

Gasification of Kentucky bluegrass (*Poa pratensis* L.) straw in a farm-scale reactor[☆]

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Abstract

A novel gasification reactor was designed for conversion of grass straw to synthesis gas. Our design goal was to improve synthetic gas yield and thermal stability at a scale suitable for on-farm use at a cost similar to that of a combine harvester. The reactor that was constructed and tested in this study follows the newly emerging design technique whereby the endothermic pyrolysis or gasification and exothermic char combustion co-exist in the same reactor. It operates in a dual mode where straw gasification occurs in the annulus of an outer tube and an inner (draft) tube. Our trials established that the dual-mode operation could be performed without material flow problems. Sustained tests demonstrated reactor stability at gasification temperatures up to 650 °C and successful gasification of Kentucky bluegrass straw utilizing combustion heat from the inner tube. Calculated equivalence ratios of combustion in the inner tube ranged from 0.3 to 0.78 indicating fuel lean combustion of residual char without slagging. Carbon conversion ranged between 35.4 and 44.8%. Energy recovery, estimated as the ratio of the heat of combustion of the gas to that of the dry-ash-free feedstock, ranged from 14.7% to 30.92%. The estimated heating value for the synthesis gas ranged from 1.27 to 2.85 MJ m⁻³. Although these conversion parameters are low, a proof of the design concept was established. They can be improved with little modification by increasing the residence time in the draft tube and complete isolation of the gaseous products of combustion and the gasification. More tests are required to evaluate the economic feasibility of the farm-scale unit.

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1. Introduction

Temperate grass seed agriculture occupies over 2200 km² in the US Pacific Northwest. Significant reductions in field burning in Oregon and Washington, with similar outcomes expected in Idaho, have left grass seed farmers with more than a million tons of straw annually [1]. Straw produced

as a byproduct of 4.5 million acres of cereal production in the region provides the potential for an additional 5–10 Mton of straw feedstock [2]. Seed and grain farmers receive little, if any, income from their straw and have interest in value-added uses including conversion of this resource to energy.

Previous attempts to convert straw to energy were not economical because of the high costs of collecting and shipping this relatively low-density material to central conversion facilities. One approach to overcome this limitation is to develop technology scaled for economical on-farm conversion. Gasification, the process of converting carbonaceous materials into gaseous products using media

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such as air, oxygen or steam, has been suggested as a cleaner alternative to combustion of low-density materials such as hulls and straw [3,4]. The chemical composition of straw feedstocks places specific demands on thermal conversion technologies because alkali, silica, chlorine and sulfur constituents in straw contribute to slag accumulation and corrosion in many previously tested reactors [5]. Pyrolysis, the first step in the biomass gasification is an endothermic process, requiring heat to drive the chemical reactions that produce synthesis gas comprising primarily carbon monoxide and hydrogen. Heat required for this process can be provided in several ways; however, the amount of energy available in the synthesis gas needs to exceed that required to produce it.

Most previous attempts to convert agricultural residues to energy focused on large-scale, high capital investment models. The concept of achieving economic viability at an on-farm scale has been largely overlooked. Our goal was to design and test a gasifier at a scale approximating the cost of a combine harvester at the \$250,000–300,000 range. Successful accomplishment of this objective requires the development of a reactor design that can overcome the challenges of using straw feedstocks e.g., operation without material bridging, slagging or corrosion and which can provide positive energy production with economic opportunities including integration into whole-farm production cycles. Another critical requirement of an ideal design is the capability to produce high-quality synthesis gas without the need for an oxygen plant to drive the pyrolysis reactions.

To ensure positive energy balance is achieved, several reactor and processing media have been proposed including designs that utilize partial combustion in pure oxygen, use of steam as a gasification medium, and techniques involving use of heat carriers to directly or indirectly heat the biomass [5–7]. Variations of the indirect heating approach to gasification and pyrolysis have been reported in recent literature [8,9]. The primary distinction among existing designs has been the manner in which heat is transferred to the feedstock material. Recent designs have employed solid media such as sand, steel, or ceramic balls as heat transfer carriers some of which have been pneumatically or mechanically conveyed to improve energy efficiency, carbon conversion, and gas yield [8,9].

The objective of this research was to employ the concepts described herein into a novel gasification process for grass straw and to evaluate the preliminary feasibility of the design for low-cost farm conversion of Kentucky bluegrass (*Poa pratensis* L.) into high-quality synthesis gas suitable for stationary power generation and the production of liquid fuels.

