Steam Gasification of Grain Dust in a Fluidized Bed Reactor

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ABSTRACT

CORN dust was gasified with steam in a 5 cm (2 in.) I.D. bench scale fluid-bed reactor. Experiments were conducted over a temperature range of 867 K to 1033 K (1100 °F-1400 °F) to examine the influence of temperature on the gas composition, yield, heating value and the energy recovery. The gas yield was found to be strongly temperature dependent and increased linearly from 0.13 m³/kg (2.0 SCF/lb) at a 867 K to 0.73 m³/kg (11.7 SCF/lb) at 1033 K. The produced gas heating value ranged from 9.4 MJ/m³ (251 Btu/SCF) at 867 K to 11.5 MJ/m³ (308 Btu/SCF) at 1033 K with maximum value of 11.6 MJ/m³ (311 Btu/SCF) at 1002 K. The principal components of the product gas were H_2, CO_2 and CO which comprised over 90 percent of the gas. The remaining 10 percent consisted of CH_4, C_2H_4 and C_2H_6. The energy recovery increased linearly from 8 to 55 percent over the temperature range studied.

INTRODUCTION

A need for alternate energy sources is rapidly emerging. Wastes from the processing of agricultural products represent an ideal source of renewable energy because these materials are concentrated at the processing site. Grain dust is one such waste product that has a potential for conversion to a fuel. Possible conversion schemes for grain dust include: direct combustion, anaerobic digestion, pyrolysis (gasification), liquefaction and hydrolysis for fermentation to ethyl alcohol. This work is concerned with the steam gasification of grain dust in a fluidized bed reactor.

Grain dust is an undesirable byproduct of grain handling. In the United States, where about 10 billion bushels of grain are handled each year, the dust is collected by dust control equipment to alleviate the potential of explosions and to provide a clean working environment. Assuming that the grain is handled only once and that it generates on the average 0.05 percent (by weight) dust, the total amount of dust potentially available is 1.4×10^8 kg/yr (1.5×10^5 tons/yr). The dust is distributed among several thousand elevators of various sizes. A large elevator, one handling 400 million bushels of grain/yr, generates 5.4×10^6 kg/h of dust (6,000 ton/yr).

Gasification is the process by which organic matter is converted to gas through thermal decomposition in an oxygen deficient environment, followed by secondary reactions of the resulting volatiles. The process also produces combustible char and tar which can be burned with air to provide the necessary energy for processing. Studies on the fluid-bed gasification of agricultural wastes have been limited in the past but data is beginning to grow rapidly. Bailie and Burton (Burton, 1972) were the first in this country to study the fluid-bed gasification of agricultural wastes. Burton's masters thesis reports the gasification of many different kinds of solid wastes including sawmill wood, chicken manure and cow manure. They used a 38 cm (I.D.) pilot plant to study the effect of temperature on composition, heating value and yield. Halligan et al., (1975) investigated the gasification of feedlot manure in a 5 cm (I.D.) fluidized-bed reactor. Their bench scale reactor was operated under partial oxidation conditions to study the effects of temperature on gasification. Howard et al., (1979) studied the pyrolysis of a wide range of wastes including manure and corncobs in a 50 cm (I.D.) fluid-bed reactor. They studied the effects of temperature, fluidization velocity, feedrate of pyrolyzable material, bed height (static) and particle size on the final products of pyrolysis. Kuester (1979) developed a dual fluid-bed system which has been used to gasify bagasse and other materials. The gas produced from his complete pilot plant facility is subsequently converted into liquid fuels via the Fischer-Tropsch synthesis. Le Pori et al., (1979) used a pilot scale, shallow fluidized-bed reactor to gasify cotton gin trash. Beck et al., (1979) conducted extensive studies with manure in a variable velocity fluid-bed reactor. Their 15 cm (I.D.) reactor evaluated the gasification of manure by partial oxidation. Walawender and Fan (1978) conducted preliminary pilot plant tests to study the gasification of feedlot manure in a fluid-bed reactor. Walawender et al., (1980) presented gasification data for a variety of carbonaceous materials including cane and rubber in the same reactor. They also (Raman et al., 1980) studied the gasification of corn stover. These works examined the effects of temperature on the gas composition, heating value and yield of produced gas.

It appears that no work has been conducted on the gasification of grain dust. In this paper we consider the steam gasification of corn grain dust in a fluidized bed. The produced gas composition, heating value, yield and energy recovery are examined as functions of the gasifier temperature.

