FLUIDIZED BED GASIFICATION OF SLUDGE DERIVED FUEL

ROGER T. HAUG
Torrance, California

F. MICHAEL LEWIS
Mountain View, California

STEVEN I. CHOTI
Pasadena, California

HARRY M. SIZEMORE
Los Angeles, California

ABSTRACT

Municipal sewage sludge can be converted to a sludge derived fuel by energy efficient removal of water prior to combustion. The City of Los Angeles is currently constructing a sludge processing system which uses the Carver-Greenfield multiple effect evaporation process to convert wet sludge cake to a “sludge derived fuel.” Energy will be recovered and air emissions controlled by fluidized bed gasification and staged combustion of the resulting fuel gas. Design criteria for the combustion process were developed from a pilot fluidized bed test program. Excess energy will be produced beyond the needs for wastewater treatment and air emissions will be reduced from present levels.

BACKGROUND

Sludge which has been dried to reduce its moisture content can serve as a useful fuel. Dried sludge is similar in fuel value to lower grade coals, refuse, and other biomass materials currently considered as alternative energy resources. Just as refuse can be processed into refuse derived fuel (RDF), sludge can be processed to the form of a sludge derived fuel (SDF) by removal of water prior to combustion.

Energy recovery from sludge is primarily a problem of moisture. Mechanically dewatered sludges usually contain too much water to support autogenous combustion. Sludge furnaces commonly support combustion by use of supplemental fossil fuels. However, the high cost of such fuels has made this practice economically undesirable in new facilities. A number of approaches are available to either reduce or eliminate the need for fossil fuels, including: (a) lowering excess air and exhaust gas temperatures; (b) use of alternative fuels to supplement combustion such as refuse, coal, or other biomass; and (c) energy efficient removal of water prior to combustion. The City of Los Angeles (CLA) is presently implementing a sludge management program which includes processing of sludge to remove water by multiple-effect evaporation prior to combustion and producing SDF with subsequent energy recovery from the fuel. The purpose of this paper is to present results of pilot SDF combustion tests, discuss scale-up and design constraints on the prototype combustor, and describe integration of the design into the wastewater treatment plant.

THE HERS PROJECT

Combustion of SDF is part of an overall program to recover energy and manage sludges produced at the City’s Hyperion Treatment Plant. A process diagram for the Hyperion Energy Recovery System (HERS) is presented in Fig. 1. About 405 dry tons per day (dtpd) of primary and waste activated sludges will be produced in year 2000. Both will be anaerobically digested and the digester gas used as fuel in a gas turbine, combined-cycle power system. Digested sludges (about 265 dtpd) will be centrifugally dewatered and then converted to SDF using the Carver-Greenfield (CG) multiple-effect evaporation process. The CG process was developed by Dehydrotech Corp., East Hanover, New Jersey, for energy efficient drying of solid suspensions and has been applied to industrial and municipal wastewater sludges. A 4-effect evaporator will be used with a thermal efficiency of about 400 Btu/lb of water removed. SDF produced by CG drying of 20 percent
FIG. 1 PROCESS DIAGRAM OF THE HYPERION ENERGY RECOVERY SYSTEM (HERS) FOR MANAGEMENT OF SEWAGE SLUDGES WITH ENERGY RECOVERY FROM THE SLUDGE ORGANICS
sludge cake will have considerably more fuel value than the energy required to operate the CG process.

The CLA investigated a number of alternative combustion systems for SDF. The fluidized bed was selected because of its (a) prior successful application in municipal sludge incineration, (b) adaptability to a wide range of fuel characteristics, (c) potential for in-bed and staged combustion NOx control, (d) ease of SDF and sludge oil fuel feed, (e) flexibility and overall ease of operation, and (f) competitive economics with other combustion alternatives.

A fluidized bed test program was sponsored by the CLA to address the problems of combusting SDF in a fluidized bed and controlling air pollutant emissions. Based on results of the test program, SDF will be gasified in a conventional, bubbling fluidized bed operated with sub-stoichiometric air addition. The resulting fuel gas will be burned in a staged combustion afterburner designed to minimize NOx, CO, and hydrocarbon emissions. Energy will be recovered in a waste heat boiler. A baghouse will be used for solid particulate control and wet scrubbing for control of SOx and condensible particulates.

Design of the HERS Project was undertaken by CLA forces and the joint venture of the Ralph M. Parsons Co. and James M. Montgomery Engineers (M-P). Foster-Wheeler Energy Corporation was a major subcontractor to M-P for design of the CG process. General design criteria for the combustion system were developed by the CLA based on the combustion test program described in this paper. The detailed design was then prepared by M-P with valuable assistance from a team of combustion and air pollution experts assembled by the City. The joint venture of Schneider Inc. and Combustion Power Co. was the successful bidder for the combustion process and is responsible for engineering review, fabrication, installation, commissioning and startup of the system. The latter is scheduled to begin in July 1985.

**FLUIDIZED BED TEST PROGRAM**

Characteristics of SDF produced by HERS are presented in Table 1. The adiabatic combustion temperature (0 percent excess air) is near 3300°F (1820°C) because of the low moisture content of the fuel [2]. Temperature control is a primary concern to avoid ash softening which begins at about 2100°F (1150°C). Fuel bound sulfur is typical of low to medium sulfur coals. However, fuel bound nitrogen at 4.5 percent (7.3 percent on a moisture-free basis) is significantly higher than that of most solid fuels. SDF produced by the CG process is powdery in form but can be pelletized. Powder, granular (1 mm nominal size, and pelleted (1 cm nominal size) forms of SDF were used in the test program.

Based on projected fuel characteristics and air emission constraints discussed later, control of reactor temperatures and control of NOx emissions were major concerns in use of SDF as an energy resource. Objectives of the combustion test program were to develop detailed criteria for design and air emission factors for subsequent air permit applications.

It was necessary to constrain the test program to a manageable number of design and operational variables to be investigated. Superficial fluidizing air velocities were limited to the range of 2 to 6 fps (0.6 to 1.8 mps) and bed depths to 3 to 6 ft (0.9 to 1.8 m), both typical of conventional, atmospheric pressure, bubbling bed designs. Almost all fluidized bed systems used for combustion of wet sludge cake operate within this range of parameters. It was considered prudent to use proven design parameters to the extent possible. Pressurized, elutriating, circulating, and dual media fluidized bed systems were not considered in the test program due to their lack of any prior application with sludge related fuels, and limited process advantages in the HERS application compared to a bubbling bed design.