2. Materials and experimental methods

2.1. Gasifier

The gasifier design follows that described by [8,9]. It consisted of the dual-stage process approach whereby the

endothermic pyrolysis reaction (reducing process) and exothermic char combustion (oxidation process) co-exist in the same vessel but occur in separate environments (Fig. 1). The reactor operates at atmospheric pressure and employs circulating solids to accomplish the oxidation/reduction cycle. The cycle follows a counter-current processing sequence whereby steel shots (the heat carrier) travel down via gravity in the annular channel outside a vertical draft tube. Biomass feedstock that travels concurrently with the steel balls is fed into the annulus. Upon pyrolysis the residual char travels down with the steel balls which eventually end up in the inner draft tube where both are entrained. Within the draft tube, the char is combusted to release heat to the balls, which, in turn, are pneumatically carried to the annulus to supply the heat needed for the endothermic biomass pyrolysis to complete the cycle.

The unit was constructed utilizing a concentric counter-current design to reduce heat loss from the system. The draft tube mounted inside the outer tube was constructed of a mild steel 5.08 cm ID by 6.7 m with a bottom structure

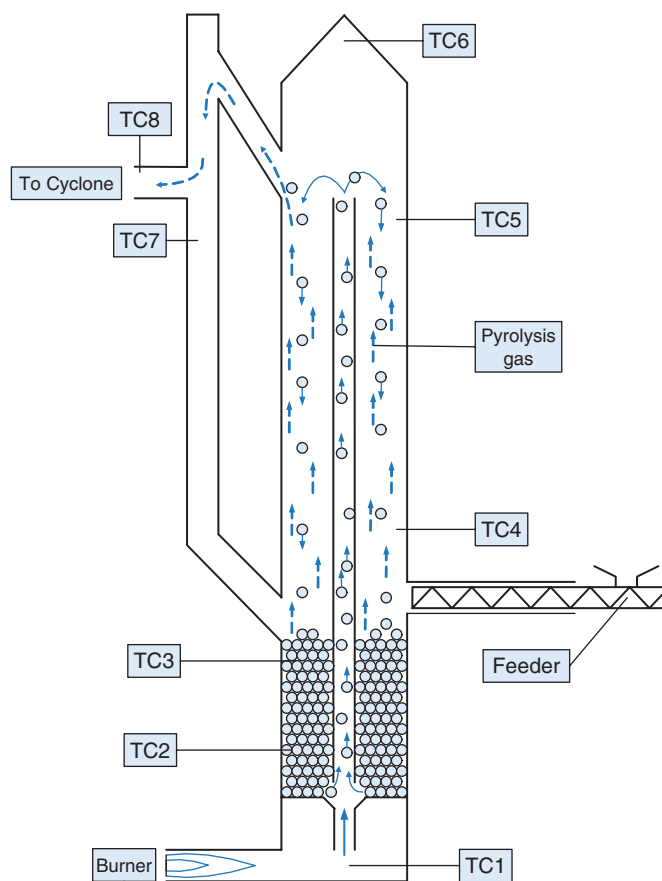


Fig. 1. Schematic of the reactor operation with thermocouple locations. TC1, Located directly at the inlet of the draft tube and overseeing combustion products; TC2, Located just above the lower draft tube-centering flange; TC3, located at 182.88 cm above the nozzle; TC4, located at 30.48 cm above the feeder; TC5, located at 274.32 cm above the feeder; TC6, located at 487.68 cm above the feeder; TC7, located on the return pipe leading to the annulus; TC8, located at the exit pipe leading to the cyclone at top of draft tube.

that allowed transition of the circulating media from the annulus into the draft tube proper. A nozzle was included at the bottom of the reactor to ensure smooth transition of a mixture of char and steel balls and to impart kinetic energy needed to transport both to the designed height. The draft tube was placed on a bowl around the orifice at the bottom of the structure to enable free upward movement to account for linear thermal expansion.

The reactor body was constructed of alternating 15.24 and 20.32 cm diameter cast iron sections bolted together with the purpose of increasing mixing among the heat transfer media and the biomass. A single exhaust gas port was provided to simplify the operation and shakedown. Synthesis gas sampling sites were provided at locations along the outer sections of the annulus (TC2–TC4 in Fig. 1). Stainless steel balls (9.5 mm diameter) were selected as the heat transfer media after initial testing of other media including ceramic balls. System heat losses were minimized by a 5.08 cm thick glass-wool insulation on the exterior portions of the reactor with a second layer applied to critical heat loss areas.

A feed system (Komar Industries, Groveport, OH) consisting of a feed box ($61 \times 61 \times 53$ cm), an auger, and a Lincoln 5HP AC motor was employed to dispense the straw at the “as received” moisture content. It was anchored on steel rails mounted to a concrete floor near the middle section of the reactor. A flexible section above the feeder was provided to compensate for overall reactor system differential thermal expansion. A remotely operated Allen-Bradley controller (Milwaukee, WI) was used to adjust feed rates.

