SEWAGE SLUDGE: A FASCINATING FEEDSTOCK FOR CLEAN ENERGY

M. Rashid Khan, M. McMahon, S. J. DeCanio
Texaco Research & Development, P.O.Box 509, Beacon, NY 12508

Sewage sludge is composed of organic and inorganic materials. The organic portion of the sludge is predominantly composed of C, H, N, and S. On a dry-basis, the heating value of sludge is greater than that of oil shale or tar sand. The volatile matter content of dry sludge can be higher than that of the high volatile bituminous coal. Available correlations in the literature, developed for coals, were applied to predict the experimentally determined heating values. In addition, the sludge compositional data (C, H, S, and ash) were used to develop a new correlation specifically for raw sewage sludge. Compared to the models tested, the new correlation developed in this study for sewage sludge provided a better fit between the measured and predicted values.

KEYWORDS: Sludge, heating value, composition, coal

INTRODUCTION AND BACKGROUND

Treatment plants receive tremendous quantities of waste-water containing dissolved and suspended solids from a variety of sources including domestic, industrial and urban-offs as well as from storm drainage. Consequently, a variety of organic and inorganic materials can be found in a waste-water treatment plant (1).

Traditionally, the solid residue or sludge, the principal product of primary and secondary treatments, has been ocean dumped or landfilled. However, current federal regulations restrict such traditional practices. The option to dispose of such materials by landfilling also suffers from psychological (e.g., "not-in-my-backyard" syndrome) and genuine environmental concerns (e.g., contamination of ground water or agricultural products and leaching). A recent survey of compositional characteristics of domestic sludges indicate that most sludges can be classified as "hazardous," and consequently not suitable for disposal by landfilling (2). Keeping these alternatives in mind, conversion of sewage sludge to clean fuels via gasification (which readily converts essentially all the organic constituents) to synthesis gas (CO and H_2) for power generation or as chemical feedstock, provides an excellent avenue to utilize this renewable resource (3).

The use of sewage sludge requires a better understanding of its physical and chemical properties. In particular, the ability to estimate its calorific value would indeed be of great importance keeping in mind that the measured heating values of sludge are generally not readily available and the reported data often suffer from a relatively large experimental variation (partly due to possible biological/chemical degradation of samples during various treatments). Correlations are important for justification and modeling of the conversion processes now being developed.

The correlations between the coal composition and heating value were reported as early as 1940. Over 20 different equations are reported in the literature which enable one to calculate the heating value of coal based on the ultimate/proximate analyses (4-9). However, essentially nothing could be found in the literature that could be readily applied to specifically estimate the heating value of sludge.

To examine the utility of existing correlations (developed for coal), the most widely used equations were tested for sewage sludge. Mott and Spooner (1940) claimed that their equation will yield heating values agreeing within 200 btu for the whole range of fuels, from peat to anthracite (4). We, however, were much less successful with this equation for dewatered sewage sludge.

Mason and Ghandi (1980) developed a correlation based on coal samples from the Pennsylvania State University coal data base (6). A comparison of the experimental results and the predicted values (based on Mason and Ghandi's equation) was made. Compared to the equation by Mott and Spooner, this equation (termed Data Base [DB] Equation) did a better job in estimating the heating value of sludge.

EXPERIMENTAL

In this study dewatered sewage sludge samples (originating in various treatment plants of the country) were dried in a lab vacuum oven under N_2. The dry samples were characterized by monitoring the following: ultimate analysis (C, H, S, N), ash content and high heating value. A selected set of samples were characterized in multiple laboratories which included the following: Huffmann Laboratories, Inc. (Golden, CO), Institute of Gas Technology (IGT, Chicago, IL), and Texaco Research & Development (Beacon, NY) to ensure that analyses in various laboratories provide comparable results. In general, the data obtained from various labs were within the variation allowed by the conventional ASTM guidelines for each analyses. All analyses for a given sample were completed relatively rapidly to minimize degradation of samples due to bacterial growth.

The data (30 observations in total) were analyzed by using the Statistical Analytical System (SAS) package developed by SAS Institute (10). The regression programs available in this package were applied to develop an empirical model.

RESULTS AND DISCUSSION:

1. The Heating Value of Sewage Sludge Compared to the Various Fossil Fuels

The mean heating value (gross) of sludge (based on 30 observations) compared to various fossil fuels is shown in Figure 1. The heating value of oil shale (Green River formation of Mahogeny zone; Colorado; 33 gal/ton, described by Khan, 1987) was 3200 Btu/lb (10). The heating value of eastern Kentucky shale can be significantly lower than the western shale considered in this study. The heating value of the Asphalt Ridge basin tar sand (Khan, 1989) was less than 2000 Btu/lb (11). By contrast, the heating value of an average sewage sludge is considerably higher (6400 btu/lb). The heating value of an industrial biosludge observed in this study to be greater than 9000 btu/lb. However, no industrial sludges were included in the data base aimed at developing the new correlation.

Compared to essentially all fossil fuels (excluding petroleum based fuels), sewage sludge has a higher H/C (atomic) ratio (Figure 2). The mean H/C ratio of sewage sludge was 1.65 (based on 30 observations), considerably higher than that of the bituminous coals (Pitt#8) with H/C ratio of 0.89 or a sub-bituminous coal (H/C of 0.96 for Wyodak coal). The H/C of the sewage sludge is comparable to tar sand bitumen (with H/C of 1.5).

In addition to the elements described above, significant amounts of chlorine and various volatile metals can be present in sewage sludge. For example, the chlorine content of one sludge was as high as 0.6% (dry basis). Other volatile inorganics identified in the this sludge included the following: Beryllium (less than 0.02 ppm), Vanadium (less than 1 ppm), and Manganese (900 ppm). However, this study did not consider the role of chlorine or vaporizable metals on the heating value of sludge.

2. Variations in the Sewage Sludge Composition

The mean, standard deviation, minimum and maximum values for the compositional analyses are presented in Figures 3 and 4. The variations in the C, H, N and S content in different samples are shown in Figure 3. The sulfur content for various sludges ranged between 0.18 and 3.61 percent with a mean value of 1.71 (with a standard deviation of 1.05 about the mean). The oxygen content of sludge ranges between 3.5 and 27.8% with a mean of 16.5 (and a standard deviation of 6.3%).

The volatile matter content for the a given sludge ranged between 45 and 62% (dry basis). These values are significantly higher than the volatile matter content of a high volatile bituminous coal (with a volatile matter content of 35%, dry basis). The H/C (atomic) for the data set used ranges between 1.44 and 1.86 with a mean of 1.65 with a standard deviation of 0.106.

Figure 4 shows that the mean heating value of the sludge was 6409 with a standard deviation of 816 (based on 30 observations). The measured values for the sludge ranged between 5261 and 8811 Btu/lb. The heating value and the compositional characteristics of sludge are dependent on the nature of sludge as well as on the degree of digestion (or pretreatment) a sludge has undergone. The minimum ash content for the sludge was 18.9% while the maximum value for the sludge was 58.68% (the mean was 40.5%). The higher ash content generally reflects that the sludge has either been digested or heat-treated to convert a large portion of the organic constituents.

The sludge composition is dependent on the nature of pretreatment a given sludge has experienced. For example, the sludge conditioned by a wet oxidation process (intermediate pressure, 300-400 psi; oxidizing atmosphere; temperature of 250-375 F) has a significantly different analysis and a lower heating value compared to an untreated sludge (low C, H but higher oxygen content compared to the untreated materials).

The compositional differences between various sludges can be significant; these differences will be discussed elsewhere. It is interesting to note that the pyritic sulfur is the dominant sulfur type for several sludge. The presence of this large concentration of pyrite is not typical of domestic sludges but suggests the formation of pyritic sulfur from organic sulfur by bacterial action.

3. Comparison of Various Correlations

Attempts were made to estimate the heating value of sludge using correlations widely reported in the literature applicable for coal (and oil shale). In particular, the equation by Mott & Spooner and the Data Base equations were compared with the newly developed correlation.

The percent variation between the measured and the predicted values were calculated for each model by the following equation:

$$\% \text{ Variation} = 100 \times (\text{Predicted-Measured})/\text{Measured}$$

The variations (between the measured and predicted values) for the three models are summarized in Figure 5. The model developed in this study provides a <u>mean variation</u> of 0.019% between the predicted and the measured values. The <u>maximum variation</u> between the measured and predicted values was never greater than 2.83%, based on the new model. In contrast, the Data Base Equation provides a maximum variation of 10.2% while the equation by Mott & Spooner yielded a maximum difference of 14.8% between the measured and predicted values.

Figure 6 compares the measured and predicted heating values based on the Data Base Model. The disagreement between the predicted and measured values in Figure 6 is much larger than those shown in Figure 7 (based on the new correlation developed in this study).

4. Evaluation of the New Correlation

Figure 7 compares the measured and predicted heating values calculated based on the correlation developed in this study. The following equation describes this model:

$$Q\ (Btu/lb) = 241.89*C + 264.26*H + 236.2*S + 20.99*Ash - 4174.68$$

R^2 for the model is 0.99. The parameter measures the proportion of total variations explained by the regression. It is calculated by dividing the sum of squares due to regression by the total sum of squares. R^2 is related to correlation coefficient, r, by the following in simple linear regression: r = square root of R^2. In addition, r, has the same sign as the slope of the computed regression.

The F value for the model was 650.6. The F ratio is the ratio produced by dividing the mean square for the model by the mean square of error. It tests how well the model as a whole (after adjusting for the mean) accounts for the behavior of the independent variable.

The P value for the model was 0.0001. P defines the "observed level of significance." In statistical terms, the level of significance, alpha, of a test is defined as the probability of rejecting the null hypothesis (i.e., no linear relationship between the dependent and independent variables) given the null hypothesis is true. The P-value gives us the largest value of alpha that would lead to the acceptance of the null hypothesis. In other words, from statistical standpoint, the correlation developed is highly significant.

SUMMARY & CONCLUSIONS

The following conclusions are derived based on this study:

- o The heating value of dry municipal sewage sludge is considerably higher than tar sand or oil shale but lower than that of bituminous coal. The atomic H/C ratio of sewage sludge, however, is higher than that of bituminous coal, but comparable to the H/C ratio of oil shale. Some industrial biosludges can have heating value comparable to that of low rank coal.

- o The volatile matter content of sludge is higher than that of coal, oil shale or tar sand.

- o The compositional characteristics (C, H, N, S and ash) of sludge can vary widely among sewage sludge of different origin. Wet oxidation of sewage sludge significantly reduces its heating value as well as its C and H content.

- o The conventional equations developed for coal are not readily applicable for sewage sludge. The equation developed in this study serve reasonably well for estimating the heating value of sludge of various origin based on its analysis (C, H, S, and ash).

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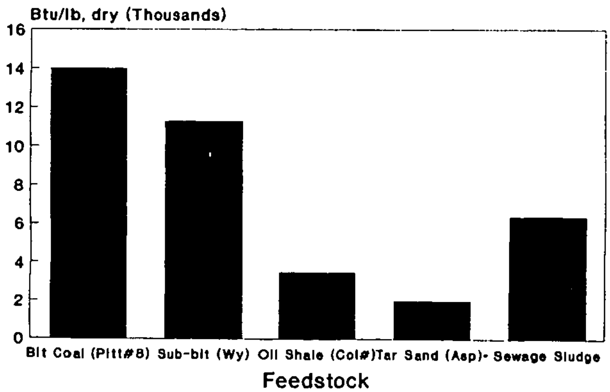
Figure 1

Comparison of Heating Values
of Various Feedstocks (Dry-basis)



#33 GPT; Khan, Energy Fuels #87/-89

Figure 2

Comparison of H/C (Atomic)
of Various Feedstocks (Dry-basis)

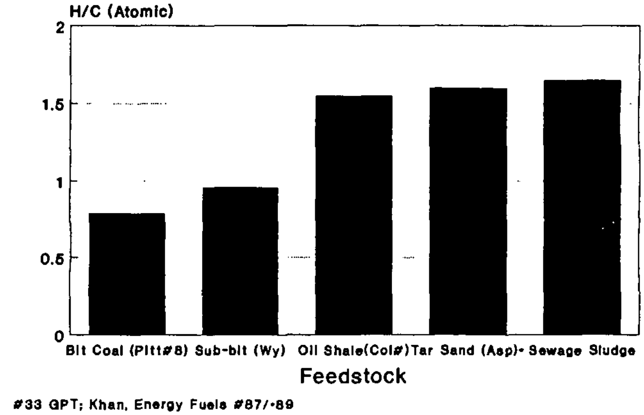


Figure 3

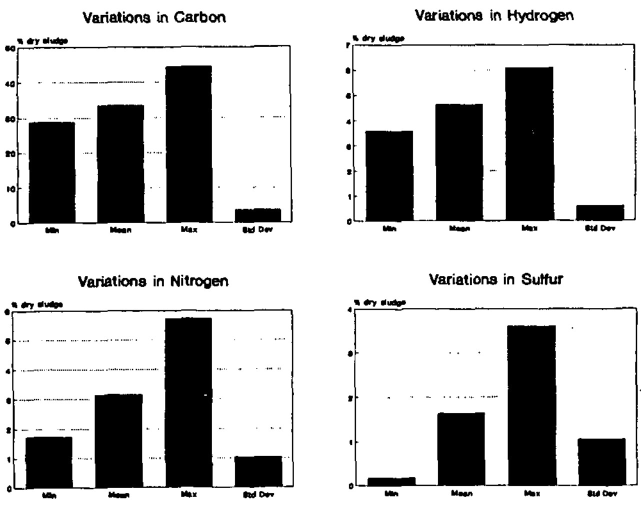


Figure 4

Variations in Sludge Heating Value   Variations in Ash Content

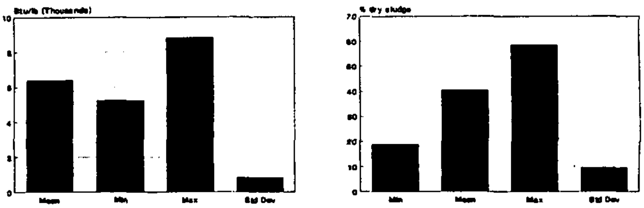


Figure 5

Comparison of Various Models

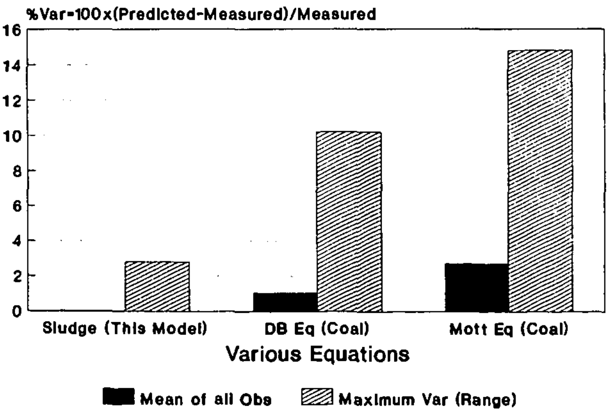
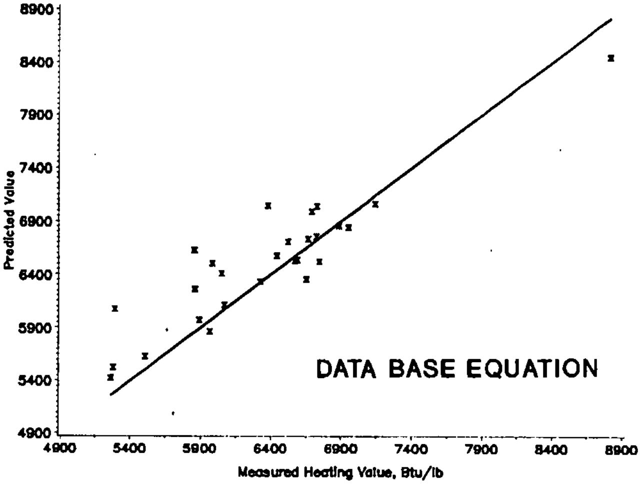


Figure 6

Predicted & Measured Heating Value For Sludge



Predicted & Measured Heating Value For Sludge

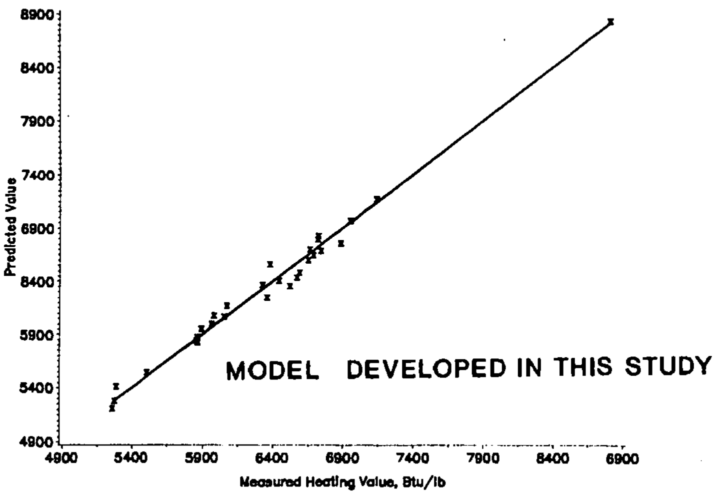


Figure 7

DRYING FUELS WITH THE CARVER-GREENFIELD PROCESS

Ernest A. Kollitides and Thomas C. Holcombe
Dehydro-Tech Corporation
6 Great Meadow Lane, E. Hanover, NJ 07046

Keywords: Drying, evaporation, solvent extraction

INTRODUCTION

Over 80 plants utilizing the Carver-Greenfield (C-G) Process ® have been licensed worldwide during the past 30 years to dry and convert a wide variety of sludges and other wastes into valuable products, such as fuel, animal feed, and fertilizer. Many of these plants generate energy from such wide-ranging feedstocks as sewage sludge, industrial bio-sludge, wool scouring waste, wood pulp mill sludge, chocolate processing waste, brewery sludge, and dye waste. Table I summarizes those C-G Process™ applications in which some or all of the dried wastes are used as fuels.