**TEST FACILITY**

The test program was conducted in a pilot fluidized bed system owned and operated by Dorr-Oliver Inc. The test facility was suitably sized for the available sludge supply and incorporated all equipment necessary to incinerate or gasify the SDF and combust the resulting fuel gas. A schematic diagram of the test facility is presented in Fig. 2. Two stainless steel reactors were used. The first-stage consisted of a 12 in. ID (0.30 m) tube 8 ft long (2.4 m), a 1 ft (0.30 m) long transition section, and a 15-in. ID (0.38 m) tube 7 ft (2.1 m) long. The second-stage was a 17.25-in. ID (0.44 m) tube, 11.5 ft (3.5 m) long. Both reactors were mounted vertically inside insulating firebrick chimneys. The annular space between the reactor and firebrick was heated by four natural gas burners. The external burners brought the reactor tubes to operating temperature, maintained desired reactor temperatures during the course of experimentation, and offset radiation losses.

The first-stage reactor was used for both fluidized bed incineration and gasification. The second-stage reactor was used to combust fuel gas produced when the fluidized bed was operated in the gasification mode. The second-stage reactor could be configured as either a second-stage fluidized bed (FBAB) or as a staged afterburner (SAB). In the SAB mode, combustion air could be introduced through two lower air ports, which were fixed in-place at the bottom of the SAB, and two upper air ports. The latter could be raised or lowered to increase or decrease the
TABLE 1 EXPECTED SDF CHARACTERISTICS FOR THE HERS PROJECT

<table>
<thead>
<tr>
<th></th>
<th>Heat Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Heat Value</td>
<td></td>
</tr>
<tr>
<td>Sludge</td>
<td>5340 Btu HHV/lb dry (2960 cal/gm)</td>
</tr>
<tr>
<td>Sludge Oil a.</td>
<td>18700</td>
</tr>
<tr>
<td>Combined</td>
<td>6410</td>
</tr>
<tr>
<td>Moisture Content</td>
<td>1% of total weight</td>
</tr>
<tr>
<td>Ash Content</td>
<td>38% of dry weight</td>
</tr>
<tr>
<td>Fixed Carbon</td>
<td>5 to 8% of dry weight</td>
</tr>
<tr>
<td>Fuel Bound Nitrogen</td>
<td>4.5% of dry weight</td>
</tr>
<tr>
<td>Fuel Bound Sulfur</td>
<td>1.3% of dry weight</td>
</tr>
<tr>
<td>Ash Fusibility</td>
<td>2100-2200 °F initial deformation (1150-1200 °C)</td>
</tr>
<tr>
<td></td>
<td>2300-2450 °F fluid temperature (1260-1340 °C)</td>
</tr>
<tr>
<td>Particle Size</td>
<td>Powder (55% by weight less than 200 microns)</td>
</tr>
</tbody>
</table>

a. Fats and oils in the sludge are extracted in the CG process and recovered as a separate process stream. CLA sludge contains about 8 to 12% oil on a dry weight basis.

gas residence time between air inputs. Exhaust flue gas was ducted out of the second-stage reactor, through a single tray impingement scrubber, and exhausted out of the building.

COMBUSTION MODES

The test program was broadly divided between two different and distinct combustion modes. In the incineration mode, all fuel and both stoichiometric and excess air were introduced into the first-stage fluidized bed. Oxidation of the fuel occurred primarily in the fluidized bed with some secondary burning in the overbed zone. In the gasification mode, primary air was introduced to the fluidized bed in sub-stoichiometric amounts. SDF was partially oxidized and gasified to a low Btu gas, termed fuel gas. In this mode, the reactor was termed a fluidized bed gasifier (FBG).

Four modes of combusting fuel gas from the FBG were tested as follows:

- Overbed Combustion (OB mode) — Secondary combustion air was introduced to the FBG freeboard. All exhaust gases were ducted from the first-stage reactor to the scrubber.
- Fluidized Bed Afterburner (FBAB mode) — Fuel gas was ducted from the FBG to the second-stage reactor which was operated as a fluidized bed. Flue gas from the FBAB was ducted to the scrubber.
FIG. 2 SCHEMATIC DIAGRAM OF THE PILOT FLUIDIZED BED AND STAGED COMBUSTION AFTERBURNER USED IN THE COMBUSTION TEST PROGRAM
TEST RESULTS

A complete discussion of all test results is available in Ref [1]. Results of the test program and conclusions and recommendations which guided the conceptual design are summarized here.

Nitrous Oxides (NOₓ)

NOₓ emissions were markedly affected by the method of combustion. Sub-stoichiometric gasification in the fluidized bed with staged combustion of the fuel gas (SAB and SAB/OB modes) significantly reduced NOₓ emissions compared to other combustion modes examined. Principal variables affecting NOₓ production were the method of secondary air staging, the residence time between initial and final air input to the afterburner, exhaust gas oxygen content, and afterburner temperature. NOₓ emissions were not significantly affected by the superficial air velocity in the range from 2 to 6 fps (0.6 to 1.8 mps), degree of sub-stoichiometric conditions in the bed over the range of 40 to 80 percent of the theoretical oxygen for combustion (TOC), fuel bound nitrogen between 3 to 6 percent by dry weight, or fuel particle size.

In the incineration mode it was concluded that no combination of feed position, reasonable excess air rate, or practical bed temperature was likely to reduce NOₓ emissions below 600-1300 ppm adjusted to 5 percent O₂. The overbed (OB) and second-stage fluidized bed (FBAB) both provided one point of supplemental air staging and NOₓ emissions were reduced to the range of 250 to 550 ppm (5 percent O₂). Additional staging of secondary air in the staged afterburner provided significant further reductions and, therefore, the remaining discussion will focus on this mode of combustion.

