CROP RESIDUE GASIFICATION

by

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Chapter I

INTRODUCTION

INTRODUCTION

A need for alternate energy sources is rapidly emerging because of rising costs and instability in the supply of fossil fuels. Agricultural wastes represent an ideal source of alternate energy because they are abundant, renewable, and low in sulfur. Furthermore, utilization of agricultural wastes for energy would help solve the disposal problems associated with some of them. Possible energy conversion schemes for agricultural wastes include direct combustion, anaerobic digestion, gasification, liquefaction, and fermentation. Gasification of biomass in a fluidized bed reactor, which has been receiving considerable attention recently, is the subject of this thesis.

The objectives of this work were: 1) to experimentally study the gasification characteristics (i.e. produced gas heating value, composition, and yield) of corn stover and sorghum stover in a fluidized bed reactor; 2) to compare the experimental results obtained with published results; and 3) to conduct a conceptual design and economic feasibility analysis for a farm-scale gasification plant.

A literature review is presented in chapter II. The chapter emphasizes experimental studies conducted in fluidized bed gasifiers. Economic studies on biomass conversion are also discussed.

Chapter III presents experimental results for the gasification of corn stover in a fluidized bed. The effect of reactor temperature on the produced gas heating value, composition, and volumetric yield is discussed. Energy recovery, carbon conversion, and mass distribution of the major products is presented also. A comparison is given between the experimental results obtained and published results for the gasification of corn stover and wheat straw.

Chapter IV presents experimental gasification results for sorghum stover. This chapter examines the effect of reactor temperature on the produced gas heating value, composition, and yield, the mass distribution of the major products, energy recovery, and carbon conversion. A statistical comparison between the experimental results and published data is also presented.

Chapter V presents a conceptual design for a farm-scale fluidized bed gasification system. The gasifier is designed to provide fuel gas for irrigation pumping and grain drying. An economic analysis of the gasification system is also presented. The return on investment and payout period are shown graphically with fuel cost and feed cost as parameters.

The major conclusions from this thesis and recommendations for future work are given in chapter VI.

Chapter II

LITERATURE REVIEW

INTRODUCTION

Substantial efforts have been made over the last decade to develop alternate energy. Biomass, which consists of agricultural and forest wastes, and other organic wastes, such as municipal and industrial wastes, are potential sources of renewable alternate energy. Both biomass and organic wastes are abundant and usually have a low sulfur content. They also have similar thermochemical behavior, and consequently, processes for converting biomass and organic wastes to fuels and chemicals are similar. However, biomass and many organic wastes have the disadvantage of low energy density, and currently, many types are not collected and transported to central locations.

There are numerous technical options for converting biomass into fuel or chemicals. The two general categories for biomass conversion processes are biochemical and thermochemical. Biochemical processes include fermentation and anaerobic digestion; thermochemical processes include direct combustion, liquefaction, pyrolysis, and gasification. This literature review focuses on the gasification process.

In the gasification process, biomass is heated to high temperatures in an oxygen deficient environment causing a series of physical and chemical changes resulting in the evolution of volatile products and a carbonaceous solid residue. The amount of volatiles produced and their composition is dependent on the following factors: heating rate, temperature, and type of feed material.

Several types of reactors are used for gasification; they include rotary kilns, fixed beds, entrained beds, moving beds, and fluidized beds. In a fluidized bed reactor, gases flow through a bed of particles at a sufficiently high velocity to fluidize the particles but not high enough to carry them out of the reactor. The fluidized bed has a uniform temperature profile, and it can handle fine feed materials that tend to plug the fixed bed reactor. A comparison of fixed-bed, fluidized-bed, and entrained-bed reactors for use in gasification is presented by Wen and Cheng (1978). Detailed reviews of fixed and moving bed systems are presented by Reed and Jantzen (1979) and Brink (1981). Raman et al. (1981) have presented a review of literature dealing with devolatilization and fluidized bed modeling. The present literature review summarizes some of the experimental work carried out in fluidized bed reactors. It also summarizes some of the economic studies focusing on biomass conversion.

In studies of biomass gasification, most investigators are interested in the composition, yield (mass and volumetric), and heating value of the gas along with the energy recovery and carbon conversion. The results can be expressed on an off-gas or produced gas basis. Off-gas includes the gas produced from the biomass feed along with the fluidizing gas; produced gas represents only the gas produced from the biomass feed. Gas composition is generally expressed on a molar or volume basis. Gas yields are normally reported in volumetric units per unit mass of dry ash-free feed. The base temperature and pressure are important when the volumetric units are employed. The base temperature and pressure for all volumetric units presented in this thesis are 101.3 kPa (1 atm) and 288 K. The higher heating value of the gas is calculated from the gas composition and the higher heating value (standard heat of combustion) of the individual components.

Energy recovery and carbon conversion are two frequently reported results. Energy recovery is the ratio of the higher heating value of the gas produced per unit mass of dry ash-free feed to the higher heating value of a unit mass of dry ash-free feed. Carbon conversion is defined as the atoms of carbon in the gas produced from a unit mass of dry ash-free feed divided by the atoms of carbon in a unit mass of dry ash-free feed. Both energy recovery and carbon conversion provide measures of the effectiveness of the gasification process.

This literature review reports some of the results of pyrolysis, gasification, steam gasification, and partial oxidation of carbonaceous materials in fluidized bed reactors. A major difference between these conversion techniques is the environment within the reactor. Pyrolysis takes place in a completely inert environment, and gasification is performed in the presence of controlled amounts of oxidizing agents (oxygen, water, carbon dioxide, or a mixture of these). In steam gasification, large amounts of steam are present inside the reactor. Furthermore, significant amounts of oxygen are used in partial oxidation. The review is divided into several sections; one section discusses the work performed in single fluidized beds, and the next section deals with dual fluidized bed gasification. In the third section, conceptual designs and economic analyses are presented.

SINGLE FLUIDIZED BED GASIFICATION

One of the earliest studies on thermochemical conversion in a fluidized bed was carried out by Morgan et al. (1953). They studied distillation of hardwood in a 0.051 m fluidized bed batch reactor. The bed material was powdered hardwood, and the fluidizing gas was preheated nitrogen. Charcoal yield was 32% by weight of the initial charge; liquid yield was 29%, and gas yield was 16% after operating for 30 minutes at 673 K. The remaining 23% loss was attributed to tar and char accumulation within the sampling train and inaccuracies in the gas yield measurement.

Bailie and Burton at West Virginia University were the first to initiate studies on the fluidized bed gasification of biomass and waste in this country. Burton's master's thesis (Burton, 1972) reported data obtained from a 0.38 m I.D. fluidized bed reactor for the gasification of ten different materials including sawdust, municipal solid wastes, feedlot manure, and sewage sludge. The bed materials were 100% silica sand or 100% limestone, and fluidizing gas was generated by the combustion of methane under starving air conditions. Burton studied the effect of temperature, over a range of 966 to 1160 K, on the produced gas composition, yield, and heating value for different feed materials. In the produced gas, H_2 composition varied from 23 to 58%; CO composition ranged from 21 to 45%; CO_2 composition ranged from 2 to 30%; and CH_4 composition varied from 3 to 12%. Gas yield increased with temperature from 0.6 to 1.2 m^3/kg DAF, and gas heating value varied from 10 to 17 MJ/m^3. Energy recovery increased with temperature from 38 to 91%, and carbon conversion increased from 34 to 97%.

Pyrolysis studies at West Virginia University were extended by Maa and Bailie (1978). They studied the pyrolysis of wood cylinders in a fluidized bed reactor over a temperature range of 713 to 1273 K. The main objective of their study was to test the shrinking core reaction model that was proposed by them for the pyrolysis of cellulosic materials (Maa and Bailie, 1973). The reaction times obtained from the pyrolysis of wood in the fluidized bed agreed well with the predicted times calculated from their model using parameters obtained from an independent TGA study. Their results also demonstrated that the pyrolysis rate increased with an increase in reactor temperature.

Researchers at Texas Tech University studied the partial oxidation of biomass for almost a decade. Halligan et al. (1975) gasified feedlot manure in a 0.05 m I.D. fluidized bed reactor. The fluidizing gas was a mixture of air and steam, and the bed consisted of the feed material only. There was some question as to the quality of fluidization in the bed because temperature gradients were noted. Over the temperature range of 966 to 1069 K, off-gas volumetric yield increased from 0.6 to 1.3 m^3/kg, and gas heating value varied from 8.7 to 9.8 MJ/m^3. Energy recovery and carbon conversion increased with temperature from 23 to 49% and 20 to 50% respectively over the same temperature range.

Studies on feedlot manure were continued by Beck et al. (1979) who reported results obtained in a 0.15 m I.D. pilot plant reactor. The operation of the pilot plant was similar to the bench-scale reactor giving similar results. Off-gas volumetric yield increased with temperature from 0.36 m^3/kg at 790 K to 1.15 m^3/kg at 909 K, and gas heating value varied from 9 to 12 MJ/m^3. The increase in heating value was attributed to an increase in the ethylene concentration of the off-gas. Energy recovery and carbon conversion also increased with temperature from 20 to 60% and 27 to 60% respectively. An axial temperature gradient as high as 500 K was noted in the pilot reactor; therefore, the quality of fluidization was questionable.

Results from the gasification of oak sawdust in the same pilot plant were reported by Beck and Wang (1980). The off-gas yield ranged from 1.1 m^3/kg at 873 K to 1.3 m^3/kg at 1073 K, and the heating value was greater than 11.2 MJ/m^3 for all temperatures. Gasification results for mesquite, corn stover, and cotton gin trash obtained from the same pilot plant facility were compared with results from gasification of oak sawdust (Beck et al., 1981). The four feed stocks were rated on several different criteria including operability, produced gas quality, calorific value of gas per pound of feed, and percentage conversion of the raw feed heating value to gas heating value. Oak sawdust had the highest rating followed by mesquite, corn stover, and cotton gin trash respectively.

Epstein et al. (1978) gasified several different crop wastes in a 0.5 m I.D. fluid-bed reactor. The bed consisted of sand, and hot fluidizing gas was generated from the combustion of natural gas. Results for the gasification of corn cobs over a temperature range of 773 to 1273 K were presented in the most detail. Produced gas yield, reported as a mass fraction of the feed, increased from 0.17 to 0.60, and tar yield decreased from 0.22 to 0.03. The gas contained large amounts of CO and H_2, and the heating value varied from 5.4 to 10.9 MJ/m^3.

Howard et al. (1979) at the Environmental Resources Company (ERCO) investigated the pyrolysis and gasification of biomass in a fluidized bed reactor. Materials studied included paper, sawdust, corn cobs, municipal solid waste, woodchips, manure, and several mixtures. They operated their 0.5 m I.D. pilot plant under a variety of experimental conditions including steam gasification, partial oxidation, combustion, pyrolysis, and steam partial oxidation. The main objective of their work was to study the effects of reactor temperature, fluidization velocity, feed rate, static bed height, and feed particle size on reactor performance. They concluded that reactor temperature had the predominate effect on reactor performance, and that the other variables did not have a significant effect on reactor performance. In the temperature range between 873 and 1273 K, the gas yield increased from 0.05 to 0.7 kg of product gas per kg of ash-free feed for all materials under pyrolysis conditions. They also developed a model to predict liquid (tar and oil) yield as a function of temperature for the pyrolysis and gasification of various biomass materials.

Groves et al. (1979) at Texas A&M University studied the fluid-bed partial oxidation of cotton gin trash in two reactors of different size. Both were shallow bed reactors and were fluidized with air. Off-gas heating value for the smaller reactor, which had a 0.05 m I.D., varied from 4.5 to 8.2 MJ/m^3 for temperatures between 922 and 1144 K. For the larger reactor, which had a 0.3 m I.D., off-gas heating value varied from 3.4 to 4.3 MJ/m^3 for the same temperature range. Energy recovery ranged from 30 to 65% in the small reactor and 27 to 53% in the large reactor. At a reactor temperature between 1033 and 1144 K, off-gas yield was 1.1 m^3/kg and 1.7 m^3/kg for the small and large reactors respectively. They also found that the gas composition was similar in both reactors.

The gasification of rice husks was studied by Chen and Rei (1980) over a temperature range of 873 to 973 K. They used an electrically heated, 0.05 m I.D. fluidized bed reactor. The bed consisted of fused alumina sand, and super heated steam was the fluidizing gas. The gas yield increased from 0.38 to 0.55 m^3/kg DAF, and the heating value varied from 16.8 to 18.5 MJ/m^3. Over the temperature range, H_2 concentration in the produced gas varied from 3.6 to 13.1%; methane varied from 14.4 to 13.5%; CO concentration varied between 52.2 and 51.1%; and CO_2 concentration varied from 23.0 and 14.6%. The balance of the produced gas was made up of higher hydrocarbons including ethane, ethylene, and propylene.

In Belgium, Schoeters et al. (1981) investigated the gasification of wood shavings in a 0.15 m I.D. bench-scale reactor. The bed consisted of sand, and the fluidizing gas was a mixture of air and steam. Experimental variables in their study were the following: air-factor, steam rate, and reactor freeboard temperature. The air-factor was defined as the ratio of actual air flow rate to the air flow rate required for complete combustion of the feed. At a reactor temperature of 1073 K, off-gas higher heating value was 4.5 MJ/m^3; off-gas yield was 2.8 kg/kg feed; and energy recovery was 60%.. They found that an increase in air-factor caused the off-gas yield to increase, and the heating value to pass through a maximum. On the other hand, an increase in the steam rate lowered off-gas yield, heating value, and energy recovery. Furthermore, freeboard temperature affected off-gas yield, heating value, and composition. The freeboard temperature was varied from 25 K above to 150 K below the bed temperature. An increase in freeboard temperature caused concentrations of CO_2 and H_2 to increase and CO to decrease.

van den Aarsen et al. (1982) studied beechwood pyrolysis in a fluid-bed reactor. They used nitrogen as the fluidizing gas and electric heaters as the heat source for the reactor. Gas yield increased from 0.6 m^3/kg feed at 973 K to 0.8 m^3/kg feed at 1273 K. The concentration of CO in the produced gas varied from 51.2% at 973 K to 45.3% at 1273 K; H_2 ranged from 15.7 to 26.3%; CO_2 varied from 14.1 to 11.2%; and CH_4 ranged from 14.1 to 12.4%. The balance of the produced gas consisted of ethane, ethylene, and propylene. Energy recovery was essentially constant at 71%.