EXPERIMENTAL FACILITIES AND PROCEDURES

Facilities

The bench scale fluidized-bed reactor used in this study is shown schematically in Fig. 1. The reactor was
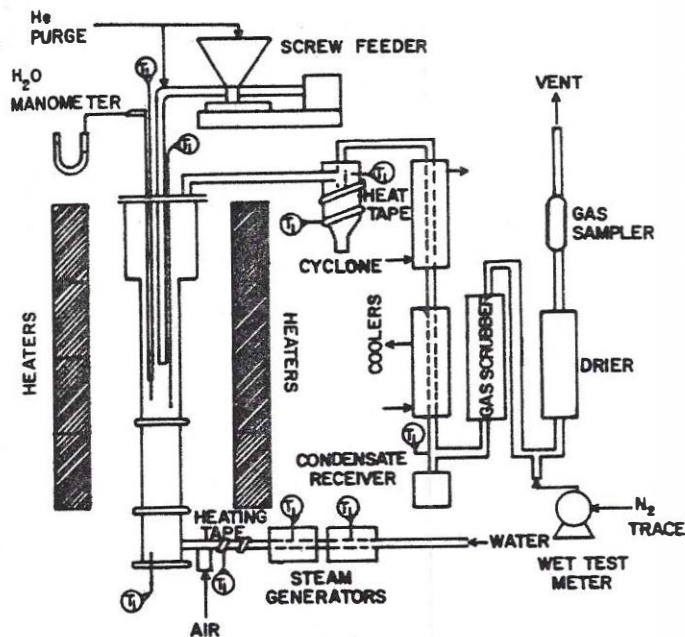
FIG. 1 Bench scale reactor.

constructed from a 5.08 cm (2 in.) I.D. by 50 cm length of schedule 40 Inconel 600 pipe, fitted on top with a 10.16 cm (4 in.) I.D. by 15 cm length pipe of the same material. The upper 15 cm of the reactor served as a disengaging zone. The lower 23 cm, which served as a gas distributor and gas preheater, was packed with aluminum oxide pellets (0.5 cm in diameter). The fluid-bed section was separated from the packed bed section by means of a 60 mesh (0.025 mm) 304 stainless steel screen. The bed material was composed of a 25 percent by weight limestone — 75 percent by weight silica sand mixture. The limestone was used to eliminate the bed ag-glomeration which typically occurs in a bed composed only of sand (Walawender et al., 1981). The limestone particle size was −7 + 50 mesh (2.82 mm-0.287 mm); the sand particle size was −30 + 50 mesh (0.595 mm-0.297 mm). The bed had a static height of 8.4 cm and an ex-panded height of about 11 cm.

The reactor was heated by means of four pairs of semi-cylindrical electrical resistance heaters, each capable of delivering up to 2300 W of power, at the maximum operating temperature of 1743 K. Voltage to each pair of heaters was controlled by a variable voltage auto-transformer. The second pair of heaters from the bottom was also controlled by a thermostat connected to a ther-mocouple inside the reactor bed. Steam, the sole fluidiz-ing gas, was generated externally in a series of two (600 and 800 W) electrical furnaces and was supplied to the gas preheater section at approximately 500 K. Heating tape was wrapped around the pipe leading from the steam generator to the preheater to prevent the steam from cooling and condensing.

The feed was introduced into the reactor by gravity flow through a vertical feedpipe which discharged 6.6 cm above the static bed. A Vibra Screw (Model SCR-20) screw feeder with a solid core flight screw was used to supply the feedpipe at a uniform volumetric rate. A purge flow of helium was used to aid flow through the feedpipe and to prevent backflow through the feedpipe and subsequent condensation of vapor in the feedpipe and feeder.

The product gases from the reactor were sent to a cyclone where entrained fines were separated from the gas. The temperature in the cyclone was maintained at about 600 K to prevent condensation of tar. This was ac-complished with a heating tape wrapped around the cyclone. The gaseous stream leaving the cyclone was cooled to approximately 300 K with two water-cooled heat exchangers in series. This resulted in the condensa-tion of the fluidizing steam, which was collected in a receiver. The resultant product gas carried with it a fine mist of condensables, which were removed by passing the gas through a packed bed of glass wool to remove tars and a column packed with Drierite (CaSO₄) to remove residual moisture. A known volumetric flow of nitrogen was introduced as a tracer gas just up stream of the dry-ing column. The product gas could bypass the column packed with Drierite and vent to a hood, or a portion could go through the drying column to a gas sample bot-tle for analysis.