Initial energy was provided by combustion of natural gas (Fig. 1). Combustion air was supplied by a compressor (Gardner/Denver, Model APOGBA, Quincy, IL) with a maximum-rated discharge pressure of 413 kPa when operated at 1072 rpm. The intake air to the compressor was supplied at a rate of 6.18 kg min^{-1} and the amount entering the system was controlled by venting airflow away from the reactor through a bypass valve prior to combustion. Calibration of the by-pass air allowed for precise estimates of the natural gas combustion air. Combustion products were directed through the draft tube which, dependent upon the combustion stoichiometry, established superficial velocities for entrainment. The reactor was instrumented with thermocouples TC1 through TC8 and temperatures were continuously logged as averages using LABVIEW data acquisition software (National Instruments, Austin, TX). Thermocouple TC1, located just below the nozzle (draft tube inlet), monitored natural gas flame temperature. TC2 was above the lower draft tube-centering flange and TC3 was 183 cm above the nozzle. TC2 and TC3 monitored the bed temperatures. TC4 was 30.5 cm above the feeder while TC5 and TC6 were 274.3 and 687.7 cm above the feeder, respectively. TC7 and TC8 were located at the exit pipe leading to the cyclone and TC6 was at the top of the draft tube. TC5–TC8 monitored freeboard temperatures. Char remaining after gasification,

elutriated, and collected by the cyclone was weighed after each trial and analyzed by loss on ignition and for ash composition in some instances. Gas samples were collected in stoppered vials and analyzed for gas content by gas chromatography.

2.2. Experimental procedure

The start-up procedure involved engaging the air compressor at full flow after temperature monitoring ensured that all thermocouple readings were within the range of standard operating conditions. The burner was lit with a flow of natural gas typically, $4.25\text{--}7.08 \text{ Nm}^3 \text{ h}^{-1}$, to provide sufficient combustion products for circulation of the heat transfer media. After the desired temperatures were established, natural gas flow was adjusted so that free circulation was maintained. When TC1 indicated 260°C , the flare was lit and straw was fed into the reactor at a slow initial rate until steady-state conditions were reached. If temperatures rose above the desired temperatures, air flow to the reactor was reduced via the air bypass and natural gas input was reduced accordingly. Gas samples were collected manually at specific intervals.

After general function of the reactor was confirmed, four 8-h trial runs were conducted. Each trial consisted of continuous operation during which gas samples were collected for analysis at specified times. Analyses included mass balance of input and output streams for the purpose of formulating empirical stoichiometry to describe the overall gasification. Because of difficulties associated with low differential pressures in measuring the exit gas stream with a Pitot tube, calculations were carried out by balancing input and output streams on the elemental level as described in [10] where the exit gas rate was estimated and the impact of feed rate, steam levels, natural gas and air flow rates were quantified.

3. Results and discussion

3.1. System design

The principle design (Fig. 1) criterion of ensuring oxidation of char in the inner tube and provide heat that would be transferred to the outer reactor where pyrolysis takes place was achieved. The solids-circulation with oxidation/reduction cycle is significantly different from traditional fluidized-bed or up-draft type gasification reactors. The stainless steel balls successfully provided the thermal mass required for heat transfer without being oxidized. The terminal velocity for the 9.5 mm steel balls used was 43 m s^{-1} and required an initial natural gas rate greater than $4.25 \text{ Nm}^3 \text{ h}^{-1}$ to produce the necessary combustion gases for particle entrainment. This flow rate left little or no margin for airflow control.

The system appeared to perform as intended by the design. Early differential pressure measurements taken with a liquid manometer at ports along the entire length of

the reactor showed little measurable differential pressure. There was no evidence of any high-pressure areas within the reactor indicating free flow of material. Upon contact with the balls that had been previously heated in the inner draft tube, straw pyrolysis was initiated at approximately 250 °C and completed by 400 °C. The residual char traveled downwards under gravity with the steel balls while devolatilized gas rose via differential pressure as intended. As the volatile fraction moved upward it contacted hotter

descending solids, assisting the cracking of tars, oils, and volatiles into low-molecular weight hydrocarbon fuel gases. In some instances, (Trial 1 and 4) steam was injected in the upper stages of the reactor to see if there would be steam reform to improve hydrogen yield. After reaching the bottom of the reactor, the steel balls and char remaining after pyrolysis were transferred into the oxidation chamber (inner draft tube) where the solids were entrained and returned to the annular section to repeat the cycle. Inside the inner draft tube, oxidation of the char was expected to generate heat to be absorbed by the balls and to raise their temperature to that required by the endothermic pyrolysis reactions. The hot air, including combustion products, served as the transport medium in the draft tube and also provided the necessary process heat.

Table 1 provides proximate and ultimate analyses of the chopped Kentucky bluegrass straw utilized as feedstock in these trials. The reactor operating conditions for the trials are shown in Table 2. The feed rate at the “as received” moisture ranged from 45.54 to 118.9 kg h⁻¹. Steam was periodically injected near TC3 at rates ranging from 0 to 13.6 kg h⁻¹. Air rate was 371 kg h⁻¹ at full compressor flow. A lower rate, typically 307 kg h⁻¹, was used when the amount of natural gas was reduced with the aid of the bypass mechanism. The amount of natural gas used to heat the reactor ranged between 5.1 and 7.8 m³ h⁻¹, resulting in equivalence ratios in the 0.14–0.24 range at the burner, a fuel lean combustion. The biomass-natural gas input energy ratio ranged between 2 and 7.6.