Fluidized bed gasification of wood has also been researched at the University of Missouri (Lian et al., 1982). Three different reactor sizes were used to gasify oak sawdust. The diameters of these reactors were 0.15 m, 0.56 m, and 1.02 m. The bed consisted of sand and gravel, and air was used as the fluidizing agent. Partial combustion of the feed supplied the heat required for gasification. The combined tar and char yield decreased linearly with temperature from 6% by weight of dry wood at 923 K to 0.5% at 1073 K. They examined several operating variables including temperature, wood feed rate, air flow rate, and residence time and concluded that the ratio of total carbon-to-nitrogen in the dry gas gave the best correlation of the important variables (carbon concentration, hydrogen concentration, and higher heating value) for the combined results from all four reactors.

At Kansas State University, researchers have studied gasification of a variety of carbonaceous materials in fluidized bed reactors. Walawender and Fan (1978) reported preliminary results on gasification of feedlot manure in a 0.23 m I.D. fluidized bed reactor similar to the system used at West Virginia University. The bed consisted of silica sand and was fluidized by a mixture of steam and flue gas produced from burning propane under starving air conditions. For a temperature range of 1000 to 1100 K, the produced gas yield increased from 0.44 to 1.02 m^3/kg DAF. Higher heating value ranged from 9 to 16 MJ/m^3, and energy recovery ranged from 18 to 62%. Agglomeration of the bed was a serious problem in gasification of manure discussed by Walawender and Fan (1978). Agglomeration of the bed occured after operating the reactor in a reducing atmosphere for about seven hours. They suspected that alkali salts in the feed reacted with the sand to form low melting silicates which caused the bed materials to fuse together. It was observed that fluidization could be re-established after operating the reactor under oxidizing conditions.

Raman et al. (1980a) continued the study of feedlot manure gasification in the same pilot plant reactor over a temperature range of 900 to 977 K. The produced gas yield increased from 0.4 m^3/kg DAF at 900 K to 0.62 m^3/kg DAF at 977 K. The heating value increased with temperature from 12.5 to 21.5 MJ/m^3, and energy recovery increased from 20 to 58%. These results differed from the preliminary results because limestone was a component of the sand bed giving improved fluidization by preventing bed agglomeration.

Walawender et al. (1981) tested the effect of limestone as a bed additive in the steam gasification of manure in a 0.05 m I.D. reactor. Limestone was found to prevent agglomeration; however, limestone also affected produced gas composition, yield, and heating value. With limestone added to the bed, produced gas concentration of H_2, CO_2, and CO and gas heating value were independent of temperature at values of 45%, 28%, 15%, and 12.4 MJ/m^3 respectively. When the limestone additive was not present, the H_2 concentration decreased from 45% at 975 K to a minimum of 38.5% at 1140 K and increased to 44% at 1370 K; CO_2 decreased from 33 to 20%; and CO increased from 17 to 25%. The heating value varied from about 10 to 12 MJ/m^3 with a maximum of 15.4 MJ/m^3 at 1150 K. Possible explanations for the differences in quality and quantity of gas produced when the additive was present were catalytic influence of the limestone, differences in the quality of fluidization, or a combination of both factors.

Studies of gasification of feedlot manure at Kansas State University were extended by examining the influence of fluidizing gas velocity and feed size fraction. Raman et al. (1980b), using the 0.23 m I.D. reactor, found that the superficial velocity did not have a significant influence on produced gas yield, composition, or heating value. Feed size fraction, however, did significantly affect gasification results. For the same temperature, produced gas yield increased with a decrease in feed size. For example, yield for the smallest size studied (-14 to +40 mesh) increased from 0.51 m^3/kg DAF at 900 K to 0.81 m^3/kg DAF at 1010 K, and for the largest size fraction (-2 to +8 mesh), yield increased from 0.1 to 0.6 m^3/kg DAF for the same temperature range. Heating value versus temperature plots for each size fraction were parabolic in shape with the maximum heating value shifting to higher temperatures with an increase in feed size. Heating value for the smallest size fraction increased from 18.0 MJ/m^3 at 900 K to a maximum of 18.3 MJ/m^3 at 910 K then decreased to 14.3 MJ/m^3 at 1010 K. For the largest sized fraction, heating value increased from 14.3 MJ/m^3 at 900 K to a maximum of 19.8 MJ/m^3 at 980 K then decreased to

12.0 MJ/m^3 at 1010 K. Produced gas composition was also significantly affected by the feed size; however, the general trend of each component was similar for each size fraction. They hypothesized that the cellulose content of the manure was different for the different size fractions because of segregation. Thus, the differences in gas quality and quantity were caused by differences in feed material make-up rather than size fraction.

Other materials gasified in the pilot plant facility included cane (sorghum), sewage sludge, and rubber tires (Walawender et al., 1980). Heating value and volumetric gas yield were dependent on feed material and reactor temperature. The gas yield for each feed material increased with temperature except for sewage sludge which was constant at about 0.5 m^3/kg DAF. Cane had the highest gas yield of the feeds tested. Gas higher heating value was largest for tires followed by sewage sludge, manure, and cane respectively.

Crop residues were also gasified in the pilot plant facility. A comparison between corn stover (Raman et al., 1980c) and wheat straw (Walawender et al., 1983) showed that corn stover gave a higher produced gas yield than wheat straw, but the gas produced from corn stover had a lower heating value. Energy recovery and carbon conversion were very similar for both feed materials. Gas yield for both feed stocks increased with temperature from 0.38 m^3/kg DAF at 830 K to 0.82 m^3/kg DAF at 1020 K with corn stover and from 0.24 m^3/kg DAF at 825 K to 0.73 m^3/kg DAF at 1030 K with wheat straw. For both feeds, the gas higher heating value was parabolic with respect to temperature and ranged from 13.0 MJ/m^3 to a maximum of 16.5 MJ/m^3 at 930 K with corn stover and from 12.5 MJ/m^3 to a maximum of 16.3 MJ/m^3 at 945 K with wheat straw. Both energy recovery and carbon conversion for corn stover and wheat straw increased with temperature; energy recovery increased from about 30 to 70%, and carbon conversion increased from about 30 to 65% for both feedstocks.

Several studies have also been conducted with a 0.05 m I.D. bench scale fluidized bed reactor at Kansas State University. Steam was used as the sole fluidizing gas, and a mixture of sand and limestone was used for the bed. Feed materials gasified in this reactor included corn grain dust (Hoveland et al., 1982), alpha cellulose (Walawender et al., 1982), cottonwood, and lignin (Singh, 1983). Produced gas yield for corn grain dust increased from 0.13 m^3/kg DAF at 867 K to 0.73 m^3/kg DAF at 1033 K, and gas heating value varied from 9.4 to 11.5 MJ/m^3. Experiments on alpha cellulose and cottonwood demonstrated that steam, when present in large excess, has a significant effect on produced gas yield and composition above 940 K. It was also demonstrated that the water-gas shift reaction was the dominant gas phase reaction above 940 K for the conditions of the experiments.

DUAL FLUIDIZED BED GASIFICATION

Tsukishima Kikai Company (1974) studied the pyrolysis of municipal waste in a dual fluidized bed reactor system (as translated by Chen et al., 1975). Char produced from gasification in the first reactor was carried to the second reactor with circulating sand. The char was burned in the second reactor to supply heat to the sand which was recycled back to the first reactor. Steam was used as the fluidizing agent in the first reactor which was operated between 723 and 1123 K. Off-gas yield increased with temperature from 30 to 85% by weight of the DAF feed. For the temperature range studied, the H_2 concentration in the off-gas increased from 4 to 35%; CO decreased from 16 to 12%; CH_4 ranged from 5 to 16%; C_3H_8 increased from 2 to 13%; and CO_2 varied from 20 to 40%. Traces of H_2S and NH_3 were also detected. Heating value of the gas varied from 19 MJ/m^3 at 723 K to 18.5 MJ/m^3 at 1123 K with a maximum of 27 MJ/m^3 at 993 K. Char and tar yields, based on weight percent of the DAF feed, decreased with temperature from 60 to 6% and 11 to 2% respectively. The advantages of the dual bed system were reported as: 1) the gas heating value is larger than that found in single bed reactors because the combustion of char occurs in a separate reactor, and hence, the combustion gas does not dilute the pyrolysis gas, 2) separation of inorganic materials in the feed is possible, and 3) heat of pyrolysis in the reactor is distributed evenly; thus, pyrolysis occurs at a relatively uniform temperature.

The Tsukishima Kikai Company continued experimentation on fluidized bed gasification of carbonaceous materials (Hasagewa et al., 1979) using three different gasifier systems. In the first system, a single 0.05 m I.D. fluidized bed reactor, they gasified municipal solid waste. The bed was composed of alumina sand, and super-heated steam was used as the fluidizing gas. Gas yield increased from 0.09 m^3/kg at 773 K to 0.8 m^3/kg at 1073 K. The second system consisted of a 0.1 m I.D. gasifier and a fluidized bed regenerator with sand circulation between the two beds. Gas yield from the dual bed system was higher than that from the first system using the same feed. Gas yield increased from 0.3 to 0.7 m^3/kg, and heating value ranged from 14.6 to 18.0 MJ/m^3 over a temperature range of 823 to 973 K. The third system was a scaled-up dual bed system with a 2 m I.D. gasifier. Organic sludge from paper and pulp plants, waste plastics, municipal solid waste, and scrap tires were gasified in this pilot plant. For municipal solid waste, gas yield was 1.3 m^3/kg, and heating value was 16.5 MJ/m^3 at 973 K. At the same temperature, the volumetric concentrations of H_2, CO, CO_2, and CH_4 were 18.5%, 34.6%, 17.0%, and 5.6% respectively, and energy recovery was 67%. Carbon conversion was 63%. The gas yield from the scaled-up dual bed system was larger than that for the other two reactors, but the gas heating value was similar for all reactors.

Kuester (1979) at Arizona State University used a dual fluidized bed reactor system to gasify various cellulosic wastes at feed rates up to 11 kg/hr. His reactor was similar to the one used by Tsukishima (1974), using inert solids for the bed and recycled product gas as the fluidizing gas. Feed materials tested included kelp residue, paper, and guayule bagasse. The maximum gas yield obtained was 95% by weight, and product gas heating value was 18.6 MJ/m^3. For a temperature range of 873 to 1073 K, the H_2 concentration in the product gas increased from 17 to 33%; CO varied from 40 to 55%; CH_4 varied from 13 to 17%; C_2H_4 ranged from 5 to 10%; and CO_2 varied from 3 to 8%. The produced gas was subsequently converted to liquid fuels through Fischer Tropsch synthesis.

Feldmann et al. (1981) used a dual bed system to gasify wood. They used a multi-solid fluidized bed (MSFB) gasifier which consisted of a 0.15 m I.D. dense-phase fluidized bed reactor coupled with a 1.0 m I.D. fluidized bed combustor. Operation involved feeding wood and a mixture of steam and recycled product gas to the dense-phase reactor composed of coarse solids and fine sand. Fine sand and char were elutriated and transferred into the fluid-bed combustor where the char was burned with air. The hot sand was then returned to the dense-phase reactor to provide heat for gasification. The entrained sand passed through the dense-phase fluidized bed in bubble-free fluidization. This provided increased heat and mass transfer in the bed.

They studied the influence of gasifier temperature, wood feed rate, steam rate, entrained-phase recycle rate, wood particle size, wood moisture content, dense-phase particle size, and dense-phase height on reactor performance. The gasifier temperature was found to be the most important variable. Conversion was limited by the maximum temperature reached by the wood and the time for the wood to dry. Thus, wood moisture content also influenced reactor performance. For a temperature range of 963 to 1171 K, carbon conversion increased from 30 to 70% for wood with a moisture content of 20%, while carbon conversion reached a maximum of 80% for wood with 6% moisture. For wood with 20% moisture, energy recovery increased from 25% at 963 K to 70% at 1171 K, and heating value ranged from 15.8 to 17.7 MJ/m^3 with a maximum of 18.1 MJ/m^3 at 1116 K. The concentrations of the principal components of the gas showed little change with temperature. At 1088 K, concentration of CH_4 was 15.6%; H_2 was 13.3%; CO was 49.2%; C_2H_4 was 5.9%; and CO_2 was 15.7%.

DESIGN AND ECONOMIC ANALYSES

One of the earliest groups to perform a conceptual design and economic analysis on a fluidized bed gasifier was Alpert et al. (1972) at the Stanford Research Institute. The proposed plant, based on studies by West Virginia University, was designed to handle 1.23 Gg (1,358 tons) per day of as received municipal refuse, which corresponded to 0.91 Gg (1,000 tons) per day of dried refuse. The capital investment was estimated at 11.7 million dollars for a dual fluidized bed system and 12.5 million dollars for a single fluidized bed process in 1972 dollars. The break even gas price was estimated to be $0.43/GJ ($0.45/MMBTU) and $0.55/GJ ($0.58/MMBTU) for the dual reactor and single reactor systems respectively. The analysis was based on a 20 year plant life.

Researchers at Kansas State University also studied the economics of commercial scale gasification of manure. Engler et al. (1975) examined manure gasification in a dual fluidized bed process. For a plant processing 3.63 Gg (4,000 tons) per day of manure with a moisture content of 50%, the gas sales price for a 14% return on investment was estimated to be $0.019/m^3$ ($0.53/MSCF) or $1.32/GJ ($1.39/MMBTU) based on a 15 year plant life and mid 1974 dollars. The break even gas price was $0.010/m^3$ ($0.29/MSCF) or $0.72/GJ ($0.76/MMBTU). For a facility processing 1.81 Gg (2,000 tons) per day of dry manure, the 14% return and break even gas prices were $0.016/m^3$ ($0.44/MSCF) or $1.04/GJ ($1.10/MMBTU) and $0.0074/m^3$ ($0.21/MSCF) or $0.52/GJ ($0.55/MMBTU) respectively. They concluded that only large-scale manure pyrolysis processes could approach the point of commercial viability.

Clark et al. (1978) examined the use of crop residues to supply energy for irrigation pumping and crop drying. They considered three different processes to convert crop residue to useable energy; anaerobic digestion, direct combustion, and fluidized bed gasification. The hypothetical problem involved fueling three 100 kW irrigation pumps. In 1980 dollars, the capital cost estimate was between $150,000 and $300,000 for a mass produced anaerobic digestion system, $250,000 and $500,000 for a mass produced fuel-bed combustion system, and $129,000 and $240,000 for a mass produced dual fluidized bed gasification facility. Based on several advantages, including economics, they concluded that the gasification process was the best process of the three processes studied.

Loewer et al. (1980) studied the economics of corn grain, cobs, and stover gasification as an energy substitute for LP gas in corn grain drying. They determined the break even investment for drying one bushel of U.S. No. 2 corn grain from 25% moisture to 15.5% moisture. They found that as much as 38.9¢ and 23.5¢ per bushel of grain dried could be invested in gasification equipment when using cobs and stover respectively.