PROCESS DESCRIPTION

The patented C-G Process uses the innovative approach of dispersing the waste in a solvent and evaporating all the water out of the suspended solids. A simplified process flow diagram of the C-G Process is shown in Figure 1. The waste is mixed with a solvent to form a slurry containing about 5 lbs solvent/lb solids. The solvent properties can be optimized for the particular application. In cases where the dried solids are to be used as a fuel, a petroleum oil is typically used as the solvent. The waste/solvent slurry is circulated through an energy-efficient evaporator system to evaporate virtually all of the water from the solids. Water levels below 5 percent are typically achieved. Either multi-effect evaporation or mechanical vapor recompression is used to achieve very low energy requirements, in the range of 300 to 500 BTUs per pound of water evaporated, versus over 2000 BTUs per pound for conventional drying processes. The solvent fluidizes the mix and creates a low slurry viscosity, thereby ensuring high heat transfer coefficients and minimal scaling/fouling of the system as the water is evaporated. While evaporation of the water is going on, the solvent extracts oil-soluble contaminants from the waste as well. Since the slurry is raised to over 250°F during processing, any pathogens or other micro-organisms present in the waste are destroyed, thereby avoiding problems with product handling and storage.

After evaporating the water, the slurry is fed to a centrifuge to separate the bulk of the solvent from the dried solids. The residual solvent is removed from the centrifuge cake by "hydroextraction", a desolventizing step involving evaporation and gas stripping. Essentially all of the solvent is recovered and reused in the process. Extracted solvent-soluble compounds can be recovered by distilling the solvent and burned as a fuel separately from the solids.

Presented at the "Clean Energy from Waste" Symposium during the ACS National Meeting, New York, August 26-30, 1991.

Through commercial experience and continuing research, a simpler, less costly and highly reliable version of the C-G Process has been developed. The new process configuration eliminates the more complex equipment used in earlier plants, such as vacuum hydroextraction and slurry-to-slurry heat exchangers. Feedstocks with widely varying compositions can be handled easily, without the need for recycling dry solids.

SLUDGE DRYING

The drying of municipal sewage sludge is an important application of the C-G Process. As shown in Table I, the City of Los Angeles, the County of Los Angeles, and the City of Tokyo have all recently built C-G Process facilities to dry sludge prior to incineration for the generation of steam or electricity. Sewage sludge typically has an energy content of 6,500 BTU per pound of dry solids, although the moisture content has a major impact on the availability of the energy. As shown in Figure 2, drying sewage sludge completely creates a fuel with a net available heat of almost 4000 BTUs per pound of solids. Drying sludge also permits high flame temperatures (2000 to 3000 °F), thereby improving the quality of the flue gas by maximizing the destruction of any toxic compounds present.

The C-G Process facility at the City of Los Angeles is designed to handle over 400 dry tons per day of sewage sludge derived from 3.5 million people living in a 600 square-mile area. The dried sludge is fed to a fluidized bed gasifier which employs recycle flue gas followed by two stages of after-burning. This staged combustion of the pyrolysis gas results in minimum NO_x formation and results in air emissions below those experienced prior to this energy-recovery project. The design energy equivalent of the dried sludge is over 1000 barrels of oil per day.

PEAT DRYING

Most of the peat used to produce energy today is sun-dried in production fields. This method is obviously highly dependent upon the weather and has a short production season, especially in northern countries such as Finland. Working with the Technical Research Centre of Finland, laboratory and pilot plant tests were conducted to demonstrate the effectiveness and efficiency of the C-G Process in drying peat.

One of the important side benefits of the C-G Process is its ability to extract compounds from the solids during the drying steps. Bitumen is a major constituent of peat which can be extracted. Extracting bitumen prior to combustion or gasification is an advantage since the bitumen can be sold as a valuable by-product and it minimizes waxy depositions in the combusting equipment.

Studies were done to determine the solvent which optimizes total process performance. Iso-octanol was found to provide the best balance between bitumen extraction, fluidization capabilities, and cost.

REFINERY WASTES

Several tests have demonstrated the ability of the C-G Process to dry and detoxify refinery wastes, such as slop oils, DAF sludges, API separator bottoms, tank bottoms, bio sludges, and primary/secondary emulsions. In one typical study, a hazardous oily sludge was taken from a refinery wastewater pond in the Northeast and tested in DTC's laboratory. As shown in Table II, the material contained 1.3 pounds of indigenous hydrocarbons per pound of solids. One sample was treated with solvent once. A second sample was treated with distilled solvent twice. The concentrations of indigenous hydrocarbons in the solids were only 2.0 and 0.2 weight percent, respectively, after treatment. The original slop oil was unsuitable for landfilling due to the presence of a variety of hazardous hydrocarbon compounds. After treatment by the C-G Process, an independent testing laboratory determined that both solids samples met all the requirements for non-hazardous landfilling as specified by the U.S. EPA. The extracted indigenous hydrocarbons can be recycled to the front end of a refinery or burned as high BTU content fuel.

CONCLUSIONS

The C-G Process, licensed by Dehydro-Tech Corporation, is a versatile process for drying and solvent extracting wastes to produce fuels and byproducts. It has been used for a wide variety of applications and is being considered for new applications, such a peat and refinery wastes. The technology should play, a growing role in the "waste to energy" markets.

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TABLE I
CARVER-GREENFIELD PROCESS WASTE TO FUEL APPLICATIONS

Location	Capacity (US DTPD)	Feed Type	Feed Solids (wt%)	Product Use	Startup Date
Los Angeles	420	Digested municipal sludge	19	Combustion for electricity generation	1992
Japan	50	Undigested municipal sludge	20	Combustion for energy and cement additive	1989
Italy	1.3	Digested municipal sludge	40	Combustion for electricity generation	1988
Los Angeles	400	Digested municipal sludge	20	Combustion for electricity generation	1987
Russia	48	Dye wastes	14.3	Fuel	1986
Italy	55	Dye waste sludge	20	Fuel	1985
Virginia	20	Bio-sludge and wool scouring waste	2	Fuel and lanolin	1985
Denmark	1	Refuse derived fuel and sewage sludge	10-50	Utility boiler fuel	1985
Washington State	50	Wood pulp mill activated sludge	13	Animal feed and fuel	1979
Indiana	24	Pharmaceutical plant wastes	2-4	Fuel	1978
Pennsylvania	0.3	Chocolate waste	2	Fuel	1978
Colorado	30	Brewery undigested sludge	4	Animal feed and fuel	1977
Japan	15	Undigested and digested municipal sludge	4.5-5	Utility boiler fuel	1976
Japan	25	Primary sewage sludge	2	Utility boiler fuel	1975
Mexico	7.2	Pharmaceutical wastes	6	Fuel	1975
Japan	1.3	Primary sewage sludge	20	Fuel	1973
Indiana	9.6	Pharmaceutical wastes	2-4	Fuel	1970
Pennsylvania	10	Undigested air floatation sludge	8	Utility boiler fuel	1964
Total	1,167.70				

TABLE II
SLOP OIL TREATMENT RESULTS

Component	Feed Composition (weight percent)	Treated Solids Composition (weight percent)	
		Treated Once	Treated Twice
Solids	12.0	97.8	99.6
Indigenous Hydrocarbons	16.0	2.0	0.2
Water	72.0	0.1	0.1
Solvent	0.0	0.1	0.1
Total	100.0	100.0	100.0

FIGURE 1
SIMPLIFIED CARVER-GREENFIELD PROCESS FLOW DIAGRAM

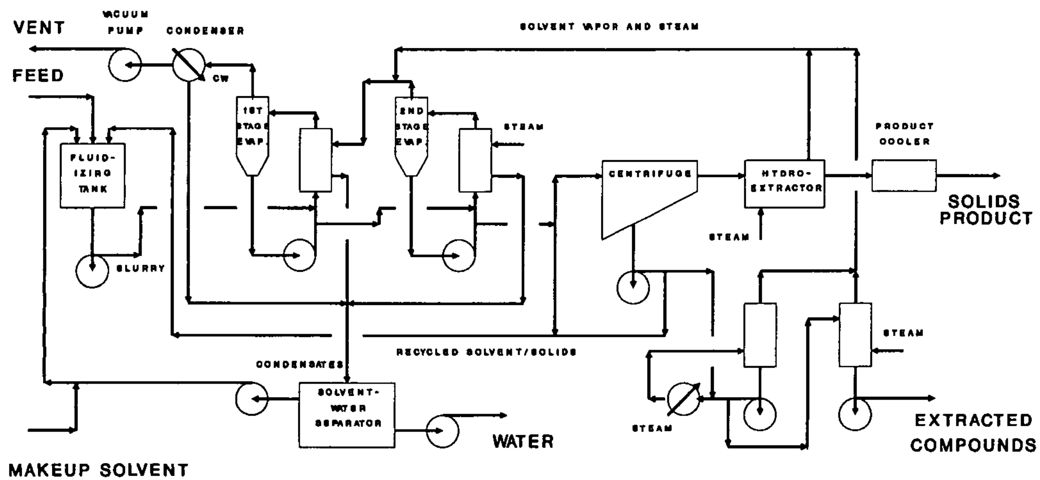
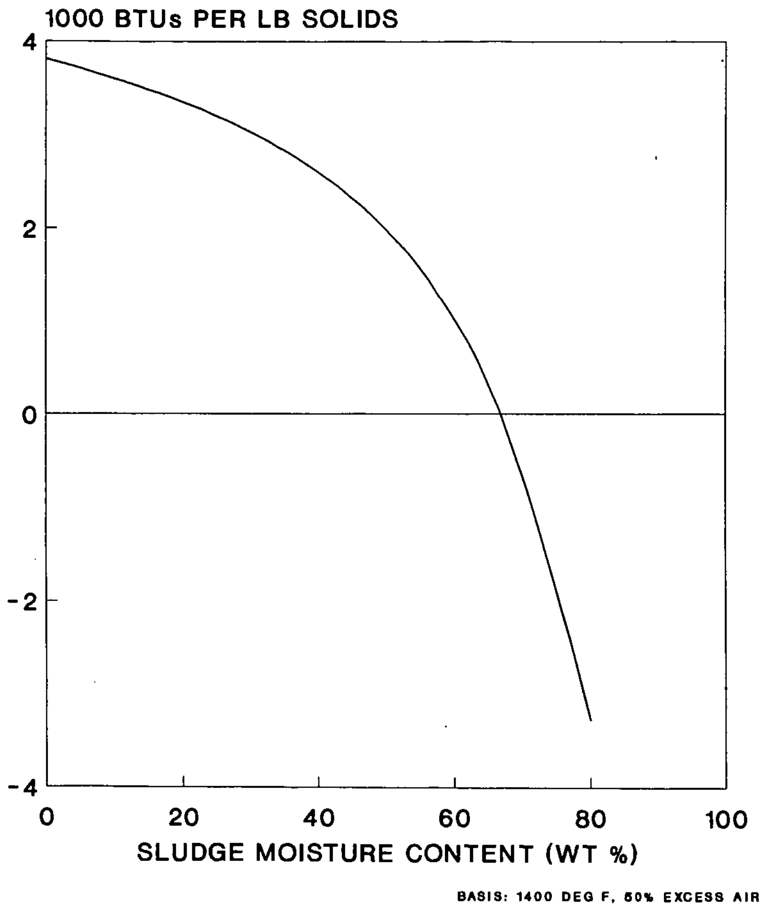


FIGURE 2
NET HEAT AVAILABLE IN SEWAGE SLUDGE



PRETREATMENT OF MUNICIPAL SEWAGE SLUDGE FOR GASIFICATION

Matthew A. McMahon, M. Rashid Khan, C. Albert and R. McKeon
Texaco Research and Development
P. O. Box 509, Beacon, NY 12508

Keywords: Sewage sludge, rheology, slurry solids loading

ABSTRACT

The flow behavior of coal water slurry is significantly degraded when untreated sludge is mixed with coal at a sufficient concentration. Various methods of treating sludge were evaluated in an effort to make coal slurries (containing more than 25 per cent sludge) or treated sludge relatively more fluid so that they could be pumped through pipes and nozzles into a pressurized gasifier. Drying sludge in commercial dryers at temperatures ranging from 180 to 400°F significantly improved its slurrying characteristics with coal. The fluid characteristics could also be improved by removing water under vacuum, dewatering with high intensity filter presses and subjecting the sludge to shear stresses. Slurry viscosity measurements were made at 70 to 212°F in viscometers.

INTRODUCTION

Over 26 billion gallons of waste water are treated by about 15,000 publicly owned treatment works in the United States serving >70% of the population (1). This treatment results in the production of 7 million metric tons per year of sewage sludge. Most of this is applied to the land while about 20% is incinerated and another 6% is dumped into the ocean. The recent ban on ocean dumping along with a decreasing number of landfills and other environmental concerns have created a need for environmentally sound sewage disposal alternatives.

The Texaco Coal Gasification Process, which has operated satisfactorily in large scale facilities for several years, appears to offer attractive features as such an alternative. Coal slurries containing about 60% coal are a usual feed for the process and concentrated sludge slurries in the form of sludge filter or centrifuge cakes containing 70 to 80% water are a usual product of water treatment plants. Sludge in this form is a low quality fuel, however, with an insufficient btu content to be gasified alone in the process. It must therefore be mixed with an auxiliary higher quality fuel such as coal, oil or gas to form a satisfactory feed for the process.

In addition to having a satisfactory heat content, slurry mixtures that are suitable feeds for the process must be pumpable at high concentration and must contain sufficient sludge to justify the incremental cost of handling it.

This paper describes the results of our efforts to characterize the fluidity properties of sludge/coal slurries and to identify a treatment process that would enable sludge concentrations in pumpable slurries with coal to be increased to practical levels.

The hydraulic transport of particulate solids has recently been reviewed (2). Campbell and Crescuolo examined the rheological characteristics of dilute sludge slurries (3). Beshore and Giampa reported on the rheological properties of concentrated coal slurries containing small amounts (up to 10%) of sludge (4). No detailed studies of the rheological characteristics of coal slurries containing high concentrations of raw or thermally treated sludge have been reported. A fuel comprised of raw (undewatered) sludge and coal has also been claimed to be pumpable and useful as a boiler fuel (5).

EXPERIMENTAL

The viscometer used for this work was developed in our laboratory and calibrated with oils of known viscosity. Usually, apparent viscosity vs. solids concentration curves were obtained from which we determined the total solids that could be included in a slurry at a given viscosity. Replicate measurements indicated that the standard deviation of measurements using this technique was 0.79. About 80 grams of slurry was required for each measurement. Slurries were prepared by mixing the desired amounts of sludge, coal and water to a measurable consistency in the measuring cup and noting the torque at a stirrer speed of 600 rpm. Measurements were then repeated as incremental amounts of water were subsequently added. Torques were related to viscosities by measurements on the oils of known viscosity. In addition, a Haake viscometer RV-100 was also used for rheological measurements.

To establish the patterns of sludge behavior, dewatering of sludge was achieved by an advanced dewatering technique, high intensity press (HIP). The HIP was simulated by distributing dewatered cake on a 4"x4" piece of filter fabric which was supported by a specially designed perforated square metal tray. This was surrounded by a square metal box. A similar piece of fabric was placed over the sample followed by an upper square tray. Pneumatic piston pressure was then applied to the upper tray forcing out entrapped water. The applied pressure was changed in various retention times (zones 1 through 4) to simulate the actual HIP zone pressures. Upon completion of the pressure cycles, the pneumatic lever was pushed up and the pressure box removed quickly. HIP applied a mechanical pressure of 125 psi (compared to only 25 psi for commonly used belt press filters) to effect dewatering. It is important to minimize contact of the dewatered sludge with the wet belt to avoid resorption of moisture by the sludge. The solid content was determined by actual measurement and the throughput was calculated by empirical equations developed by Andritz, the manufacturer of the device.

