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Biomass Gasification and Power Generation Using Advanced Gas Turbine Systems

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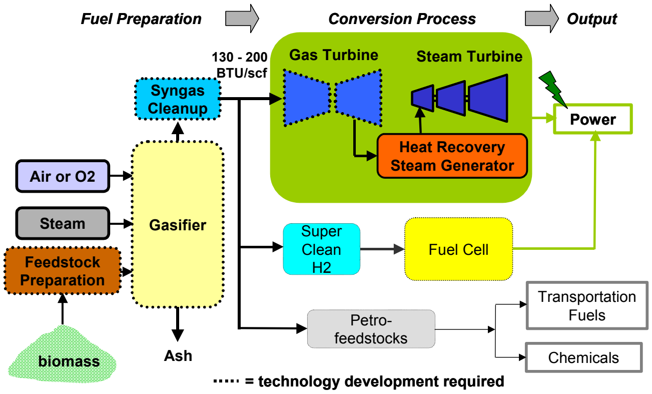
Biomass Gasification and Power Generation Using Advanced Gas Turbine Systems

EXECUTIVE SUMMARY

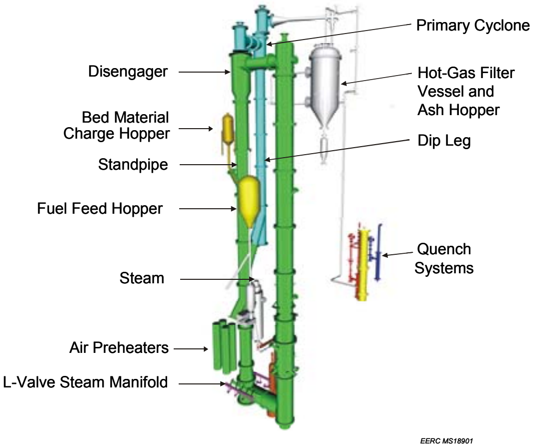
A multidisciplined team led by the United Technologies Research Center (UTRC) has identified a biomass integrated gasification combined-cycle (BIGCC) plant that is both **technically feasible <u>and</u> economically viable.**

- ➢ overall integrated system efficiency of **45%** (HHV)
- ➢ emission levels **less than half** of New Source Performance Standards (NSPS)



The key system attributes are:

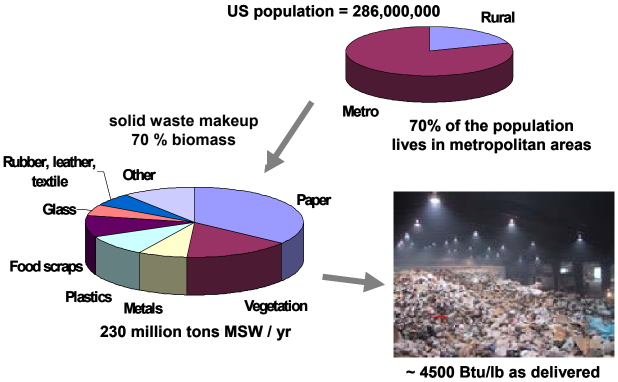
- An air-blown, high-pressure, fast-circulating fluidized-bed Advanced Transport Gasifier (ATG) having **wide fuel flexibility and high gasification efficiency.**



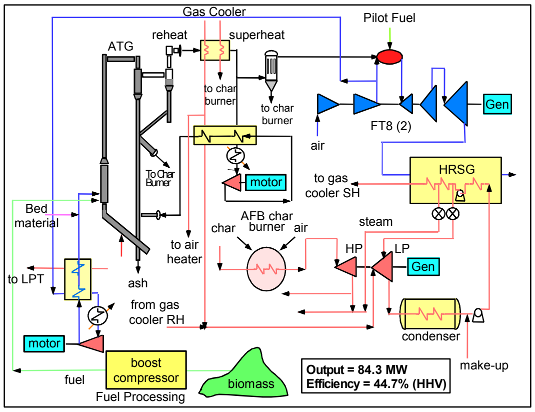
The reduced size of the ATG lowers its cost and allows use of standard off-the-shelf components which can be economically scaled as needed.

- An FT8 TwinPac™-based **combined-cycle of approximately 80 MWe**

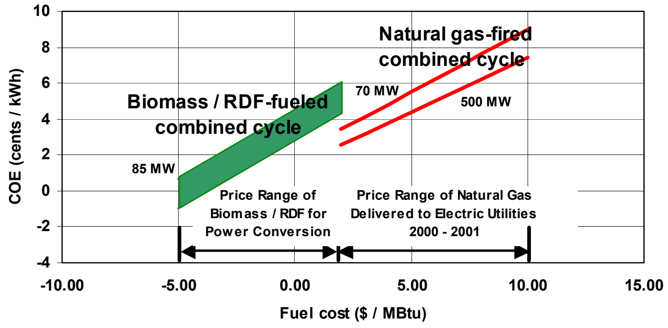
- **Sustainable low cost biomass primary fuel source** with widespread availability – refuse-derived fuel (RDF).