A conceptual design and economic analysis on a fluidized bed wood gasifier was performed by Lian et al. (1982). The proposed plant would gasify 91 Mg (100 tons) per day of sawdust with a 36% moisture content. The plant was depreciated over a 15 year period with a 48% tax rate. For a 15% rate of return before tax, cost of the gas was estimated to be \$3.48/GJ (\$3.66/MMBTU) in 1980 dollars.

CONCLUSIONS

A review of the literature has shown that a number of groups have studied the gasification and pyrolysis of a variety of carbonaceous materials in fluidized bed reactors. All groups have found that reactor temperature is the major factor affecting the gasification process. Their results show that as the reactor temperature increases, the gas yield increases, and gas heating value decreases. Furthermore, economic studies on the gasification of biomass show that gasification is marginally competitive with conventional energy sources.

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Chapter III

CORN STOVER GASIFICATION

INTRODUCTION

In the United States, approximately 1.4×10^{11} kg (154 million tons) of corn stover was produced in 1981 (Agricultural Statistics, 1982). If all this corn stover were collected and gasified in a fluidized bed reactor, about 1.14×10^{12} MJ (1.1×10^{15} BTU) of producer gas could have been obtained. This would have accounted for 1.4% of the total energy requirement for the United States in 1979. However, most of the corn stover is currently left in the field and returned to the soil. In most cases, about 30 to 40% of the corn stover should be left on the field to protect the soil from wind and water erosion. Furthermore, any crop residue left in the field provides a source of organic nitrogen. It could be economical, in the near future, to utilize the energy potential from the excess crop residues on a farm scale in certain regions of the country.

Gasification of crop residues such as corn stover, corn cobs, and wheat straw has been investigated by several groups. Howard et al. (1979) gasified corn cobs and a mixture of corn cobs and manure in a 0.51 m I.D. fluidized bed over a temperature range of 773 to 1273 K. They found that the gas yield (weight percent) increased linearly with temperature up to 1173 K where in levelled out at 60%. Liquid yield decreased rapidly with temperature to 15% at 973 K and then decreased slowly to 5% at 1273 K. Gasification of corn stover in a 0.23 m I.D. fluidized bed reactor was studied by Raman et al. (1980b). They found that the concentration of H_2 in the dry produced gas increased from 35 to 41%, and concentration of CO varied from 20 to 29% over a temperature range of 840 to 1020 K. Concentration of CH_4 was virtually constant at 10%. The higher heating value of the dry produced gas increased from 13 MJ/m^3 at 840 K to 16.5 MJ/m^3 at 930 K and then decreased to 14 MJ/m^3 at 1020 K. Produced gas yield increased from 0.38 to

0.82 m^3/kg DAF over the temperature range studied. They did not report the mass yield of the produced gas, energy recovery, or carbon conversion. Also, no statistical analysis was performed on their data. Walawender et al. (1983) using the same reactor gasified wheat straw over a temperature range of 825 to 1030 K. Produced gas yield increased with temperature from 0.24 to 0.73 m^3/kg DAF, and gas heating value was parabolic with temperature and ranged from 12.5 to 13.5 MJ/m^3 with a maximum of 16.3 MJ/m^3 at 945 K. The concentration of H_2 in the produced gas increased from 11.6 to 31.5%; while the concentration of CH_4 ranged from 10.6 to 12.0% with a maximum of 13% at 950 K. Beck et al. (1981) gasified corn stover in a pilot scale, countercurrent, fluidized bed reactor. They used a temperature range of 873 to 1073 K and found that off-gas yield varied from 0.8 to 1.5 m^3/kg DAF.

The use of producer gas from crop residue gasification as an internal combustion engine fuel has been studied by Parke et al. (1981). They conducted preliminary tests with an internal combustion engine powered by producer gas obtained from corn stover gasified in a fluidized bed. In this study, maximum engine power for the producer gas was about 50% less than maximum engine power for natural gas. In addition, the use of gas produced from corn cobs in an up-draft gasifier to dry grain has been researched by DeKalb (Bozdech 1980). They presently have gasifiers operating at several plants.

The objectives of this study were to examine the feasibility of gasifing corn stover in a fluidized bed reactor and to determine the effect of reactor temperature on produced gas composition, higher heating value, yield, product mass distribution, energy recovery, and carbon conversion.

EXPERIMENTAL

Facilities

The pilot scale gasifier consisted of the following major components: 1) reactor, 2) feed bin and screw feeder, 3) cyclone separator, 4) venturi scrubber, 5) afterburner, 6) control panel, and 7) gas sampling system. Figure 1 shows a schematic diagram of the pilot plant.

The reactor body was made of 330 R stainless steel and had a 0.23 m I.D. (9 in) which expanded to 0.41 m I.D. (16 in) in the freeboard section. Hot fluidizing gas was generated in the plenum burner by burning propane under starving air conditions. Water was also injected into the plenum burner to help control the burner temperature and to provide additional fluidizing gas. Gas was distributed through a 3 mm thick perforated distributor plate.

The plate was covered with a layer of limestone gravel. The gravel was used to prevent the sand bed from percolating through the distributor plate holes into the plenum. Ninety percent by weight of the bed was silica sand, approximately 45 kg, with a particle size of -14 to +50 mesh. The balance of the bed was limestone (-20 to +50 mesh) which prevented agglomeration of the sand (Walawender et al., 1981). Supplemental heating was provided by the radiant burner, a jacketed heater. Kao-Wool TM blanket insulation was used to reduce reactor heat loss.

Feed material was introduced just above the sand bed from a pressurized feed bin by a variable speed screw feeder via a 0.075 m (3 in) diameter feed pipe. Feeding was aided by a nitrogen purge which prevented gas back flow and condensation in the feed pipe. A slide valve prevented gas back flow when the screw feeder was not in use.

Off-gas and entrained char were sent through a hot cyclone for char separation and then to a venturi scrubber. There was no appreciable elutriation of sand from the bed. The scrubber quenched the gas and removed most of the tar. Finally, misty gas from the scrubber was incinerated in the afterburner.

Pressure was monitored in the freeboard section and the feed bin. Temperature was monitored in the following locations in the plant facility: 1) plenum burner, 2) radiant burner, 3) bottom of the sand bed, 4) top of the sand bed, 5) freeboard section, and 6) afterburner. These temperatures were recorded by a strip chart recorder located on the control panel. Also mounted on the control panel were rotameters for monitoring the flow of fuel and air into the burners and controls for the burners and screw feeder.

Gas sample ports were located in both the plenum burner and the off-gas line. Samples of either fluidizing gas or off-gas were sent through the sample train and then to the gas chromatograph for monitoring dry gas composition. The gas sampling train consisted of a tar filter, a condenser with condensate trap, a cold trap, and a wet test meter. A more detailed description of the pilot plant facility has been given by Raman et al. (1980a).

Procedure

Experiments were initiated by heating the reactor to the desired operating temperature with the plenum and radiant burners. Several samples of the plenum gas were taken by an on-line gas chromatograph. Water content of the fluidizing gas was calculated from the amount of condensate for a measured volume of plenum gas. When consistent readings of plenum gas composition were obtained from the gas chromatograph, the nitrogen purge was started. Gas samples were then taken to provide the composition of the off-gas with purge included. The amount of nitrogen purge was determined by nitrogen balance between off-gas with purge and fluidizing gas. This calculation usually yielded a value in close agreement with the rotameter reading for nitrogen purge gas. Therefore, leaks in the system were minimal.

After feeding of corn stover into the reactor was started, starting time and flow rates of propane, air, injection water, and nitrogen were recorded along with the temperature and pressure of each stream. Temperature change in the reactor was recorded by the strip chart recorder. In approximately 20 minutes, temperature in the reactor stabilized. Continuous feeding was maintained for approximately one hour, and off-gas samples were taken during this period. Dry off-gas composition was determined by the gas chromatograph. The weight of tar collected on the filter and weight of condensate collected for a measured volume of off-gas were recorded for calculating the total liquid content of the off-gas. Corn stover feed rate was determined from the difference between the initial and final weights of the material in the feed hopper. Samples of the feed and char were reserved for analysis. Char production rate was determined indirectly by an ash balance between the feed and char. A more detailed description of the procedure has been given by Raman et al. (1980a).

Chemical Analysis

An Applied Automation (optichron 2100) on-line gas chromatograph was used to determine compositions of the dry fluidizing gas and dry off-gas. Gas components determined were H_2, CO, CO_2, CH_4, N_2, O_2, C_2H_4, C_2H_6, C_3H_6, C_3H_8, and C_4. A Perkin-Elmer model 240b elemental analyzer was used to conduct ultimate analyses of the feed and char samples. Moisture and ash content of the feed and char samples were determined by standard ASTM procedures.

Feedstock

Corn stover used in the experiments was ground in a hammer mill to pass a 0.64 cm (¼ in) screen. The average elemental make up of the dry corn stover, based on 24 samples, was 39.0% Carbon, 5.1% Hydrogen, 44.2% Oxygen, 1.5% Nitrogen, and 10.3% ash. As-received moisture content of corn stover was 9.0%. The standard deviation of the elemental composition was 3.66% for Carbon, 0.58% for Hydrogen, 3.98% for Oxygen, 0.35% for Nitrogen, 2.18% for ash, and 0.62% for moisture content. Higher heating value of the feed, calculated by the Dulong formula using average elemental analysis, was 14.3 MJ/kg DAF (6,130 BTU/lb DAF).

CALCULATIONS

Material balance calculations were performed on both the plenum and reactor sections. First, plenum gas analysis coupled with a nitrogen balance based on the metered combustion air allowed dry plenum gas rate to be calculated. The amount of condensate per unit volume of dry plenum gas was used to calculate the water input rate. Then a material balance on the reactor section was performed. Dry off-gas rate was calculated from the analysis of dry off-gas coupled with a nitrogen balance. Condensate rate was calculated from the mass of tar collected on the filter plus the mass of condensate per unit volume of dry off-gas. Feed rate was determined from the weight of feed introduced over the course of the experiment, and char rate was determined by ash balance between the feed and char. It was assumed that the amount of nitrogen produced in the reactor was negligible, and that the plenum gas components did not react in the gasifier. The above items allowed the evaluation of the material balance closure.

The produced gas yield, composition, and heating value were determined from the above information. Produced gas was defined as the difference between off-gas and the combined flow of plenum gas and nitrogen purge. Produced gas volumetric yield was calculated from the difference in the volumetric rates of these streams divided by the dry ash-free (DAF) feed rate. The basis for the volumetric calculations was 289 K and 101.3 kPa. Produced gas composition was calculated from the difference between the components in these streams followed by normalization. Higher heating value of the produced gas was calculated from produced gas composition and the standard heats of combustion of the individual gas components.

Energy recovery, defined as the ratio of the higher heating value of the gas produced per unit mass of DAF feed to the standard heat of combustion of a unit mass of DAF feed, was calculated from the product of the volumetric gas yield and gas heating value divided by the heat of combustion of dry ash-free feed. Carbon conversion represents the percentage of carbon in the produced gas with respect to carbon in the feed. Carbon conversion was calculated by multiplying the stoichiometric coefficient for carbon with the amount of gas produced for each component, summing, and dividing by the amount of carbon in the feed. The mass distribution of the feed into gas, liquid, and char was calculated from the material balance information on a dry ash-free basis. Mass of dry produced gas and mass of dry char were extracted from the material balance, and liquid yield was determined by difference. Raman et al. (1980a) has given additional details on the material balance calculations.

RESULTS

The main objective of this study was to determine the dependence of produced gas composition, higher heating value, and volumetric gas yield; product mass distribution; energy recovery; and carbon conversion of the gas produced from corn stover on the gasifier operating temperature. Operating temperature was defined as the average of the freeboard temperature and the temperature at the top of the sand bed. All experimental data reported had material balance closures between 85 and 100%.

Statistical Analysis System (SAS) computer programs released by SAS Institute Incorporated were used to find the "best fit" model for the dependent variables, gas composition, product mass distribution, higher heating value, produced gas volumetric yield, energy recovery, and carbon conversion as functions of temperature. First, second, and third order polynomial models were fitted to the data points. Criteria used to select the best fitting model were model F-test, parameter significance level, and multiple correlation coefficient, R^2. Multiple correlation coefficient expresses the fraction of data explained by the model. All of the models selected were significant at the 0.004 probability level, and all model parameters selected were significant at the 0.013 probability level.

The results of the statistical analysis is summarized in Table I. Dependent variables are noted by y, correlation coefficient by R^2, F-test value by F-value, and coefficients for the selected regression models by b_0, b_1, and b_2. The analysis was based on 86 data points.

Produced Gas Composition

The major components of the gas were CO, CO_2, H_2, and CH_4. All of the major components were best described by second order polynomials except CO, which was virtually independent of temperature. Figure 2 shows the concentrations of the major components versus temperature. The 95% confidence limits are also shown for several points. Carbon monoxide varied inversely with temperature. Concentration of CO decreased from 32.4% at 820 K to 28.4% at 1020 K. Concentration of CO_2 varied from 42.9% at 820 K to 21.3% at 1020 K and passed through a minimum of 20.7% at 990 K. On the other hand, the H_2 concentration increased from 11.9% at 820 K to 36.4% at 1020 K. Concentration of CH_4 varied from 10.9% at 820 K to 10.4% at 1020 K with a maximum of 11.5% at 904 K. The minor gas components which made up the balance of produced gas were C_2H_4, C_2H_6, and C_3H_6. Concentrations of all three minor components were best described by second order polynomials. Figure 3 shows the concentrations of minor components versus temperature along with the 95% confidence limits. Concentration of ethylene (C_2H_4) varied from 0.7% at 820 K to 3.8% at 1020 K with a maximum of 3.8% at 1016 K. The ethane (C_2H_6) concentration varied from 0.9 to 0.6% in the temperature range studied and had a maximum of 1.0% at 892 K. Finally, the propylene (C_3H_6) concentration varied from 1.0% at 820 K to 0.4% at 1020 K with a maximum of 1.5% at 899 K.

Produced Gas Yield and Heating Value

Figure 4 shows the plot of produced gas volumetric yield and higher heating value versus temperature along with the 95% confidence limits. The produced gas volumetric yield varied linearly with temperature. Gas yield increased from 0.17 m^3/kg DAF at 820 K to 0.74 m^3/kg DAF at 1020 K. Higher heating value for produced gas was best described by a second order polynomial. Heating value ranged from 11.7 MJ/m^3 at 820 K to a maximum of 15.6 MJ/m^3 at 950 K then back to 14.6 MJ/m^3 at 1020 K.

Product Mass Distribution

The mass distribution into gas, liquid, and char as a function of temperature is shown in Figure 5 along with the 95% confidence limits. Gas mass percentage was directly proportional and liquid mass percentage was inversely proportional to temperature. Char mass percentage was independent of temperature. Gas mass percentage increased from 25% at 820 K to 64% at 1020 K; whereas, liquid mass percentage decreased from 68 to 29% over the temperature range studied. Char mass percentage was constant at 7%.