RESULTS AND DISCUSSION

1. Sludge Characteristics

The digested sludge used for most of the measurements made in this study was obtained from water treatment plants in Los Angeles County, Los Angeles City and San Bernadino County in California. The as-received filter cakes were amorphous, fibrous materials containing 20 to 30% total solids. These materials were not pumpable but could be made so by diluting to about 15% total solids. Polymeric flocculating agents were employed in their preparation. Their composition is compared with coal and peat in Table 1. Digested sludge solids generally contain: 30-60% volatile solids, 5-20% grease and fats, 5-20% protein, 10-20% silica and 8-15% cellulose (6).

2. As Received Sludge/Coal Slurries

For the purposes of this work we needed to know the maximum amount of sludge that could be incorporated into a pumpable slurry with coal. Economics dictated that commercially viable sludge containing feeds should include at least 25% sludge and have a total solids content of above 50%. Experience with coal slurries indicates that slurries having apparent viscosities of about 1000 cP are pumpable. Results of viscosity measurements on a number of sludge/coal slurries containing varying amounts of Los Angeles sludge in Utah-Sufco coal showed that the amount of total solids which can be incorporated into a pumpable slurry decreases with increasing sludge content. It was also apparent that in the 1000 to 2000 cP range, small increases in total solids content of slurries effect large increases in viscosity. None of these mixtures was considered a satisfactory fuel because either their sludge or total solids content was too low at the 1000 cP pumpable viscosity.

3. Rheology of Evaporatively Dried Sludge

Thermal treatment of sludge improves its slurrying characteristics. Heat treated sludge is readily dewatered on filters to solids concentrations of 30 to 50%, while unheated sludge is usually dewatered to about 20 to 30% only with the aid of polymeric or inorganic conditioning agents (6,7,8).

Dilute sludge slurries containing less than 5% solids are thermally treated on a commercial scale throughout the country in different types of dryers and thermal treatment units. Thermal treatment over the wide temperature range encompassed by these processes did, in general, improve slurrying properties. The products of various treatment processes are physically and compositionally different from raw sludge (**Table 2**). Some of the products are dry homogeneous powders somewhat coal-like in appearance while others are quite fibrous containing about 60% moisture. The moist products could not be ground into a powder satisfactorily for slurry testing without drying them first.

Results of viscosity tests on slurries prepared from these materials are summarized in **Table** 3. We usually measured the slurrying characteristics of the neat sludges and of mixtures containing 30% sludge and 70% coal (dry basis). Clearly, all of these materials demonstrated slurrying characteristics far superior to those of untreated sludge. Some treated sludges were coal-like in slurry behavior hardly affecting the fluidity of the coal at low concentrations, while the other materials degraded the fluidity of the slurries to varying extent. This behavior is not unexpected since these materials have not only been heated at different temperatures but under different conditions. For example, sludge is heated while suspended in oil in one process in a manner that oil soluble compounds can be extracted from it, while in others, organic components in sludge are simply volatilized. Overall, the rheological characteristics of sludge and the sludge we treated in various ways were found to be very consistent. The rheology of a thermally treated sludge, as a function of shear rate is shown in **Figure 1.**

Thermally dried sludge could also be slurried in oil but the viscosity of the slurry oil determined to some extent the amount of slurry solids that could be included in a pumpable mixture.

Another factor that influences the rheology of sludge slurry is the applied shear rate. At relatively low shearing rates, the untreated sludge filter cakes could be reduced from an intractable solid with no meaningful viscosity to pourable liquids. This behavior is consistent with the thixotropic nature of sludge. As would be expected, more of this sheared product could be incorporated into a pumpable slurry than the "as received" sludge. This phenomenon is attributable to at least two factors: the first of these is the simple shearing of the cellulosic and polyamide flocculating polymers in the sludge. The second is a consequence of the shearing stresses on the flocculated colloidal particles. The drag forces and unfolding of the flocculated particles probably release trapped water and make it available as a carrier fluid with a consequent decrease in viscosity (**Figure** 2).

In an effort to explore various means for treating sewage sludge for improving its slurrying characteristics with coal, we have found that virtually any means of removing water trapped in the raw sludge filter cake (including thermal treatment, advanced dewatering, shearing, vacuum drying and simple air drying) appear to improve the slurrying characteristics.

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Figure 1

Apparent Viscosity vs. Shear Rate for 38.02 wt%
Thermally Dried Sludge
at 30, 60 and 90 C.

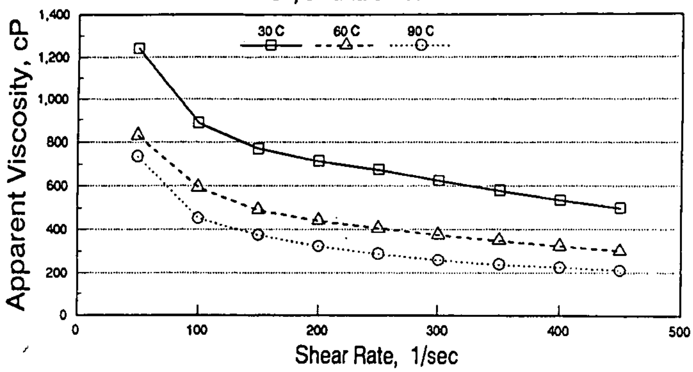

Figure 2

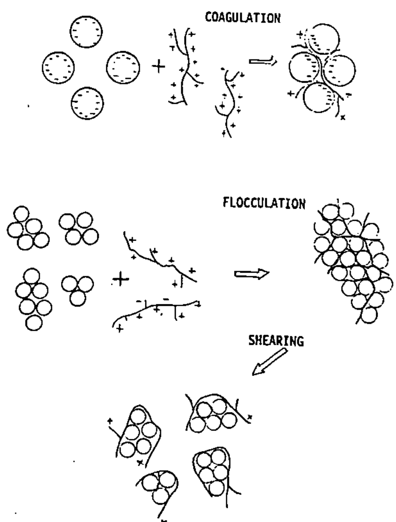

TABLE 1 TYPICAL ANALYSES OF SEWAGE SLUDGE AND OTHER SOLID FUELS								
FUEL	MOISTURE %	ASH	C	H	N	O BY DIFF	S	BTu/lb (DRY)
SEWAGE SLUDGE	80	36	31	4.8	3.9	22.1	1.7	6400
PEAT	83.7	3.4	47.1	5.4	1.4	42.6	0.1	
LIGNITE, TX	29.3	21.5	55.7	4.5	1.0	15.8	1.4	9788
SUBITUMINOUS C, WYOMING	28	7.8	68.1	4.9	1.1	17.2	0.6	11840
BITUMINOUS PITTSBURGH 8	0.8	8.6	76.5	5.1	1.4	5.8	2.5	13765

TABLE 2 ANALYSES OF DRIED SLUDGE FROM VARIOUS PROCESSES					
	LOS ANGELES AS RECEIVED	PROCESS A	PROCESS B	PROCESS C	PROCESS D
% MOISTURE	80.2	3.03	7.93	2.79	9.2
% ASH	36.3	50.91	32.29	31.38	44.0
% C	31.3	30.48	34.71	33.8	28.9
% H	4.84	4.49	4.99	5.32	3.81
% N	3.92	3.96	5.52	2.57	3.21
% S	1.66	1.6	.71	0.62	1.11
% O (BY DIFF)	22.00	8.6	21.8	26.3	19.0

TABLE 3 SLURRYING CHARACTERISTICS OF COMMERCIALLY AVAILABLE THERMALLY TREATED SLUDGES					
PROCESS	MAX TEMP, F	% ASH	SLURRY COMPOSITION SLUDGE %	COAL %	TOTAL SOLIDS AT 1000 CP
A	250	50.9	30 100	70 -	51.0 36
B	1200	32.29	30 100	70 -	58.5 48.2
C	365	33.8	30 100	70 -	45 30
D	358	44.0	30 100	70 -	62 53

PRODUCTION OF CLEAN MEDIUM Btu GAS FROM GASIFICATION OF SLUDGE WASTES

by

K. Durai-Swamy, Prent C. Houck, Qiao D. Feng,
13080 Park Street, Santa Fe Springs, California 90670

Momtaz N. Mansour
5570 Sterrett Place, #206, Columbia, Maryland 21044

Experimental tests were conducted to assess performance of an indirectly-heated, fluidized bed, sludge gasification process employing pulse enhanced heat transfer. Feedstocks included sewage sludge, RDF, lignite, sub-bituminous coal, mild gasification char, SRC residue and black liquor and wastes from pulp mills. Feedstocks were reacted with steam at temperatures of 1150°F - 1500°F to produce a clean medium Btu fuel gas with a higher heating value of approximately 400 Btu/SCF. Heat for the gasification reaction was supplied by means of an integrated pulse combustion system consisting of multiple firetubes immersed within a fluidized bed. This process is being scaled up from laboratory scale (20 lb/hr) to field test demonstration units (2,000 to 6,000 lb/hr). This unique gasification process does not require an oxygen plant to produce medium-Btu gas, thus reducing capital costs significantly. This paper presents the detail of results and progress of scale-up activities.

INTRODUCTION

The potential for recovering energy from renewable sources and organic waste products has been recognized for many years. In recent years and with the realization that fossil fuels (oil and gas in particular) are unrenewable and being depleted at an accelerated rate, the need for an effective technology for utilizing renewable organic sources of energy is a prime consideration for a U.S. Energy Policy.

The paper mill biomass waste is representative of materials discharged from virgin pulp mills and recycle mills located throughout the United States. These sludges typically contain 70 percent moisture as delivered from a belt press. Many mills are currently installing screw presses to reduce the moisture content to 50 percent. The presence of chlorinated organics (such as dioxins) in the sludges from virgin pulp mills, and plastics in the sludges from recycling mills poses serious problems for the landfilling of these wastes. The Hazardous and Solid Waste Act of 1984 (HSWA) restricts land disposal and requires pretreatment at the source prior to final disposal. Increasing disposal costs, diminishing landfill sites, and social and environmental factors are forcing mills to find unique solutions.

MTCI has developed a unique process that reduces the volume of the solid wastes, destroys the chlorinated organics such as dioxins and produces clean fuel gas for use in the mill replacing natural gas.

BACKGROUND

A pulse-enhanced, indirectly heated fluidized bed gasifier system was constructed and tested by MTCI during 1985-1986 for gasification of waste feedstocks under the Phase II of a DOE/SBIR Program. The results of these tests confirmed the technical feasibility of the steam gasification of various feedstocks using the resonance tubes of the MTCI pulse combustor technology as an in-bed heat exchanger. In fact, the system demonstrated a capability for generating a medium-btu product gas of a quality superior to that attainable in air-blown, direct gasification system. The system's overall simplicity, due to the compact nature of the pulse combustor, and the high heat transfer rates attainable within the pulsating flow resonance tubes, provided a decided and near-term potential economic advantage for the MTCI system when compared to alternative direct or indirect gasification systems.

Under Phase II of this DOE/SBIR grant, testing of the gasifier was limited to biomass feedstocks only. In early 1987, Weyerhaeuser Paper Company expressed an interest in testing the MTCI gasifier using black liquor feedstocks. The pulp and paper industry has an ongoing and substantial interest in developing new black liquor recovery methods since the existing technology has significant economic, safety, and environmental shortcomings.

Preliminary feasibility tests were conducted in a 33 lb/hr reactor, which verified the feasibility of the MTCI gasifier with black liquor feedstocks. In order to further develop this technology for mill sludge waste gasification and black liquor recovery, MTCI received funding from the Department of Energy; the Weyerhaeuser Company; and the California Energy Commission's Energy Technologies Advancement Program (ETAP). A scale-up reactor (200 lb/hr) was constructed and tested for verification of the technology. The tests emphasized the collection of definitive process data on conditions and with a broad range of feedstocks.

These projects were completed in 1989 and have yielded extremely successful results confirming the commercialization potential of the MTCI technology to process a wide spectrum of biomass feedstocks as well as the ability to process mill biomass waste from recycling operations. In these tests, the MTCI gasifier was operated using samples of paper mill biomass waste provided by the Gaylord Container Corporation. This waste is currently being landfilled at a significant expense to Gaylord. Despite the high moisture content and presence of plastic material in these waste products, the MTCI gasifier operated without problems.

DESCRIPTION OF THE PULSE COMBUSTOR

The indirect gasifier tested under this program is supplied heat through the resonance tubes of a pulse combustor. The use of pulse combustion significantly enhances the performance and economics of the indirectly-heated gasifier since heat transfer coefficients can be obtained which are 3 to 5 times that achievable in steady-flow systems.

The physical principles involved in the operation of a pulse combustor are essentially the same as those which govern the undesirable pulsations that sometime plague conventional combustion, however, pulsations or combustion-driven oscillations are induces by design and are intended to improve combustion rates, heat transfer, and system performance. The frequency, intensity, and nature of these self-induced combustion oscillations depends upon the precise geometry of the pulse combustion apparatus.

The pulse combustor consists of three main components: an air inlet valve, a combustion chamber, and a tailpipe or resonance tube. The pulse combustor operates over a natural oscillation cycle as shown in Figure 2. In the first step (Figure 2-1), fuel and air ignite within the combustion chamber. Ignition is spontaneously triggered from hot gases of the previous cycle and is not controlled by a spark plug or other external means. In the second step (Figure 2-2), the combustion-induced pressure rise forces the burning mixture to expand outward toward the tailpipe exit. Although some gases escape through the air inlet, the fluidic design of the air valve significantly impedes flow in this direction. In the third step (Figure 2-3), the momentum of the outrushing combustion gases creates a suction within the combustion chamber. This results in the purging and recharging of the chamber with fresh air and fuel. The fourth and final step (Figure 2-4) involves recompression of the fuel/air mixture within the chamber. This is followed by spontaneous ignition to repeat the natural pulsation cycle.

Chamber pressure fluctuations are commonly achieved in pulse combustors. These pressure fluctuations are translated into velocity fluctuations of 600 ft/sec within the resonant tailpipe. The velocity fluctuations, which occur at a frequency characteristic of the combustor (30-300 Hz), intensely scrub the convective boundary layer inside the tube surface, thereby enhancing heat transfer rates.

In the conceptual commercial gasifier designed for atmospheric operation, as shown in Figure 3, the pulse combustors are constructed in modules which can be inserted into the side-wall of a refractory-lined containment vessel. The pulse combustor modules can be installed using techniques employed for conventional heat exchange tube bundles. Each pulse combustor will comprise an individual combustion chamber connected to a multitude of resonant heat exchange tailpipes. Since the pulse combustor is of a modular construction, scale-up of the gasifier is anticipated to be straightforward. In fact, MTCI has tested individual pulse combustors at firing rates of 6 MMBtu/hr which are comparable to the size expected for use in commercial systems.

Using the resonance tubes as a firetube bundle substantially alleviates heat transfer limitations on the flue gas side. This is due to the presence of a vigorous oscillating flow field contained within the resonance tubes. The oscillatory flow field, which causes periodic flow reversal, induces a significant level of turbulence in the boundary layer on the inner walls of the firetubes. This, in turn, gives rise to effective heat transfer coefficients (40 to 50 Btu/hr/ft^2/°F) which is about five times higher than that for conventional firetubes. Thus, the characteristics for the pulse-enhanced, indirectly-heated gasifier overcomes many of the limitations of the state-of-the-art gasifier systems.

GASIFICATION TESTS

Test Systems

Two separate pulse-enhanced gasifier systems were constructed under this program. The first unit consisted of a 20 cm fluid bed reactor (35 lb/hr) enclosing two pulse combustor resonance tubes. This unit was employed to define essential gasification process data. The second gasifier consisted of a 48 cm reactor (90 Kg/hr) containing eight pulse combustor resonance tubes. The larger unit was intended to provide scale-up design criteria for integration of large, multi-tube, heat exchange bundles. Both of these reactors shared essentially similar basic designs.

The main components of the gasifier test rigs used until 1989 are shown in Figure 3.

Steam was supplied to the fluidized bed reactor (R-1) from a boiler (H-1) where it was injected at the base of the bed through a series of sparge tubes. The fluidized bed consisted of sized calcium carbonate (limestone) particles. The pulse combustor (X-2) was fixed to the base of the reactor with its resonance tubes positioned to penetrate the bed in a U-tube arrangement. Combustion gases were then vented through an induced draft fan (F-2). The feed was injected into the fluid bed using a screw feeder. The feeder was charged periodically with a lock hopper.