The minimum fluidization velocity (u_{mf}) in the draft tube was estimated as 5.04 m s⁻¹ with a terminal velocity (u_t) of

Table 1
Thermal analysis of Kentucky bluegrass straw

Proximate analysis	As rec'd wt%	Dry wt%	DAF ^a wt%
Moisture	8.35		
Ash	5.03	5.49	
VM	74.24	81.00	85.71
Fixed C	12.38	13.51	14.29
Total	100.00	100.00	100.00
Heating Value (kJ g ⁻¹)	16.65	17.46	18.48
Ultimate moisture	8.35		
H	5.02	5.48	5.80
C	44.87	48.96	51.80
N	11.84	12.92	13.67
S	0.31	0.34	0.36
O	24.58	26.82	28.38
Ash	5.03	5.49	
Total	100.00	100.00	100.00

^aDry-ash-free.

Table 2
Operational conditions for four independent gasification trials

	Feed rate (kg h ⁻¹)	Comb. air rate (kg h ⁻¹)	Steam rate (kg h ⁻¹)	Nat. gas rate (m ³ h ⁻¹)	Biomass: nat. gas energy ratio	Equiv. ratio ^a Φ_1 [-]	Equiv. ratio ^b Φ_2 [-]
<i>Trial 1</i>							
1114-101-1356	58.7	371.2	9.1	5.66	3.384	0.144	0.301
<i>Trial 2</i>							
1114-102-1358	88.8	371.2	13.6	5.66	5.131	0.144	0.508
1114-102-1431	88.8	371.2	13.6	5.66	5.131	0.144	0.508
<i>Trial 3</i>							
1114-103-1002	43.6	371.2		7.08	2.020	0.179	0.338
1114-103-1035	58.7	371.2		7.08	2.713	0.179	0.404
1114-103-1113	73.6	371.2		7.08	3.417	0.179	0.471
1114-103-1237	88.8	306.9		7.79	3.735	0.239	0.667
1114-103-1242	88.8	306.9		7.79	3.735	0.239	0.667
<i>Trial 4</i>							
1114-105-1041	88.81	371.2		6.51	4.456	0.165	0.525
1114-105-1111	88.81	371.2		6.51	4.456	0.165	0.525
1114-105-1218	88.81	371.2		5.95	4.885	0.151	0.514
1114-105-1223	88.81	371.2		5.95	4.885	0.151	0.514
1114-105-1256	118.93	306.9		5.95	6.543	0.182	0.782
1114-105-1359	118.93	306.9		5.10	7.646	0.156	0.761
1114-105-1425	118.93	306.9	13.6	5.10	7.646	0.156	0.761

^aActual fuel-to-air ratio/stoichiometric fuel-to-air ratio for combustion at the natural gas burner.

^bActual fuel-to-air ratio/stoichiometric fuel-to-air ratio for combustion in the inner tube.

43.13 ms^{-1} . Superficial velocities in the tube ranged between 46 and 85 ms^{-1} , depending on the firing conditions. The ratio $u_{\text{mf}} u_t^{-1}$ was within the design norm [11] at about 0.12 . The firing conditions at the burner and in the draft tube are expressed in terms of equivalence ratio defined as the ratio of the actual fuel-to-air ratio to stoichiometric fuel-to-air ratio (Φ). The purpose of introducing equivalence ratio was to estimate combustion power since it normally gives simple maxima where combustion temperature is highest. In calculating these, we distinguish between the equivalence ratio for natural gas combustion at the burner prior to entry into the inner tube, (Φ_1) which establishes TC1, and that based on char combustion in the draft tube as Φ_2 if any combustion takes place at all. The latter is based on the estimated excess air in the draft tube assuming complete combustion at the burner which will establish the circulating media temperature and, in turn, bed and freeboard temperatures.

3.2. Reactor temperatures

Reactor temperatures recorded during the four trials are presented in Fig. 2 where combustion air temperature was represented by TC1, bed temperature by TC2, and the temperature that the feed straw first encountered, by TC3. Temperatures were held steady for each test day indicating a steady operational system. The bed temperatures ranged from 500°C (Trial-1) to 650°C (Trials 2 and 4), typical of packed-bed gasification [12] and slightly lower than the 705°C design material limit. These temperatures were sustained for at least 2 h without compromising reactor integrity or encountering fusion of the ash on the circulating steel balls. The freeboard temperatures (TC4–TC6) averaged between 300 and 500°C .