The temperature in the fluidized bed was monitored by a thermocouple placed inside the bed. The tempera-ture in the disengaging zone was monitored by using a sliding (91 cm long) thermocouple, which measured tem-peratures from the bottom of the fluidized bed to the top of the disengaging zone. The temperature was also moni-tored in the steam generators, cyclone, heat exchanger outlet, reactor inlet and the two heating tapes. A pres-sure probe, connected to a water manometer, was used to monitor the bed pressure and as a check on the quality of fluidization. A second pressure probe on the feedpipe was connected to a water manometer and was used to monitor the reactor pressure.

Feed Stock

The feed stock was corn grain dust treated with oil (0.025 percent by weight) obtained from the U.S. Grain Marketing Research Laboratory, Manhattan, KS. As a convenience for introducing the feed into the bed, the grain dust was pelletized without binders and subse-quently ground and sieved prior to introduction to the reactor. The size fraction used was −14 + 40 mesh (1.41 mm-0.42 mm). The grain dust contained 10.8 per-cent (by weight) moisture. The ultimate analysis of the feed (in weight percent) on a dry basis was as follows: C, 39.2; H, 5.7; N, 1.0; 0, 42.9; ash, 11.2. The heating value on a dry ash free basis, estimated using the Dulong formula, was found to be 15.4 MJ/kg DAF. Pelletizing had no influence on the elemental composition of the dust.

Procedure

For start-up of the experimental system, the reactor heaters, steam generators, cyclone and inlet heating tapes were turned on. Initially, air was used as the fluid-izing agent and as the feedpipe purge gas. During the heat-up period, the steam generators, cyclone heater and inlet section heater were brought to their operating tem-peratures of 670 K, 920 K, 750 K and 700 K, respective-ly. When the effluent from the steam generators reached about 800 K, the air flow was replaced by steam. The volumetric flow rate of the steam required to maintain the selected fluidization velocity was measured both by collecting condensate downstream of the heat exchangers and by metering the water flow into the steam generators.

When the reactor temperature stabilized at the selected operating temperature, the air purge was re-

placed with helium. For this study the same temperature was maintained in the bed and through the disengaging zone. About 1/2 h after the helium purge was started, feeding and nitrogen tracer were initiated. The nitrogen tracer was introduced just upstream of the drying column at a known rate which averaged about 60 percent of the flow rate of the dry produced gas. The total start-up time was about 2 h from a cold start.

Any change in the temperature of the reactor due to feeding the dust was corrected automatically by the thermostat. A gas sample was taken 15 min after feeding began and a second sample was taken after 25 min. The condensate and nitrogen rates were measured over 10 min time intervals as the gas samples were being taken. After all of the pertinent data were taken, feeding of the dust was terminated and the reactor temperature was elevated for another run. Feeding was started after the temperatures in the bed and disengaging zone had stabilized. This procedure was repeated for each run in a set of experiments.

The feed rate was evaluated by disconnecting the lower section of the feed pipe and weighing the effluent collected for 10 min time intervals at a given setting on the feeder. The feed rate was evaluated several times before and after each set of experiments.

It was not possible to measure the total char produced in an experiment because of the hold-up of char in the bed. Nor was it possible to measure the total tar produced because of hold-up on the inside of the water cooled heat exchangers.

Chemical Analysis

The chemical analysis included the ultimate analysis of the feed as well as the quantitative analysis of the product gas. Ultimate analysis was performed with the aid of a Perkin-Elmer Model 240 B Elemental Analyzer. Ash analyses were performed on the feed material according to the standard ASTM procedure in a muffle furnace and the moisture contents were determined by drying in an oven for 10 h at 423 K.

The product gas was analyzed by using a Packard Model 417 Becker dual column gas chromatograph equipped with thermal conductivity detectors. The gas components of interest included H_2, CO, CO_2, CH_4, C_2H_4, C_2H_6, O_2 and N_2. A 1.83 m × 0.0032 m (6 ft × 1/8 in.) column with No. 5A molecular sieve packing (−80 + 100 mesh) was used to separate and analyze H_2, O_2, N_2, CH_4 and CO. The remaining components were analyzed by using a 1.83 m × 0.0032 m (6 ft × 1/8 in.) column with a −80 + 100 mesh Porapak Q packing preceded by a short lead section of Porapak R. The Porapak R was used to shift the retention time of the