Product gas from the reactor entered a hot cyclone (V-2) and disengaged particulate matter which was collected in drum (V-3). The product gas was then incinerated (H-2) and scrubbed (V-4) prior to being vented to the atmosphere. In 1990 a quench scrubber and a condensate heat exchanger were added to remove the water from product gas.

In a commercial operation, a portion of these medium Btu product gases would be delivered to the pulsating heat exchanger as a fuel source. Combustion of these gases provides the heat necessary for the indirect gasification process.

Waste heat at the exit of the pulsating heat exchanger is utilized to superheat the process heat. Additional waste heat is available for feedstock drying, if necessary, or for generation of export steam to be used elsewhere in an integrated plant.

Gas samples were extracted downstream of the hot cyclone. The sample gas was cooled in a condenser and knock-out pot prior to being analyzed by gas chromatography. The condensate was collected and sent to an independent laboratory for analysis of tars and oils. In addition, a screw sample valve, located on the reactor shell, allowed continuous monitoring of the bed carbon content.

Test Results

System testing of the indirect gasifier on biomass/waste feedstocks was conducted. The test results provided detailed information on gas compositions, char and tar/oil yields, and bed carbon inventory levels.

Tests were performed using three different biomass feeds - pistachio shells, woodchips, and rice hulls; two different sludge waste products from a recycle paper mill; and a Kraft mill sludge (the two sludge wastes differed primarily in their plastic content). Table 1 summarizes the ultimate analysis for mill waste and other biomass feedstocks employed in the test program. The ash content of the sludge waste exhibited a high degree of statistical variation.

The waste paper sludge was obtained from a mill located in Northern California. The sludge fraction is composed of short fiber and plastic reject material which is recovered from a clarifier. The dilute waste stream is dewatered in a belt press and delivered to a pile where some additional draining occurs. Currently, the product is hauled by truck to a landfill site for disposal. Disposal costs represent a significant expense for the mill and exceeds $600,000 annually. These sludge wastes are representative of high moisture waste materials which are generated in similar mills located throughout the United States.

Table 2 summarizes the operating conditions for the various test runs. Temperatures were varied over the range of approximately 1215°F to 1450°F. Steam to biomass ratios varied from approximately .75 to 2.6. Test run durations typically ranged from four to ten hours. No process operating problems were encountered for any of the runs, including those with rice hulls which have a high ash content and low ash fusion point.

Selected gas compositions for the various feedstocks are summarized in Table 3. The methane content appears to be relatively constant (8 percent to 12 percent) over the range of feeds and processing conditions tested. Higher hydrocarbons show a decreasing trend with increasing temperature and a concomitant increase in hydrogen yields. The ratio between carbon monoxide and carbon dioxide appear relatively constant. The dry gas heating value typically ranged from 370 to 418 Btu/scf.

As seen in Table 3, carbon conversion to dry gas ranged from 92 percent to 94 percent for pistachio shells and woodchips. Char and tar/oil yields for pistachio shells diminished noticeably with increasing temperature (1.3 percent at 987°K). Significantly higher char yields were obtained for both rice hulls and paper mill sludge wastes. The increase in char yields for rice hulls is probably due to the associated high ash content of this feed which tends to inhibit the gasification reaction. The higher char yields for the sludge waste is believed to be due to high rates of entrainment exhibited by this fine, fibrous material. Also, since the sludge waste contained high moisture levels, vaporization of the feed moisture resulted in gas superficial velocities within the fluid bed which were generally higher than for the other feeds tested, thus further exacerbating the entrainment problem. It is anticipated that closer control of gasifier superficial velocities and modest recycle of fly ash can significantly enhance carbon utilization rates. It should be noted that the char and tar/oil levels obtained in these tests are comparable to those achieved for similar feeds in other gasifiers which operate at significantly higher temperatures.

CONDENSATE

Effluent from the gasifier includes condensed steam which contains tar/oil and particulate products. The Biological Oxygen Demand (BOD) level for the condensate was measured at 3920 mg/L. However, a significant portion of the BOD is in the form of carbon particulate which can be clarified or filtered. Further investigation will be needed to determine the achievable quality of wastewater discharge from the gasification plant.

DIOXIN REDUCTION

A key objective of the sludge gasification trial was to determine the dioxin destruction efficiency of the indirect gasifier. Dioxin, which is present in the feed sludge, has been attributed to the pulp-bleaching operation which includes a chlorination step. Dioxin levels were measured for the feed sludge, cyclone ash, condensate, and gasifier bed material. The dioxin feed and effluent results are summarized in Tables 4 and 5. As seen in Table 4, the feed sludge contains a total dioxin concentration of 1543 ppt. By comparison, the gasifier ash effluent contains only 100.2 ppt of dioxin. Including the effect of mass reduction, the net dioxin destruction efficiency is approximately 97.5 percent. It should be noted that dioxin concentrations in the gas product were not measured. However, the gas condensate showed very low dioxin levels and any dioxin that might be present in the gas phase is likely to be destroyed with high efficiency upon subsequent combustion.

The high destruction efficiency for the indirect gasifier is thought to be attributed to the use of calcium material within the fluid bed which serves to absorb HCl released during gasification, and the absence of oxygen in the reducing environment of the reactor.

Recent studies (Ref. Hagenmaier, H., et al) on the occurrence of dioxin in fly ash from waste incinerators have implicated a metal-catalyzed dioxin formation reaction which is facilitated by the presence of surplus oxygen at a temperature regime of approximately 575°F. Above 1100°F dioxin destruction rates exceeded dioxin formation rates. However, at somewhat lower temperatures, it was postulated that a Deacon-type mechanism was responsible for the release of molecular chlorine as shown below:

$$\begin{array}{l} CaCl_2 + \frac{1}{2}O_2 \rightarrow CaO + Cl_2 \\ CaO + 2HCl \rightarrow CaCl_2 + H_2O \\ \hline 2HCl + \frac{1}{2}O_2 \rightleftharpoons H_2O + Cl_2 \end{array}$$

Under this assumption, oxygen in the incinerator flue gas reacts with $CaCl_2$ or other metal chlorides contained in the fly ash to form chlorine which subsequently gives rise to the formation of organochlorine compounds and finally dioxins.

These studies further showed that no such dioxin formation occurred in an oxygen-deficient (nitrogen) environment. On the other hand, dioxin formation increased in an air stream spiked with HCl.

Based on this evidence, there is reason to conclude that gasification of chlorinated waste material may avoid the dioxin-forming reactions that contribute to dioxin emissions from incineration processes.

ASH TOXICITY

EP toxicity extract tests were performed on a typical feed, bed and cyclone ash materials. The results of tests are summarized in Table 6. The results indicate that cyclone char/ash disposal should not present a problem.

CONCLUSIONS

The test results confirmed the ability of the MTCI indirect gasifier to handle a wide range of biomass feedstocks including those with high moisture content, low ash fusion temperature, and high plastic materials content. Also, product gas quality was shown to be quite insensitive to feedstock moisture level.

The gasifier does not require any special feedstock preparation such as pelletization. The gasifier produces a medium-Btu gas without the consumption of oxygen. The reactor is easily scaled, since the pulse combustor tube bundles are constructed in modules. The gasifier produces a gas with a hydrogen to carbon oxide ratio considerably higher than oxygen-blown systems, thus making it particularly attractive for methanol production. The gasifier integrates well with both methanol and high purity hydrogen plants.

The results of these tests have provided a significant data base for preparing designs at the 50 ton/day level. MTCI will field demonstrate the gasifier at a commercial paper mill in the fall of 1991.

SCALE-UP AND COMMERCIALIZATION PLANS

MTCI is currently engaged in the design of scaled up units for 2 ton/hr (TPH) black liquor gasification and 1 TPH mill sludge gasification. The 2 TPH black liquor gasifier will be constructed and field-tested in 1991-1992, and the sludge gasifier will be constructed and operated in 1991-1992. The list of sponsors of these field test plants includes California Energy Commission, Department of Energy, Weyerhaeuser Paper Company, James River Paper Co., and Mead Paper Company. There are other paper companies interested in MTCI technology.

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TABLE 1
ANALYSIS FOR FEEDSTOCKS TESTED IN PULSE-ENHANCED INDIRECT GASIFIER

ULTIMATE (MAF Basis)	Pistachio Shells	Wood Chips	Rice Hulls	Recycle Mill Fiber Waste	Kraft Mill Sludge	RDF Sand Bed	MSW Sand Bed
Carbon	49.51	49.33	49.09	50.00	59.36	46.96	58.20
Hydrogen	6.18	6.74	6.17	6.55	6.90	7.58	8.42
Oxygen	43.96	43.67	44.19	42.76	28.02	43.84	26.45
Sulfur	0.11	0.16	0.04	0.31	1.04	0.86	1.63
Nitrogen	0.24	0.10	0.51	0.38	4.68	0.77	5.30
Total	100.00	100.00	100.00	100.00	100.00	100.00	100.00
PROXIMATE (As rec'd wt %)							
Moisture	8.74	20.46	8.67	49.50	62.40	17.04	5.47
Ash	0.41	0.18	20.48	2.80	7.10	12.66	32.58
Volatile	N/M	N/M	N/M	N/M	N/M	60.23	60.82
Fixed Carbon	N/M	N/M	N/M	N/M	N/M	10.07	1.13
Total	9.15	20.64	29.15	52.30	69.50	100.00	100.00
HHV (Btu/lb), dry	8334	8334	8334	8850	10353	7515	6607

TABLE 2
OPERATING AND PROCESS CONDITIONS FOR BIOMASS TEST RUNS

FEEDSTOCK	TEMP. (F)	AVERAGE FEED RATE (lb/hr)	STEAM RATE (lb/hr)	STEAM TO BIOMASS (lb/lb)	TOTAL FEED (lbs)
Pistachio Shells	1317	35.5	26.0	0.73	337.1
Pistachio Shells	1216	30.6	31.5	1.03	115.3
Wood Chips	1286	22.9	31.4	1.37	205.7
Rice Hulls	1326	30.8	26.0	0.84	185.5
Recycle Paper Mill Sludg	1250	17.6	36.5	2.07	118.8
Kraft Mill Sludge Waste	1250	17.6	36.5	2.07	299.6
RDF (sand bed)	1450	11.0	29.0	2.64	66.0
MSW (sand bed)	1410	12.0	28.0	2.33	62.0
MSW (limestone bed)	1306	15.2	27.0	1.78	84.0

TABLE 4
DIOXIN LEVELS IN PAPER MILL SLUDGE
FEED AND GASIFIER EFFLUENTS
(In parts per trillion, ppt)

	DIOXIN	TCDD	TCDD	PCDD	HXCDD	HPCDD	OCDD
FEED SLUDGE	1543	74	33	69	580	150	670
BED MATERIAL	10.0	N/D	N/D	N/D	N/D	2.9	7.2
CYCLONE ASH	100.2	53	27	14	14	9.5	9.7
CONDENSATE	0.33	0.23	0.07	N/D	N/D	N/D	0.33

TABLE 5
DIOXIN AND FURAN ANALYSIS OF RDF FEEDSTOCK AND CYCLONE CAUGHT ASH IN THE DECEMBER 7, 1990 TEST (ng/g)

	RDF FEEDSTOCK		CYCLONE CAUGHT ASH	
	DETECTION LIMIT	CONCEN-TRATION	DETECTION LIMIT	CONCEN-TRATION
DIOXINS				
TOTAL TCDD	0.56	ND	0.089	ND
TOTAL PeCDD	0.76	ND	0.13	ND
TOTAL HxCDD	0.11	ND	0.091	ND
TOTAL HpCDD	*	0.27	0.23	ND
TOTAL OCDD	*	1.70	0.21	ND
FURANS				
TOTAL TCDF	0.30	ND	0.29	ND
TOTAL PeCDD	0.22	ND	0.13	ND
TOTAL HxCDD	0.30	ND	0.20	ND
TOTAL HpCDD	0.23	ND	0.21	ND
TOTAL OCDD	0.48	ND	0.13	ND

* = Not Supplied

TABLE 6
EP TOXICITY METALS ANALYSIS FOR A TYPICAL GASIFIER FEED AND EFFLUENT

ELEMENT	FEED	BED	ASH
	(mg/L in the EP Tox Extract)		
Ag	<.01	<.01	<.01
As	<.1	<.1	<.1
Ba	<.5	<.5	<.5
Cd	<.01	<.01	<.11
Cr	0.01	<.01	<.01
Hg	<.0005	<.0005	<.0005
Pb	<.05	<.05	<.05
Se	<.1	<.1	<.1

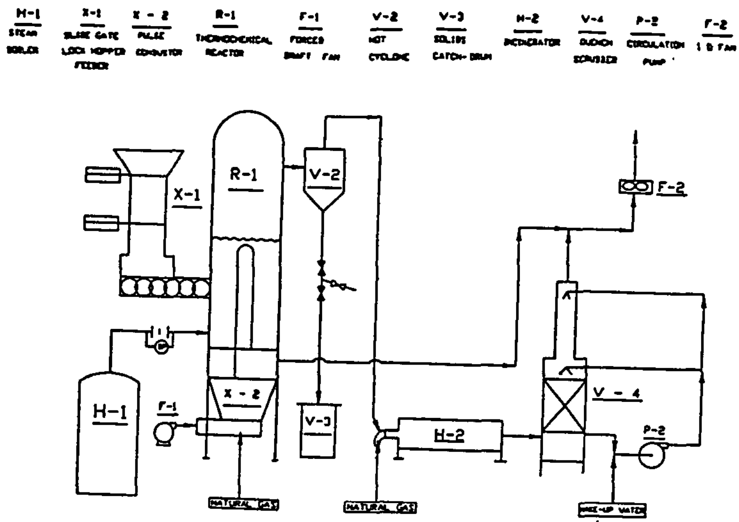

FIGURE 3. PROCESS FOR INDIRECT GASIFICATION PILOT PLANT

TABLE 3
SELECTED GAS COMPOSITIONS AND PRODUCT YIELDS FOR BIOMASS AND MILL SLUDGE TESTS CONDUCTED INPULSE-ENHANCED INDIRECT GASIFIER

COMPOSITION (VOL%)	PISTACHIO SHELLS	PISTACHIO SHELLS	WOOD CHIPS	RICE HULLS	RECYCLE MILL FIBER WASTE	RECYCLED WASTE PAPER W/PLASTIC	KRAFT MILL SLUDGE	RDF SAND BED	MSW SAND BED	MSW LIMESTONE BED
H_2	37.86	35.04	48.11	42.83	38.86	50.50	52.94	45.54	55.21	54.40
CO	18.84	23.43	22.91	19.67	23.34	19.26	11.77	25.26	28.10	25.46
CO_2	28.73	25.20	20.18	24.40	23.27	20.10	21.94	14.51	5.95	5.66
CH_4	10.65	11.31	8.32	11.56	8.31	8.42	8.95	8.30	5.00	5.86
C_2+	3.92	5.02	0.48	1.54	6.40	1.72	3.00	6.38	5.74	8.62
TOTAL	100.00	100.00	100.00	100.00	100.00	100.00	98.60	100.00	100.00	100.00
HHV (Btu/scf)	370	406	329	367	412	364	372	418	374	448
TEMP (°F)	1317	1216	1286	1326	1250	1326	1250	1450	1410	1306
YIELD (% CARBON)	94.1	92.1	93.0	N/A	86.8	N/A	56.0	83.6	93.7	83.8

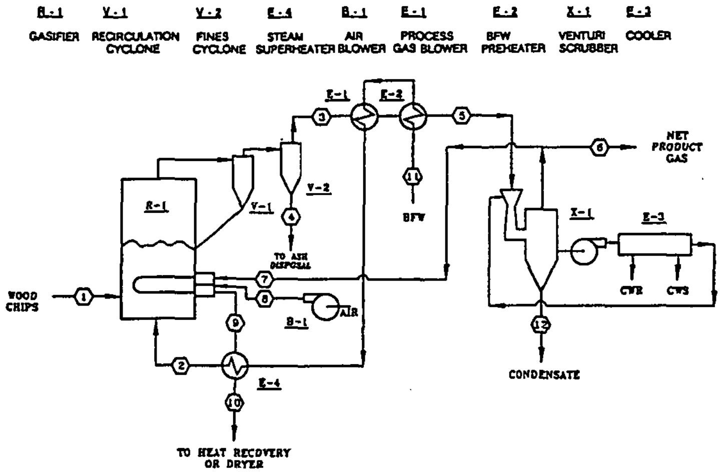

FIGURE 1. PROCESS FLOW DIAGRAM FOR INDIRECT GASIFIER

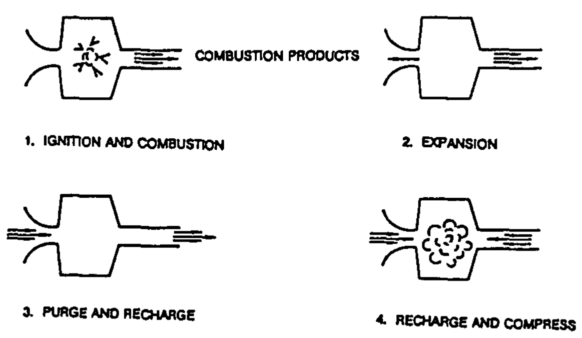

FIGURE 2. OPERATING PHASES OF PULSE COMBUSTOR

BIOMASS FUELED GAS TURBINE DEVELOPMENT

Joseph T. Hamrick
Aerospace Research Corporation
5454 Aerospace Road
Roanoke, Virginia 24014

Keywords: Biomass Fuel, Gas Turbines, Power Generation

INTRODUCTION

Research and development on a 3000 kw wood burning gas turbine power generating system has progressed to the production stage. A system using a General Electric aircraft derivative gas turbine is being prepared for installation at Huddleston, Virginia. The generated power will be sold to the Virginia Power Company. The R&D system located at Red Boiling Springs, Tennessee, will be upgraded for operation with the General Electric engine. Tests were conducted with sugar cane bagasse with good results. Sorghum and sugar cane promise to be major sources of fuel in the future.