Energy Recovery and Carbon Conversion

Energy recovery was defined as the ratio of the higher heating value of produced gas to the standard heat of combustion of the DAF feed. Energy recovery varied linearly with temperature. Figure 6 shows energy recovery as a function of temperature along with the 95% confidence limits. Energy recovery increased from 14.9% at 820 K to 81.0% at 1020 K. Energy recovery continued to increase after 955 K despite the drop in higher heating value of produced gas because the produced gas volumetric yield continued to rise. Carbon conversion was defined as the percentage of carbon in produced gas relative to the carbon in feed. The plot of carbon conversion versus temperature is also shown in Figure 6 along with the 95% confidence limits. Carbon conversion was directly proportional to temperature and increased from 20% at 820 K to 63% at 1020 K. The increase in carbon conversion was also a direct result of the continuous increase in produced gas yield.

DISCUSSION

The results for corn stover from this work are compared to those for corn stover obtained by Raman et al. (1980b) and for wheat straw obtained by Walawender et al. (1983). The produced gas higher heating value obtained in this work is in agreement with that obtained by Raman et al. (1980b); but the produced gas volumetric yield in the former is slightly lower than that obtained in the latter. However, the data of Raman et al. (1980b) have a large standard deviation (±0.26 m^3/kg DAF) while the present data, which have a relatively small standard deviation of 0.06 m^3/kg DAF, lie within the bounds of the data of Raman et al. (1980b). The CH_4 concentration in the produced gas is similar, but the H_2, CO, and CO_2 concentrations are different. Concentration of H_2 is significantly lower at the low temperatures and slightly lower at high temperatures compared to the results of Raman et al. (1980b); whereas, concentration of CO_2 is higher at low temperatures and lower at high temperatures. Concentration of CO was always higher. Differences in the results could be due to differences in feed material composition and/or the method of gas analysis. Energy recovery and product mass distribution for the data of Raman et al. (1980b) were reported by Walawender et al. (1983). Energy recovery is similar; however, the produced gas mass percent is larger, and liquid mass percent is smaller than the present work, especially at higher temperatures. The char mass percent is similar.

Regression models for the data of this work are compared to those for wheat straw found by Walawender et al. (1983). Regression models for the produced gas volumetric yield, higher heating value, and mass distribution (gas, liquid, and char) are not significantly different at the 5% probability level. Also, regression models for the concentrations of CO_2 and CH_4 in the produced gas are not significantly different. Regression lines for energy recovery are significantly different at a 1% probability level, possibly because the heats of combustion for corn stover and wheat straw are different (14.3 and 14.8 MJ/kg DAF respectively). Energy recovery for corn stover is lower at low temperatures and higher at temperatures above 910 K. Since the variance and degrees of freedom are different for corn stover and wheat straw, Satterthwaite's adjusted t-test (Snedecor and Cochran, 1980) has been used to test slopes and intercepts of the regression models. The calculated t-value for each test is presented in Table II. The major gas components H_2 and CO could not be tested because models for corn stover are of different order than models for wheat straw. In the case of H_2, the quadratic fit was better than the linear fit for corn stover data. When the linear fit is used for corn stover, the regression lines are not significantly different. The CO concentration exhibited the most experimental variability in the data compared to the other major gas components for both corn stover and wheat straw. Wide scatter in the data points could be the reason that the regression models for CO are different for corn stover and wheat straw.

CONCLUSIONS

Corn stover was gasified in a 0.23 m I.D. fluidized bed reactor. Gas yield, energy recovery, and carbon conversion increased with increasing reactor temperature; whereas, liquid yield decreased with increasing reactor temperature. Concentration of hydrogen in the produced gas increased while concentrations of carbon monoxide and carbon dioxide decreased with increasing temperature. Concentration of methane in the produced gas passed through a maximum of 11.5% at 904 K. Higher heating value of the produced gas passed through a maximum of 15.6 MJ/m^3 at 950 K. Produced gas volumetric yield increased from 0.17 to 0.74 m^3/kg DAF as temperature increased from 820 to 1020 K.

Produced gas volumetric yield, higher heating value, energy recovery, and CH_4 concentration from corn stover gasification results are similar to the results obtained by Raman et al. (1980b) for gasification of corn stover. However, the results for H_2, CO, and CO_2 concentrations in the produced gas are different from theirs. Finally, the results for corn stover gasification agree with the results obtained for wheat straw gasification by Walawender et al. (1983). Results for produced gas volumetric yield, mass distribution (gas, liquid, and char), higher heating value, CO_2 concentration, and CH_4 concentration are not significantly different at the 5% level of probability. Regression lines for energy recovery are significantly different at the 1% probability level.

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Table I
Regression Models

Dependent Variable y	R^2	F-value	Regression Model Parameters $y = b_0 + b_1T + b_2T^2$		
			b_0	b_1	b_2
Gas Yield (m^3/kg)	0.84	443.5	-2.170	0.00285	
Higher Heating Value (MJ/m^3)	0.75	124.9	-179.4	0.4083	-2.138×10^{-4}
Produced Gas Composition (mole %)					
H_2	0.85	237.2	-386.5	0.7780	-3.563×10^{-4}
CO_2	0.85	236.4	769.7	-1.512	7.634×10^{-4}
CO	0.10	8.8	49.03	-0.02022	
CH_4	0.13	6.0	-57.39	0.1524	-8.432×10^{-5}
C_2H_4	0.83	199.9	-77.92	0.1608	-7.918×10^{-5}
C_2H_6	0.42	30.3	-22.81	0.05347	-2.997×10^{-5}
C_3H_6	0.80	168.4	-57.37	0.1310	-7.285×10^{-5}
Energy Recovery (%)	0.87	555.9	-256.0	0.3304	
Carbon Conversion (%)	0.71	202.2	-158.6	0.2175	

Table I
(continued)

Dependent Variable			Regression Model Parameters $y = b_0 + b_1T + b_2T^2$		
y	R^2	F-value	b_0	b_1	b_2
Product Mass Distribution (%)					
Gas	0.76	264.5	-140.6	0.2015	
Liquid	0.69		233.9	-0.2015	
Char			6.71		

Table II
t-Values for Corn Stover and Wheat Straw

$t_{.05,50} = 2.01$ $t_{.01,50} = 2.68$

Dependent Variable		Parameter Value Corn Stover	Wheat Straw	Calculated t-Value
Gas Yield (m^3/kg)	b_0	-2.170	-1.708	-1.37
	b_1	0.002851	0.002362	1.39
Higher Heating	b_0	-179.4	-216.5	0.97
Value (MJ/m^3)	b_1	0.4083	0.4943	-1.05
	b_2	-2.138×10^{-4}	-2.627×10^{-4}	1.13
Produced Gas Composition (mole %)				
H_2[1]	b_0	-86.91	-68.96	-1.46
	b_1	0.1237	0.09748	2.02
CO_2	b_0	796.7	633.2	0.69
	b_1	-1.512	-1.268	-0.58
	b_2	7.634×10^{-4}	6.66×10^{-4}	0.44
CH_4	b_0	-57.39	-92.81	0.70
	b_1	0.1524	0.2188	-0.62
	b_2	-8.432×10^{-5}	-1.141×10^{-4}	0.52
Energy	b_0	-256.0	-151.8	-2.81
Recovery (%)	b_1	0.3304	0.2177	2.91
Product Mass Distribution (%)				
Gas	b_0	-140.6	-135.4	-0.15
	b_1	0.2015	0.2024	-0.03
Liquid	b_0	233.9	224.8	0.27
	b_1	-0.2015	-0.1992	0.06
Char	b_0	6.71	7.43	-0.28

[1]The linear equation for Corn Stover was used instead of the second order equation.

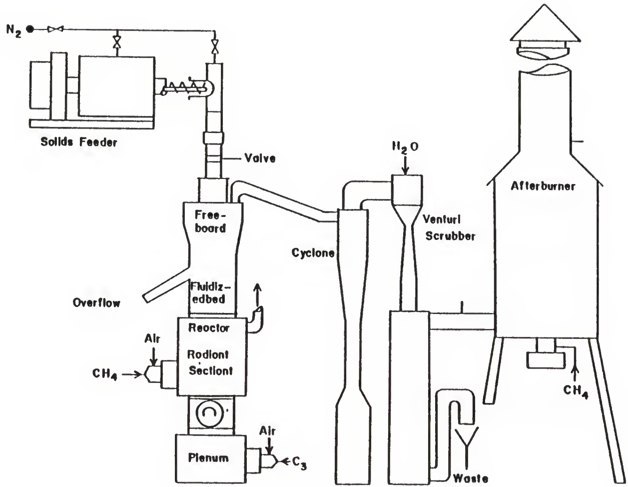

Figure 1. Schematic diagram of the Pilot Plant.

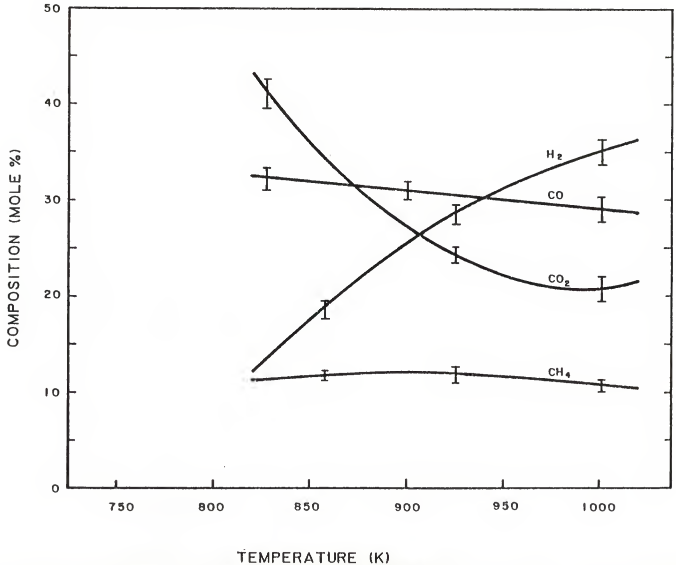

Figure 2. Gas Composition vs. Temperature.

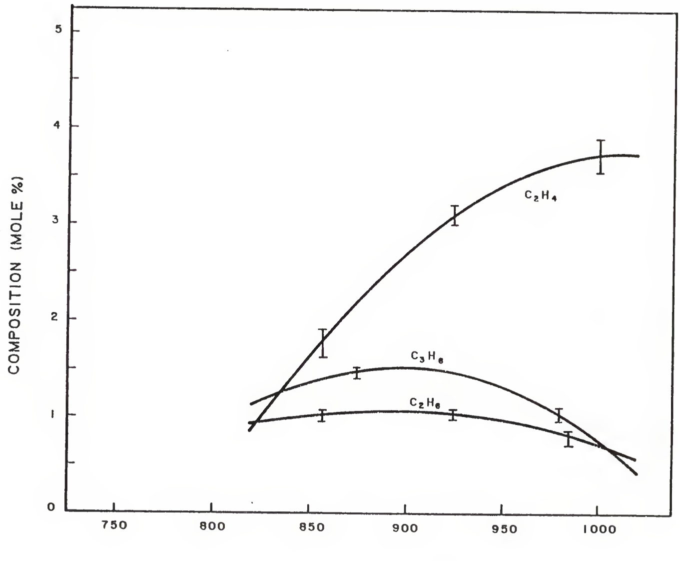

Figure 3. Minor Gas Components vs. Temperature.

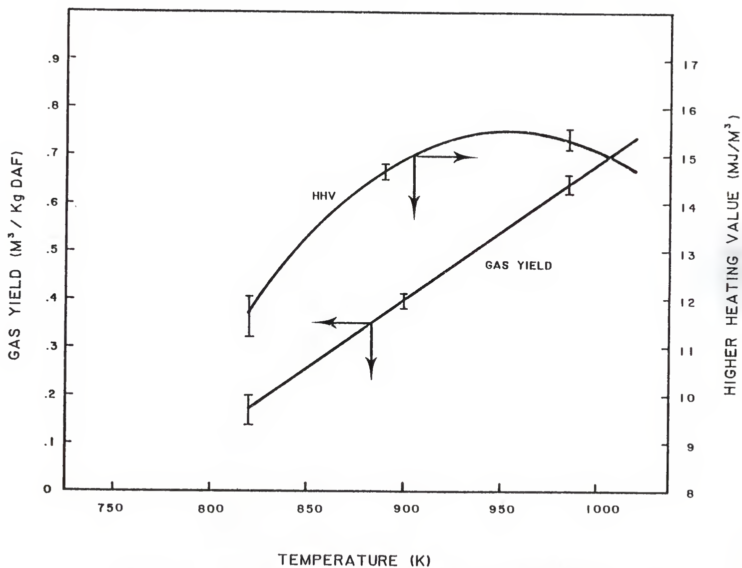

Figure 4. Gas Yield and Higher Heating Value vs. Temperature.

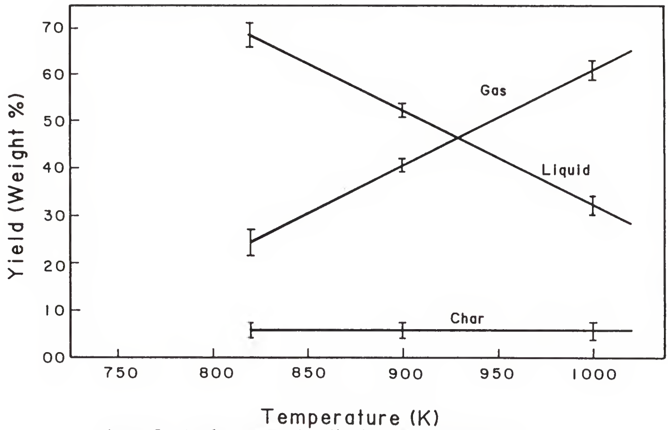

Figure 5. Product Mass Distribution vs. Temperature.

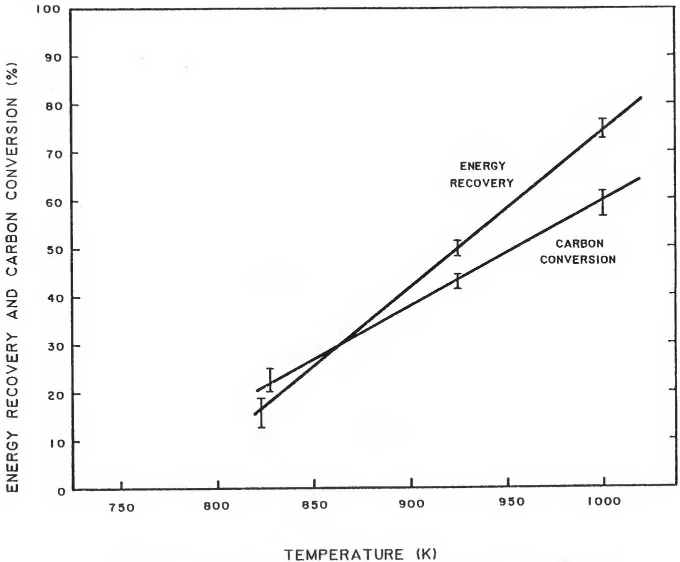

Figure 6. Energy Recovery and Carbon Conversion vs. Temperature.

