required to estimate energy recovery, was estimated using atom balance on reactants (biomass, air, and steam) and products (tar composition, char, water in syngas, and dry syngas composition). Table 7 presents dry tar free energy recovery, which refers to energy recovered in syngas for each unit biomass gasified. The remaining energy is in tar, char, and energy lost by radiation and convection on the outside walls of the reactor.

From the results presented in Table 7, it can be concluded that increased ER tends to reduce energy recovery, whereas energy recovery increases with a higher SF. In general, the energy recovery ranged between 14% and 62%; the remaining percentage of energy is in tar, char, and latent heat of syngas.

4. Conclusions

- (i) The temperature in the combustion zone ranged from 521°C to 960°C . Lower values of peak temperature are because of higher SFs, which favor the endothermic char reforming reaction ($\text{C} + \text{H}_2\text{O} \rightarrow \text{CO} + \text{H}_2$). Above the combustion zone (4 cm up to the grate), temperature decreases because of the presence of endothermic reactions in the reduction, pyrolysis, and drying zones, with the exception of the shift reaction ($\text{CO} + \text{H}_2\text{O} \rightarrow \text{CO}_2 + \text{H}_2$).
- (ii) Increasing both ER and SF produces mixtures rich in H_2 , CH_4 , and CO_2 , but poor in CO , due to the fact that H_2 and CO_2 are competitive species with CO , via the shift reaction ($\text{CO} + \text{H}_2\text{O} \rightarrow \text{CO}_2 + \text{H}_2$). Also, CO and CO_2 , produced by char oxidation via reactions $\text{C}_2 + (1/2)\text{O}_2 \rightarrow \text{C} + \text{CO}$ and $\text{C} + \text{O}_2 \rightarrow \text{CO}_2$, are competitive. H_2 content in the syngas ranged from 7.54% to 19.07%, whereas CO content in syngas was from 7.77 to 13.49%. On the other hand, the maximum CH_4 content was of 4.10%.
- (iii) HHV of syngas varied from $3112 \text{ kJ}/\text{SATP m}^3$ to $5085 \text{ kJ}/\text{SATP m}^3$. Those values are similar to the ones reported in previous gasification studies. That is, high heating values (HHV), ranging from $3268 \text{ kJ}/\text{SATP m}^3$ to $4581 \text{ kJ}/\text{SATP m}^3$, were reported by [9] Gordillo and Annamalai, for a syngas obtained from air-steam dairy biomass gasification.

- (iv) The quality of syngas produced highly depends on operating parameters such as ER and SF. If rich- H_2 syngas mixtures are desired, gasification must be carried out under higher ER and SF ratios.
- (v) Air-steam gasification of coffee husk is possible without heat addition under $1.54 < \text{ER} < 6.11$ and $0.23 < \text{SF} < 0.87$. At $\text{ER} > 6.11$ and $\text{SF} > 0.87$, gasification of CH tends to be near global pyrolysis, since at these conditions the temperature at the combustion zone is only 521°C , which is not much higher than the ignition temperature of C (400°C).

Additional Points

Highlights. (i) Biomass gasification for treatment of agroindustrial wastes is used. (ii) Syngas is generated through an updraft gasifier using coffee husk as fuel. (iii) Syngas HHV based on both air and steam as gasifying agents is improved.

Conflicts of Interest

The authors declare that they have no conflicts of interest.

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