With this reactor design, bed temperatures were dependent upon the natural gas and char combustion rates. The char produced was, in turn, a function of the straw feedstock feed rate and the bed temperature. From Fig. 2 the reactor temperatures and the fuel or straw feed rates can be established as follows: For Trial-1 which represented most fuel lean conditions for natural gas, TC1 measured $470 \pm 0^\circ\text{C}$. In Trial-3 when gas rate increased TC1 increased to $580 \pm 30^\circ\text{C}$. In contrast, when the gas rate was lowered in Trial-4, TC1 dropped to $550 \pm 30^\circ\text{C}$. The bed temperature (TC2) was $480 \pm 20^\circ\text{C}$ in Trial-1. When straw feed rate increased, providing more char to burn, TC2 increased to $540 \pm 10^\circ\text{C}$. However, it must be noted that natural gas flow also increased which raised TC1 accordingly. When increased feed rates were used, TC2 was recorded as $550 \pm 20^\circ\text{C}$ (Trial-3) and $650 \pm 5^\circ\text{C}$ (Trial-4). Fig. 3 relates these temperatures to the respective equivalence ratios. Φ_1 ranged from 0.14 to 0.24 indicating lean combustion with plenty of excess air to establish TC1. The amount of excess air and the combustion products determined the inner tube equivalent ratio (Φ_2), velocities, char residence time, and char combustion. As seen in Fig. 3, the bed and freeboard temperature increased almost

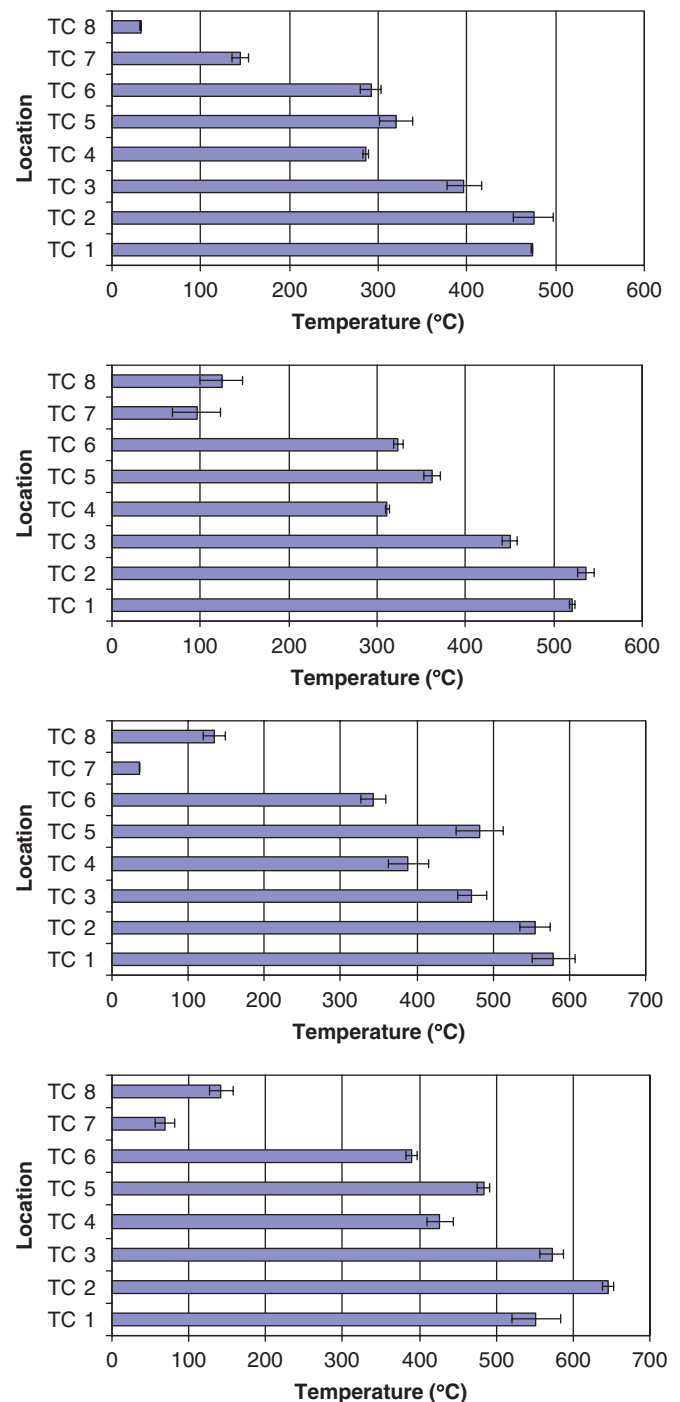


Fig. 2. Trials 1–4 steady state temperatures at eight thermocouple (TC) sites within the gasification reactor. Horizontal bars at the tip indicate the standard error of the mean.

linearly with increasing equivalence ratio as more feed and char were introduced into the reactor. However, the extent of char combustion is equally dependent on the dwell time. At these temperatures, it was unlikely that much freeboard homogeneous reactions could occur within the bed after initial pyrolysis. Nonetheless, the effect of steam injection was investigated.