Chapter IV

SORGHUM STOVER GASIFICATION

INTRODUCTION

Considerable effort has been made over the past ten years to develop biomass energy resources. Sorghum stover is one such source of biomass. Approximately 2.2×10^{10} kg (24.3 million tons) of sorghum stover is produced annually in the United States with about 27% of this amount being grown in Kansas (Agricultural Statistics, 1982). Sorghum stover alone would not play a major role in the energy picture; however, all of the crop residues added together constitute a substantial amount of energy.

This study reports on the gasification of sorghum stover in a fluidized bed. One objective of this study was to determine the effect of reactor temperature on produced gas composition, higher heating value, yield, product mass distribution, energy recovery and carbon conversion. A second objective was to compare the gasification results with results from gasification of wheat straw (Walawender et al., 1983) and corn stover (Chapter 3) in the same reactor.

EXPERIMENTAL

Operating Conditions

The experimental facilities, procedure, and calculations are the same as discussed in Chapter III. The reactor was operated over a temperature range of 740 to 1020 K. The gas superficial velocity varied from 0.29 to 0.51 m/s.

Feedstock

Sorghum stover used in the experiments was ground in a hammer mill to pass a 0.64 cm (1/4 in) screen. The average elemental make up of the dry sorghum stover, based on 23 samples, was 35.4% Carbon, 4.7% Hydrogen, 46.7% Oxygen, 2.2% Nitrogen, and 11.0% ash. As-received moisture content of sorghum stover was 8.2%. The standard deviation for the elemental components was 2.88% for Carbon, 0.37% for Hydrogen, 3.55% for Oxygen, 0.64% for Nitrogen, 1.05% for ash, and 0.84% for moisture content. The higher heating value of the feed, calculated from the Dulong formula using the average elemental analysis, was 11.7 MJ/kg DAF (5,028 BTU/lb DAF).

RESULTS

The main objective of this study was to determine the effect of temperature on the gas composition, product mass distribution, higher heating value of the gas, produced gas volumetric yield, energy recovery, and carbon conversion for the gasification of sorghum stover. The temperature used in the figures to follow was the average of freeboard temperature and the temperature at the top of the sand bed. The data presented here had material balance closures between 87 and 110%.

Statistical Analysis System (SAS) computer programs released by SAS Institute Incorporated were used to find the "best fit" model for the dependent variables; namely gas composition, product mass distribution, higher heating value, produced gas volumetric yield, energy recovery, and carbon conversion as functions of temperature. First, second, and third order polynomial models were fitted to the data points. The criteria used to select the best fitting model were model F-test, parameter significance level, and multiple correlation coefficient, R^2. The multiple correlation coefficient expresses the fraction of data explained by the model. All of the models chosen were significant at the 0.0001 probability level, and all model parameters chosen were significant at the 0.01 probability level.

The results of the statistical analysis are summarized in Table I. The dependent variables are noted by y, the correlation coefficient by R^2, F-test value by F-value, and coefficients for the selected regression models by b_0, b_1, b_2, and b_3. The number of samples used to fit the regression models was 107.

Produced Gas Composition

The variations in the concentrations of H_2, CO, CO_2, and CH_4 with temperature are shown in Figure 1. The 95% confidence limits are also shown at several temperatures. These four components comprised approximately 95% of the produced gas. Three of the components (H_2, CO_2, and CH_4) were best described by second order polynomials; the CO concentration was best described by a third order polynomial. Concentration of CO did not vary appreciably and ranged from 29.6% at 750 K to 25.5% at 1020 K with a minimum of 25.0% at 855 K. The CO_2 concentration decreased with temperature from 53.0% at 750 K to 20.0% at 1020 K. Concentration of H_2, on the other hand, increased with temperature from 7.3% at 750 K to 37.4% at 1020 K. Concentration of CH_4 did not vary appreciably and ranged from 8.6 to 10.2% with a maximum of 10.5% at 946 K. Minor components of the produced gas were ethylene (C_2H_4), ethane (C_2H_6), and propylene (C_3H_6). Variations in the concentrations of minor produced gas components with temperature are shown in Figure 2 along with the 95% confidence limits. Ethylene and propylene concentrations were both best described by second order polynomials; ethylene concentration increased from 0.4% at 750 K to 3.9% at 1020 K. Propylene concentration varied from 0.54 to 0.44% in the same temperature range with a maximum of 1.4% at 881 K. Concentration of ethane decreased linearly with temperature from 1.4% at 750 K to 0.7% at 1020 K.

Produced Gas Yield and Heating Value

Figure 3 shows the variations of the produced gas volumetric yield and higher heating value with temperature along with the 95% confidence limits. Volumetric yield and higher heating value were best described by second and third order polynomials respectively for the temperature range of 750 to 1020 K. Gas volumetric yield increased from 0.19 to 0.91 m^3/kg DAF. Higher heating value varied from 9.97 to 13.96 MJ/m^3 passing through a maximum of 14.90 MJ/m^3 at 960 K. The arithmetic average heating value was 13.36 MJ/m^3 for the entire temperature range.

Product Mass Distribution

The variations in the mass distribution of dry ash-free feed into gas, liquid, and char with temperature are presented in Figure 4 along with the 95% confidence limits. Over the temperature range of 750 to 1020 K, gas mass yield increased linearly from 21.9 to 71.9%. Char mass yield decreased slowly with temperature from 750 to 850 K and was virtually constant for temperatures above 850 K. This trend was not noticed in the results for corn stover gasification; however, reactor temperature for corn stover gasification was never less than 820 K. The devolatilization of sorghum stover was not complete in the low temperature range of the experiments. Hence, the amount of char diminished with increasing temperature. The char yield decreased linearly from 13.4% at 750 K to 9.7% at 850 K and then remained constant at 9.7%. Antal et al. (1978) obtained similar results. They stated that pyrolysis occured up to 873 K. Above 873 K, only cracking and reforming of the volatile matter occured. Because liquid yield was determined by difference, it also had a break point at 850 K. Liquid yield decreased linearly with temperature from 64.6% at 750 K to 49.8% at 850 K and then decreased linearly to 18.3% at 1020 K.

Energy Recovery and Carbon Conversion

Figure 5 shows the variations in both energy recovery and carbon conversion as functions of temperature along with the 95% confidence limits. Energy recovery was best described by a linear function of temperature. Energy recovery increased from 4.7% at 750 K to 103% at 1020 K. The low values for energy recovery at low temperatures are caused by high CO_2 concentration, low H_2 concentration, and low gas yield. Gas yield and heating value for sorghum stover were similar to those for corn stover; however, the heat of combustion of sorghum stover was significantly smaller than that of corn stover. The small heat of combustion for sorghum stover caused the energy recovery to be large. This was evident at high temperatures. Carbon conversion also increased linearly with temperature from 17.4 to 73.8% over the same temperature range.

DISCUSSION

The results for the sorghum stover data agree quite well with the results for other crop residues gasified in the same reactor. A t-test is used to compare the regression model parameters for sorghum stover with their counter parts for wheat straw (Walawender et al., 1983) and corn stover (Chapter III). Satterthwaite's adjusted t-test (Snedecor and Cochran, 1980) has been used to calculate the t-values because the variance and degrees of freedom for each data set are different. Regression model parameters and calculated t-values are presented in Table II and Table III.

Regression model parameters for H_2 and CH_4 concentrations in the produced gas, carbon conversion, and energy recovery for sorghum stover are not significantly different from respective regression model parameters for corn stover at the 5% probability level. Thus, the lines for these variables lie within the variance of the respective lines for the two feedstocks. Also, regression models for the product mass distribution (gas, liquid, and char) are not significantly different at the 5% probability level. The regression models used in the t-tests for sorghum stover char and liquid mass yields are based only on temperatures greater than 850 K. Regression models for CO_2 concentration are significantly different at the 1% probability level. The decrease in CO_2 concentration with increasing temperature between approximately 820 and 930 K for sorghum stover is less than that for corn stover; however, the general trends are similar for each feed. Regression models for the produced gas volumetric yield, higher heating value, and CO concentration for sorghum stover are of different order than the models for corn stover. The quadratic fit for volumetric gas yield is better than the linear fit for the sorghum stover data. When the linear fit is used for the sorghum stover data, the slopes for the regression models for the two feed materials are not significantly different at the 5% probability level.

The present data are compared with the wheat straw data of Walawender et al. (1983). Regression models for the product mass distribution (gas, liquid, and char) and CH_4 concentration of the present work are not significantly different at the 5% probability level from the wheat straw data of Walawender et al. (1983). Again, the regression models used in the t-test for sorghum stover char and liquid mass yields are based only on the data obtained in the temperature range of 850 to 1020 K. In the case of CO_2 concentration, the intercept and first order parameter with respect to temperature are not significantly different, but the second order parameter is significantly different at the 5% probability level. Also, the CO concentration and energy recovery regression models are significantly different at the 1% probability level. In the case of energy recovery, the energy recovery for sorghum stover is always larger and increases with temperature at a significantly greater rate than that for wheat straw. the regression models for sorghum stover gas volumetric yield, higher heating value, and H_2 concentration are of different order from those for wheat straw; thus, they could not be compared statistically. When linear fits for volumetric yield and H_2 concentration are used for the sorghum stover data, regression models for both feed materials are not significantly different at the 5% probability level.

Statistical Model Building

The gasification of biomass is a complex process dependent upon many factors including temperature. Therefore, the stepwise model building procedure was used to attempt to identify additional factors affecting the produced gas composition, volumetric yield, higher heating value, and mass distribution. A simple linear model was used with reactor temperature, superficial gas velocity, moisture content of the feed, and carbon content of the feed as the independent variables. Later, temperature squared and temperature cubed were added as independent variables. Of the independent variables used, temperature had the most profound effect on the dependent variables. No distinct trends among the other independent variables was observed; however, there was a large amount of variability in the data with respect to the other independent variables tested because the experiments were not specifically designed for model building.

CONCLUSIONS

Sorghum stover was gasified in a 0.23 m I.D. fluidized bed reactor over a temperature range of 750 to 1020 K. Produced gas volumetric yield increased from 0.19 to 0.91 m^3/kg DAF, and gas higher heating value varied from 9.97 to 14.9 MJ/m^3. The H_2 concentration in the produced gas increased, and CO_2 concentration decreased with increasing temperature. Methane concentration stayed close to the average value of 10.0%. Energy recovery, carbon conversion, and gas mass percentage all increased linearly with increasing temperature; liquid and char mass percentages both decreased linearly with increasing temperature from 750 to 850 K. Above 850 K, liquid yield decreased with increasing temperature, but char yield remained constant.

Results for sorghum stover are statistically similar to those for corn stover and wheat straw. In comparison with corn stover, regression models for the concentrations of H_2 and CH_4 in the produced gas, carbon conversion, energy recovery, and product mass distribution (gas, liquid, and char) are not significantly different at the 5% probability level. In comparison with the wheat straw data obtained by Walawender et al. (1983), regression models for the concentrations of CO_2 and CH_4 in the produced gas and product mass distribution (gas, liquid, and char) are not significantly different.

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Table I
Regression Models

Dependent Variable			Regression Model Parameters $y = b_0 + b_1T + b_2T^2 + b_3T^3$			
y	R^2	F-value	b_0	b_1	b_2	b_3
Gas Yield (m^3/kg)	0.87	351.2	3.462	-0.009511	6.867×10^{-6}	
Higher Heating Value (MJ/m^3)	0.92	374.7	325.3	-1.196	0.001481	-5.958×10^{-7}
Produced Gas Composition (mole %)						
H_2	0.90	473.9	-258.0	0.5316	-2.373×10^{-4}	
CO_2	0.93	710.6	300.0	-0.4818	2.032×10^{-4}	
CO	0.23	10.5	1093.5	-3.526	0.003865	-1.406×10^{-6}
CH_4	0.29	21.7	-32.62	0.09108	-4.814×10^{-5}	
C_2H_4	0.89	405.7	-27.75	0.05567	-2.417×10^{-5}	
C_2H_6	0.82	468.7	3.614	-0.002899		
C_3H_6	0.79	190.0	-36.54	0.08608	-4.885×10^{-5}	
Energy Recovery (%)	0.87	679.4	-268.6	0.3644		
Carbon Conversion (%)	0.73	284.7	-139.4	0.2090		

Table I
(continued)

Dependent Variable			Regression Model Parameters $y = b_0 + b_1T + b_2T^2 + b_3T^3$			
y	R^2	F-value	b_0	b_1	b_2	b_3
Product Mass Distribution (%)						
Gas	0.78	380.7	-117.0	0.1852		
Liquid ($T < 850$)	0.35		175.4	-0.1477		
Liquid ($T > 850$)	0.67		207.2	-0.1852		
Char ($T < 850$)	0.22	8.1	41.52	-0.03746		
Char ($T > 850$)			9.73			

Table II
t-Values for Corn Stover and Sorghum Stover

$t_{.05,80} = 1.99$ $t_{.01,80} = 2.64$

Dependent Variable		Parameter Value Corn Stover	Sorghum Stover	Calculated t-Value
Gas Yield[1] (m^3/kg)	b_0	-2.170	-1.816	-2.22
	b_1	0.002851	0.002564	1.64
Produced Gas Composition (mole %)				
H_2	b_0	-386.5	-258.0	1.51
	b_1	0.7780	0.5316	1.31
	b_2	-3.563×10^{-4}	-2.373×10^{-4}	-1.16
CO_2	b_0	796.7	300.0	5.82
	b_1	-1.512	-0.4818	-5.81
	b_2	7.634×10^{-4}	2.032×10^{-4}	5.75
CH_4	b_0	-57.39	-32.62	-0.98
	b_1	0.1524	0.09108	1.10
	b_2	-8.432×10^{-5}	-4.814×10^{-5}	-1.19
Energy	b_0	-256.0	-268.6	0.70
Recovery (%)	b_1	0.3304	0.3644	-1.71
Carbon	b_0	-158.6	-139.4	-1.05
Conversion (%)	b_1	0.2175	0.2090	0.43
Product Mass Distribution (%)				
Gas	b_0	-140.6	-117.0	-1.66
	b_1	0.2015	0.1852	1.05
Liquid[2]	b_0	233.9	207.2	1.29
	b_1	-0.2015	-0.1852	-0.73
Char[2]	b_0	6.71	9.73	-0.94

[1]The linear equation for Sorghum Stover was used instead of the second order equation for comparison.