TABLE 1. REACTOR OPERATING PARAMETERS

Reactor temperature range	867 K - 1033K
Fluidizing gas	Steam
Superficial velocity	0.22 - 0.32 m/s
Feed rate	1.34 - 1.96 g/min
Grain dust particle size	1.41 mm - 0.420 mm (−14 + 40 mesh)
Bed particle size	
Sand	0.595 mm - 0.297 mm (−30 + 50 mesh)
Limestone	2.82 mm - 0.297 mm (−7 + 50 mesh)

water peak so that it could be easily separated from the other components. The gas chromatograph was operated isothermally at 350 K with helium as the carrier gas.

Operating Conditions

The operating conditions for all the experimental runs are summarized in Table 1. The principal experimental variable was the operating temperature. The superficial velocity of the fluidizing gas corresponded to about two times the minimum fluidization velocity and 20 percent of the terminal velocity of the feed. It corresponded to about three times the minimum fluidization velocity of the bed material.

RESULTS AND DISCUSSION

Calculations and Analyses

From the GC analysis and the nitrogen tracer rate, the volumetric flow rate of the dry produced gas (nitrogen free) was calculated with the aid of a nitrogen balance. The small amount of nitrogen (less than 2 percent of the produced gas) resulting from the gasification of the feed can be neglected without serious error for the conditions of the experiments. The dry produced gas analysis (nitrogen-free basis) was used to compute the higher heating value of the dry produced gas using the standard heat of combustion for each component. The volumetric gas yield per unit mass of feed was calculated from the volume flow rate of the dry produced gas (at 288 K and 101.3 kPa) and the mass flow rate of the dry ash-free feed. Because of the hold-up of char and tar in the experimental system, it was not possible to conduct an overall material balance on the bench scale reactor. Consequently, only the yield and properties of the dry produced gas are considered in this report.

Regression analysis was used to determine the relationship between the independent variable, reactor temperature, and the dependent variables: the produced gas compositions, the produced gas heating value, the produced gas yield and the energy recovery. An analysis of variance was also performed on the data to determine the significance of the regression model. The parameters determined by the regression analysis were accepted as being significant on the basis of the F-test at the 10 per-

TABLE 2. STATISTICAL ANALYSIS

Dependent variable y	Sample size n	The square of the correlation coefficient R^2	F-value	Probability > F α	Significant regression model
Gas yield*	8	0.918	66.88	0.0002	$y = -3.01 + 0.00362T$[†]
Heating value ‡	8	0.730	6.76	0.0379	$y = -112. + 0.247T - 0.00012T^2$
mole % H_2	8	0.464	5.21	0.0627	$y = 1.05 + 0.0464T$
mole % CO_2	8	0.655	4.75	0.0699	$y = 1020.5 - 2.08T + 0.00109T^2$
mole % CO	8	0.829	29.20	0.0017	$y = 89.7 - 0.0754T$
mole % CH_4	8	0.704	14.28	0.0092	$y = -10.3 + 0.0155T$
mole % C_2H_4	8	0.863	37.89	0.0008	$y = -12.3 + 0.0154T$
mole % C_2H_6	8	0.240	1.89	0.2180	—

* m^3/kg
† T is reactor temperature in Kelvin
‡ MJ/m^3

cent significance level.

The statistical analysis is summarized in Table 2. For each dependent variable, the number of data points used for the analysis (n), the square of the correlation coefficient (R^2), the F-test statistical value (F-value), the probability of falsely rejecting the proposed regression model (α) and the significant regression model are tabulated.

Produced Gas Composition

The variations in the concentrations of H_2, CO_2 and CO, the major components of the produced gas, as functions of temperature are shown in Fig. 2. The mole percent concentration of CO_2 varied parabolically over the temperature range studied. The concentration ranged from 33.4 percent at 867 K to 31.9 percent at 1033 K with a minimum of 25.1 percent at 954 K. The concentration of CO decreased linearly over the temperature range studied, and ranged from 24.3 percent at 867 K to 11.8 percent at 1033 K. The concentration of H_2 increased linearly over the temperature range studied, and ranged from 41.3 percent at 867 K to 49.0 percent at 1033 K.