The effect of flue gas O₂ content on NOₓ emissions in the SAB mode is presented in Fig. 3. NOₓ concentrations increased sharply if exhaust gas O₂ content increased above 7 to 8 percent. Below 6 to 7 percent, NOₓ concentrations were relatively unaffected by O₂ content. This relationship was qualitatively observed in several other test sequences. CO concentrations were observed to increase if the O₂ concentration decreased below about 2 to 3 percent. Therefore, an O₂ content between 4 to 6 percent was used in most tests to reduce CO emissions but still remain on the plateau of low NOₓ emissions shown in Fig. 3.

The effect of gas residence time between initial and final air input to the SAB on outlet NOₓ concentrations is presented in Fig. 4. If all SAB air ports are at the same elevation, the residence time between air ports is zero and only a single staging occurs in the afterburner. NOₓ emissions varied between 370 and 540 ppm (5 percent O₂) under these conditions. This range is consistent with the general range of NOₓ emissions observed in the OB and FBAB combustion modes which also used only a single staging of secondary air. The upper air ports in the SAB were raised or lowered to increase or decrease the gas residence time between combustion air inputs. Additional staging of air to the SAB (increased gas residence time between air addition ports) resulted in a significant further decrease in NOₓ concentration. Concentrations below 100 ppm (5 percent O₂) were achieved with a 2.0 to 2.5 sec residence time between initial and final air input. Data shown in Fig. 4 include tests in which the SAB air was split evenly between upper and lower ports, both upper air ports were at the same elevation, afterburner temperature was constant, and no overbed air was used.

As shown in Fig. 4, residence times on the order of seconds are required to complete the gas phase reactions responsible for reducing NO back to N₂. Discussion of the theoretical mechanisms of NOₓ formation and reduction is beyond the scope of this paper. However, the general range of residence times is consistent with the laboratory and theoretical work of Beer, et al. [3, 4, 5] on free-board destruction of NOₓ by reduction with char and CO. Thus, there is theoretical support for the range of residence times observed in this study.

Tests were conducted in which the air distribution was varied between lower and upper afterburner ports. It was observed that minimum NOₓ concentrations occurred when the fuel gas was brought to a near stoichiometric condition in the first part of the afterburner as shown in Fig. 5. This suggests that air input to the SAB should be such that a stoichiometric condition is achieved at one point in the afterburner.

The effect of afterburner temperature on NOₓ concentration was determined by maintaining the FBG temperature at 1750°F (954°C) and bringing the SAB to almost 1900°F (1040°C). SAB temperature was then decreased gradually to about 1425°F (774°C) over a 90 minute period by decreasing the external burner heat.
FIG. 3 EFFECT OF $O_2$ CONTENT ON NO$_x$ CONCENTRATIONS IN THE SAB COMBUSTION MODE
FIG. 4 EFFECT OF RESIDENCE TIME BETWEEN INITIAL (LOWER AIR PORTS) AND FINAL (UPPER AIR PORTS) AIR ADDITION ON OUTLET NOx CONCENTRATION FROM STAGED AFTERBURNER (SAB MODE)
rate. NOx concentrations decreased slightly with decreasing afterburner temperature. NOx averaged about 150 ppm above 1850°F (1010°C) and from 100-120 ppm from 1400 to 1700°F (760-930°C). No evidence of a temperature window in which thermal de-NOx reactions occur was evident in these results. Rather, NOx increased slightly with temperature over the range from 1400 to 1900°F (760-1040°C).

Additional air staging by use of partial overbed air addition (SAB/OB mode) was observed to further reduce the NOx concentration compared to use of the staged afterburner alone. This is illustrated by the data on Fig. 6. NOx concentrations ranged from 105-160 ppm with an average of about 130 ppm with the system in the SAB mode. The air pattern was then changed to the SAB/OB mode. This change required only about 30 seconds to accomplish in the pilot system. NOx concentrations responded almost instantaneously, decreasing to about 65 ppm. NOx ranged from about 40 to 85 ppm (average 65 ppm) over the remainder of the test. The effectiveness of additional air staging to further reduce NOx emissions was verified on four other occasions using several SDF fuel types. The effect was independent of fuel type.

For the data shown in Fig. 6, about 62 percent of TOC was supplied to the fluid bed, 21 percent to the overbed, 26 percent to the lower SAB air ports, and 26 percent to the upper ports. About 109 percent of TOC was supplied up to and including the first stage SAB air. Gas residence time between lower and upper air ports of the afterburner was 1.6 sec. The latter two conditions were less than optimum based on the previous test work, but NOx concentrations nevertheless decreased with the additional air staging.

Based on these results, fluidized bed gasification and staged combustion of the fuel gas were recommended for the full-scale design because of the low NOx emissions achieved. Multiple points for secondary air injection were recommended in the afterburner, each with a variable and controllable air supply. With these design provisions, air staging to the SAB could be optimized after system startup to minimize NOx production. The system for introduction of secondary air should be designed to assure rapid and uniform mixing of fuel gas and air and to minimize local zones of oxidative conditions. Air addition to the afterburner should bring the gas to a near stoichiometric condition at one point. Exhaust gas O2 content from the afterburner should be maintained between 3 and 6.5 percent to assure low concentrations of both NOx and CO. Detention time between the point of initial and final air input in the SAB should be 2.0 to 2.5 sec to minimize NOx production.

Sulfur Oxides (SOx)

A large fraction of the fuel bound sulfur was converted to SOx regardless of the combustion mode. The data suggest that uncontrolled emissions would be essentially the same regardless of the combustion mode and that an assumption of complete conversion of fuel sulfur should be the basis for design of a flue gas desulfurization system.

Limestone was added to the fluidized bed to determine the potential for in-bed sulfur retention. 80-90 percent sulfur retention was achieved in the incineration mode with a Ca/S molar ratio of 4. However, sulfur retention decreased dramatically in the gasification mode to about 50 percent retention at a Ca/S of 4.

It was concluded that SOx retention by limestone addition is enhanced by maintaining oxidative conditions in the fluidized bed. Conversely, NOx control is favored by reducing conditions in the bed and staged combustion of the resulting fuel gas. Therefore, control of NOx by gasification and staged combustion and control of SOx by limestone addition are not entirely compatible control strategies. Flue gas treatment for SOx control was recommended because control of NOx emissions is more constraining to the HERS design, and flue gas desulfurization processes are well proven compared to flue gas denitrification.