[2]The regression model parameters for temperatures greater than 850 K were used for Sorghum Stover.

Table III
t-Values for Sorghum Stover and Wheat Straw

$t_{.05,50} = 2.01$ $t_{.01,50} = 2.68$

Dependent Variable		Parameter Value: Sorghum Stover	Parameter Value: Wheat Straw	Calculated t-Value
Gas Yield[1] (m^3/kg)	b_0	-1.816	-1.708	-0.33
	b_1	0.002564	0.002362	0.59
Produced Gas Composition (mole %)				
H_2[1]	b_0	-75.61	-68.96	-0.58
	b_1	0.1143	0.09748	1.39
CO_2	b_0	300.0	633.2	-1.78
	b_1	-0.4818	-1.268	1.97
	b_2	2.032×10^{-4}	6.66×10^{-4}	-2.19
CH_4	b_0	-32.62	-92.81	1.29
	b_1	0.09108	0.2188	-1.28
	b_2	-4.814×10^{-5}	-1.141×10^{-4}	1.25
Energy Recovery (%)	b_0	-268.6	-151.8	-3.17
	b_1	0.3644	0.2177	3.78
Product Mass Distribution (%)				
Gas	b_0	-117.0	-135.4	0.56
	b_1	0.1852	0.2024	-0.49
Liquid[2]	b_0	207.2	224.8	0.26
	b_1	-0.1852	-0.1992	0.08
Char[2]	b_0	9.73	7.43	0.69

[1]The linear equation for Sorghum Stover was used instead of the second order equation for comparison.

[2]The regression model parameters for temperatures greater than 850 K were used for Sorghum Stover.

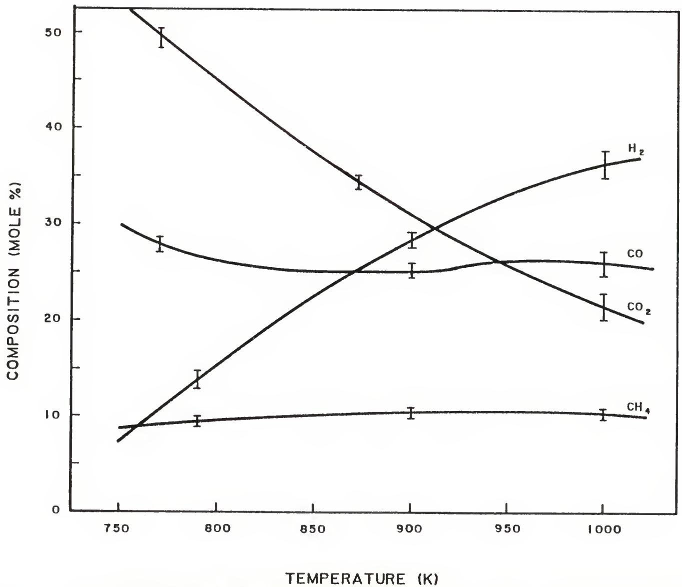

Figure 1. Gas Composition vs. Temperature.

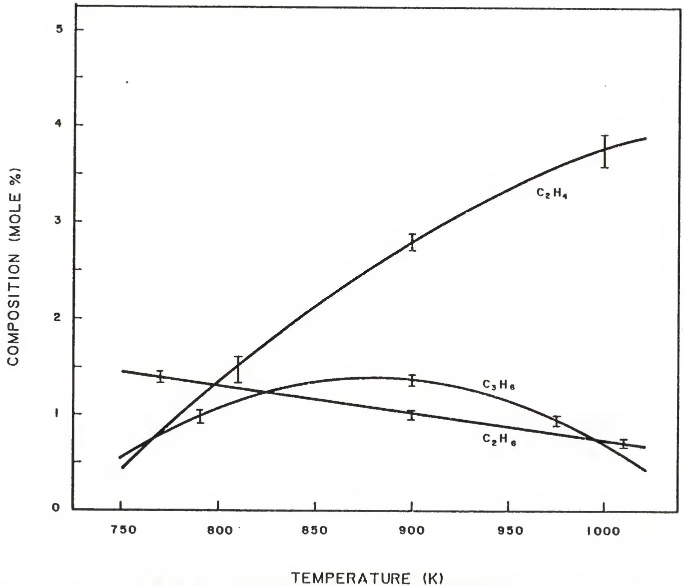

Figure 2. Minor Gas Components vs. Temperature.

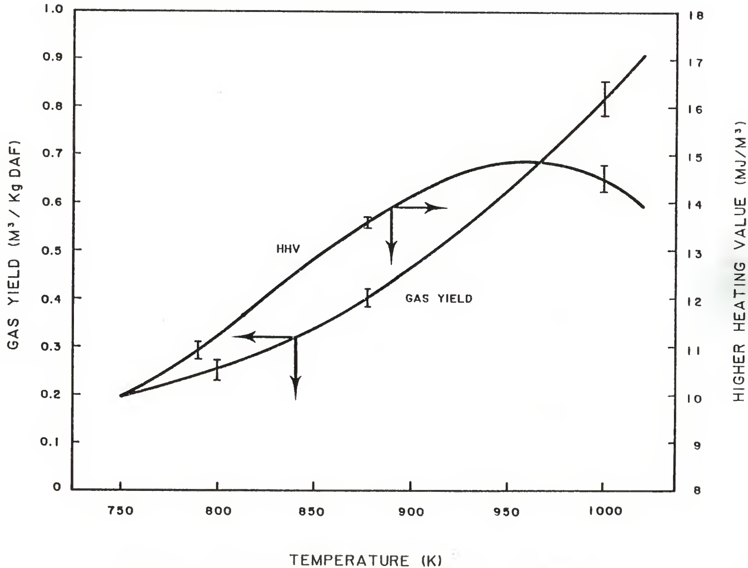

Figure 3. Gas Yield and Higher Heating Value vs. Temperature.

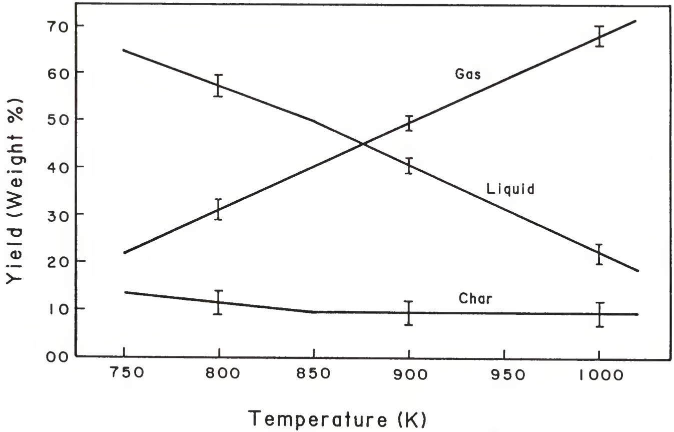

Figure 4. Product Mass Distribution vs. Temperature.

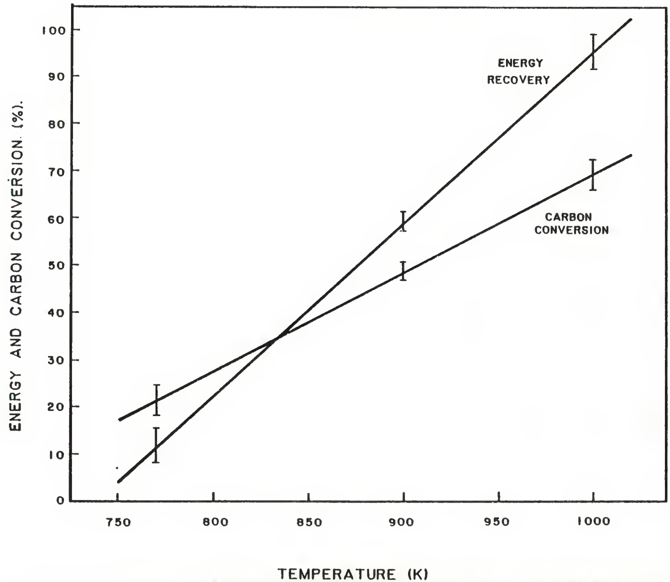

Figure 5. Energy Recovery and Carbon Conversion vs. Temperature.

Chapter V

CONCEPTUAL DESIGN AND ECONOMIC ANALYSIS

INTRODUCTION

Energy is very essential to agricultural productivity. The high cost and fluctuating supply of energy, especially at critical times, has prompted research into alternate energy sources for agriculture. Irrigation pumping and grain drying constitute a major portion of the energy use in agriculture. Eleven states consume 89% of the energy used for irrigation with Texas, Nebraska, Kansas, and Arizona being the top four in energy use (Clark et al., 1978). The major crops being irrigated are corn, grain sorghum, cotton, alfalfa, and wheat in decreasing order. Furthermore, grain drying uses a substantial amount of energy with corn drying making up the largest part of the energy use.

Most irrigation pumps and grain dryers are fueled by natural gas. Producer gas, made from the gasification of crop residue in a fluidized bed reactor, offers an alternative to natural gas. Parke et al. (1981) has demonstrated that the technology for operating an engine with producer gas from a fluidized bed exists. The purpose of this chapter is to present a conceptual design and feasibility analysis for a farm-scale fluidized bed gasifier for producing fuel for irrigation pumping and grain drying.

Alpert et al. (1972) at Stanford Research Institute examined a hypothetical fluidized bed plant to gasify 1.23 Gg (1,358 tons) per day of as-received municipal waste. The capital investment was estimated at 11.7 millon dollars for a dual fluidized bed system and 12.5 million dollars for a single fluidized bed process in 1972 dollars. These capital investments convert to $36,600 and $39,300 per MMBTU/hr output for the dual and single fluidized beds respectively. Break even gas price was estimated at that time to be $0.43/GJ ($0.45/MMBTU) and $0.55/GJ ($0.58/MMBTU) for the dual reactor and single reactor systems respectively. These costs were based on a 20 year plant life.

Engler et al. (1975) at Kansas State University studied the economics of commercial scale manure gasification in a dual fluidized bed. Break even gas price was estimated to be $0.010/m^3 ($0.29/MSCF) or $0.72/GJ ($0.76/MMBTU) for a plant processing 3.63 Gg (4,000 tons) per day of manure with 50% moisture and $0.017/m^3 ($0.47/MSCF) or $1.17/GJ ($1.23/MMBTU) for a facility processing 0.91 Gg (1,000 tons) per day of manure. A substantial drop in gas price was obtained if dry manure was used. Cost estimates were based on 15 year plant life and mid 1974 dollars. The total capital investment was estimated to be 14.9 and 6.5 million dollars for the 3.63 Gg and 0.91 Gg per day plants respectively. This converts to $22,000 per MMBTU/hr output for the large plant and $38,000 per MMBTU/hr output for the small plant. They concluded that only large scale manure gasification processes could approach the point of commercial viability.

Economic analyses on the use of fuel derived from crop residues to supply the energy needs for irrigation pumping and grain drying was performed by Clark et al. (1978). The hypothetical problem involved fueling three 100 kW irrigation pumps. The capital cost estimate for a mass produced, dual fluidized bed facility was estimated to be between 129 and 240 million dollars in 1980 dollars. This converts to $40,000 per MMBTU/hr output. In a comparison between anaerobic digestion, fuel-bed combustion, and fluidized bed gasification, they concluded that fluidized bed gasification was the best process to provide fuel for irrigation pumping and grain drying.

Loewer et al. (1980) studied the economics of corn grain, cobs, and stover gasification as an energy substitute for LP gas in corn grain drying. They determined the break even investment for drying one bushel of U.S.

No. 2 corn grain from 25% moisture to 15.5% moisture. They found that as much as 38.9¢ and 23.5¢ per bushel of grain dried could be invested in gasification equipment when using cobs and stover respectively.

A conceptual design and economic analysis on a fluidized bed wood gasifier was performed by Lian et al. (1982). The proposed plant would gasify 91 Mg (100 tons) per day of sawdust with a 36% moisture content. The total capital investment was estimated to be $874,800 in 1980 dollars. This converts to $25,700 per MMBTU/hr output. The plant was depreciated over a 15 year period with a 48% tax rate. For a 15% rate of return before-tax, cost of the gas was estimated to be $3.48/GJ ($3.66/MMBTU).

DESIGN BASIS

The reports published so far have been based on assumed material balances from limited experimental data. The conceptual design presented in this chapter is based on extensive experimental results for corn stover gasification in a pilot-scale fluidized bed reactor (Chapter III).

Sizing of the farm-scale gasifier is based on the irrigation requirements for a 480 acre farm yielding 11.31 m^3/ha (130 bu/acre). The net crop water required for irrigation pumping is 0.46 m (18 in). A water use efficiency of 70% and a pump efficiency of 75% are assumed. Three irrigation wells are used to supply the water required; each well has a 100 kW engine. A 25% engine efficiency and a 67% gasification efficiency are used. The gasifier will also be used to supply heat for drying corn from 25% to 15% moisture. The system will be operated 1404 hours for irrigation pumping and 311 hours for grain drying a year.

The corn stover feed for the gasifier will be collected from the field. The residue yield is assumed to be 1 kg of residue per kg of grain (1 lb residue per lb grain). It is assumed that as much as 75% of the available corn stover can be collected, and that the storage loss is 15%.

Qualitatively, the overall gasification reaction is as indicated below:

$$\text{Corn Stover} \xrightarrow{\text{Heat}} \text{Gas} + \text{Tar} + \text{Char} + \text{Water}$$

The process heat requirements are supplied by the combustion of tar and char. The amount of char produced from gasification is constant, and the amount of tar produced decreases with increasing temperature. In the pilot plant studies, the composition of corn stover, char, and tar are determined from elemental analyses. Gas composition is determined from gas chromatograph readings. Flow rates of feed, gas, and char are determined from the experimental measurements and calculations. This data is presented in Chapter III. An elemental balance, based on the pilot plant data, is performed to determine the quantities of tar and water. The resulting empirical stoichiometric relations for gasification, char combustion, and tar combustion are given below:

Corn Stover Gasification

$$C_{3.64}H_{5.71}O_{3.06}N_{0.115} \longrightarrow 0.544\ CO_2 + 0.764\ CO + 0.875\ H_2 + 0.288\ CH_4 + 0.095\ C_2H_4 + 0.022\ C_2H_6 + 0.027\ C_3H_6 + \text{Char} + \text{Tar} + 0.674\ H_2O$$

Char Combustion

$$C_{0.476}H_{0.182}O_{0.0464}N_{0.0065} + 0.5\ O_2 \longrightarrow 0.091\ H_2O + 0.476\ CO_2 + 0.0033\ N_2$$

Tar Combustion

$$C_{1.25}H_{0.605}O_{0.487}N_{0.108} + 1.66\ O_2 \longrightarrow 0.303\ H_2O + 1.25\ CO_2 + 0.054\ N_2$$

The above stoichiometry is used to conduct the energy balance on the reactor and to determine the char and tar requirements for the process to be self sustaining. Standard heats of combustion for corn stover, char, and tar are calculated using the Dulong formula; they are 14.3 MJ/kg DAF (6,130 BTU/lb DAF), 30.8 MJ/kg DAF (13,200 BTU/lb DAF), and 18.4 MJ/kg DAF (7,900 BTU/lb DAF) respectively.