- An overall integrated system that **exceeds the DOE goal of 40% efficiency** at emission levels well below the DOE suggested limits at an estimated capital cost of $1500/kW.



- An advanced biofueled power system whose levelized **cost of electricity can be competitive** with other new power system alternatives.



The conceptual design of the BIGCC plant consists of a fluidized-bed Advanced Transport Gasifier, warm gas cleanup, and a PWPS FT8 TwinPac™ aeroderivative gas turbine operated in combined-cycle to produce ~80 MWe. The biofueled power system was found to have a levelized cost of electricity competitive with other new power system alternatives including natural gas combined-cycle.

The higher process efficiency of combined-cycle technology reduces consumption of fuel currently used for power generation, including that from foreign sources. In addition, the gasification process can be used for coproduction of fuels and chemicals, as well as power, to help the Nation achieve its energy goals.

To minimize development risk and allow early demonstration, the BIGCC plant maximizes the use of advanced technology, commercially available equipment and of components in DOE-sponsored advanced development programs. There are, however, near term requirements for R&D to adapt areas of the technology to the unique characteristics of biomass feedstock:

- Feedstock preparation for pressurized feed
- Warm gas cleanup for nitrogen compounds and trace elements
- Combustor modifications allowing low-NOx combustion of low calorific value fuels in compact combustors

None of these development areas present showstoppers but prompt attention to them would allow a meaningful scale (>20MW) demonstration to be undertaken within three years.

The addition of BIGCC to the national power generation portfolio would have significant Public benefits:

- ✓ Increased diversity of energy resources because the fuel flexible ATG can process a variety of feedstocks – low rank coals to biomass (energy crops, wood wastes, tires, RDF and others).

- ✓ Rapid attainment of renewable energy utilization goals because MSW/RDF is a low-cost sustainable biomass resource with the potential of generating 10 GW of clean power in the near term.

- ✓ Significant reductions in greenhouse gases, both because of biomass use and, when using RDF, because landfills and their associated emissions, particularly methane, would be greatly reduced.

- ✓ Increased security of power systems because BIGCC would be located within metropolitan areas without need of long distance transmission of natural gas or of the resultant electricity.

- ✓ Opportunity to increase biomass based energy utilization through combined heat and power systems.

- ✓ Potential for application to Vision 21 systems producing both high value-added chemicals and clean fuels as well as efficient electric power.

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LIST OF ACRONYMS AND ABBREVIATIONS

AEP	analytical electron microscopy
ASU	air separation unit
ATG	advanced transport gasifier
BGL	British Gas Lurgi
BIGCC	biomass integrated gasification combined cycle
C&D	construction and demolition
CCT	clean coal technology
CHP	combined heat and power
COE	cost of energy
CRRA	Connecticut Resources Recovery Authority
DOE	U.S. Department of Energy
EDF	Electricite de France
EERC	Energy & Environmental Research Center
FP&L	Florida Power & Light
FT8	PWPS gas turbine consisting of a two-shaft gas generator and power turbine
GRI	Gas Research Institute
GT	gas turbine
GTI	Gas Technology Institute
HAT	humid air turbine
HGFV	hot-gas filter vessel
HHV	high heating value
HRSG	heat recovery steam generator
HTW	High-Temperature Winkler
IGCC	integrated gasification combined cycle
IGT	Institute of Gas Technology
IPP	Industrial Power Producers
KRW	Kellog–Rust–Westinghouse
kWS	kraftWork Systems, Inc.
LHV	lower heating value
LRC	low-rank coal
MBtu	million British thermal units
MSW	municipal solid waste
MW	megawatt = million watts
MW_e	megawatt electrical
Northeast	Eleven Northeast US States = Connecticut, Delaware, Maine, Maryland, Massachusetts, New Hampshire, New Jersey, New York, Pennsylvania, Rhode Island, Vermont
NSPS	New Source Performance Standards
O&M	operating and maintenance
PG&E	Pacific Gas & Electric
PHI	fuel-air equivalence ratio