Carbon Monoxide

The effect of temperature on outlet CO concentration was determined for both the FBAB and SAB modes. In each case, the afterburner was stabilized at normal operating temperature and with sufficient residual O2 and residence time to give initial CO concentrations below 10 ppm. The afterburner temperature was then decreased gradually to about 1400°F (760°C) over a period of about 90 min. Results are presented in Fig. 7. In each case CO emissions began to increase below a temperature of about 1600°F (870°C) with the increase becoming significant below 1500°F (815°C).

It was concluded that CO emissions can be reduced to levels below 10 to 20 ppm if the following conditions are maintained from the point of final air input to the afterburner outlet: (a) a temperature of 1700°F (930°C) or above, (b) residual O2 content above 3 percent, (c) a gas residence time of 2 sec, and (d) a high level of gas phase turbulence. Equally low CO emissions may be achievable with less severe conditions than those described above. However, the stated conditions were verified by the test program and have a high confidence level of achieving the desired emissions.
FIG. 5 EFFECT OF BED AIR AND 1ST STAGE SAB AIR STOICHIOMETRY ON NO\textsubscript{X} CONCENTRATION. FRACTION OF THEORETICAL OXYGEN REQUIREMENTS IS THAT ADDED BY BED AIR PLUS LOWER AIR PORTS OF AFTERBURNER. BED AIR WAS 59.6-64.2 PERCENT OF TOC AND AVERAGE EXHAUST OXYGEN 5.0-6.2 PERCENT FOR ALL TESTS.
Hydrocarbons

When combustion conditions were maintained to achieve CO concentrations of 20 ppm or less, total hydrocarbons (THC) were also quite low. Among 85 samples collected under these conditions, THC values ranged from 11.9 ppm as CH₄ to 32.0 ppm, with an average 23.6 ppm. An average 17 percent of the total measured hydrocarbons were non-methane hydrocarbons (NMHC). NMHC averaged 4.0 ppm with a range from 2.0 to 5.4 ppm as CH₄.

An attempt was made to correlate hydrocarbon emissions with SAB temperature and CO concentrations. SAB temperatures were raised to 1900°F (1040°C) and then decreased gradually to about 1425°F (774°C) over a 90 min period. CO emissions increased from less than 10 ppm to a range of 70-100 ppm at the lower temperature. Interestingly, there was no apparent increase in THC over the temperature range examined even though CO emissions did increase. On two other occasions, THC samples were drawn when CO emissions were over 500 ppm due to unstable operating conditions. THC concentrations of 400...
FBAB MODE. 2 SEC EMPTY BED CONTACT TIME IN FBAB AND ABOUT 2.8 SEC CONTACT TIME IN FREEBOARD OF FBAB. 5.8% OXYGEN

SAB MODE. 2.9 SEC DETENTION TIME AFTER LAST AIR INPUT, 5.8% OXYGEN

FIG. 7 EFFECT OF COMBUSTION TEMPERATURE ON CARBON MONOXIDE CONTENT OF EXHAUST GAS
and 1800 ppm were then recorded. Thus, it is likely that THC emissions eventually increase with increasing CO, but perhaps not until CO emissions exceed a threshold somewhere above 100 ppm. No correlation between CO and THC was noted below 100 ppm CO in these tests.

It was concluded that conditions defined above for CO burnout would also assure control of hydrocarbon emissions.

Particulates

Mass balances verified that the majority of fuel ash was exhausted from the bed as fly ash regardless of the fuel particle size. Some buildup of ash in the bed was observed with granular and pelletized fuel forms, but not with the powder form. Uncontrolled particulate loadings can be estimated on the basis of complete ash carryout. Therefore, attention was focused on the concentration of condensible particulates and the removal efficiency provided by wet scrubbing.

With stable combustion conditions, inorganic condensibles averaged 0.011 gr/dscf (25 mg/Nm³) in the afterburner exhaust and organic condensibles 0.0086 gr/dscf (19.7 mg/Nm³). Wet scrubbing was effective in removing both inorganic and organic condensibles. Inorganic condensibles in the scrubber exhaust averaged 0.0034 gr/dscf (7.8 mg/Nm³), an 84 percent reduction from the median of all values ahead of the scrubber. Organic condensibles were reduced to an average 0.0043 gr/dscf (9.8 mg/Nm³), about a 56 percent reduction from the average of all upstream values. Scrubber outlet temperatures ranged from 102 to 119°F (39 to 48°C) during these tests. It was concluded that wet scrubbing with reduction of flue gas temperature to below 120°F (49°C) can significantly reduce the concentration of both organic and inorganic condensibles.

Particle Size Distributions

Particle size distributions of fly ash were remarkably similar regardless of whether the SDF was in powder, granular, or pellet form. Medium particle diameter was between 12 and 20 microns. The sub-micron size fraction ranged from about 0.5 to 2.0 percent of the total fly ash.

Based on the similarity of ash particle size distributions, it was concluded that selection and operation of a particulate control device would not be significantly affected by the particle size distribution of the starting fuel. Approximately the same particle size distribution will be “seen” by the particulate control device whether a powder or pellet form of the fuel is fed to the fluidized bed.

Fly Ash Characteristics

Ash samples collected with the fabric filter were analyzed for ash softening temperatures in accordance with ASTM procedures. Initial deformation begins at about 2200°F (1200°C) under oxidative conditions and 2100°F (1150°C) under reducing conditions. Ash will be exposed to both conditions if the fluidized bed is operated in a gasification mode. It was concluded that combustion temperatures should be maintained below about 1900°F (1040°C) to assure minimal problems from ash softening.

Silicates account for about 40 percent of the total ash weight, reflecting the typical presence of fine sand and grit in the SDF. Together, oxides of aluminum, calcium, and iron represent a fraction about equal to that of the silicates.

Fuel Gas

H₂, CO, CH₄, and condensible tars and oils were the principle fuel components of the fuel gas and provided about 42, 29, 17, and 12 percent, respectively, of the total HHV. With an average 66 percent of the stoichiometric oxygen supply to the bed, HHV of the gas was determined to be about 84 Btu/wscf (750 Kcal/m³).