The variations in the concentrations of CH_4, C_2H_4 and C_2H_6, the minor components of the produced gas, as functions of temperature are shown in Fig. 3. No relationship could be found between the mole percent concentration of C_2H_6 and temperature. However, the amount present was almost negligible. Its average value was 0.1 percent and it varied from 0.4 percent to 0.0 percent. The concentration of CH_4 and C_2H_4 increased linearly with an increase in temperature. The mole percent of C_2H_4 ranged from 1.0 percent at 867 K to 3.6 percent at 1033 K, while CH_4 ranged from 3.1 to 5.7 percent.

The principle gas phase reaction in this system is the water-gas shift reaction'

$$CO + H_2O \rightleftharpoons CO_2 + H_2.$$

In our experiments, water was present in large excess relative to the other components (about two orders of magnitude). Because of this, we produce a gas that is rich in hydrogen. Equilibrium constants were evaluated from the concentration ratio for each experiment and compared with the theoretical equilibrium constant for each temperature studied. This comparison showed that our system is far removed from equilibrium with the con-
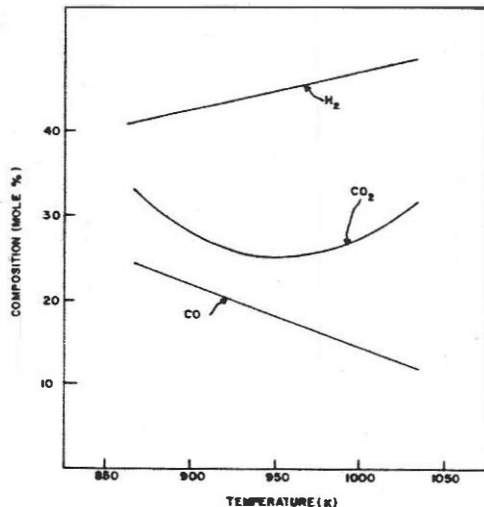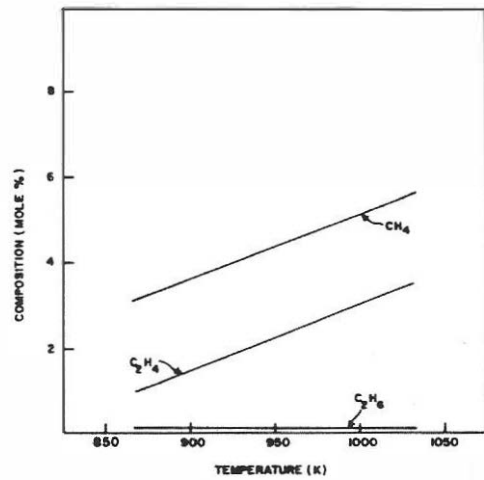

FIG. 3 Gas composition vs. temperature.

centration ratio being one to two orders of magnitude smaller than the theoretical equilibrium constant.

Produced Gas Heating Value

The variation in the produced gas heating values as a function of temperature is shown in Fig. 4. From the analysis of variance, it was found that the produced gas heating value varied parabolically over the temperature range studied and ranged from 9.4 MJ/m³ (251 Btu/SCF) at 867 K to 11.5 MJ/m³ (308 Btu/SCF) at 1033 K with a maximum of 11.6 MJ/m₃ (311 Btu/SCF) at 1002 K. The overall effect of the concentrations of H_2, CH_4 and C_2H_4 increasing and the concentration of CO decreasing is a parabolic curve with a maximum value at about 1000 K.

Produced Gas Yield

The variation in the volumetric yield of the produced gas (on a dry ash-free basis) as a function of temperature is shown in Fig. 5. The yield of dry produced gas was a linear function of temperature and ranged from 0.13 m³/kg (2.0 SCF/lb) at 867 K to 0.73 m³/kg (11.7 SCF/lb) at 1033 K.

Energy Recovery

The effectiveness of the gasification process can be examined by considering the ratio of the heating value of the produced gas per unit mass of feed, on a dry-ash free basis, to the heat of combustion of a unit mass of feed,