PSDF	power systems development facility
PSR	perfectly stirred reactor
PW	Pratt & Whitney (a division of UTC)
PWPS	Pratt & Whitney Power Systems (a division of UTC)
R&D	research and development
RDF	refuse-derived fuel
SCS	Southern Company Services
SMUD	Sacremento Muncipal Utility District
SOAPP	State-of-the-Art Performance Program
SOFC	solid oxide fuel cell
TPD	tons per day
TRDU	transport reactor development unit
UND	University of North Dakota
UTC	United Technologies Corporation
UTRC	United Technologies Research Center (a division of UTC)
WTE	waste-to-energy
XRFA	x-ray fluorescence analysis

Biomass Gasification and Power Generation Using Advanced Gas Turbine Systems

INTRODUCTION

At this time, there is an increasing awareness that the energy resources of this country are being affected by a number of factors that either constrain or actually threaten their availability for power generation. For example, our most plentiful and least expensive resource, coal, has many environmental issues, not the least of which is an accelerating global concern of greenhouse gas emissions. Natural gas, while currently available, is facing a growing transport infrastructure problem plus an inherent vulnerability to disruption, natural or man made. Oil, used only in a relatively few power plants, is costly and its supply subject to potentially unstable international sources. These factors have increased the interest in renewable resources including the use of biomass in its many forms as a potential energy source for both power generation and chemical production. Biomass has a number of potential benefits for the Nation including improved strategic security, improved rural economic development, improved environmental quality, reduced greenhouse gases, and an overall diversification of energy resources and a sustainable resource supply.

Currently, the majority of biomass-based power generation plants are small and relatively inefficient, e.g., < 2 MW with efficiencies generally in the ~20-25% range. These plants use agricultural wastes, landfill methane, wood wastes, etc. in various energy conversion devices such as steam turbines, internal combustion engines, and, more recently, fuel cells. Larger biomass combustion systems used in industrial plants, generally in the wood products industry, and in municipal solid waste disposal can range from 5- 75 MW, but again are not very efficient. With these factors in mind, the Department of Energy/National Energy Technology Laboratory, requested a program to identify methods to more efficiently use the biomass resources to generate power. DOE/NETL specified that gasification of biomass be integrated with the power plant into a system that offers significant improvements in thermal efficiency and environmental performance. Specifically, they requested that for plants under 100MW, a total power production efficiency of at least 35% with the clear potential to evolve to 45% be identified. In addition, emissions shall not be greater than one-half the amount allowed by current New Source Performance Standards for coal-fired electric power generating stations. The cost of power from these facilities must be competitive with local circumstances.

To meet these ambitious goals, the five-partner team has broken the project down into subtasks that are discussed in the following pages. We have identified and focused on a power system consisting of components that are either commercially available or in an advanced development stage, thereby reducing technical risk and advancing the time to commercialization. These components will form a highly efficient, and relatively inexpensive power system. Because of the unique requirements of biomass gasification, however, each of the major elements will require some additional development. For example: (1) the feedstock processing system, (2) the gasifier, (3) syngas cleanup, and (4) low-Btu gas turbine, have components that must be modified to perform effectively. The team feels strongly, however, that it has identified a sustainable biofueled system that can be commercialized and, subsequently, supply power economically to areas of the United States serving some 70% of the population.

The content of this report includes discussion of following highlighted topics.

- Economical power production from biomass depends on feedstock availability at low cost as well as sustainability of that supply. Since feedstock type will undoubtedly vary with plant location, a gasification technology that is highly feedstock-flexible is required. The experience of the EERC with the Advanced Transport Gasifier has been key in the gasifier selection process.

- Power system identification and performance evaluation are being done by kWS. The simulation model has been verified by comparison to EERC data, and performance characteristics have been explored.

- Combustion modeling at UTRC has shown that minimal FT8 hardware modification will be required to handle the syngas.

- Biomass evaluation in Connecticut and the Northeast focuses on wood and RDF as low-cost feedstocks. Team member CRRA operates a 2200-ton/day waste-to-energy facility and has provided data that indicate the viability and sustainability of RDF as a fuel in metropolitan regions.

- Emission and emission control technology that indicates that the biogas fueled power system can reach emission levels only one-tenth of the New Source Performance Standards for equivalent coal-fired systems

- Economic analysis by kWS shows that capital costs combined with fuel costs for a BIGCC system allow the cost of electricity to be comparable to natural gas-fired systems.