Combustion Efficiency

Combustion efficiency is defined as heat released in the combustion process divided by the heat value of the fuel. Non-combusted or partially combusted materials, such as CO, THC, and fixed carbon in the fly ash, represent a loss of fuel value and reduce combustion efficiency. CO and THC, with concentrations less than 20 and 30 ppm, respectively, accounted for less than 0.05 percent of the fuel value of the feed SDF. Fixed carbon in the fly ash was generally less than 1 percent by weight, giving a total combustion efficiency of about 99 percent.

Fixed carbon burnout is defined as fixed carbon oxidized in the combustion process divided by fixed carbon contained in the fuel. Fixed carbon burnout ranged from about 93 to 95 percent. No significant differences in the efficiencies of combustion or fixed carbon burnout were noted for the various SDF fuel types tested. It was concluded that recycle of ash material would not be necessary because of the high efficiency of combustion and fixed carbon burnout achieved with single pass operation in the bubbling bed.

FULL SCALE GASIFIER DESIGN

Conversion of the test program recommendations into a workable design was a demanding process requiring a
number of key decisions. Development and evolution of the full-scale system will be described in the following section. A profile of one train of the final combustion system is shown in Fig. 8. The system is designed with three complete and independent trains, each capable of processing 50 percent of the year 2000 design loading of 265 dry tons per day of SDF.

AIR EMISSION CONSTRAINTS

The HERS Project must comply with all applicable air pollution regulations. The Hyperion Treatment Plant is situated in a critical air basin and compliance with New Source Review (NSR) and Prevention of Significant Deterioration (PSD) provisions of the Clean Air Act are the most constraining regulatory requirements in terms of design. In addition, the environmental impact documents established a goal that air emissions from the treatment plant should not be increased as a result of the project.

Present plant emissions are summarized in Table 2. Present emissions result primarily from use of existing digester gas in reciprocating diesel engines. As part of the HERS project these engines will be retired from service and replaced by a gas turbine, combined cycle system. Projected emissions from the latter facility under year 2000 conditions are also presented in Table 2. The difference in emissions between present plant sources and the combined cycle system are available to implement the fluidized bed combustion system.

Based on results of the combustion program, NO\textsubscript{X}, SO\textsubscript{X}, and particulates were the most constraining of the pollutants. Of these, particulates require flue gas treatment for control. Considering NO\textsubscript{X} and SO\textsubscript{X}, the combustion program clearly showed that reducing conditions in the fluidized bed were not compatible with limestone addition for SO\textsubscript{X} control. Furthermore, the exhaust SO\textsubscript{X} concentration necessary to stay within the available offsets would require a very high Ca/S ratio. On the other hand, available flue gas denitrification processes are very expensive compared to use of combustion controls. Therefore, NO\textsubscript{X} became the governing pollutant and will be controlled by use of the combustion parameters outlined previously. This decision also meant that SO\textsubscript{X} will be controlled by flue gas desulfurization.

FUEL PREPARATION

It was initially recommended that the combustion system be designed to feed SDF in both the powder and pelleted form. SDF produced by the CG drying process is powdery in nature and pelletizing equipment is necessary to produce the pelleted form. Pelleted fuel could be fed by one or more overbed “flingers” or spreader stokers. SDF in the powder form must be fed through a number of underbed or inbed ports.

Both alternative feeding systems have been used in practice with varying degrees of success. Overbed feeding itself is easier from a design and operating standpoint, but involves significant capital and O&M costs for pelleting the fuel. Pneumatic conveyance through underbed or radial in-bed feed pipes appeared to be a practical and reasonably proven design. The decision was made to accept the added difficulty of feeding fuel in the powder form to avoid what were judged to be great difficulties and costs for pelleting.

TEMPERATURE CONTROL

Temperature profiles in the combustion test program were maintained at about 1750°F (954°C) in the bed, overbed, and afterburner zones. NO\textsubscript{X} concentrations were only mildly influenced by temperatures below 1900°F (1040°C). Nevertheless, it was considered prudent to limit combustion temperatures to a maximum of 1900°F (1040°C) to maintain a close simulation with the pilot studies, to limit any unnecessary increase in NO\textsubscript{X}, and to minimize problems of ash slagging.

Five methods are available for control of combustion temperature: (a) dilution with excess air; (b) heat extraction by means of heat transfer surfaces; (c) addition of a cool diluent gas such as steam or recycled flue gas; (d) use of water sprays to absorb heat primarily by the heat of vaporization; and (e) control of oxygen supply in the sub-stoichiometric region (starved air combustion). Dilution with excess air was not a viable option for HERS because of the use of staged combustion and because NO\textsubscript{X} emissions had been shown to increase dramatically with O\textsubscript{2} concentrations above about 8 percent. Use of water sprays is generally limited to emergency quenching, trimming where only small temperature changes are required, or complete quenching of gas temperature. In addition, the heat of vaporization cannot be recovered in a waste heat boiler which reduces the potential for waste heat recovery. Starved air combustion for temperature control is viable for the initial gasification step, but not for subsequent combustion of the fuel gas. Therefore, temperature control options were reduced to use of heat transfer surfaces or addition of a diluent gas. Relative advantages/disadvantages of the alternative control strategies are summarized in Table 3.