A reactor temperature is selected, and the amount of tar produced is calculated. Then an energy balance is performed on the reactor to determine if the appropriate amount of tar is present to satisfy the energy needs. Iterations are conducted to find a suitable operating temperature.

PROCESS DESCRIPTION

A process flow diagram for the conceptual farm-scale gasifier system is shown in Figure 1. Major components of the system are air compressor, gasifier, heat exchanger network, and fiber bed filter. The gasifier produces approximately 32,600 m^3/day (1.15 MMSCF/day) of gas having a higher heating value of 3.77 MJ/m^3 (101 BTU/SCF) from 10.9 metric tons (12 tons) of dry ash-free feed per day.

Ground corn stover is conveyed into the storage bin (V-1) which is large enough to hold about two days' supply of feed. The corn stover is transferred to the feed hopper (V-2) by a bucket elevator (CV-1) and is fed to the reactor (R-1) by a screw feeder. The hopper (V-2) contains enough feed for approximately five hours.

The reactor (R-1) consists of a bed of fluidized sand which is maintained at 978 K (1300 F) and 55 kPa (8 psig). Heat required for gasification is supplied by burning recycled tar in the plenum section of the reactor and char in the lower portion of the reactor. The plenum section is maintained at 1644 K (2500 F) and 69 kPa (10 psig) by injection of recycled water. The air compressor (C-1) supplies air needed for combustion of tar and char, and the combustion gases together with the injected water provide sufficient flow rate to fluidize the sand bed.

Off-gas is sent to a cyclone (CY-1) to remove entrained ash. Ash is collected in a hopper (V-3) large enough to hold approximately two days' production of ash. Then, off-gas is sent through a series of three heat exchangers. Heat exchanger E-1 is an air cooled exchanger which reduces the gas temperature to 644 K (700 F). Heat exchanger E-2 reduces the temperature of the off-gas to 422 K (300 F). Tar condenses inside this exchanger and is recycled back to the plenum burner. Water is condensed in heat exchanger E-3 and is recycled back to the plenum burner. This heat exchanger reduces the off-gas temperature to 325 K (125 F). Finally, the off-gas is sent to a fiber bed filter (F-1) to remove any tar mist before off-gas is sent to the irrigation pump engines.

The overall flow rates into and out of the gasifier plant are given in Figure 2. The recycle flow rates are also given in this figure.

With the exception of the electric power for the compressor and pumps, no additional energy is required after start-up. The energy input into the process is less than 0.1% of the energy in the product gas.

ECONOMIC ANALYSIS

The preceding conceptual design has been used to evaluate the capital investment, operating costs, and profitability for a farm-scale gasifier. All monetary values are based on third quarter 1983 dollars unless stated otherwise.

Equipment Costs

Equipment costs are presented in Table I. The major items of equipment are the reactor (R-1), compressor (C-1), and fiber bed filter (F-1). All equipment costs except for the costs of the tar storage tank (V-4), water storage tank (V-5), and fiber bed filter (F-1) have been estimated using data and techniques reported by Guthrie (1969). Costs for storage tanks V-4 and V-5 have been estimated by means of a procedure reported by Peters and Timmerhaus (1980). Both sets of costs have been converted to third quarter 1983 dollars by using Marshall and Stevens Installed-Equipment Indexs. Price of the fiber bed filter (F-1) has been obtained from a direct price quotation from a vendor. A stand-by filter is included so that the gasifier does not have to be shut down for filter cleaning.

Total Capital Investment

The total capital investment has been evaluated to be $112,000. Details for the total capital investment are presented in Table II. These estimates have been calculated by a procedure similar to one reported by Peters and Timmerhaus (1980). All direct and indirect costs have been estimated as percentages of the carbon-steel purchased equipment cost estimate except the cost of piping. The piping cost estimate is based on the actual purchased equipment cost estimate. The carbon-steel equipment cost estimate has been calculated by assuming that all items of equipment are made of carbon-steel; it has been evaluated to be $75,000. The percentages used are smaller than those reported by Peters and Timmerhaus (1980) because the gasifier is a small and close coupled system that can be factory assembled. No buildings, yard improvements, service facilities, or land will be required.

A contingency of ten percent of the total direct and indirect costs is used. The mass produced cost for the gasifier system is assumed to be 60% of the fixed capital investment. The contractor's fee is set at ten percent of the mass produced cost.

Operating Costs

Estimates of the annual operating expenses are presented in Table III. These costs have been estimated using information from Peters and Timmerhaus (1980). Parameters affecting the total product cost are feed cost and gasifier service life. Feed material cost is assumed to range between $11 and $44.1 per metric ton ($10 and $40 per ton). Approximately 907 metric tons (1,000 tons) of corn stover are gasified per year. Plant service life is selected as ten and fifteen years, and straight line depreciation with no salvage value was used.

Operating labor cost has been calculated assuming two man hours per day at a cost of eight dollars per man hour. The cost of utilities has been determined from an estimate for the work load of the compressor, pumps, and motors. The cost of electricity is taken to be $0.055/kWh. Maintenance and repairs, operating supplies, and insurance have been estimated to be 3%, 0.5%, and 1% of the total capital investment respectively. Water from the irrigation wells will be used for cooling water and then will be returned to be used for irrigation; hence, cost for cooling water is negligible. The costs associated with direct supervision and overhead are also neglected. Calculations are on a before-tax basis; thus, taxes are not included. The annual operating expenses have been estimated to be $15,900 plus feed material cost and $19,600 plus feed material cost for a 15 year and 10 year plant life respectively.

Profitability Analysis

For this preliminary study, only the simpler method of determining profitability is used. Discounted cash flow or present worth could be used for a more in-depth profitability study. Profitability of the gasifier is dependent upon operating cost and fuel savings. Fuel savings is defined as the amount of energy replaced in one year multiplied by the fuel cost. The range used for fuel cost is $3.8 to $9.5 per GJ ($4 to $10 per MMBTU). Approximately 7,200 GJ (6,800 MMBTU) per year are replaced for irrigation pumping and grain drying. Net savings is defined as fuel savings minus operating cost.

Payout period as a function of fuel cost for different feed costs is shown in Figure 3. Payout period is defined as the length of time to recover the depreciable capital investment in the form of cash flow to the project. Cash flow is defined as net savings plus depreciation. For the before-tax basis used, payout period is independent of plant life. As expected, payout period increases with increases in feed material cost and/or decreases in fuel costs. For a payout period of seven years, a fuel cost of $4.7, $6.2, $7.6, and $9.0 per GJ is required for a feed cost of $11, $22, $33, and $44 per metric ton respectively. This converts to a fuel cost of $5, $6.5, $8, and $9.5 per MMBTU for a feed cost of $10, $20, $30, and $40 per ton respectively.

Return on investment as a function of fuel cost for different feed costs is shown in Figure 4 for a 15 year plant life and in Figure 5 for a 10 year plant life. Return on investment is defined as the net savings divided by the total capital investment and multiplied by 100. A before-tax basis is used for the return on investment. An increase in feed cost or a decrease in plant life decreases the return on investment for a given fuel cost. A decrease in fuel cost also decreases the return on investment. For a seven percent before-tax return on investment on a plant with a 15 year life span, the fuel cost must be \$4.7, \$6.1, \$7.5, and \$8.9 per GJ for a feed cost of \$11, \$22, \$33, and \$44 per metric ton respectively. This converts to fuel costs of \$5, \$6.4, \$7.9, and \$9.4 per MMBTU for feed costs of \$10, \$20, \$30, and \$40 per ton respectively. On a break even basis, fuel costs must be \$3.6, \$5, \$6.4, and \$7.8 per GJ (\$3.8, \$5.3, \$6.8, and \$8.2 per MMBTU) for respective feed costs of \$11, \$22, \$33, and \$44 per metric ton (\$10, \$20, \$30, and \$40 per ton) for a plant life of 15 years. The fuel cost is \$0.47/GJ (\$0.5/MMBTU) higher for a plant with a ten year service life for a given feed cost and return on investment.

DISCUSSION

The cost of the gasifier system is comparable to the costs reported by Electric Power Research Institute (EPRI) (1983). They have reported total capital investments for three different wood gasifiers presently available to be between \$20,000 and \$25,000 per MMBTU/hr output (\$19,000 and \$23,700 per GJ/hr output) in mid-1982 dollars. The cost for the conceptual design has been estimated to be \$28,000 per MMBTU/hr output (\$26,600 per GJ/hr output) in third quarter 1983 dollars. This amount is only slightly larger than those for the wood gasifiers. Cost of feed material is also similar to that reported by EPRI (1983) for wood. They have given a range for feed cost from \$1.9 to \$2.4 per GJ (\$2 to \$2.5 per MMBTU). The range of corn stover cost of \$11 to \$33 per metric ton (\$10 to \$30 per ton) converts to \$0.95 to \$2.8 per GJ (\$1 to \$3 per MMBTU).

One of the reasons that the gasifier discussed in this chapter has a high capital cost to energy output ratio is because the gasifier has been designed to produce a small amount of energy, 4.2 GJ/hr (4 MMBTU/hr) compared to 26.3 GJ/hr (25 MMBTU/hr) for the wood gasifiers. It also has a low operating factor (approximately 2.5 months per year) which acts to limit the profitability. Increasing the operating factor will improve the economics of the process. This could be done by using the producer gas to make liquid fuel in the off season. This requires an addition feed source because irrigation pumping and grain drying uses about 90% of the available corn stover based on 75% removal from the field and 15% storage loss. However, a variety of feed sources are possible. For example, manure, wheat straw, or wood could be gasified in the same reactor.

Slight modifications could be made in the conceptual design. One change would be to use off-gas to preheat the air fed to the plenum burner. Trade off between the price of a gas-gas heat exchanger and increases in gasification efficiency exists. Gasification efficiency increases because the reactor temperature is increased since less tar needs to be burned in the plenum. The increase in reactor temperature will increase gas production, and thus, gasification efficiency. Furthermore, the air flow rate will decrease, and the heating value of the off-gas will increase because there will be less nitrogen present. Consequently the compressor size will decrease also.

CONCLUSIONS

The conceptual design and economic feasibility study for a farm-scale fluidized bed gasifier has been conducted. The gasifier produces approximately 32,600 m^3/day (1.15 MMSCF/day) of gas having a heating value of 3.77 MJ/m^3 (101 BTU/SCF) from 10.9 metric tons (12 tons) of DAF corn stover a day. The producer gas will be used to fuel three 100 kW irrigation pumps and to dry grain. The fluidized bed reactor will be operated at 978 K (1300 F) and 55 kPa (8 psig). Except for the electricity needed for the compressor, pumps, and motors, the reactor is virtually energy self-sufficient.

Total capital investment for the gasification system has been estimated to be $112,000 in third quarter 1983 dollars. For a before-tax payout period of seven years, fuel costs of $4.7, $6.2, $7.6, and $9 per GJ are needed for feed costs of $11, $22, $33, and $44 per metric ton respectively. Break even fuel costs have been estimated to be $3.6, $5, $6.4, and $7.8 per GJ for feed costs of $11, $22, $33, and $44 per metric ton respectively for a gasifier with a 15 year service life. The fuel cost is $0.47/GJ higher for a plant with a ten year life span for a given feed cost and return on investment. The gasifier appears to be competitive with propane gas and new natural gas; however, the gasifier is not competitive with old natural gas.

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Table I
Equipment List

Name		Symbol	Size		Cost
Reactor	Shell	R-1	1.22 m dia by 1.83 m	bed	14 000
			1.83 m dia by 1.83 m	freeboard	
	Refractory		32.71 m		3 600
Air Compressor		C-1	12.74 m^3/min at 69 kPa		16 800
Cyclone		CY-1	42.5 m^3/min	ss	7 000
Tar Injection Pump		P-1	0.11 m^3/hr	ss	3 200
Water Injection Pump		P-2	0.04 m^3/hr	ss	2 100
Air-Cooled Exchanger		E-1	9.29 m^2	ss	1 500
Tar Condenser		E-2	6.50 m^2	ss	3 300
Water Condenser		E-3	8.36 m^2	ss	3 400
Water Pumps					
	Tar condenser	P-3	4.86 m^3/hr	bronze	2 000
	Water condenser	P-4	2.52 m^3/hr	bronze	1 800
Storage Tanks					
	Tar	V-4	0.28 m^3	ss	2 900
	Water	V-5	0.15 m^3	ss	2 200
Ash Bin		V-3	6.37 m^3		400
Feed Hopper		V-2	14.16 m^3		800
Storage Bin		V-1	133.8 m^3		5 700
Screw Feeder		CV-3	0.15 m dia by 1.52 m	ss	6 600
Elevator		CV-1	9.14 m		1 800
Ash Conveyor		CV-2	1.52 m		3 600
Fiber Bed Filter		F-1	12.74 m^3/min		12 900
					$95 600

ss denotes stainless steel
prices are based on third quarter 1983 dollars

Table II
Total Capital Investment

Direct Costs					
Purchased Equipment		100% PE	95 600		
Installation		15% CSPE	11 400		
Instrumentation & Controls		10% CSPE	7 600		
Piping		5% PE	4 800		
Electrical		3% CSPE	2 300		
Indirect Costs					
Engineering		40% CSPE	30 300		
Construction Expenses		3% CSPE	2 300		
Direct and Indirect Costs				154 300	
Contingency	10% Direct & Indirect Costs			15 400	
Fixed Capital Investment				169 700	
Mass Production	Discounted 40%				101 800
Contractor's Fee	10%				10 200
Total Capital Investment					$112 000

Costs are in third quarter 1983 dollars
CSPE denotes carbon-steel purchased equipment

Table III
Operating Costs

Raw material Corn Stover	\$11 - 44/metric ton	
Operating Labor	2 man hours/day @ \$8/hr	1 200
Utilities	38 200 kWhr @ \$0.055/kWhr	2 100
Maintenance & Repairs	3% TCI	3 400
Operating Supplies	0.5% TCI	600
Insurance	1% TCI	1 100
Depreciation	15 year plant life	7 500
	10 year plant life	11 200

Total Operating Cost	
15 year plant life	\$15 900 + feed cost
10 year plant life	\$19 600 + feed cost

Costs are in third quarter 1983 dollars
TCI denotes Total Capital Investment

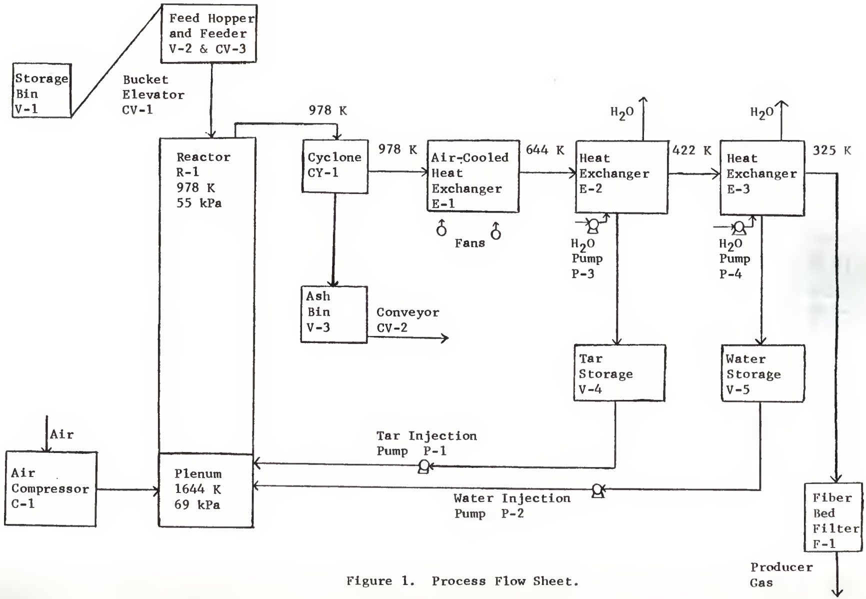

Figure 1. Process Flow Sheet.