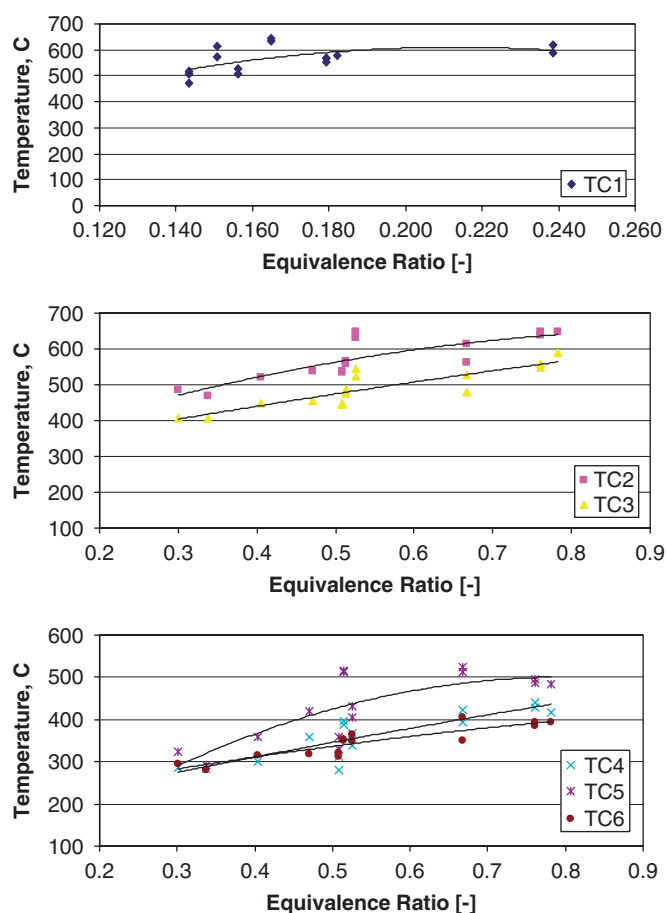


Fig. 3. Reactor temperatures as function of equivalence ratios. TC1 based on natural gas combustion at burner. Bed temperatures (TC2–TC3) and freeboard temperatures (TC4–TC6) based on char combustion with excess air at TC1 within inner tube.

3.3. Steam injection

There was little difference between the composition of gas samples taken when steam was injected into the gasifier and when the grass feedstock was fed as indicated by Trial-2 and the last two samples in Trial-4 (Table 3). It was expected that the presence of water with the nickel surfaces of the stainless steel and the moderate temperatures might lead to higher hydrogen concentrations. However, gas chromatography analyses did not support that hypothesis. Hydrogen concentrations were 1 to about 3 percent in almost all gas samples. It was likely that temperatures would have to be significantly higher in portions of the gasifier for steam reforming to occur at any measurable rate.

3.4. Gas composition

Selected components of the synthesis gas produced during all runs and experimental conditions are presented in Table 3. Nitrogen constituted the highest concentration,

ranging from 64 to 74 vol%. Mansaray et al. [13] reported nitrogen concentrations between 56 and 64 vol% in an air fluidized bed gasifier at slightly higher temperatures over that used in the present work. The higher nitrogen content we observed may have resulted from some blending of the combustion gases produced within the draft tube since we showed oxygen concentrations of 0–4% in some sample gases. The CO_2 concentration ranged from 15.6 to 19.3 vol%.

In terms of suitability for liquid fuel production, the most important combustible gases detected were H_2 , CO and CH_4 . Hydrocarbon gases up to C_3 were detected, but were present in low concentrations. Of the combustible gases measured, CO showed the highest concentration ranging from 5.2 to 12.9 vol%. This was followed by H_2 (1.12–2.85 vol%), and CH_4 (0.69–2.32 vol%). Fig. 4 shows the relationship between the bed temperature (TC2) and the combined CO and H_2 (typical syngas composition). No discernable effect was observed as indicated by the severe scatter, probably due to the mixing of combustion products. However, these numbers agree with air gasification results of similar scale [12]. The effect of steam injection was not evident, perhaps due to the lower freeboard temperatures (Table 3).