GASIFICATION TECHNOLOGY

Introduction

Gasification is a partial-oxidation process for the conversion of carbonaceous feedstocks to combustible gas mixtures consisting primarily of carbon monoxide (CO), hydrogen (H_2), and methane (CH_4). This conversion occurs at elevated temperatures and pressures according to several competing reactions. These reactions are as follows:

$C + O_2 \leftrightarrows CO_2$ (combustion reaction; highly exothermic at 14,167 Btu/lb C) [a]
$C + CO_2 \leftrightarrows 2CO$ (Boudouard reaction; endothermic at 6016 Btu/lb C) [b]
$C + H_2O \leftrightarrows CO + H_2$ (carbon-steam reaction; endothermic at 4863 Btu/lb C) [c]
$CO + H_2O \leftrightarrows CO_2 + H_2$ (water-gas shift reaction; mildly exothermic at 1152 Btu/lb C) [d]

Another reaction that can take place at temperatures less than 1093°C (2000°F) and at very high operating pressures is:

$C + 2H_2 \leftrightarrows CH_4$ (carbon hydrogenation reaction; exothermic at 3282 Btu/lb C) [e]

Plus, whenever the carbonaceous feedstock can generate significant quantities of volatile matter, methane can be formed by thermal cracking of the volatile matter according to the qualitative reaction:

$C_mH_n \leftrightarrows n/4CH_4 + (m-n)/4C$ (thermal cracking reactions; endothermic) [f]

Gasification of carbonaceous feedstocks is typically done in the presence of an oxidant (air or O_2) and steam/water vapor in order to conduct the gasification reactions shown above. Injection of CO_2 and H_2 into the gasifier in order to promote reactions [b] and [e] have also been investigated but is not done commercially. Gasifiers which indirectly supply the heat to promote the endothermic gasification reactions have also been developed. These types of gasifiers do not add any oxidant to the gasifier, thereby improving the fuel gas heating value.

Second-Generation Advanced Power Systems

Goals for future market-driven power systems used by DOE planners are listed in Table 1. Capital costs under $1000/kW can be met only by highly simplified systems based on hot-gas cleanup, with minimum heat recovery and reheat duty. Combined-cycle efficiencies of 45%–50% will require that pristine fuel or combustion gases are delivered to gas turbines operating at elevated expander temperatures above 1204°C (2200°F). These future performance goals will have to be met under more stringent environmental requirements, including 95%–99% sulfur control and NO_x standards as low as 0.1 lb/MBtu. Air toxic emissions pose additional, but as yet unknown, requirements. First-generation systems cannot meet all of these requirements, and the future of advanced power generation depends, in considerable measure, on the success of emerging second- and third-generation systems.

Table 1: U.S. DOE Goals for Future Market-Driven Coal-Fired Power Systems

Capital Cost	Under US$1000/kW
Efficiency	45%–50%
Sulfur Control	95%–99%
NO_x Control	0.1–0.3 lb/MBtu
Air Toxics	Meet 1990 CAAA regulations

Simplified IGCC systems [1,2] designed to minimize capital cost will be necessary to keep the process economics reasonable for a smaller-scale (under 100MW) gasification system required by biomass-based power systems. Such a system, shown schematically in Figure 1, will likely consist of a gasifier, a hot-gas cleanup module, and the gas turbine/steam turbine power system in order to achieve 45% system efficiency (solicitation requirement). System configurations are predicated on the type of gasifier used, which ideally would be air-blown to avoid an expensive air separation unit (ASU) and match the temperature requirement of the hot-gas cleanup module. Gasifier exit gas temperatures vary from as high as 1400°C (2552°F) in the Texaco entrained-flow gasifier down to as low as 200°C (392°F) in a moving-bed gasifier operating on a high-moisture fuel.