Heat Transfer Surfaces

Use of heat transfer surface has been demonstrated in several fluidized bed boiler facilities. For the HERS system, heat transfer surface would be required in the three
FIG. 8 HERS SYSTEM FOR FLUIDIZED BED GASIFICATION OF SDF, STAGED COMBUSTION AFTERTURNING, WASTE HEAT RECOVERY, AND AIR POLLUTION CONTROL
TABLE 2 AIR EMISSION BALANCE BEFORE AND AFTER HERS
(Pounds Per Day)

<table>
<thead>
<tr>
<th></th>
<th>NO\textsubscript{x} (As NO\textsubscript{2})</th>
<th>SO\textsubscript{x} (As SO\textsubscript{2})</th>
<th>Total Suspended Particulates (TSP)</th>
<th>Non-Methane Hydrocarbons (As CH\textsubscript{4})</th>
<th>CO</th>
</tr>
</thead>
<tbody>
<tr>
<td>1. Existing diesel engines using digester gas (to be retired from service)</td>
<td>1,800</td>
<td>600</td>
<td>225</td>
<td>650</td>
<td>1,740</td>
</tr>
<tr>
<td>2. HERS (year 2000 conditions)</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>a. Gas turbine combined-cycle system using digester gas</td>
<td>680</td>
<td>20</td>
<td>55</td>
<td>25</td>
<td>715</td>
</tr>
<tr>
<td>b. Thermal processing of SDF</td>
<td>395</td>
<td>145</td>
<td>55</td>
<td>20</td>
<td>65</td>
</tr>
<tr>
<td>c. Total HERS (a+b)</td>
<td>1,075</td>
<td>165</td>
<td>110</td>
<td>45</td>
<td>780</td>
</tr>
<tr>
<td>3. NET REDUCTION FROM PRESENT CONDITIONS</td>
<td>725</td>
<td>495</td>
<td>115</td>
<td>605</td>
<td>960</td>
</tr>
</tbody>
</table>
### TABLE 3 COMPARISON OF ALTERNATIVE METHODS FOR TEMPERATURE CONTROL

<table>
<thead>
<tr>
<th>Control Method</th>
<th>Advantages</th>
<th>Disadvantages</th>
</tr>
</thead>
<tbody>
<tr>
<td>Dilution by Excess Air</td>
<td>Well proven, Simple to operate and control.</td>
<td>Higher NOx emissions, Large quantities of flue gas, Reduced potential for waste heat recovery.</td>
</tr>
<tr>
<td>Heat Transfer Surfaces</td>
<td>Used in recent fluid bed designs, Lower flue gas volume.</td>
<td>Limited control with fixed surface area, Metallurgy not well proven in a reducing atmosphere, Reduce turndown capability due to constant heat extraction, Requires accurate design calculations.</td>
</tr>
<tr>
<td>Addition of Recycle Flue Gas</td>
<td>Well proven, Simple to operate and control, Effective NOx control strategy, Turndown limited only by fluidization in bed and fan characteristics, Requires less accurate design calculations.</td>
<td>Increased flue gas volume, Increased reactor size, Increased baghouse size, Increased energy usage for RFG fan.</td>
</tr>
<tr>
<td>Water Sprays</td>
<td>Simple in concept, Lower flue gas volume</td>
<td>Reduced heat recovery potential, Difficult to control, Not generally used as primary control.</td>
</tr>
<tr>
<td>Sub-stoichiometric Oxygen Control</td>
<td>Simple to control and operate</td>
<td>Applicable only to initial gasification stage, Pyrolysis reactions make heat balances difficult to predict.</td>
</tr>
</tbody>
</table>
zones where heat is liberated, namely the fluidized bed, freeboard, and the afterburner up to the point where excess air is added. Determining the surface area required in each zone requires rather precise knowledge of the applicable heat transfer coefficients. Once transfer surfaces are installed in the reactor, a nearly constant heat rate is extracted since the surface area is constant and temperature differences nearly constant. Thus, maintenance of a specific temperature profile requires a precise calculation of the surface area. Any error in calculation will be reflected by an increase or decrease from the desired temperature.

Heat transfer in the fluidized bed can be adjusted while the bed is in operation by two methods. One, bed depth can be adjusted to increase or decrease the tube area exposed to the high heat transfer rates in the fluidized bed. Two, air supply to portions of the bed can be stopped, "slumping" that portion of the bed and thereby reducing heat transfer. Both strategies have been used successfully in full-scale systems.

While heat transfer rates within the bed are amenable to operational controls, the same is not true in the overbed and afterburner regions. Heat transfer area and transfer coefficients cannot be adjusted "on the run". If temperature profiles are not correct or exceed limits established for the process, the system must be "brought down" and transfer surface added or deleted, a difficult prospect at best. This becomes especially difficult in a "fuel-following" system where burning rates may vary from 50 to 100 percent of design capacity.

Another disadvantage of using heat transfer surfaces is that the majority of full-scale experience is with oxidizing atmospheres. Concern was raised over potential erosion/corrosion of boiler tubes in a reducing atmosphere. Experts were consulted and no clear consensus of opinion emerged from these discussions. There appeared to be significant uncertainties over the effect of reducing conditions on the service life of metals used for heat transfer.

Flue Gas Recycle

Addition of a diluent gas as a cooling medium offers several advantages compared to use of heat transfer surface. Temperature can be controlled at any point in the combustion process by the quantity of diluent added. Also, the control strategy is not dependent on exact design calculations. It is only necessary to know the maximum probable air and diluent quantities for sizing of ducting, blowers, and nozzles. Diluent gas addition can then be adjusted "on the run" to maintain a set point temperature and compensate for unknowns in the design calculations.

Steam and recycled flue gas were both evaluated as possible diluents. Compared to steam, recycled flue gas (RFG) offers improved energy production, reduced boiler feedwater makeup, and essentially no impact on the steam power circuit. The major concern with RFG was the effect of fine particulate recycle and the potential for build-up of high concentrations of fly ash fines. This potential problem was resolved by recycling from downstream of the particulate control system (baghouse). The size of the baghouse was increased by downstream recycle but this was judged as an acceptable trade-off.

The RFG system as initially proposed included recycle to all combustion zones where heat is generated, including the fluidized bed. These are the same regions that would otherwise require heat transfer surface as discussed previously. Flue gas recycle to the overbed and afterburner zones was adopted for the system design. Recycle to the fluidized bed itself was deferred in favor of sub-stoichiometric oxygen control.

Sub-Stoichiometric Oxygen Control

In the sub-stoichiometric region, heat release is proportional to the quantity of oxygen added for combustion. This allows a control strategy whereby combustion air to the fluidized bed is controlled by bed temperature, decreasing with temperature rise above set point, and vice-versa. This control approach has been used successfully for temperature control in both fluidized bed and multiple hearth furnaces. The principal advantage to the HERS system is that flue gas recycle is no longer required in the fluidized bed. Bed surface area is also reduced because stoichiometric conditions will be lower than the 60 percent of TOC nominally used in the test program. Heat balances predict that the fluidized bed should equilibrate at 1750°F (954°C) with about 28 to 45 percent of TOC. This rather large range of estimate results from uncertainties in assumptions for the chemical equilibrium of the gasification reactions, a subject not definitively treated in the literature. However, the advantage of maintaining bed temperature by sub-stoichiometric oxygen control is that the system will adjust itself "on the run." Air supply blowers and the bed surface area must be designed to span the range of expected stoichiometry. The system can then adjust itself in operation and less demanding process calculations are required.