Sweet sorghum and sugar cane juices are readily converted to alcohol by yeast fermentation. Sweet sorghum can be grown throughout the United States as well as the tropic and temperate zones of the earth. These plants have the highest conversions of solar energy into biomass of any of the species in the plant kingdom, substantially greater than trees. With the use of bagasse as a fuel for gas turbines in the generation of power, it is possible for the income from power sales to reduce the cost of ethyl alcohol well below that for gasoline. The sorghum grain can be used for fermentation or food. The high volume, high temperature exhaust gases from the turbine can be used to concentrate the juice, make alcohol, dry the bagasse or generate steam for injection into the turbine. There is adequate heat to concentrate the juice and dry the bagasse for year-round use during the harvest period.

Growth of sugar cane and sorghum on the 66.4 million acres of land taken out of production in the U.S. between 1981 and 1988 can supply enough energy to generate 34 percent of the power that was generated in 1986, enough to supply increased power demands into the next century. At the current rates paid by Virginia Electric Power Company for power generated with renewable fuels, 25.4 billion gallons of alcohol can be produced from the profits earned on power sales, enough to supply gasohol to the entire nation.

The system, which can be located at any point where there is a power distribution line and a sorghum or sugar cane source, can provide jobs in the area and an alternative crop for farmers while saving billions of dollars on set-aside payments. At $20/barrel, approximately $8 billion would be saved on the trade imbalance by the reduction of oil imports.

BACKGROUND INFORMATION

Research on wood burning gas turbines was started by Aerospace Research Corporation in 1978. It culminated in the operation of an Allison T-56 gas turbine power generating system at a facility located in Red Boiling Springs, Tennessee. Over two million dollars in U.S. Department of Energy funds and a matching amount of private funds were spent in carrying out the program. In addition, gas turbine engines were furnished by the Air Force and Naval Air Systems Command. A view of the research and development facility is shown in figure 1.

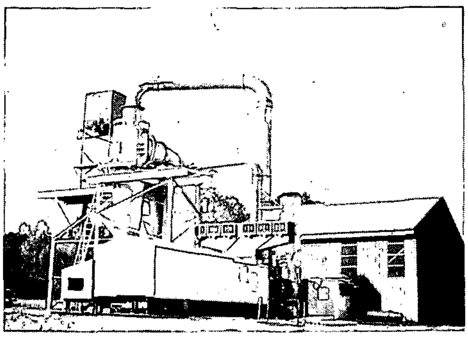

FIGURE 1 View Of Facility At Red Boiling Springs, Tennessee

Operational difficulties which resulted in learning curves peculiar to the system such as wood processing, conveying, drying, combustion, ash removal, engine starting, synchronization with the TVA power distribution grid, and development of emergency procedures are covered in reference 1.* Feeding a pulverized solid into a high pressure chamber and dealing with turbine blade fouling presented the greatest challenge. An anticipated problem that was most feared at the outset, eroding of the turbine blades, never materialized. In over 1500 hours of operation, no erosion has been detected. The measures taken to resolve the two problems and the approach taken with the General Electric LM 1500 gas turbine in meeting the problems are presented.

Modern aircraft engines which require very high power to weight ratios are designed for high turbine inlet temperatures and high compressor discharge pressures. As turbine blade cooling techniques, advanced materials, and more sophisticated design methods have become available the pressure ratios and allowable turbine inlet tempatures have increased to high levels. As a result, the modern aircraft deriviative gas turbines are less suitable for operation with biomass than the earlier models. The current need for low turbine inlet temperature and low combustor pressure with biomass makes the earlier models more compatible. The LM 1500 gas turbine fits well into the biomass picture.

THE ROTARY AIR LOCK FEEDER

The rotary air lock feeder is also referred to as a rotary valve. A schematic view of a rotary feeder is shown in figure 2.

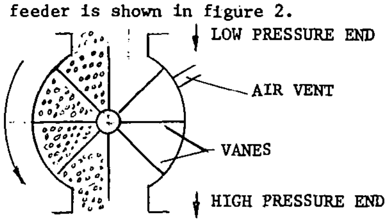

FIGURE 2 Schematic Of A Rotary Air Lock Feeder With Eight Wiper Vanes

* A list of references is included at the end of the paper.

Referring to figure 2, the tips and sides of the vanes are fitted with seals that compartmentalize particles fed into a low pressure sector for movement around to a zone of high pressure and thence into the combustor. A major effort was directed toward development of long lasting sealing methods and materials. Sawdust is an extremely abrasive material that requires special techniques that were developed in the program. To meet the 130 psig requirement of the R&D installation two air lock feeders operating in series proved adequate. To provide conservative design margins, it is planned to use two feeders in series for the 90 psig pressure requirements of the Huddleston installation as well as in succeeding installations up to 6000 kw.

TURBINE BLADE FOULING

At the outset of the R&D program reports on work involved with solids fueled gas turbine systems (References 2 through 5) were reviewed at length. The primary problem with coal fired turbines was erosion of the turbine blades. A secondary problem was fouling of the blades. In work performed by the Coal Utilization Research Laboratory at Leatherhead, England (Ref. 2) it was determined that single cyclones in series adequately cleaned the ash from the combustion gases to prevent erosion. Therefore, it was decided to use only single cyclones in the wood burning program. As a result, there has been no erosion of the turbine blades in the more than 1500 hours of operation with the gas turbines used in the R&D program. In the R&D performed by the Australians (Ref. 3) on brown coal it was found necessary to limit the turbine inlet temperature to 1200°F to avoid deposition of ash on the turbine blades. In the R&D performed at Leatherhead, England with stationery blades there was no significant deposition at 1450°F after 1000 hours of operation with black coal. In tests with pine sawdust in early operation at Roanoke with a small Garrett turbine no significant deposition occurred at 1450°F in 200 hours of operation. In tests with the Allison T-56 at Red Boiling Springs it was found necessary to periodically clean the turbine blades with milled walnut hulls when firing with a mixture of oak and poplar sawdust at 1450°F turbine inlet temperature. Above 1450°F the particles adhered to the blades snd could be removed only by scraping. The 1248°F turbine inlet temperature needed to produce 4000 kw with the LM 1500 gas turbine in the Huddleston installation is well below any problem zone for deposition with sawdust.

DISCUSSION OF LM 1500 GAS TURBINE PERFORMANCE AND FACTORS FAVORING ITS SELECTION

When the advancement was made from the Garrett 375 kw gas turbine to a larger engine, the Allison T-56-9 gas turbine selection was made on the basis of its perceived easy adaptability to the system and the availability of used engines from the U.S. Air Force. As the R&D program advanced, it became clear that the turbine inlet temperature would have to be restricted to 1450°F to avoid excessive turbine blade fouling. The turbine inlet temperature of the T-56-9 is 1700°F at its normal rated overall electrical output of 2332 kw. With a 1450°F turbine inlet temperature the output drops to 1500 kw, a value too low for economical operation.

A search for a more suitable gas turbine from standpoints of availability, adaptability to wood fueling, and electrical output led to selection of the General Electric J-79 gas generator and companion power turbine. The combined gas generator and power turbine was given the designation LM 1500 by General Electric. For aircraft propulsion the hot gases leave the engine at high velocity, propelling the airplane forward. For use in power production the hot gases are ducted to a power turbine. A favorable feature of a two shaft arrangement, such as this one, is that the gas generator can operate efficiently at part load by adjusting its speed downward while the power turbine operates at the required constant speed for power generation. The compressor efficiency is high over a broad range. This is made possible by adjustment of variable stators in the first six stages of the compressor. By adjustment of the stators to match the compressor speed and air flow, rotating stall is avoided and good compressor efficiency is maintained. Rotating stall is a phenomenon associated with flow separation on the compressor blades as the angle of attack on the blades increases with changes in rotative speed and air flow. Compressor efficiency over the speed range results in economical operation over a wide range of power production. Detailed information on turbine inlet temperature, compressor discharge pressure, and wood feed rate as a function of power output was derived from General Electric specification MID-S-1500-2.

Turbine Inlet Temperature - Figure 3 shows a straight line relationship between

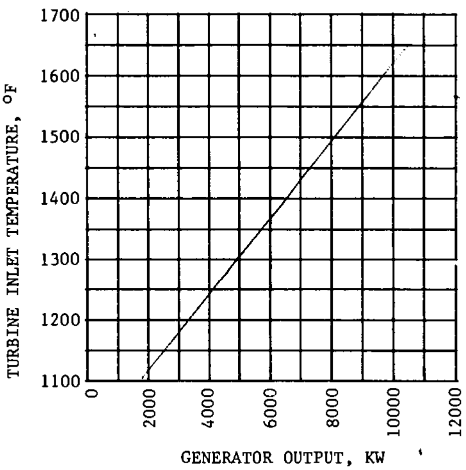

FIGURE 3 Plot Of Turbine Inlet Temperature Versus Generator Output With The G.E. LM 1500 GAS TURBINE At 1000 ft. Altitude And Compressor Inlet Air At 70°F

turbine inlet temperature and generator output. This characteristic provides a significant amount of latitude in operation with untried species of plants or sources of fuels such as clean waste. For example, it can be safely predicted that in the worst case the turbine inlet temperature of 1200°F required for a 3400 kw output will not result in excessive or difficult to clean accumulations on the turbine blades. Mininum performance guarantees would be warranted in such cases. With most wood species a 7000 kw output probably can be tolerated.

Compressor Discharge Pressure - Figure 4 shows a straight line relationship between compressor discharge pressure and generator output. The primary concern with pressure is the feeding of solid fuel into the combustion chamber. The demonstrated maximum sustained pressure in the R&D system is 130 pounds per square inch. Thus, the ability to sustain feeding of sawdust from 3000 kw to approximately 7500 kw is assured. The pressure required to produce the 4000 kw projected for the production facility now being prepared for installation at Huddleston, Virginia is only 90 psig.

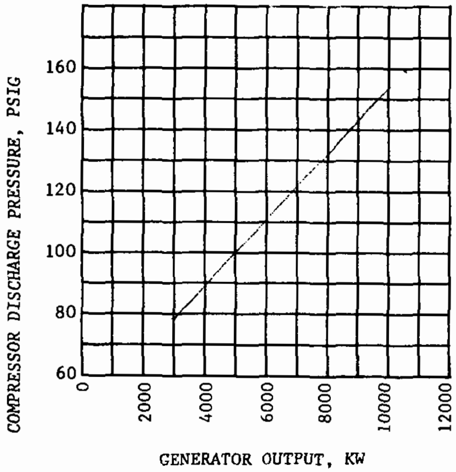

FIGURE 4 Plot Of Compressor Discharge Pressure Versus Generator Output With The G.E. LM 1500 GAS TURBINE At 1000 ft. Altitude And Compressor Inlet Air At 70°F

<u>Wood Feed Rate</u> - The wood feed rate in figure 5 is based upon a heat value of 8,200 Btu/lb for sawdust. The heat value ranges from 8100 Btu per pound for oak to 8,600 Btu per pound for yellow pine. Green sawdust as delivered from the mill averages

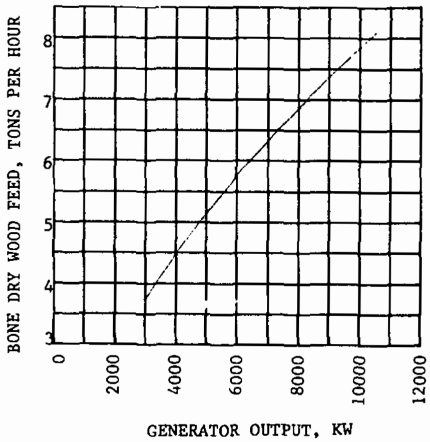

FIGURE 5 Plot Of Wood Feed Versus Generator Output With G.E. LM 1500 GAS TURBINE At 1000 ft. Altitude And Compressor Inlet Air At 70°F

approximately 45 percent water content. Trailers 40 ft. long normally deliver on the order of 25 tons of green sawdust per load. For the 4000 kw output projected for the Huddleston facility five trailer loads per day will be required. Shelter for approximately fifty truck loads will be needed to assure continued operation in the winter months when sawdust delivery can be erratic due to weather conditions.

ENGINE DURABILITY

A question frequently arises as to the life of an aircraft derivative gas turbine in stationary power applications. The answer is that the lower power output and lower turbine inlet temperature that are projected for this application make for very favorable longevity for the LM 1500 gas turbine. A twelve year life or greater before overhaul is predictable. The gas generator in the stationary application is never exposed to the extreme power requirements and high turbine inlet temperatures that exist during airplane take off. The primary requirement for long engine life in stationary applications is adequate filtration of the air entering the compressor.

SUGAR CANE AND SWEET SORGHUM FUELS

Sweet sorghum which is highly drouth resistant can supply two to three times as much fiber energy per acre as trees in some areas in addition to the sugar produced for alcohol and grain for food. The yield from sugar cane, in the areas where it can be grown, is even higher than from sweet sorghum. A further advantage is that there is no stigma attached to its use as a fuel, as there is with trees. This renewable fuel will result in a zero net increase in carbon dioxide. Based on the published research results (Reference 6) for sweet sorghum, the 66.66 million acres taken out of production between 1981 and 1988 can supply the energy to generate 34 percent of the power generated in 1986 in the U.S. The annual payment for setting land aside is estimated to be over $5 billion. Much more additional acreage can be easily devoted to sorghum as an alternative crop. Besides providing fuel for electric power the grain and sugar can produce in excess of 25.4 billion gallons of ethanol which equals fifteen percent of the energy supplied from imported oil. Intensive cultivation of sugar cane and sorghum in states bordering the Gulf of Mexico can result in tripling these outputs.

SUMMARY

The General Electric LM 1500 gas turbine has been chosen for use in the wood burning power production system because of its highly compatible performance characteristics, the ease with which it can be mechanically adapted to the system, and its ready availability. Salient points are as follows:

1. The 4000 kw power output projected for the production system being readied for installation at Huddleston, Virginia can be achieved with a 1250°F turbine inlet temperature and compressor discharge pressure at 90 psig. Both are well below the 1450°F turbine inlet temperature and 130 psig compressor discharge pressure found acceptable in the R&D program.

2. Power outputs up to 7500 kw can be achieved with oak sawdust while remaining below the 1450°F turbine inlet temperature and 130 psig compressor discharge pressure found acceptable in the R&D program.

3. There is adequate distance between the compressor and turbine to adapt the engine to the external burner required for wood and other biomass fuels.

4. J-79 gas generators are readily available on the overhaul and used market. New power turbines are available from manufacturers. In addition, a limited number of

serviceable complete LM 1500 sets are available for immediate use.

5. Both the Red Boiling Springs and Huddleston facilities are ideally located for demonstration of combined electrical power and fuel alcohol production from sweet sorghum.

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FUEL EVALUATION FOR THE U-GAS® FLUIDIZED-BED GASIFICATION PROCESS

A. Rehmat and A. Goyal
Institute of Gas Technology
Chicago, Illinois 60616

Keywords: Gasification; U-GAS Process; Fuel Evaluation

ABSTRACT

The gasification characteristics of a solid carbonaceous fuel in the U-GAS fluidized-bed gasification process can be predicted by laboratory examination of the fuel, which includes chemical and physical characterization, and thermobalance and agglomeration bench-scale tests. Additional design information can be obtained by testing the feedstock in the U-GAS process development unit or the pilot plant.