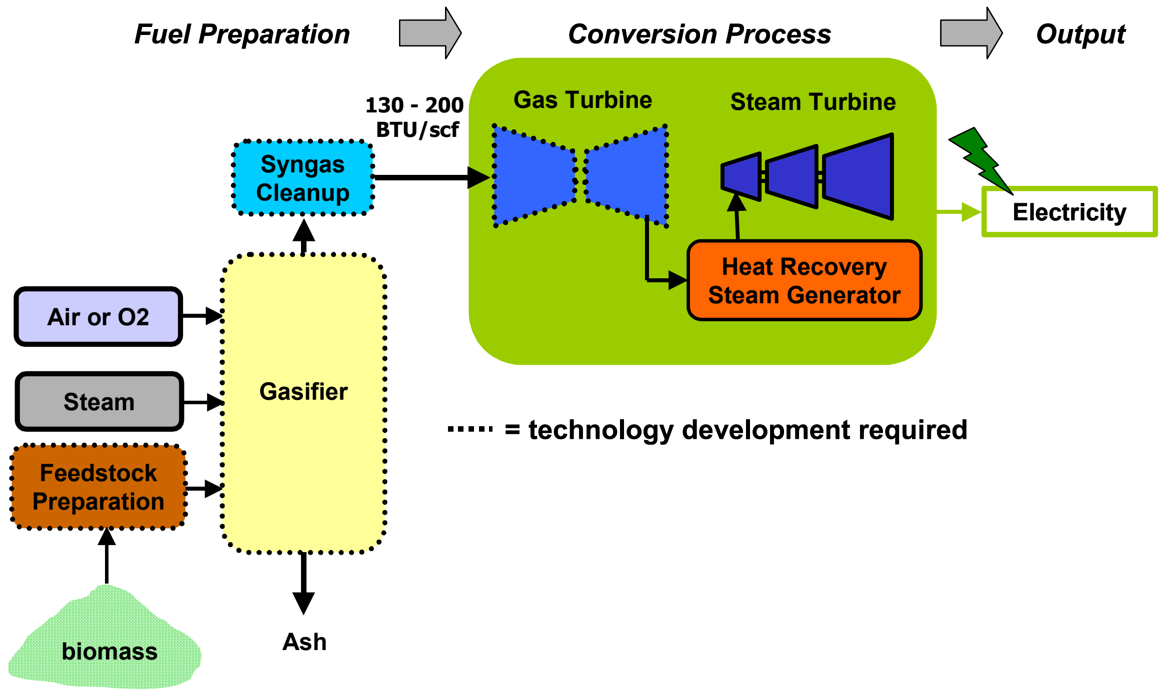

Figure 1: Simplified Schematic of a Biomass Integrated Gasification Combined Cycle System

The second-generation IGCC systems that are the principal focus of this evaluation are entering their demonstration phase of development. The IGCC demonstrations involve different combinations of air- or oxygen-blown gasification along with hot- or cold-gas cleanup for particulates and H_2S and combustion control for NO_x. The three major types of gasifiers (classified by the way fuel flows in the device) are:

1) Fixed-bed gasification systems (British Gas/Lurgi and Lurgi)
2) Entrained-flow (Texaco, E-Gas (Destec), Shell, and Prenflo)
3) Fluidized-bed (Advanced Transport Gasifier (ATG), Kellogg–Rust–Westinghouse (KRW), High-Temperature Winkler (HTW), Gas Technology Institute RenuGas, FERCO SilvaGas

These specific systems are discussed in more detail in Appendix A.

Fixed-Bed Gasifiers

Fixed-bed gasification has the inherent advantages of high thermal efficiency and relatively low offgas temperature because of the countercurrent flow of fuel and gaseous reactants. A constant-depth bed of fuel is maintained in the gasifier by adding fuel at the top and discharging ash or slag through a grate or taphole at the bottom. The fuel fed at the top is progressively heated and reacted as it moves slowly down through drying, devolatilization, gasification, and oxidation zones. The air or oxygen and steam used as the gasification medium converts fuel to synthesis gas (CO and H_2) by partial oxidation, steam gasification, and water-gas shift reactions. The rising high-temperature gases serve to dry and carbonize the fuel in the upper portion of the bed, causing the raw offgas to contain all of the fuel moisture and devolatilization products, including methane, hydrocarbons, and heavy tars. The process steps required to condense and separate this tar, oil, and gas liquor add to the complexity and cost of fixed-bed gasification systems, compared to other types of gasification where volatile products are consumed in the gasifier; however, the separated tar and oil can also provide a source of potentially valuable by-products. Because of the importance of maintaining gas flow and limiting fuel dust carryover in the product gas, the fuel feed to a fixed-bed gasifier must be double-screened to remove fines and provide a minimum particle size of about 6.4 mm (1/4 inch).

Entrained-Flow Gasifiers

Entrained-flow gasifiers rapidly convert pulverized fuel to synthesis gas in a short residence time by partial oxidation typically with oxygen at high temperatures of 1370° to 1925°C (2500° to 3500°F). Molten ash produced at these elevated operating temperatures is continuously water-quenched and removed as a fritty glassy slag. Fuel feed to pressurized units is accomplished either by pumping a fuel–water slurry or injecting dry fuel in a dense phase of transporting nitrogen. The principle advantages of entrained-flow gasifiers are in their conceptually simple design, good tolerance of caking fuels, very high throughput, high carbon conversion efficiency, and thermal flexibility for increasing operating temperatures well beyond the melting point of ash. Disadvantages are the large amount of gas cooling and heat recovery necessitated by the high exit gas temperature, the limited opportunities for in-gasifier sulfur capture, the complex feeding systems required for pressure operation, and the necessity for close control of oxygen feed rate for safe operation.