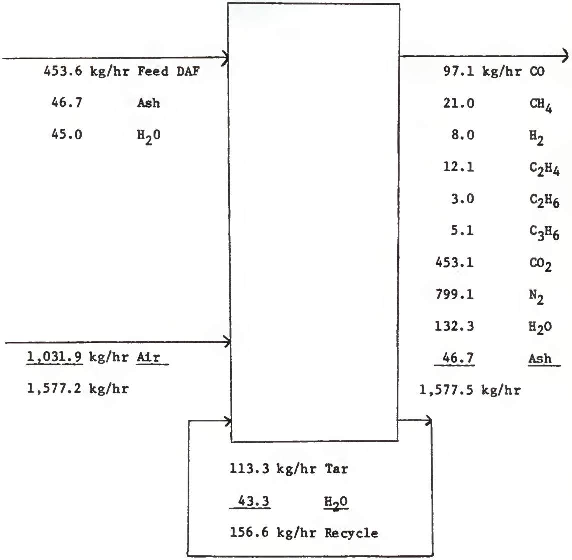

Figure 2. Overall Material Balance Flow Rates.

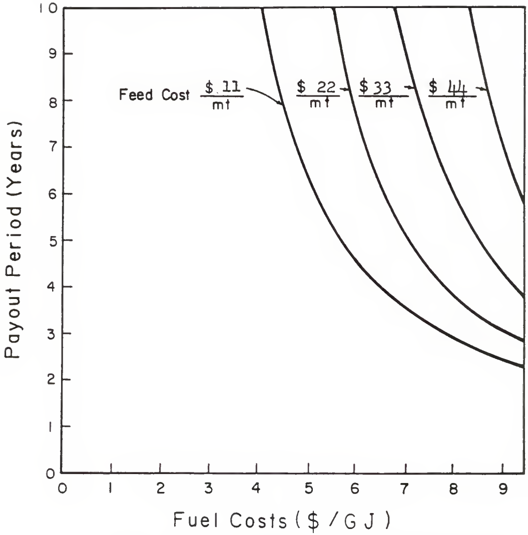

Figure 3. Payout Period vs. Fuel Cost with Feed Cost as a parameter.

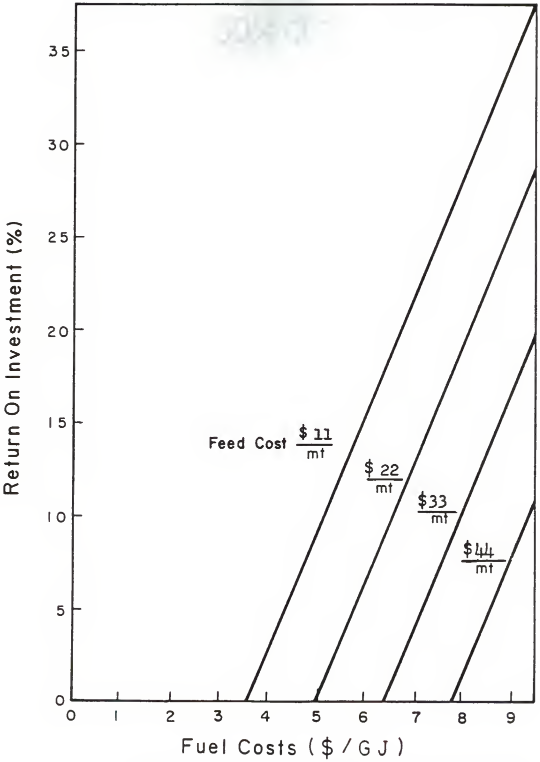

Figure 4. Return on Investment vs. Fuel Cost for a fifteen year plant life.

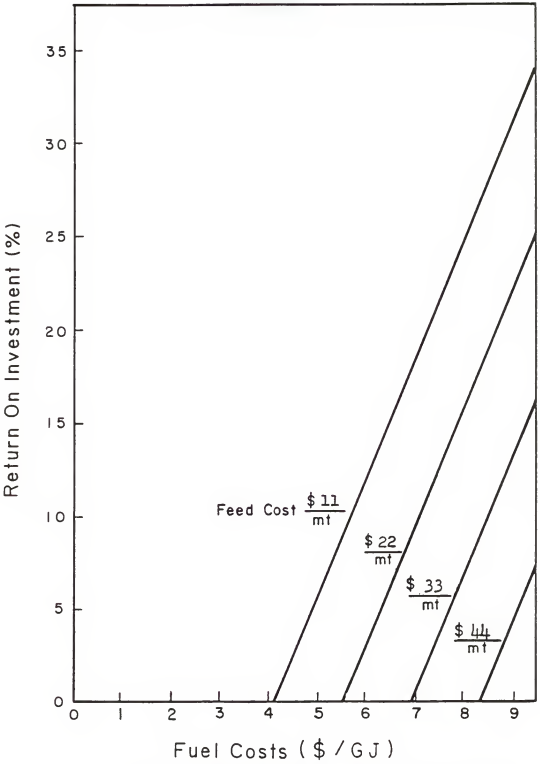

Figure 5. Return on Investment vs. Fuel Cost for a Ten year plant life.

Chapter VI

CONCLUSIONS AND RECOMMENDATIONS

CONCLUSIONS

Corn stover and sorghum stover were gasified in a 0.23 m I.D. fluidized bed reactor. The temperature ranges studied were 820 to 1020 K for corn stover and 750 to 1020 K for sorghum stover. The effect of reactor temperature on produced gas volumetric yield, composition and higher heating value was studied. The mass distribution of the major gasification products, carbon conversion, and energy recovery were also evaluated.

For corn stover, gas yield, energy recovery, and carbon conversion increased with increasing reactor temperature; whereas, liquid yield decreased with increasing reactor temperature. Produced gas volumetric yield increased from 0.17 to 0.74 m^3/kg DAF feed in the temperature range of 820 to 1020 K. Higher heating value ranged from 11.7 to 14.6 MJ/m^3 with a maximum of 15.6 MJ/m^3 at 955 K. Hydrogen, carbon dioxide, carbon monoxide, and methane made up over 90% of the produced gas. Concentration of hydrogen in the produced gas increased, and concentration of carbon dioxide decreased with increasing reactor temperature. Methane concentration was almost constant at 11.3%.

For sorghum stover, produced gas volumetric yield increased from 0.91 m^3/kg DAF feed at 750 K to 0.91 m^3/kg DAF feed at 1020 K. Produced gas higher heating value varied from 9.97 to 14.0 MJ/m^3 with a maximum of 14.9 MJ/m^3 at 960 K. The produced gas consisted mainly of hydrogen, carbon monoxide, carbon dioxide, and methane. Hydrogen concentration in the produced gas increased, and carbon dioxide concentration decreased with increasing temperature. Methane concentration was virtually constant at 10.0%. Energy recovery, carbon conversion, and gas mass yield all increased linearly with temperature.

Regression model parameters for the dependent variables (i.e. produced gas yield, composition, and heating value) for corn stover are compared to the respective model parameters for sorghum stover using a t-test. Regression model parameters for hydrogen and methane concentrations in the produced gas, produced mass distribution (gas, liquid, and char), carbon conversion, and energy recovery are not significantly different at the 5% probability level. Thus, these lines are within the 95% confidence limits of the respective line for the other feedstock.

A conceptual design and economic feasibility study for a farm-scale fluidized bed gasifier is conducted. The gasifier is designed to produce approximately 32,600 m^3/day (1.15 MMSCF/day) of gas having a heating value of 3.77 MJ/m^3 (101 BTU/SCF) from 10.9 metric tons (12 tons) of DAF corn stover per day. This gas will be used to fuel three 100 kW irrigation pumps and to dry corn grain. The reactor is operated at 978 K (1300 F) and 55 kPa (8 psig). Except for the electricity needed for the compressor, pumps, and motors, the reactor is virtually energy self-sufficient. The char and tar produced from gasification are burned to supply the energy requirements for gasification. Total capital investment for the system is estimated to be $112,000 in third quarter 1983 dollars. For a before-tax payout period of seven years, fuel costs of $4.7, $6.2, $7.6, and $9 per GJ are needed for feed costs of $11, $22, $33, and $44 per metric ton respectively. Break even fuel costs are estimated to be $3.6, $5, $6.4, and $7.8 per GJ for feed costs of $11, $22, $33, and $44 per metric ton respectively for a gasifier with a 15 year service life. The process appears to be competitive with propane gas and new natural gas, but it does not appear to be competitive with old natural gas.

RECOMMENDATIONS

A simplified statistical model building procedure was attempted to describe the gasification data. The stepwise model building procedure was used in an attempt to identify additional factors affecting produced gas composition, volumetric yield, higher heating value, and mass distribution of the major gasification products. A linear model was used with reactor temperature, superficial gas velocity, moisture content of the feed, and carbon content of the feed as the independent variables. Later, temperature squared and temperature cubed were added as independent variables. Of the independent variables examined, reactor temperature had the most significant effect on the dependent variables. Some dependence upon other independent variables was indicated; however, because the experiments were not specifically designed for this model building approach, the results were not conclusive. The model building approach should be extended in order to develop a model for the gasification of biomass from which predictions can be made. Properly designed experiments are needed preferably for a bench-scale reactor where the independent variables are easily controlled. Additional independent variables such as cellulose content of the feed, reactor water to feed ratio, residence time, and parameter interaction terms could also be examined. Furthermore, models that are more complex than the simple linear model should be tested.

Modifications to the conceptual design should be considered. One such change is to use the off-gas to preheat the air fed to the plenum burner. A trade off between the price of a gas-gas heat exchanger and increased gasification efficiency exists. Gasification efficiency will increase because reactor temperature will be increased since less tar will need to be burned in the plenum. The increase in reactor temperature increases gas production, and hence, gasification efficiency. Furthermore, the air flow rate will be decreased; thus, the heating value of the off-gas will increase because it will contain less nitrogen. In addition, compression requirements will be reduced.

CROP RESIDUE GASIFICATION

by

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B. S., Iowa State University, 1981

AN ABSTRACT OF A MASTER'S THESIS

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MASTER OF SCIENCE

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Corn stover and sorghum stover were gasified in a 0.23 m ID fluidized bed reactor. The temperature ranges studied were 820 to 1020 K for corn stover and 750 to 1020 K for sorghum stover. The effect of reactor temperature on the produced gas volumetric yield, composition and higher heating value was studied. The major product mass distribution, carbon conversion, and energy recovery were also determined.

For corn stover, gas yield, energy recovery, and carbon conversion increased with increasing reactor temperature; whereas, liquid yield decreased. Produced gas volumetric yield increased from 0.17 m^3/kg DAF feed over the temperature range studied. Higher heating value ranged from 11.7 to 14.6 MJ/m^3 with a maximum of 15.6 MJ/m^3 at 955 K. Hydrogen, carbon dioxide, carbon monoxide, and methane made up over 90% of the produced gas. The concentration of hydrogen in the product gas increased, and the concentration of carbon dioxide decreased with increasing reactor temperature. Methane concentration was virtually constant.

For sorghum stover, produced gas volumetric yield increased from 0.19 m^3/kg DAF feed at 750 K to 0.91 m^3/kg DAF feed at 1020 K. Produced gas higher heating value varied from 9.97 to 14.0 MJ/m^3 with a maximum of 14.9 MJ/m^3 at 960 K. The produced gas consisted mainly of hydrogen, carbon monoxide, carbon dioxide, and methane. The hydrogen concentration in the produced gas increased and the carbon dioxide concentration decreased with increasing temperature. The methane concentration was virtually constant. Energy recovery, carbon conversion, and gas mass yield all increased linearly with increasing temperature. A statistical t-test showed that experimental results obtained from the gasification of sorghum stover were statistically similar to results for corn stover and wheat straw.

A conceptual design and economic feasibility analysis for a farm-scale fluidized bed gasifier is conducted. The gas is to be used as an alternate fuel for irrigation pumping and grain drying. The gasifier is designed to produce approximately 32,600 m^3/day (1.15 MMSCF/day) of gas having a heating value of 3.77 MJ/m^3 (101 BTU/SCF) from 10.9 metric tons (12 tons) of DAF corn stover per day. Total capital investment for the system is estimated to be $112,000 in third quarter 1983 dollars. Payout period and return on investment are calculated for different values of feed cost, fuel cost, and gasifier service life. Feed cost ranges from $11 to $44 per metric ton and fuel cost ranges from $3.8 to $9.5 per GJ. For a before-tax payout period of seven years, fuel costs of $4.7, $6.2, $7.6, and $9 per GJ are required for feed costs of $11, $22, $33, and $44 per metric ton respectively. On a break even analysis (before-tax), fuel costs are estimated to be $3.6, $5, $6.4, and $7.8 per GJ for respective feed costs of $11, $22, $33, and $44 per metric ton for a gasifier with a 15 year service life. The fuel cost is estimated to be $0.47/GJ higher for a reactor with a ten year life span.