3.5. Quantification of uncombusted char

The uncombusted char was estimated by weighing the residual char collected after each trial. The reported values were for the entire run. Since one goal of the reactor design was to provide heat by combustion of char produced in the draft tube (combustion chamber), the amount of uncombusted char was also an indication of the extent of that reaction in the draft tube. Fig. 5 shows that the amount of uncombusted char decreased as the equivalence ratio in the draft tube increased. The increase from 0.3 to 0.78 suggests trends towards stoichiometric combustion ($\Phi = 1$) and, therefore, possible increased char combustion and consequently increased bed temperatures. However, the extent of char combustion in the draft tube and its pyrolysis in the annulus are not clear because chars were not analyzed for every trial run. Nonetheless, the fact that the terminal velocity required for the steel ball circulation was high (43 m s^{-1}) and required higher superficial velocities to entrain it (typically 62 m s^{-1}), the residence times for char in the draft tube were on the order of fractions of a second which is likely to be lower than the time scales required for the kinetics of heterogeneous char-gas reactions. Design modifications and revised heat transfer material selection are needed to improve this condition.

A representative char ash analysis (Table 4) shows the presence of high concentrations of silica with a silica ratio of about 69%. Although high silica and alkaline contents can present operational problems such as ash fusion with the heat transfer medium, little or no slagging was encountered at the temperatures used in this study. This is an important finding because grass straw feedstocks are

Table 3
Composition of gas produced during four independent gasification trials

	TC2 ^a (°C)	Equiv. ratio ^b [-]	Gas composition (mol%)							
			H ₂	N ₂	O ₂	CO	CH ₄	CO ₂	C ₂ H ₆	C ₃ H ₈
<i>Trial 1</i>										
1114-101-1356	486	0.301	1.215	70.404	2.033	6.183	0.729	19.104	0.296	0.037
<i>Trial 2</i>										
1114-102-1358	538	0.508	1.318	70.848	3.825	5.495	0.763	17.420	0.318	0.012
1114-102-1431	533	0.508	1.284	71.991	4.299	5.241	0.631	16.554	0.000	0.000
<i>Trial 3</i>										
1114-103-1002	470	0.338	1.870	73.157	2.954	8.441	1.226	12.351	0.000	0.000
1114-103-1035	520	0.404	2.117	65.428	1.150	12.207	1.906	17.192	0.000	0.000
1114-103-1113	537	0.471	2.610	64.228	0.595	12.957	2.175	17.417	0.000	0.017
1114-103-1237	562	0.667	2.055	71.676	0.563	5.380	1.111	19.032	0.183	0.000
1114-103-1242	614	0.667	2.376	64.179	1.247	12.733	2.316	17.132	0.000	0.016
<i>Trial 4</i>										
1114-105-1041	632	0.525	2.846	74.732	0.038	4.833	0.709	16.829	0.000	0.012
1114-105-1111	647	0.525	2.246	72.529	0.344	4.713	1.019	19.148	0.000	0.000
1114-105-1218	558	0.514	1.687	74.032	3.244	8.961	1.641	10.435	0.000	0.000
1114-105-1223	567	0.514	1.498	68.990	0.305	8.640	1.075	19.476	0.000	0.017
1114-105-1256	648	0.782	1.793	71.589	0.250	6.432	0.688	18.986	0.261	0.000
1114-105-1359	648	0.761	2.007	73.181	0.485	7.675	1.063	15.589	0.000	0.000
1114-105-1425	637	0.761	2.225	69.121	0.030	7.955	1.041	19.329	0.285	0.015

^aBed temperature.

^bEquivalence ratio (Φ_2)—actual fuel-to-air ratio/stoichiometric fuel-to-air ratio for char combustion in the inner tube.

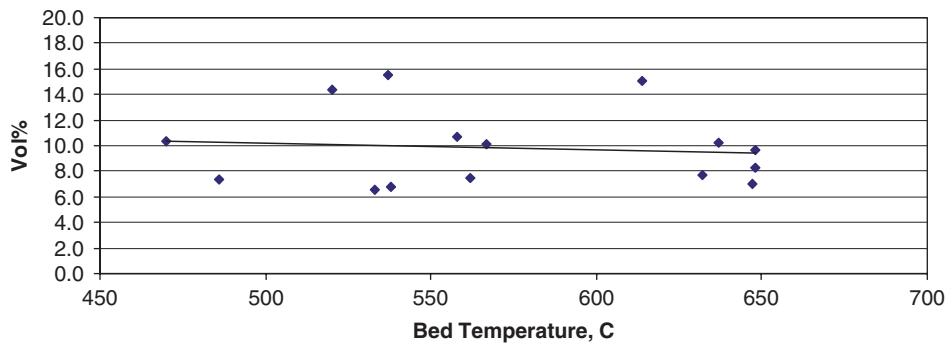


Fig. 4. Effect of temperature on syngas (H₂ + CO) produced.