Commercial entrained-flow gasifiers are offered by Texaco, Dow, Shell, and Prenflo for oxygen-blown operation and are described in detail in Appendix A. All of these demonstrations of advanced IGCC systems based on an entrained-flow gasifier have utilized cold-gas cleanups. However, only the Prenflo plant approaches the 45% thermal efficiency goal established for this project because of its use of cold-gas cleanup technology. Wabash River, Buggenum, and Puertollano all use hot- or warm-gas filter systems to recover ash for recycle to the gasifier; however, these plants still use subsequent water scrubbers and amine-based scrubbers for H_2S removal. The use of hot-gas cleanup technologies with these technologies is possible and has been demonstrated in part by operations of pilot-scale and slipstream systems.

For biomass feedstocks, slurry feed gasifiers are not feasible because of the inability to slurry biomass to very high solids concentrations owing to the hydrophilic nature of most biomasses. Even the use of dry-feed entrained-flow gasifiers presents some operating issues since the small feedstock particle-size requirements may be very hard to achieve cost-effectively with most biomass materials. The high capital and operating costs associated with the need for an air separation unit for the oxygen feed to these gasifiers is also a concern given the smaller size range (30 to 80 MW) at which a biomass-fed system is going to be operating. Internal diameter information on these types of gasifiers is not readily available.

Fluidized-Bed Gasifiers

Fluidized-bed gasification operates on the principle of suspending fuel, along with other solids present in the reactor, in turbulent motion in the high-velocity upward flow of reactant gas. The turbulent environment provides excellent gas–solid contact, which rapidly heats entering reactants to the bed temperature and facilitates their intimate mixing and reaction.

The advantages of fluidized-bed gasifiers are in their ability to incorporate in-bed sulfur capture using limestone and their reduced gas-cooling requirement. Up to 90% sulfur removal can be accomplished in the bed at temperatures around 900°C (1652°F), where the limestone is substantially calcined (mixed metal oxide sorbents such as zinc ferrite, or zinc titanate, can remove 99% of sulfur). But in-bed sulfur removal adds complexity by requiring another combustion unit to convert unstable calcium sulfide waste produced in the gasifier into calcium sulfate that is suitable for disposal, and to make use of unburned carbon in gasifier char. Operation of the gasifier above 900°C (1652°F) with high-alkali fuels also raises concern over bed agglomeration and the increased need for alkali gathering to reduce alkali vapor carryover into the gas turbine. High carbon conversions at low char recycle rates can be achieved at temperatures below 900°C (1652°F) when reactive fuels such as low rank coal or biomass are used, provided that a lower level of in-bed sulfur capture is acceptable.

Fluidized-bed gasification systems have been designed for a wide range of operating conditions involving different temperatures, gas velocities, gaseous reactants, and bed materials. Systems can be configured to include two or more beds in series to facilitate sequential reaction steps under optimum conditions. The inherent advantages of fluidized-bed gasification include 1) design flexibility for a wide range of fuel feeds, including caking fuels; 2) high specific gasification rates resulting from high rates of heat and mass transfer; 3) in-bed sulfur removal using a limestone bed; 4) good control of gasification temperature and other reaction conditions in one bed or a series of beds to accomplish particular stages of reaction and 5) high-product gas uniformity resulting from the highly turbulent mixing. Disadvantages are in the carryover of significant carbon in the gasifier fines and limited turndown capability because of the need to maintain fluidizing velocities.

Fluidized-bed designs with application for high reactivity fuels include the advanced transport reactor, KRW gasifiers, high-temperature Winkler, the GTI RenuGas, and the FERCO SilvaGas. Demonstrations of IGCC based on fluidized-bed gasifiers include an advanced transport gasifier (ATG) at the EERC and PSDF, the Piñon Pine IGCC project (KRW), GTI RenuGas in the United States, and the HTW in Germany.

Gasifier Recomendation

After review of the biomass gasification solicitation objectives of 45% thermal efficiency with greater than 95% biomass thermal energy input and less than half the NSPS emission rates, it was clear that this project should focus on larger-scale (greater than 30 MW), pressurized gasifiers that would have flexibility in handling a variety of biomass feedstocks. As shown in Table 2, a fluid-bed gasifier is the best choice for meeting these objectives. Entrained-flow gasifiers are at a disadvantage to fluid-bed gasifiers because of their requirement for either slurried or finely pulverized feedstocks, both of which are hard to achieve with biomass feedstocks. Fixed-bed gasifiers cannot process fuels with fines less than 1/4 inch, so fuel processing to make pellets would be required for this type of gasifier. Fixed-bed gasifiers make significant levels of tars. Unless an alternative use for these tar/oils is available, only so much of these organics can be recycled to the hot zone of the slagging fixed-bed gasifier. Very little tar can be recycled to the dry-bottom fixed-bed gasifiers.