Excess Air Preheat

One conclusion of the pilot combustion program was that air input to the staged afterburner should bring the gas to a near stoichiometric condition at one point. Once
full stoichiometry is reached, however, little fuel value will remain when the excess air is added. With little or no fuel value in the combustion gas, excess air will act as a diluent and reduce the gas temperature.

A temperature of 1700°F (930°C) in the second-stage afterburner was recommended to reduce CO and THC emissions. Based on projected mass balances and assuming excess air input at 80°F (27°C), temperature in the first-stage SAB would have to rise to 2300°F (1260°C) to maintain the second-stage at 1700°F (930°C). This condition is undesirable because of the likelihood of increased NOx emissions and the certainty of ash softening.

Preheat of excess air will be used to reduce the first stage afterburner temperature. Excess air is preheated against the second-stage afterburner exhaust prior to introduction into the bottom of the first-stage afterburner. This is a desirable location for the preheater since tube metal surfaces are exposed only to oxidative conditions. Flue gas temperature is also reduced, which further reduces concern over ash softening and possible adhesion to superheater and boiler tubes.

PROCESS CONTROL

The HERS combustion system is "fuel-following" as opposed to "load-following." The fuel feed rate is expected to be constant with once/day adjustments based on the fuel inventory. Fuel feed rate will not be adjusted to meet an outlet demand for process steam or power. During normal operation the process controls simply operate to maintain proper temperatures and stoichiometry for the established fuel feed rate.

Because the fluidized bed is sub-stoichiometric, an increase in bed temperature will signal for a decrease in air supply; similarly, a decrease in temperature will signal an increase in air supply. RFG to the overbed is controlled by overbed temperature. Outlet temperature from the second-stage afterburner controls the RFG flow to the first-stage afterburner. In each of these cases the control loop adjusts RFG flow to maintain the set point temperature.

O₂ content in the flue gas from the second-stage afterburner controls the excess air supply. Again, the control loop adjusts the excess air supply to maintain the desired set point O₂ content.

The combustion air blower supplies additional oxygen to both the overbed and first-stage afterburner regions. The control circuit is designed to maintain set point stoichiometry (i.e., percent of TOC) in both the overbed and first-stage afterburner. Stoichiometric set points will nominally be 80 percent of TOC in the overbed and 100 percent in the first-stage afterburner. The proper combustion air supply is determined by first summing all input air flowrates to the combustion train. The mass flowrate of oxygen to reach full stoichiometry is then determined from this value and the oxygen content in the exhaust gas. The mass flow of O₂ from RFG and bed fluidizing air are subtracted to determine the additional O₂ required from secondary air in each case. Control signals are then sent to control valves on the combustion air blower.

Only the basic control logic is described above. A number of additional control loops are included to check the basic control logic, provide corrective or comparative signals, provide alarm signals if process variables exceed predetermined levels, and for startup and shutdown procedures.

A major advantage of the control system described above is that the combustion process can be adjusted "on the run" to optimize the temperature and stoichiometry profiles. Also, the control system can adjust the process to compensate for unknowns in the engineering design. This is considered an important advantage since it reduces the need for exact design calculations.

AIR QUALITY CONSIDERATIONS

AIR POLLUTION CONTROL SYSTEM

As discussed previously, NOₓ, CO, and THC will be controlled by the temperature and stoichiometric conditions maintained in the combustion process. Particulate emissions will be controlled by a multiclone (multiple cyclone) in series with a fabric filter baghouse. A three-stage wet scrubbing system will be used for control of acid gases (i.e., SOₓ, HCl, HF) and condensible particulates.

The multiclone is designed to reduce ash loading in the baghouse and remove any large particulates which may not be fully combusted. The presence of such "sparks" is considered unlikely based on the fuel particle size distribution and gas residence time in the combustion system. Nevertheless, the multiclone adds an additional measure of safety to assure that hot particulates to not burn holes in the bag material.

The baghouse is a conventional, multi-compartment, pulse jet type with a maximum net air/cloth ratio of 3.0 acfm/ft² bag area. Bag material is woven teflon on an expanded teflon membrane. Baghouse technology is considered as Best Available Control Technology (BACT) by local regulatory agencies. Use of wet scrubbers or electrostatic precipitators for particulate control was not seriously considered for this reason. A guaranteed outlet loading of 0.0035 grain/scf (8.0 mg/Nm³) (front-half EPA Method 5) is required from the baghouse supplier.
Two venturi scrubbers and a tray tower in series comprise the three-stage, wet, flue gas desulfurization system. The system is designed to maximize use of wastewater secondary effluent and centrate from dewatering, both of which are readily available within the treatment plant. Secondary effluent is a high quality water which contains about 200 mg/l alkalinity. Centrate is less desirable from a quality standpoint but contains about 4500 mg/l alkalinity. The venturi scrubbers are supplied with sufficient secondary effluent to neutralize expected mass emissions of acid gases. Two venturi scrubbers are used to reduce the liquid/gas ratio required when using secondary effluent for neutralization. Centrate will be used as necessary to augment the alkalinity.

The tray tower is used as a polishing scrubber and will use a sodium sulfite/bisulfite scrubbing liquor with pH maintained by NaOH addition. Since plant effluent will neutralize most of the acid gases, use of NaOH is not cost prohibitive. Another advantage is that the liquid sidestream, composed primarily of soluble sodium bisulfite, can be easily disposed with the treatment plant effluent. Use of lime or limestone for scrubbing would result in a solid precipitate that would have to be land disposed.