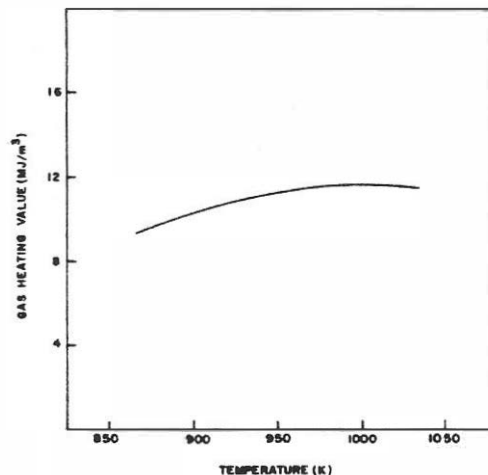FIG. 2 Gas composition vs. temperature.
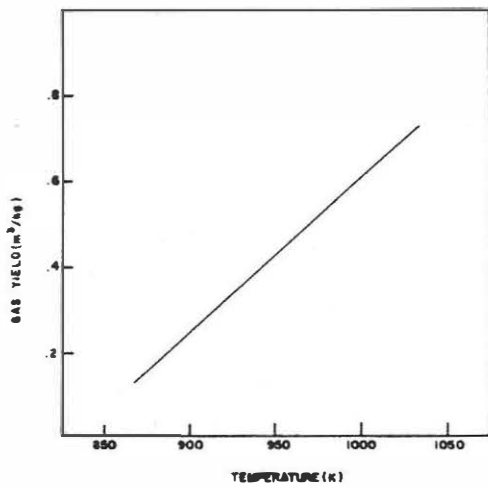

FIG. 4 Gas heating value vs. temperature.
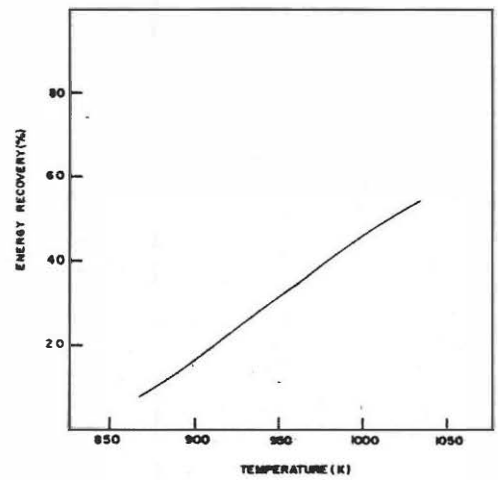
FIG. 5 Gas yield vs. temperature.

FIG. 6 Energy recovery vs. temperature.

also on a dry-ash free basis. The ratio, termed the energy recovery, was computed by multiplying the significant regression models of the gas yield by the gas heating value and dividing by the heating value of the feed. A plot of energy recovery vs. temperature is presented in Fig. 6.

The energy recovery increased from 8 percent at 867 K to 55 percent at 1033 K. The increase observed in the energy recovery is a direct consequence of the increase in the produced gas yield and the relatively constant produced gas heating value with respect to temperature.

If the produced gas were converted to electricity via direct combustion, it would provide about 36 percent of a grain elevator's electricity needs. This estimate is based on the assumptions of a 50 percent energy recovery ratio for gasification, a 25 percent electricity generation efficiency, a 0.05 percent (by weight) generation of grain dust from the grain handled and an electricity need of 2 kWh per 100 bu of grain handled.

CONCLUSION

Steam gasification experiments were conducted with corn dust in a 5.08 cm fluidized-bed reactor. The results indicated that the yield of produced gas increased linearly with an increase in the temperature of the reactor and ranged from 0.13 m^3/kg (2.0 SCF/lb) at 867 K to 0.73 m^3/kg (11.7 SCF/lb) at 1033 K. It was also found that the produced gas heating value ranged from 9.4 MJ/m^3 (251 Btu/SCF) at 867 K to 11.5 MJ/m^3 (308 Btu/SCF) at 1033 K, with a maximum of 11.6 MJ/m^3 (311 Btu/SCF) at 1002 K. The results also indicated that the concentrations of H_2, CH_4 and C_2H_4 increased linearly with an increase in the temperature over the temperature range studied. The concentration of H_2 ranged from 41.3 percent at 867 K to 49.0 percent at 1033 K; that of CH_4 ranged from 3.1 percent at 867 K to 5.7 percent at 1033 K; and that of C_2H_4 ranged from 1.0 percent at 867 K to 3.6 percent at 1033 K. The concentration of CO which decreased linearly as temperature was increased

ranged from 24.3 percent at 867 K to 11.8 percent at 1033 K. The concentration of CO_2 ranged from 33.4 percent at 867 K to 31.9 percent at 1033 K, with a minimum of 25.1 percent at 954 K; that of C_2H_6 averaged 0.1 percent over the temperature range studied. The energy recovery ranged from 8 percent at 867 K to 55 percent at 1033 K. If the produced gas were converted to electricity it would provide about 36 percent of a grain elevator's needs.

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