Table 2: Comparison of Gasifier Characteristics

Gasifier Type	Advantages	Disadvantages
Fixed-bed	• High thermal efficiency • Low volatilization temperature	• Produces tars that require separation • All feedstock must be >0.25” – biomass will require pelletization
Entrained-flow	• Simple design • Tolerant to caking feedstocks • High C conversion efficiency • Thermal flexibility	• Large volume of process gas • Oxygen operation generally required • Large vessel size • Water cooling required • Must pulverize feedstock • Cannot slurry biomass
Fluidized-bed	• air-blown operation • can handle all feedstock sizes • uniform temperature distribution minimizes production of tar and oils – high gasification rate • sorbents and bed materials can be changed to match fuel	• Carryover of carbon • Limited turn-down

The project team selected the advanced transport reactor (Figure 2) for the BIGCC plant because of its three major advantages:

1) The ATG's ability to separate the fuel feed point from the oxidant, thereby allowing better conversion of the volatile matter to fuel gas;

2) The ATG's higher fuel throughput allowing a significantly smaller diameter gasifier to be constructed than any other gasifier investigated. This smaller diameter allows for less expensive off-site construction and transport to the commercial site and allows for easier installation at the site;

3) The ATG requires less feedstock processing. Several biomass feedstocks, including RDF, have been successfully tested in other fluid-bed gasifiers, so feedstock type is not expected to be a problem in the ATG.

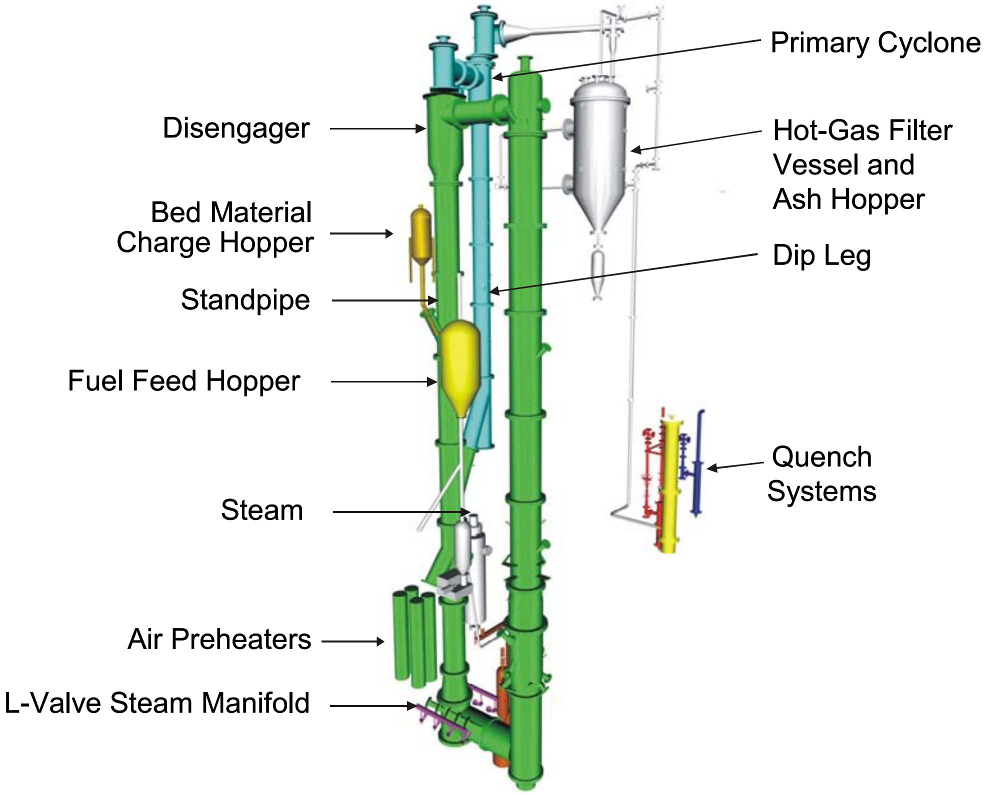

Figure 2: EERC Advanced Transport Gasifier

POWER SYSTEM TECHNOLOGIES

Introduction

A simulation model of the ATG was constructed by kWS and validated using data supplied by the EERC for operation of the ATG using lignite coal. The simulation is a heat and mass balance around the gasifier with gas-phase equilibrium obtained using typical reformer reactions. This approach is very accurate for high-temperature gasifiers where water-gas shift and methane conversion reactions are near equilibrium. The module was used to predict the performance of an ATG/FT8 TwinPac™ combined cycle using a biomass feedstock. Preliminary estimates for this system project an overall HHV efficiency of 44.7% and a power output of 84.3 MW_e.