INTRODUCTION

The Institute of Gas Technology (IGT) has developed an advanced, single-stage, fluidized-bed gasification process, the U-GAS process, to produce a low- to medium-Btu gas from a variety of solid carbonaceous feedstocks, such as coal, peat, wood/biomass, sludge, etc. The development of the process is based on extensive laboratory testing of these feedstocks as well as large-scale tests in a low-pressure (50 psig) pilot plant and a high-pressure (450 psig) process development unit conducted over a period of several years. Up to 98% feedstock utilization with long-term steady-state operation has been achieved. The testing has provided information related to the effect of various gasification parameters, such as pressure, temperature, and steam-to-carbon feed ratio, on gasification characteristics of the feedstocks. The concept of *in-situ* desulfurization by simultaneous feeding of dolomite/limestone has also been established. Reliable techniques have been developed for start-up, shutdown, turndown, and process control. The process represents the fruition of research and development in progress at IGT since 1974. The product gas from the process will be a low-Btu gas that is usable as a fuel when operating with air, and a medium-Btu or synthesis gas when operating with oxygen. The medium-Btu or synthesis gas can be used directly as a fuel, can be converted to substitute natural gas, or can be used for the production of chemical products such as ammonia, methanol, hydrogen, and oxo-chemicals. The low- and medium-Btu gas can also be used to produce electricity generated by a combined cycle or by fuel cells.

On the basis of the operational results with numerous feedstocks, IGT has developed an experimental program for the evaluation of a solid carbonaceous fuel for use in its fluidized-bed gasification technology.

U-GAS PROCESS

The U-GAS process employs an advanced single-stage fluidized-bed gasifier (Figure 1). The feedstock, which is dried only to the extent required for handling purposes, is pneumatically injected into the gasifier through a lockhopper system. Within the fluidized bed, the feedstock reacts with steam and air or oxygen at a temperature dictated by the feedstock characteristics; the temperature is controlled to maintain nonslagging conditions of ash. The gases are introduced into the gasifier at different compositions at different points at the bottom of the gasifier. The operating pressure of the process depends on the ultimate use of the product gas and may vary between 50 and 450 psi. Upon introduction, the feedstock is gasified rapidly and produces a gas mixture of hydrogen, carbon monoxide, carbon dioxide, water, and methane, in addition to hydrogen sulfide and other trace impurities. Because reducing conditions are maintained in the bed, nearly all of the sulfur present in the feedstock is converted to hydrogen sulfide.

The fines elutriated from the fluidized bed are separated from the product gas in two stages of external cyclones and are returned to the bed where they are gasified to extinction. The product gas is virtually free of tars and oils due to the relatively high temperature of the fluidized-bed operation, which simplifies the ensuing heat recovery and gas cleanup steps. The process yields a high conversion, especially because of its ability to produce ash agglomerates from some of the feedstocks and selective discharge of these agglomerates from the fluidized bed of char.

FUEL EVALUATION

Three steps are required to evaluate the suitability of a potential feedstock for the process:

1. Laboratory analyses

2. Bench-scale tests

3. Process development unit (PDU) or pilot plant gasification test.

Laboratory Analyses

Table 1 lists those fuel properties that are normally determined for assessing a solid fuel for use in the process. Additional analyses are performed as required with unusual feedstocks. For example, run-of-mine coals with a high mineral content may require mineral identification and evaluation of the effect of high mineral content on ash fusion properties.

The bulk density, heating value, ash content, and elemental composition of the organic portion of the feedstock usually have no direct effect on the behavior of the feedstock in fluidized-bed gasification. However, they do influence the oxygen requirement, the gas yield, and the gas composition. The higher heating value (HHV) is a measure of the energy content of the feedstock. It relates, with other factors, to the amount of oxygen needed to provide the desired gasification temperature levels. If a feedstock has a low HHV, more oxygen is needed to maintain the gasifier temperature at an acceptable operating level. If the HHV is higher, less oxygen will be required to maintain the desired temperature levels.

The ash fusion temperature reflects the ease of agglomeration of the ash in the gasifier. The free swelling index (FSI) indicates the caking tendency of the feedstock; for highly caking feedstocks, a proper distribution of the feed material, as it enters the gasifier, is critical. In the U-GAS process, the Pittsburgh No. 8 bituminous coal with an FSI of 8 has been successfully gasified and agglomerated with overall coal utilization of 96%. The feedstock is generally sized to 1/4-inch x 0 before it is fed to the gasifier. If a finer size is available, the fluidization velocity is reduced accordingly.

To utilize a feedstock today, one needs to know a great deal about it prior to purchase. It is essential to know the sulfur content to comply with airborne emissions standards and the ash content and its constituents to ensure compliance with solid waste regulations. Other standards are still evolving as new environmental and energy legislation is enacted.

The range of various properties of the feedstocks that have been tested in the U-GAS process development unit or pilot plant is given in Table 2.

Bench-Scale Tests

Three types of bench-scale tests are conducted to evaluate the fuel. These bench-scale tests establish a range of operating conditions that can be used to plan tests in the process development unit or the pilot plant facility, and to perform material and energy balances for the gasifier and estimate its throughput. These tests are described below.

Thermobalance Tests

The gasification of a solid carbonaceous fuel consists of two major steps: 1) initial rapid pyrolysis of the feedstock to produce char, gases, and tar and 2) the subsequent gasification of the char produced. (In addition, some combustion reactions take place if gaseous oxygen is present; these reactions are very rapid.) Because the rate of the second step is much slower than that of the first step, the volume of a gasifier (or the carbon conversion in the gasifier) is primarily dependent on the gasification rate of the char. Due to the relatively well-mixed nature of a fluidized-bed gasifier, the char particles undergoing gasification are exposed to gases consisting primarily of CO, CO_2, H_2, H_2O, and N_2.

The thermobalance testing is performed to determine a relative reactivity constant for the feedstock for comparison with the reference coal, Western Kentucky No. 9 bituminous coal, which has been extensively tested in the thermobalance (Goyal *et al.*, 1989) as well as in the U-GAS process. In the thermobalance, a small quantity of the feedstock is continuously weighed while being gasified at a specific temperature, pressure, and gas composition. This measured weight loss data versus time and the thermobalance operating conditions and analyses of feed and residue are used to calculate the specific relative reactivity constant for the feedstock. The kinetic data, in conjunction with the reference coal information, are used to plan tests in the PDU or pilot plant. As an example, Figure 2 shows the gasification rate for maple hardwood char, peat char, and bituminous coal char, as determined by the thermobalance.

Ash-Agglomeration Tests

Prior to the large-scale testing, the ash-agglomeration tests are conducted in the laboratory to determine the possibility of agglomerating the feedstock ash in the gasifier. These tests are performed in a 2-inch fluidized-bed reactor capable of operating at temperatures up to 2200°F. Several tests have successfully demonstrated that ash agglomerates can be produced in this bench-scale unit at conditions that can be related to the pilot plant operating conditions. The 2-inch reactor has a unique grid design that allows close simulation of the pilot plant fluidized-bed dynamics and mixing characteristics, which are essential for proper ash agglomerate formation, growth, and discharge. The tests are generally conducted at different temperatures, superficial velocities, gas compositions, and operating times to evaluate conditions favoring ash agglomerate formation and growth. The results are quantified using size distribution curves of feed, residue, and fines to show size growth of particles. Visual evaluation of the agglomerates includes separation of the +8 mesh fraction (normally 100% agglomerates) in the residue and, if required, separation of agglomerates by float-sink techniques for each size fraction. The agglomerates thus separated can be easily photographed or examined petrographically. An example of the test results with different coal samples is given in Table 3.

Fluidization Test

A fluidization test in a glass column at ambient conditions may also be conducted to determine the minimum and complete fluidization velocities of the material. This information is then translated into the necessary operating velocity in the PDU or pilot plant test. The fluidization test is conducted only if the feedstock is unusual or if the feedstock size is different than that typically used (1/4 in. x 0) in the process. This test is conducted with the char produced from the feedstock.

Process Development Unit (PDU) or Pilot Plant Test

IGT has two continuous U-GAS gasification units: 1) The 8-inch/12-inch dual-diameter high-pressure process development unit, which can be operated at up to 450 psig and has a nominal capacity of 10 tons per day (at 450 psig operation), and 2) The 3-foot-diameter low-pressure pilot plant, which can be operated at up to 50 psig and has a nominal capacity of 30 tons per day. In addition, a 2-foot/3-foot dual-diameter high-pressure pilot plant has recently been constructed at Tampere, Finland, and its shakedown has begun. Plans are under way to further test various coals, peat, wood and bark waste, and pulp mill sludge in this unit.

A test in the PDU or pilot plant provides the following information:

- It confirms the suitability of the candidate feedstock for the U-GAS process.
- It establishes the base design operating conditions as well as an operating window for the gasifier.
- Design data for fines characteristics, ash agglomeration characteristics, and gas characteristics are obtained.
- Estimates for gas quality, gas yields, and process efficiency are established.

- Necessary environmental data to define the environmental impact are taken.

- Various samples, such as bed material samples, ash discharge samples, fines samples, and wastewater samples, are collected and saved and provided as needed for use during detailed engineering.

The PDU testing is recommended where high-pressure gasifier operation would be required. Each test in the PDU usually consists of 2 days of operation, whereas one 5-day-duration test is usually conducted in the pilot plant with the candidate feedstock. During the test, the gasifier is operated in ash-balanced, steady-state conditions, during which most of the design data are procured. A detailed test plan is generally prepared based on a comparison of the feedstock with a similar feedstock or from information obtained from bench-scale testing. Depending on the feedstock characteristics, the gasifier is operated at a temperature of up to 2000°F and a superficial velocity of up to 5 ft/s.

Numerous solids samples are collected regularly during the test run so that accurate material balances can be prepared. Process sample points include the coal feed, fluidized bed, ash discharge, and cyclone diplegs (for the pilot plant). Samples from the fluidized bed are also collected and analyzed hourly during the test to help the operators determine and maintain steady-state operation.

All process solids and gas flow streams are measured and recorded. Temperatures are recorded for all process streams and at several locations within the reactor. Redundancy is provided for the reactor pressure taps used for bed density and height.

A product gas sample stream is drawn continuously from the gasifier freeboard for chromatograph analysis. The chromatograph system provides accurate on-line analysis for CO, CO_2, CH_4, H_2, N_2, H_2O and H_2S. The chromatograph sequencing is microprocessor-controlled for flexibility in the scope and frequency of the analysis. The product gas samples are also collected in gas bombs for later laboratory analyses.

Special sampling and instrumentation are available for complete chemical characterization (organic compounds as well as trace elements) of the product gas and wastewater streams. Test results of this nature are necessary to satisfy environmental permitting requirements and for proper design of downstream processing equipment. Equipment is also available for determination of product gas dust loading after one, two, or three stages of cyclone separation.

These units use a microprocessor-based data acquisition system to ensure accurate, timely, and reliable collection of all process data of interest. About 85 process data points (temperature, pressure, flow, etc.) for the pilot plant and about 40 data points for the PDU are scanned repeatedly throughout the test. A full scan is completed in approximately 10 seconds and is repeated at 3-minute intervals. The reactor operating status, including various flows, pressures, temperatures, and velocities (grid, venturi, bed, freeboard, cyclone), bed density, bed height, etc., is calculated and displayed on the computer CRT screen. The data are stored on magnetic tape in both raw signal and converted form. The converted data are averaged hourly, and an hourly average report of all data points and the operating status is automatically printed in the control room; in addition, a shift report is printed every 8 hours to allow a shift engineer to review the operation of a previous shift. Particular emphasis is placed on the use of the data acquisition system as an operating tool. Specialized programs have been developed to aid the operators in the approach to and confirmation of steady-state operation. This results in more steady-state operating time, and therefore more useful design data, per test run.

The details of the PDU system and some test results are given by Goyal *et al.*, (1989, 1991). The details of the pilot plant system and some test results are given by Goyal and Rehmat (1984, 1985).

In September 1989, IGT entered into a licensing agreement with the Power Industry Division of Tampella, Ltd., Tampere, Finland, which will result in the commercial application of the process. Tampella selected the pressurized fluidized-bed technology because of its versatility and applicability to a wide variety of feedstocks, including coal, peat, forestry waste, etc. As a first step toward commercialization, a 10-MW pressurized (450-psi) pilot plant has been designed and constructed at Tampella's R&D Center in Tampere, Finland. Various coals, peat, wood and bark waste, and pulpmill sludge will be tested in this unit. To demonstrate the application of the technology to the Integrated Gasification Combined Cycle (IGCC), a detailed engineering of a 650-ton/day plant will begin in 1991 and will be commissioned in 1993-1994. Tampella also plans to demonstrate its IGCC process in other parts of the world, including the United States.

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Table 1. LABORATORY ANALYSES OF THE FUEL

- Proximate Analysis
- Ultimate Analysis
- Higher/Lower Heating Value
- Bulk Density
- Particle-Size Distribution
- Grindability
- Equilibrium Moisture
- Free Swelling Index
- Ash Fusion Temperatures (Reducing Atmosphere)
- Ash Mineral Analysis

Table 2. RANGE OF FEEDSTOCK PROPERTIES TESTED IN THE U-GAS PROCESS

Moisture,* %	0.2 to 41
Volatile Matter,** %	3 to 69
Ash,** %	6 to 78
Sulfur,** %	0.2 to 4.6
Free Swelling Index	0 to 8
Ash Softening Temperature, °F	1980 to 2490
Higher Heating Value,** Btu/lb	2,330 to 13,630

* As received
** Dry basis

Table 3. 2-INCH ASH-AGGLOMERATION TESTS WITH VARIOUS FEEDSTOCKS

Coal Sample	Run Temp., °F	Char Initial Ash, %	Run Time, h	Fluidizing Velocity, ft/s	Comments
FC-1	1985	31.5	2.0	1.0	Sinter particles plus some agglomerates
FC-1	2100	31.5	1.5	1.5	Agglomerates formed, little or no sinter
FC-2	1990	45.5	1.0	1.5	Small agglomerates present
FC-2	1988	45.5	1.3	1.5	Larger agglomerates found
FC-3	2080	45.4	2.5	1.5	Agglomerates formed
FC-4	1960	15.5	1.0	2.1	No agglomerates found
FC-4	1920	20.9	3.0	1.5	Small agglomerates found
FC-4	2000	15.5	2.5	1.6	Greater number of large agglomerates
KY #9	2000	51.0	1.3	1.5	Many agglomerates produced

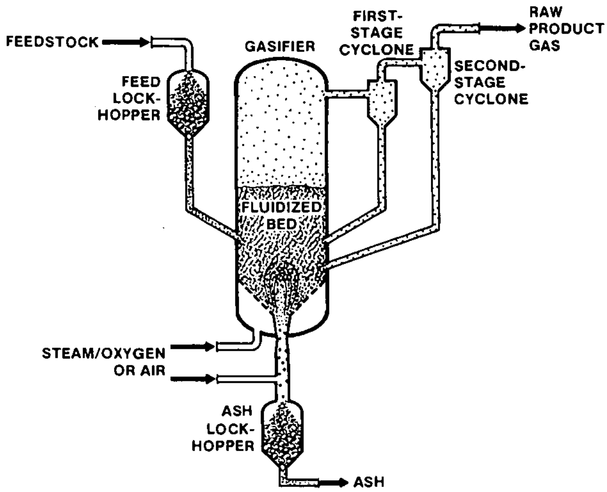


Figure 1. SCHEMATIC DIAGRAM OF THE U-GAS GASIFIER

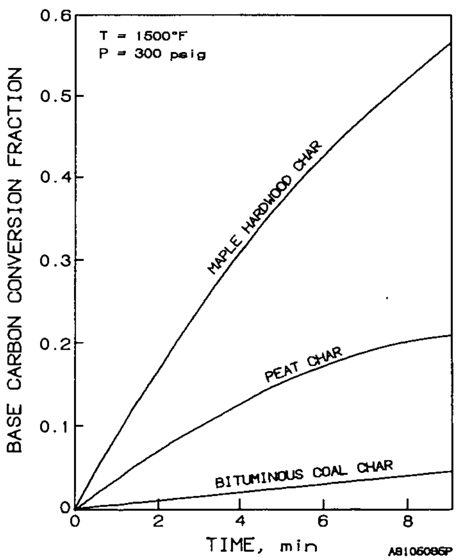


Figure 2. GASIFICATION RATES FOR MAPLE HARDWOOD CHAR, PEAT CHAR, AND BITUMINOUS COAL CHAR

TREATMENT OF MUNICIPAL SOLID WASTE (MSW) BY THE HYDROCARB PROCESS

Meyer Steinberg
Process Sciences Division
Department of Applied Science
Brookhaven National Laboratory
Upton, NY 11973

Keywords: Solid waste, carbon, methanol.