The wet scrubbing system will also cool the flue gas to within 10°F (5.5°C) approach to the inlet water temperature. An exit flue gas temperature below 100°F (38°C) is expected from the tray tower. During the combustion tests, flue gas cooling and wet scrubbing were shown to be effective in removing condensible particulates. The HERS design provides a high liquid/gas ratio and the maximum cooling possible with the available water supply.

Sub-cooling below the saturation vapor temperature in the scrubbing system will remove considerable water from the flue gas and aid in steam plume suppression. Cooled flue gas from the tray tower is reheated about 35°F (19°C) in the ID fan and then mixed with 250°F (120°C) exhaust from the gas turbine, combined cycle system. The combined gas flow is discharged through a single stack at about 220°F (104°C).

AIR EMISSION BALANCE

Air emission factors for the HERS combustion system are presented in Table 4. These factors are believed to be realistically achievable in continuous operation. They are based on results of the combustion test program and the expected efficiency of the air pollution control system.

A summary of air emission projections for HERS is presented in Table 2. Emission reductions from retirement of existing diesel engines were previously discussed. The mass emission rates expected from the HERS facilities under year 2000 conditions are less than the available offsets for all primary air pollutants. Implementation of HERS will result in a net decrease in air pollutant emissions from existing conditions. All necessary “permits to construct” from air regulatory agencies have been issued for the project.

HERS ENERGY BALANCE

A process diagram for the energy recovery elements of HERS is presented in Fig. 9. Projected digester gas production in year 2000 is 6.25 million standard cubic feet per day (0.177 MNm³/day), equivalent to 4060 million Btu/day (1020 MKcal/day). About 265 dtpd of SDF will be produced, equivalent to 3400 million Btu/day (860 MKcal/day). The gas turbine generators will produce about 12.6 MW. Each gas turbine will exhaust to a heat recovery steam generator (HRSG) producing superheated steam at a nominal 870°F/1000 psig (470°C/7600 kPa) and low pressure saturated steam at 30 psig (210 kPa). Superheated steam will be expanded through a backpressure turbine producing about 2.4 MW. Expanded steam at 165 psig (1140 kPa) and low pressure steam at 30 psig (210 kPa) will satisfy all process demands for digester heating and CG drying.

Waste heat boilers in the FBG/afterburner combustion trains will produce about 100,000 lb/hr (45,400 kg/h) of superheated steam at 750°F/620 psig (340°C/4290 kPa). The steam will be expanded in a condensing turbine with three stages of automatic extraction for deaeration and feedwater heating. About 10.1 MW will be produced. Secondary effluent from the HTP will be used for condenser cooling.

With the above fuels, HERS will produce about 25 MW of electrical power and all process steam requirements. The Hyperion Treatment Plant is expected to consume a total of about 15 MW for both wastewater treatment and solids processing, leaving about 10 MW of power for export to the electrical utility grid.

ACKNOWLEDGMENT

The authors gratefully acknowledge the support and encouragement of the City’s Board of Public Works and the Office of the City Engineer and the many people who contributed to the successful conduct and completion of the test program and the system design. The contributions of the Montgomery-Parsons design team, too numerous to mention individually, Mr. Clarence Wall of Dorr-Oliver, Inc. and special consultants Mr. Dale Bergstedt and Mr. Konrad Semrau are especially noteworthy.
FIG. 9 PROCESS DIAGRAM FOR THE ENERGY RECOVERY ELEMENTS OF THE HERS PROJECT. ENERGY PRODUCTION ESTIMATES ARE FOR YEAR 2000 CONDITIONS WITH 6.25 MILLION SCFD OF DIGESTER GAS (4060 MILLION BTU/DAY) AND 265 DTPD OF SDF (3400 MILLION BTU/DAY).
### TABLE 4 FACTORS USED FOR ESTIMATING EMISSIONS FROM HERS SDF THERMAL PROCESSING FACILITIES

<table>
<thead>
<tr>
<th>Process Description:</th>
<th>Fluidized bed gasification of sludge derived fuel with staged combustion of the fuel gas.</th>
</tr>
</thead>
<tbody>
<tr>
<td>Emission Controls:</td>
<td>Combustion controls for NO\textsubscript{x}, CO, and hydrocarbons. Baghouse for solid particulates and wet scrubbing for acid gases and condensible particulates.</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Pollutant</th>
<th>Factor</th>
<th>Remarks</th>
</tr>
</thead>
<tbody>
<tr>
<td>NO\textsubscript{x}</td>
<td>75 ppm as NO\textsubscript{2} in dry gas exiting scrubber at 5.5% O\textsubscript{2}</td>
<td>Emissions equal to or less that this were achieved in the combustion test program.</td>
</tr>
<tr>
<td>SO\textsubscript{x}</td>
<td>20 ppm as SO\textsubscript{2} in dry gas exiting scrubber at 5.5% O\textsubscript{2}</td>
<td>Two stage venturi scrubbing with plant effluents followed by a polishing tray tower with caustic.</td>
</tr>
<tr>
<td>CO</td>
<td>20 ppm as CO in dry gas exiting scrubber at 5.5% O\textsubscript{2}</td>
<td>Afterburner design incorporates conditions to achieve this level of emissions based on results of the combustion test program.</td>
</tr>
<tr>
<td>NMHC</td>
<td>10 ppm as CH\textsubscript{4} in dry gas exiting scrubber at 5.5% O\textsubscript{2}</td>
<td>Afterburner design incorporates conditions to achieve this level of emissions based on results of the combustion test program.</td>
</tr>
</tbody>
</table>

**PARTICULATES**

| Solids - | 0.002 gr/dscf (4.6 mg/Nm\textsuperscript{3}) at 3% O\textsubscript{2} in baghouse exhaust | Based on fly ash particle size distribution and expected removal efficiencies. 0.0035 gr/dscf guaranteed by baghouse supplier. |
| Condensibles - | 0.008 gr/dscf (18 mg/Nm\textsuperscript{3}) at 3% O\textsubscript{2} in scrubber exhaust | Equivalent to levels achieved in combustion test program. Efficient combustion, flue gas cooling, and wet scrubbing will be used to control condensibles. |
| Total - | 0.01 gr/dscf (23 mg/Nm\textsuperscript{3}) at 3% O\textsubscript{2} in scrubber exhaust | |

718
REFERENCES