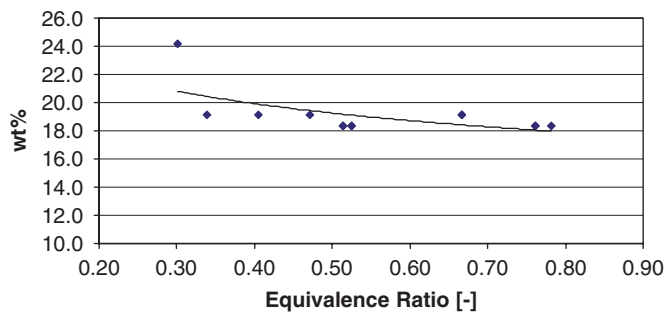


Fig. 5. Weight percent of feed char discharged at all test conditions defined by equivalence ratio of combustion in the inner tube.

generally high in Si and Cl that have restricted the use of these kinds of feedstocks in conventional up-draft and fluid bed gasification reactors in the past [5].

3.6. Gas analyses

Following simple material balance calculations based on the input streams and estimated exit streams carbon conversion, gas yield, energy recovery, and heating values were computed (Table 5). Carbon conversion, defined as the ratio of the carbon atoms in the gas to the carbon atoms in the dry-ash-free feed ranged from 35.4% to 44.8%. Mansaray et al. [13] reported values for rice hulls ranging from 55% to 81%. In contrast, Boateng et al. [3] reported 31–46% carbon conversion when rice hulls were used as feedstock in a fluidized-bed steam-gasification. It is likely that carbon conversion could be improved with increased residence time in the draft tube, along with the complete isolation of the combustion and the gasification products.

Energy recovery, estimated as the ratio of the heat of combustion of the gas to that of the dry-ash-free feedstock [3], ranged from 14.7% to 30.92%. Given the reactor conditions utilized in these trials with a temperature below 700 °C, these results can be considered encouraging for the preliminary results reported here. Boateng et al. [3] reported 37.2–56.5% for steam gasification of rice hulls in a fluidized bed reactor when temperatures between 700 and 800 °C were used. The estimated heating value for

synthesis gas produced in these trials ranged from 1.27 to 2.85 MJ m⁻³.

4. Conclusions

The concept of using separate oxidizing and reducing chambers to indirectly heat and gasify biomass via a recirculating heat transfer media was demonstrated. Sustained tests indicated consistent reactor temperature stability at temperatures up to 650 °C that enabled successful gasification of Kentucky bluegrass straw without slagging of the residual char. At this preliminary stage, carbon conversion close to 45% was achieved with energy recovery reaching a modest 15%. Although synthesis gas quality estimated by heating value was low, it was attributed to the preliminary design choice we used. Synthesis gas produced in the reducing chamber and combustion gas from the oxidizing chamber were directed to the same exhaust port, effectively mixing the two gas streams. With relatively modest design changes, these two streams can be separated and a synthetic gas of higher quality should result. After successfully testing this hypothesis, a reactor can be developed to fit the farm-scale. The economics of farm-scale synthesis gas production can then be determined.

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Table 4
Representative ash composition of char from Kentucky bluegrass straw gasification

	wt% of ignited ash
SiO ₂	34.84
Al ₂ O ₃	1.00
Fe ₂ O ₃	9.66
CaO	4.09
MgO	1.92
Na ₂ O	0.07
K ₂ O	23.39
TiO ₂	0.10
MnO ₂	0.20
P ₂ O ₅	3.24
SrO	0.02
BaO	0.05
SO ₃	1.57
Other	19.85
% Acid	47.88
% Basic	52.12

Table 5
Evaluation of four independent gasification trials

Trial/sample #	Equiv. ratio ^a [-]	Carbon conversion ^b (%)	Syngas yield (%)	Energy recovery ^c (%)	Syngas HHV (MJ m ⁻³)
<i>Trial 1</i>					
1114-101-1356	0.301	37.80	14.80	14.32	1.47
<i>Trial 2</i>					
1114-102-1358	0.508	36.31	15.30	14.82	1.40
1114-102-1431	0.508	35.41	15.73	14.05	1.08
<i>Trial 3</i>					
1114-103-1002	0.338	44.83	17.47	21.78	1.79
1114-103-1035	0.404	42.95	15.02	27.54	2.57
1114-103-1113	0.471	42.08	4.46	32.35	2.85
1114-103-1237	0.667	37.40	14.34	21.20	1.51
1114-103-1242	0.667	41.85	14.48	30.92	2.85
<i>Trial 4</i>					
1114-105-1041	0.525	35.79	14.41	25.59	1.27
1114-105-1111	0.525	35.79	14.09	21.99	1.29
1114-105-1218	0.514	41.49	15.62	21.90	2.00
1114-105-1223	0.514	37.12	13.44	18.70	1.73
1114-105-1256	0.782	35.36	13.31	18.60	1.50
1114-105-1359	0.761	36.52	13.48	21.92	1.65
1114-105-1425	0.761	36.12	13.19	23.61	1.92

^aEquivalence ratio—actual fuel-to-air ratio/stoichiometric fuel-to-air ratio for inner tube combustion.

^bRatio of carbon atoms in produced gas to carbon atoms in dry-ash-free feedstock.

^cRatio between the heating value of produced gas to the heating value of the dry-ash-free feedstock.

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