INTRODUCTION

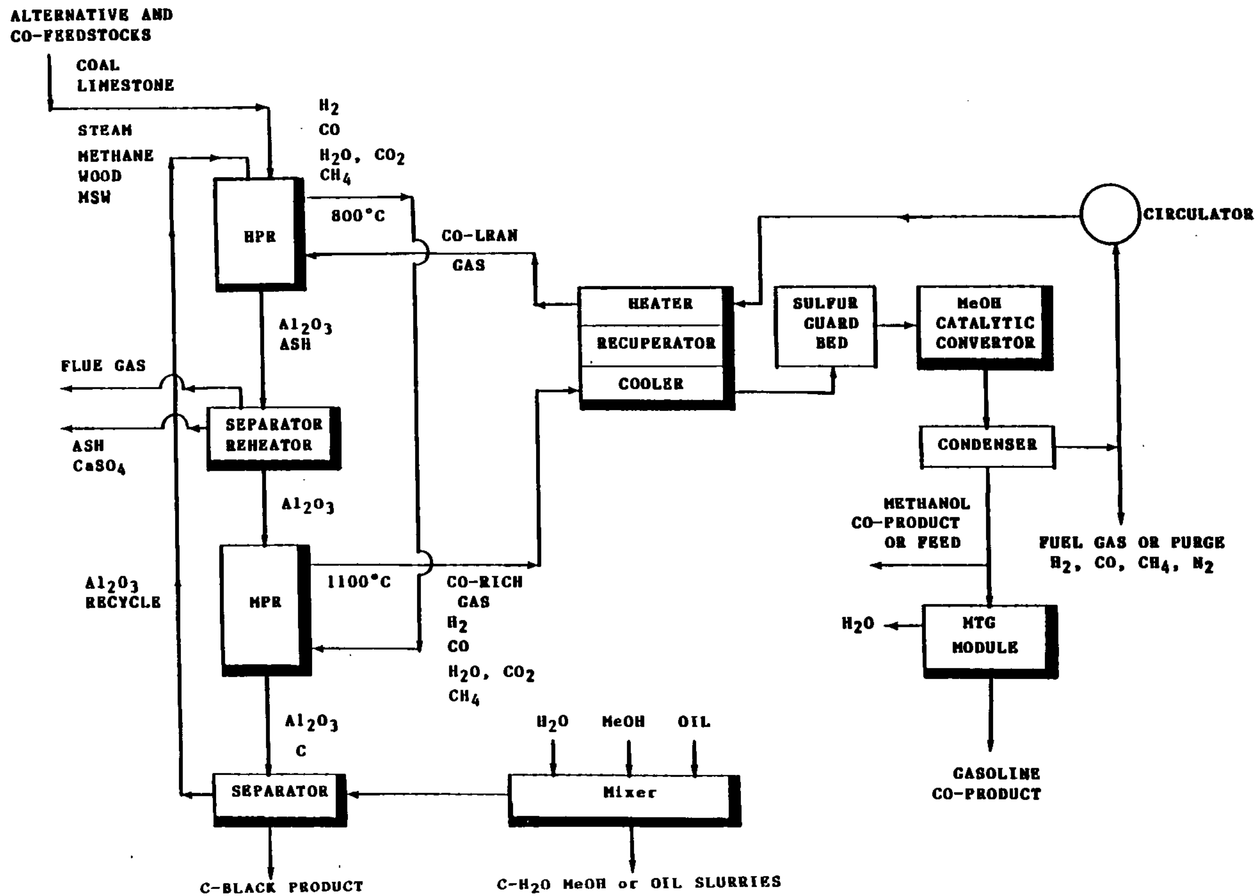
It is now generally known that the municipal solid waste problem has become an ever increasing problem in populated areas in the U.S. The increase in the standard of living manifested by a vast array of consumer goods has added to the problem of disposal of industrial and municipal solid waste (MSW). The land-fill disposal sites around metropolitan areas have become exhausted so that tipping fees are soaring. Municipalities are opting for more waste incineration or mass-burn plants. Legislation is being passed to require separation of waste for recycling and resource recovery. Because separated waste is market demand dependent, the cost of recycling is highly time and location variable. In fact, there are a number of municipalities that pay carters to remove and transport recyclable the waste to other locations which instead of becoming s source of income becomes a liability. MSW roughly consists of 50% paper and plastic and the remainder being glass, metal and kitchen waste. Industrial waste includes paper, wood and used rubber tire discard.

The most traditional waste disposal method is incineration. The modern and improved method for the same process is now termed mass-burn. In some cases, the energy generated is used to produce steam for electricity generation which can be sold, and therefore constitutes a positive value. The problem here is that the mass-burn plant generates potentially polluting gaseous and solid residue effluents. In the gaseous effluent, dioxin has been one of the most elusive and worrisome pollutants and has caused the shutting down of a number of incinerator plants. There are other gaseous pollutants, including volatile refractory organics, chlorine containing compounds, and particulates from plastic and organic waste. The chemical and biological activity in the remaining solid ash residue from incinerators is also a problem which still requires landfilling or other methods of disposal. There is concern that leachates from incinerated ash will eventually contaminate the aquifers. Municipalities are also passing legislation forbidding the use of materials which do not degrade and tend to remain in long-term storage in the landfill, such as plastics. A number of communities are outlawing disposable plastic products and appear to be returning to paper bags and containers. Much effort is also going into developing biodegradable plastics. Whether this is a sound environmental solution is yet to be determined.

HYDROCARB WASTE PROCESS

The HYDROCARB Process offers a viable alternative. The process was originally conceived for the purpose of processing our vast resources of coal to produce a clean carbon fuel.[1,2] However, the process can operate as well with virtually any carbonaceous raw material and certainly a large fraction of MSW qualifies as a carbonaceous material. The process is new and unique and the products formed can be used primarily as premium clean fuels as well as for the commodity market. The process depends on two basic steps, (1) the hydrogenation of coal to form a methane-rich gas while leaving the ash behind and (2) the thermal decomposition of the methane-rich gas to form carbon black and hydrogen which is recycled. The excess hydrogen and oxygen from the co-products can be a hydrogen-rich co-product which can either be hydrogen, methane, methanol or water. Figure 1 shows a

Figure 1
HYDROCARB PROCESS
CLEAN CARBON FROM CARBONACEOUS FEED STOCK WITH METHANOL AND GASOLINE CO-PRODUCT



schematic flow with alternative feedstocks, coal, wood or MSW and with co-feedstock additions.

Figure 1 gives a schematic of the process listing various feedstock materials, additives and co-products. The process can be made very efficient because the only raw material used is the carbonaceous material. The energy required to operate the process is relatively small. The overall reaction is thermally neutral. The primary product is always carbon black which can be used as a clean burning fuel and can also supply the market for vulcanization of rubber for automotive tires, pigment for inks and paints and for lubricants. The co-product hydrogen-rich gas can primarily be used as a burner fuel and the methanol as an automotive fuel, or as a commodity chemical, or can even be converted to gasoline. The process is fundamentally different than mass-burn in that it operates in a reducing atmosphere rather than in an oxidizing atmosphere and it is run in a closed system under pressure. Temperature conditions are about the same or perhaps even somewhat lower than in mass-burn incinerator plants. Because of the elevated operating pressure and reducing atmosphere, no dioxin can be formed and all the oxygen containing organic material is reduced to carbon and methane and any metals that may be present in the waste are kept in their reduced state as opposed to mass-burn where the metals can become oxidized. The following describes how the process can be effectively used in processing MSW and the economic dynamics of the process.

The process can be used with either separated or non-separated waste. To simplify the example and avoid discussion of head-end costs, we will give examples of the process operating on separated waste. Thus, the main MSW feedstock is paper and plastic and we can include rubber tires for this example. Since paper is essentially produced from wood, the process can be represented by the following chemical stoichiometric formula, limiting the products to carbon and methanol.

$$\underset{\text{paper}}{CH_{1.44}O_{0.66}} + \underset{\text{plastic}}{0.44CH_{1.5}} + \underset{\text{nat. gas methane}}{0.14CH_4} = \underset{\text{carbon black}}{0.92C} + \underset{\text{methanol}}{0.66\ CH_3OH}$$

Notice that the formula for plastic contains only C and H, like rubber and methane. The oxygen containing material in paper is in the form of hemi-cellulose. The above equation is based on an assumed MSW composition such that the amount of plastic is 25% of the weight of paper. This can be changed for specific sites and the mass balance adjustment can be made by varying the amount of natural gas added. The gas can be purchased from the local gas company in the particular area where the waste is being processed. We now have to set the production capacity of the plant. Mass-burn incinerator plants have been built in the 2,000-3,000 T/D capacities in and around metropolitan areas. Of course, around New York, for example, it might be worthwhile building a 10,000 T/D or more of waste paper and plastic HYDROCARB plant. However, for this and generally more widespread applications, we will fix on a 3,000 T/D MSW processing capacity which would contain 2,400 T/D paper and 600 T/D plastic.

We now calculate that to run this plant, we have to add 226 T/D of natural gas from the natural gas pipeline company's distributing company. This natural gas is equivalent to 10.7 million SCF/D of methane, which must be purchased from the gas company. The separated MSW is thus co-processed with natural gas.

ECONOMICS

We now must estimate the capital investment of the plant. We can obtain this estimate by scaling down from a large plant we estimated in detail, operating on 25,000 T/D of coal. Because this is a volumetrically controlled process, we can scale it by the well known 0.6 power factor of capacity. The 25,000 T/D plant making carbon and methanol from coal is estimated to cost $\$800 \times 10^6$. Thus, the 3,000 T/D waste plant will cost:

$$800 \times 10^6 \times \left(\frac{3000}{25000}\right)^{0.6} = \$200 \times 10^6$$

We can now calculate a selling price for the carbon black fuel and methanol co-product. The financial parameters operating on the capital investment are as follows: capitalization 80% debt/20% equity, 20 yr depreciation, 11% interest on debt, 25% return on equity (ROE) and 38% tax on ROE before taxes. This results in a 21.9% annual fixed charge operating on the total capital investment.

We assume a high natural gas cost from the gas company of \$5.00/MSCF which equals a cost of \$0.119/lb CH_4. We then add operation and maintenance cost and the 21.9% fixed charges on the \$200 million capital investment. We can now calculate the <u>G</u> price of the MSW value of the waste taken from the municipality, which can range from a negative value, in which case the community pays the processor to take the waste away, to a positive value in which case the processor pays the community to acquire the waste for processing. We shall first calculate a breakeven G price for the waste in \$/Ton in Table 1, assuming we obtain \$5.00/MMBtu for the resulting fuel products.

TABLE 1
HYDROCARB WASTE PROCESSING PLANT
Plant Factor 90%, Efficiency 90%, capacity 3,000 T/D
Production Capacity of Fuel - 11,000 Bbl/D Fuel Oil Equivalent

Production Cost			\$/Day
Waste Cost		=	3,000 T/D x \$G/Ton
Nat. Gas	= 0.119 x 226 x 2,000	=	53,000
Op & Maint	= $\frac{3,000}{25,000} \times 120,000$	=	20,000
Fixed Charges	= $\frac{0.219 \times \$200 \times 10^6}{328}$	=	133,000
			206,000 + 3,000 G

Thus,

$$206,000+3,000\ G = 0.9 \times (3000+226\ T/D) \times \frac{22.9\ MMBtu}{Ton} \times \frac{\$5.00}{MMTU}$$

Solving for G = \$41.50/Ton; this is what the processor can afford to pay the town for taking the MSW for processing and while still obtaining a 25% return on equity.

The above is based on a fuel value for a C-methanol composition makeup mixture of 34.3% carbon in 65.7% methanol by weight. The plant produces 700,000 gal/Day of this C-methanol slurry which is equivalent to 11,000 Bbl/D of fuel oil equivalent.

If we assume the processor obtains the waste from the town free, so that G = \$0/Ton, we can then calculate the selling price of \$3.10/MMBtu for both <u>co-products</u> carbon and methanol. This is equivalent to \$18.70/Bbl oil or \$0.44/gal.

Now if the town pays the processor \$25/Ton to cart the waste away (as some towns on Long Island have already done), then the selling price for carbon and methanol can come down to only $2.00/MMBtu which is equivalent to $12.00/Bbl fuel oil equivalent or $0.28/gal while maintaining a reasonable return on the investment equity.

At $2.50/MMBtu which is highly competitive with oil at $15.00/Bbl, the town would only have to pay $13.50/Ton to a processor to take it away.

The conclusion is that even at a waste capacity of 3,000 T/D and an investment of $\$200 \times 10^6$, the processor can sell the carbon and methanol as a clean burner fuel for domestic and industrial boilers, as well as for diesel and turbine engines at an economically attractive price. Additional return can be obtained by the processor selling the methanol and carbon at a higher price to the chemical commodity market so that the cost of waste disposal would even bring a profit to the town by selling the waste to the processor at a higher price.

The above indicates that the HYDROCARB Process for the disposal of MSW is highly attractive and should be taken up for development on a fast track schedule. Because this process utilizes natural gas for co-processing waste in a reducing atmosphere, not only is the process environmentally acceptable but is potentially economically attractive and thus it should be worthwhile to develop this process in conjunction with a municipality that is generating the waste.

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COMMERCIAL APPLICATION OF WASTE-FUEL FIRED HYBRID FLUIDIZED BED BOILERS

Robert H. Vander Molen and
Robert Rossi
Combustion Power Company
1020 Marsh Road
Menlo Park, CA 94025

Keywords: Fluidized Bed Combustor, Waste Fuels, Boilers

INTRODUCTION

In the early 1980's, developments in California created opportunities for aggressive developers. The California Public Utilities Commission (CPUC) ordered the states utilities to sign power contracts with anyone willing to pay a nominal fee. A range of contract types were ordered. Amongst them was the Standard Offer No. 4 (SO4) type contract which calculated payments on avoided costs that utilities would have incurred had they installed new capacity. The SO4 contract resulted in very attractive payments for both capacity and energy. Before the CPUC rescinded the SO4 contract, nearly 12000 MW were signed up for by developers. Many of these contracts were for small power plants less than 50 MW in size. By staying below 50 MW, review of permits and need by the California Energy Commission was not required. And in most cases, plants burning solid fuels sized less than 30 MW could avoid US EPA review under the Prevention of Significant Deterioration (PSD) regulations.

Combustion Power entered this market with a hybrid circulating fluidized bed boiler. The system uses the positive attributes of a bubbling/turbulent bed combined with a bed-solids circulating loop to develop system performance and emissions profiles which possess substantially enhanced performance characteristics. First commercial operation of the technology was December 1986. At present, a total of 12 units are operating or under construction for a combined steam generation capacity of about 2,500,000 lb/hr (300 MWe). These FI CIRC™ boilers are designed to burn coal and/or a variety of waste fuels having sulfur levels of up to 5.0% and nitrogen up to 3.0%. SO_2 is controlled in typical fluidized bed combustor fashion by introducing limestone into the bed. NOx control, where required such as in California, is attained using a Combustion Power proprietary ammonia injection technique which was developed specifically for the California applications.

DESIGN CONCEPT

The hybrid approach was created by the need in California during the early 1980's to design a solid fuel fired boiler that could meet the very stringent emissions standards in that state. The emissions standards for NOx and SO_2 were typically 0.05 lb/mmBTU for both pollutants. These levels are 5% and 8% respectively of today's Federal EPA New Source Performance STandards for SO_2 and NOx.

Combustion Power has acquired significant fluid bed R&D and operational experience since 1970. This experience, combined with the experience of others, convinced us that the best technical approach for a "California" fluid bed boiler was a hybrid combination which could meet the following design criteria:

- Meet or exceed the most stringent SO_2, NOx and CO emissions standards.
- Have reliable scale-up from pilot plant data.
- Be largely shop-fabricated for low cost field fabrication.

- Designed for ease of maintenance and minimal erosion potential.

Figure 1 is a chart that is often used to classify the three major classes of fluid bed technologies by degree of fluidization. The chart is a plot of superficial gas velocity versus bed expansion with a comparison made between mean gas velocity (gas passing unhindered through the solids bed) and mean solids velocity. As the differential (slip velocity) decreases between these two velocities, the degree of bed expansion (i.e., solids carryover) increases to the point where the solid particle passes unhindered through the process.

Over the range of this expansion there is Type A, classical bubbling bed; Types B and C, circulating fluid beds; and Type D, transport reactors. As depicted, there are four types of beds in these three categories, with two being circulating beds.

At low velocities the Type "A" bed has a defined fluidized-solids/freeboard interface that is readily visible. A very high proportion of product solids is discharged from the bed via a gravity overflow or underflow technique. This is the well known bubbling bed concept.

Higher gas velocity creates a higher rate of particle elutriation (carryover). Also the fluidized-solids/freeboard interface becomes less defined. This is a Type "B" bed, often called a turbulent layer bed which is the bed closest to a FI CIRC™ combustor. This bed typically elutriates over 50% of its incoming solids to the recirculation cyclone for enhanced particle reprocessing. In a fluid bed boiler this reprocessing means improved carbon burnout and lime sulfation.

The Type "C" circulating bed differs in that much higher gas velocities are used. Extensive product recycling creates a condition known as particle "clustering". These clusters have much higher transport velocities which explains the greater slip velocities at increasing solids throughput.

Both Type "B" and "C" beds are considered true circulating fluid beds due to their greater solids elutriation and subsequent extensive recycling via cyclones when compared to Type "A". It is important to remember, however, that both "B" and "C" still retain a dense lower bed relative to the lower density upper zone. In some respects this dense lower bed is similar to a bubbling bed but with much greater gas-solids heat and mass transfer.

A key difference between the FI CIRC™ design and Type "C" beds is the utilization of multiple small-diameter, high-efficiency cyclones to recycle a larger cut of fine particles contained in the elutriated solids. Type "C" beds typically utilize only one or two large diameter cyclones, which do not have as small a fine particle Dp_{50} as a smaller diameter cyclone. Typical Dp_{50} for these larger cyclones is over 75 microns, whereas the FI CIRC™ cyclone Dp_{50} is 15 microns.

Since a significant portion of unsulfated calcium oxide and carbon particles are found in the larger particle cuts, the FI CIRC™ cyclones will recycle the majority of those particles that need to be recycled. The net result is that FI CIRC™ attains the same or better degree of fuel carbon burnout (98 to 99.5+) and lower Ca/S ratios (1.6 to 1.8 at 90% SO_2 suppression) at much lower solids recirculation rates than the Type "C" bed.

The cyclones are critical in FI CIRC™, as with any circulating bed design, since they greatly affect carbon burnout and lime utilization. But with FI CIRC™ they serve the extra function of constantly "doping" the lower bed with a large portion of small particles. It is this "doping" that creates the fine bed particle size average of 300 to 600 microns. With this fine bed average particle size a significant portion of the bed is elutriated, even at velocities of 3-6 ft. per second.

This "doping" also creates a dense bed having a fluidized bulk density of 40 to 60 lbs/ft^3. It is well know that decreasing solids fluidized bulk density decreases the bed side heat transfer coefficient. Increased fluidized bulk density increases the heat transfer coefficient. At these high densities, the bed side heat transfer film coefficient is typically over 100 Btu/hr-ft^{2}F.

FI CIRC™, with its higher heat transfer rate and denser combustion bed using in-bed tube bundles, controls steaming rate by pneumatic removal of bed solids into an external bed material silo. By dropping bed level and exposing tube surface, the steaming rate drops. A one-inch drop in bed level has a noticeable effect in steaming rate.

Because small bed level changes result in noticeable steam rate change, it is possible to maintain boiler MCR (Manufacturer's Continuous Rating) SO_2, CO and HC emissions levels as low as 60% MCR. All of these pollutants require residence time (bed depth) in an oxidizing mode to be thoroughly abated. Since bed level does not greatly change, FI CIRC™ can therefore maintain excellent emissions levels at low MCR ratings.

In summary, the FI CIRC™ fine particle recycling technique gives this design the flexibility to:

- Obtain very turbulent dense bed heat and mass transfer equivalent to higher velocity circulating bed designs without the high velocities or increased solids recirculation ratio and consequent erosion potential:

- Capture the fine particles and introduce them into the lower dense bed, thereby enhancing carbon burnout and greatly improving sulfur capture while also reducing lime consumption.

- Introduce a very high and stable bed side heat transfer coefficient which allows for accurate steaming rate control over a wide MCR range without sacrificing fuel efficiency or emissions control. The design overall heat transfer coefficient (Uo) is greater than 65 BTU/hr-ft^2-F and is constant within 60-to-100% of MCR.

- Permit combustion of low ashing fuels such as petroleum coke and wood waste without the need for either supplemental or costly bed makeup.

- Create a fine particle bed which at lower velocities is much less likely to erode critical in-bed metal components and refractory.

Figure 2 is a cross-sectional view of a typical FI CIRC™ fluid bed boiler and depicts many of the key design features. These features are:

- A standardized modular design, whereby each design "module" consists of:

    - one cyclone and dipleg
    - five fuel guns
    - four in-bed tube bundles
    - 112 ft^2 of tuyere area.

  Each "module" is capable of generating 25 KPPH to 40 KPPH of steam, depending on fuel and emissions standards. Each boiler, regardless of size, utilizes these "building block" modules to reduce scale-up variances. The FI CIRC™ pilot plant has a 4 ft^2 bed area so the maximum process scale-up factor for any size boiler is 112 ft^2 ÷ 4 ft^2 or 28:1.

- Under bed above-stoichiometric air introduction to ensure uniform fluidization and complete combustion burnout in the lower zone where boiler water evaporation and sulfur removal control are critical. An oxidizing lower bed eliminates reducing gas initiated metal corrosion and ensures complete sulfur capture.

- Patented flat plate directional flow tuyeres which continuously move unfluidized particles to the gravity bed drain. This ensures optimal air introduction, fuel/air mixing and elimination of gas "jetting" with subsequent solids impingement on metal surfaces.

- Pneumatic fuel guns which positively and accurately introduce fine sized solids fuels under-bed, thereby enhancing fuel/air contact.

- Air swept spreaders or gravity drop pipes, which evenly distribute coarse particle fuels above-bed.

- Flanged, removable in-bed heat exchanger tube bundles which are located in the upper region of the dense bed to ensure uniform oxidation and minimize erosion. The upper bed placement also allows for quick and easy startup and fine steam rate control "on-the-run". They are also designed to be removed, turned over, and rotated, thereby extending wear life.

- A high freeboard region allows for coarse particle disengagement, thereby reducing refractory erosion in the cyclones. The freeboard also provides for final burnout and temperature stabilization which is required for good NOx removal without proprietary ammonia injection system.

- Multiple, small diameter, high efficiency cyclones which recycle fine particles (Dp_{50} = 15 microns) of unburned fuel and undersulfated lime particles into the lower dense bed via gravity diplegs.

- Forced circulation water flow through the tube bundles to ensure nucleate boiling in the evaporators. The net result is a low height profile, bottom supported fluid bed boiler.

FUEL PROPERTIES

Since 1977, Combustion Power has converted waste or low grade fuels to useable energy using fluidized bed combustors. Previous to then, the company was primarily doing research and development in the area of waste-to-energy conversion. Earliest work was with conversion of municipal solid waste processed to refuse derived fuel (RDF). A 100 TPD pilot plant was operated during the 1970's with continued work in a smaller unit through 1985. The pilot plant work consisted of pressurized fluidized bed combustion, atmospheric fluidized bed boiler development, and oxygen blown gasification. Typical RDF properties were as shown on Table I. Numerous other waste and low-grade fuels were burned in various Combustion Power Company pilot scale fluidized bed combustors, including high sodium North Dakota lignite, Texas lignite, pulp and paper industry sludges, wood wastes, waste oils, aircraft paint stripping wastes, etc.

These experiences with waste fuel combustion influenced design of the FI CIRC™ boiler. Low fluidizing velocities (high residence times) and high rates of solids recirculation are two examples of this. Currently operating or in construction boilers are being fired with or designed for, ponded coal fines, delayed petroleum coke, fluid petroleum coke, pat, mixed industrial wastes, log yard debris, pulp and paper sludge, municipal sludge, and rubber tires. Typical properties of some of these fuels are shown on Table II.

EMISSIONS CONTROL

Aside from particulates, there are four major gaseous species which are of most concern today: SO_2, NOx, HC and CO.

The FI CIRC™ system utilizes the principle of temperature, time and turbulence to effect extensive and cost efficient reduction of all these pollutants. SO_2 is controlled by the addition of calcium carbonate (limestone), bed temperature control, excess air, and dense bed height. CO and HC are controlled by operating temperature, excess air, and dense bed height. NOx is controlled by the addition of ammonia with gas temperature control.

All of these pollutants have temperature control in common and it is within the FI CIRC™ combustion region that the optimum range of temperatures is found. The optimum temperature range to remove each pollutant is as follows:

NOx with NH_3 reductant:	1500 - 1750°F
SO_2 with limestone:	1500 - 1750°F
CO with excess air:	above 1600°F
HC with excess air:	above 1550°F

Therefore the best temperature operating range is 1500 to 1600°F in the combustor bed and 1550 to 1650°F in the freeboard. The uniform temperatures long residence times at the proper temperatures and good solids-gas mixing are the key reasons why fluid bed combustors outperform older solid fuel combustion techniques, such as stokers and pulverized coal systems.

SO_2 is controlled by the chemical absorption of SO_2 with active calcium oxide primarily in the fluidized dense combustion bed. Some removal also occurs in the freeboard region. There are several simultaneous chemical reactions which occur to effect SO_2 removal.

CxHyOzNiSj + O_2	--->	$CO_2 + H_2O + N_2$
Fuel Combustion	--->	CO + NOx + SO_2/SO_3 + HC + ash and char

The SO_2 portion is absorbed by calcium oxide after the calcium oxide is first formed by calcination of limestone:

Ca CO_3	--->	CaO + CO_2
CaO + SO_2 + 1/2 O_2	--->	$CaSO_4$

The conversion to $CaSO_4$ does not occur unless oxygen is present. Also it is critical that the dense bed temperature remain below 1700°F to minimize calcium consumption. With FI CIRC™, the important Ca/S ratio is optimized via several techniques.

- Dense bed temperature control below 1700°F using in-bed heat exchange tube bundles.
- A dense bed high enough and operating at a low enough gas velocity to ensure excellent gas/solids contact time.
- Excess oxygen in the combustion zone to "drive" the sulfation reaction.
- Extensive fine particle recycling to ensure that elutriated and unreacted CaO is given another opportunity to sulfate.

The above techniques yield not only very low SO_2 emissions levels, as high as 98% suppression, but do so at very economical Ca/S ratios. Figure 3 depicts the relationship in a FI CIRC system between Ca/S ratio and SO_2 suppression.

In fluidized bed combustors operating at temperatures of 1700F or below, NOx is generated by the combustion of fuel-bound nitrogen. This NOx is destroyed by ammonia injected into the freeboard region at the proper residence time and temperature according to the following reactions:

$$4\ NH_3 + 4\ NO + O_2 \quad \text{--->} \quad 4\ N_2 + 6\ H_2O$$
$$4\ NH_3 + 2\ NO + 2\ O_2 \quad \text{--->} \quad 3\ N_2 + 6\ H_2O$$

Excessive ammonia introduction will create "ammonia slip" which is free ammonia leaving the stack as well as increasing CO production. This later relationship is as depicted on Figure 4. NOx control, as practiced by Combustion Power, is proprietary. This is due to the careful control required to ensure that NOx is indeed destroyed without undesired side effects, such as CO production and ammonia slip.

Both CO and hydrocarbon emissions are controlled by temperature and excess air:

$$HC + O_2 \quad \text{--->} \quad CO_2 + H_2O$$
$$CO + 1/2\ O_2 \quad \text{--->} \quad CO_2$$

As with SO_2 control, these pollutants are controlled in the FI CIRC™ boilers by:

- Dense bed temperature control using in-bed heat exchange tube bundles.
- A dense bed high enough and operating at a low enough gas velocity to enhance gas/ gas reaction times.
- Excess oxygen in the combustion zone to essentially complete the oxidation reactions.
- Careful ammonia injection control to minimize CO production.

This extensive emissions control technology has resulted in California FI CIRC™ plants being permitted at very low SO_2, NOx and CO levels. Of the 33 fluid bed boiler plants in California, 1/3 are FI CIRC™ systems.

Table III shows permitted and actual emissions from six petroleum coke fired plants in southern and northern California.

REFERENCES

1. G. Goldbach, et.al, Program to Develop MSW Fired Fluidized Bed Boiler, CPC Final Report, U.S. DOE sponsored project under Contract No. AC02-80CS24321, June 1983.

2. Ebbe Skov, et.al, Low Emissions Profile from Efficient Fluidized Bed Power Plants Using Petroleum Coke, Paper presented at ASME Industrial Power Conference, St. Louis, MO, October 1990.

3. Michael A. O'Hagan, The California Applications of the Fines Circulating Fluidized Bed Technology, Paper presented at ASME Industrial Power Conference, Hartford, CT, October 1989.

TABLE I: REFUSED DERIVED FUEL CHARACTERISTICS (composition on a dry basis*)

	Proximate Analysis (ASTM D 3172) wt % (mean)	(σ)		Ultimate Analysis (ASTM D 3176) wt % (mean)	(σ)
Ash	18.4%	(2.54%)	Ash	18.36%	(2.54%)
Volatile	71.5%	(2.27%)	C	41.66	(1.25%)
Fixed Carbon	10.1%	(2.86%)	H	5.41	(0.23%)
			N	.74	(0.17%)
			Cl	.37	(0.10%)
Heating Value	7272 Btu/lb	(283 Btu/lb)	S	.27	(0.09%)
			O	33.20	(2.12%)

*Based on 11 samples taken over 6 mo. period with average moisture content of 26%

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TABLE II: TYPICAL WASTE FUEL PROPERTIES

Property	Ponded Coal Fines	Delayed Pet Coke	Fluid Pet Coke	Peat	Mixed Ind. Wastes	Wood Wastes	Municipal Sludge	Rubber Tires
Moisture, wt%	15.0	3.1	0.2-0.8	37.6	35	50	70	0.5
Volatiles, wt%	NA*	9.0	4.7-6.4	35.9	35	45	NA	62.3
Ash & Inerts, wt%	25.5	0.2	0.2-0.3	1.9	8	1.5	NA	5.7
Ultimate Anal, wt%								
C	45.2	86.0	90-92	NA	NA	25.5	58.3	83.2
H	3.3	3.6	1.8-2.2	NA	NA	2.9	8.5	7.1
O	9.7	0.5	0.4-1.2	NA	NA	19.8	22.9	2.5
N	0.6	1.4	2.3-3.0	NA	2	0.1	8.8	0.3
S	0.6	5.2	2.7-3.5	0.3	NA	0.2	1.6	1.2
Cl	0.1	0.04	0.4-1.2	NA	NA	NA	NA	NA
Higher Heating Value, Btu/lb	8,150	14,770	14,550	6,481	6,423	4,500	3,600	16,329

*NA = Not Available

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TABLE III: FI CIRC™ EMISSIONS BURNING PETROLEUM

Plant	Fuel	SO_2[1]	NOx[1]	CO[1]	Total Part[2]
Torrance, CA	Delayed pet coke	21/13	19/8	190/13	7.8/1.7
Contra Costa County, CA (5 plants)	Fluid pet coke	54/20	50/25	117/25	3.0/1.8

(1) Values in table are given as Permit/Actual in ppm, dry, standard conditions, corrected to 3% O_2. Data taken from emissions compliance test conducted to obtain operating permit.

(2) Total particulate values are given as Permit/Actual in lb/hr.

TYPES OF FLUIDIZED BEDS

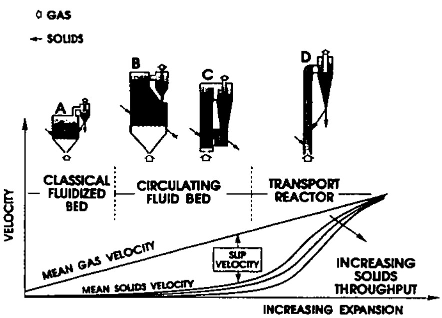

Figure 1

FICIRC™ FLUIDIZED BED BOILER SEGMENT

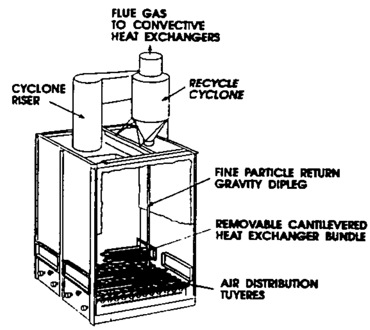

Figure 2

SO_2 CONTROL BY CALCIUM ADDITION

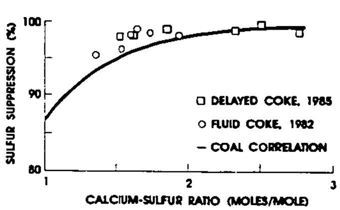

Figure 3

TYPICAL CARBON MONOXIDE PRODUCTION UPON AMMONIA INJECTION

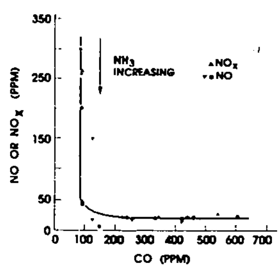

Figure 4