Gasifier Model

A simulation model of the ATG was constructed using the State-of-the-Art Performance Program (SOAPP), a proprietary simulation program first developed by the Pratt & Whitney Division of United Technologies for use in the analyses of the performance and design of aircraft gas turbine power systems. Over the years, the capability of SOAPP has been extended by UTRC and PWPS to ground-based power systems of all types including gas turbines, steam turbines, fuel cells, etc., and all the ancillary equipment and heat exchangers/boilers that make up modern power systems. In addition, UTRC and later kWS have developed models of the basic types of gasifiers such as air- and oxygen-blown entrained-flow, fixed-bed, and fluid-bed. Modules representing both high- and low-temperature cleanup systems have also been developed.

The simulation module representing the ATG is based on establishing a heat and mass balance around the gasifier and using equilibrium gas-phase reactions and reformer reactions. The module is shown in Figure 3. Fuel, air, steam, and makeup bed material are injected in the base of the riser. (Note: The exact number and location of the air and steam injection points are considered sensitive by Kellogg Brown and Root, the designers of the ATG in use at the PSDF, Wilsonville, Alabama. Thermodynamically, it does not make a difference exactly where injection occurs). Fuel gas is recycled to the gasifier, again exact location undefined, to serve as a carrier for the spent material at the base of the standpipe.

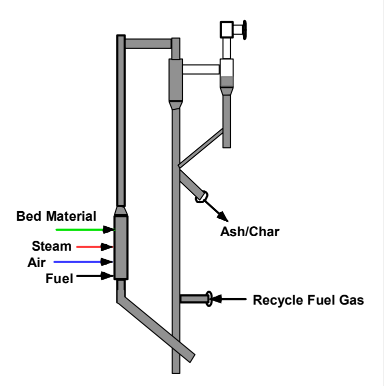

Figure 3: ATG Module used for Simulation

In the initial ATG model used in this phase of the program, contributions due to reactions involving the bed materials have been ignored. Once a module has been developed, it needs to be validated by comparison to actual operational data. The EERC at UND has operated a 2.4 - ton/day pilot-scale ATG on a variety of lignite, subbituminous, and bituminous coals. Comparison of the typical heating values and compositions of lignite and RDF showed remarkable agreement. Thus for the initial model, data for lignite gasification could be used both to validate the model and also to indicate roughly the potential performance with RDF. The results of the validation are shown in Table 3.

Table 3: Comparison of SOAPP Results and Gasifier Test Data.

	Simulation Results			**Gasifier Data**	
	Wet %	Dry %		Wet %	Dry %
H_2	16.00	19.54		15.95	19.45
CO	9.94	12.14		9.23	11.25
CH_4	2.59	3.16		2.59	3.16
C_2H_6				0.42	0.51
CO_2	12.34	15.07		13.68	16.68
N_2	39.67	48.44		38.93	47.48
Ar	0.44	0.54		0.97	1.18
H_2S	0.90	1.09		0.15	0.18
COS	0.02	0.02		0.00	
NH_3				0.09	0.11

As seen in Table 3, there is very good agreement between the simulation results and the operating data. The model does not predict any higher hydrocarbons since they generally do not have a material effect on the overall performance. Also, as mentioned previously, bed material reactions are ignored; thus sulfur compounds are present in the simulation at higher levels than in the actual data.

Power System Configuration

The power system is based on the PWPS FT8, a nominal 25-MW_e aeroderivative gas turbine widely used for peaking and midrange generation in the United States and worldwide. Many of the installations use two gas turbines arranged in a TwinPac™, as shown in Figure 4. This also serves as the base of a combined-cycle system having a nominal gas turbine output of 54.8 MW_e and a dual-pressure, 427°C (800°F)/ 87 bar (1250 psig) non-reheat steam turbine with a nominal output of 20.6 MW_e. The heat rate for this system is 7022 Btu/kWh. This combined cycle was used as the starting point for the integrated biomass gasification system.

Figure 4: PWPS FT8 TwinPac™ Installation

A flow sheet showing the components and the flow stations is shown in Figure 5. Air from the compressor discharge is cooled, further compressed, and then regenerated before going into the gasifier. RDF (886 TPD – 12% moisture) is further processed before being screw fed to the base of the gasifier. Steam and makeup bed material are also added at this point. After gasification, the hot gas from the gasifier passes through a cyclone that removes the majority of particulate carryover, then through a series of heat exchangers that drop the temperature to 370°C (700°F), the highest that can be tolerated by the gas turbine fuel control system. At this temperature, it then goes into a candle filter to remove essentially all the remaining particulates as well as alkali metals that would have condensed on these particulates. A slipstream of the fuel gas is cooled, compressed, and regenerated for use as a recycle gas to assure fluidization in the base of the gasifier and the "J" leg. By elevating the temperature of the recycle gas, the oxidant (air), and the steam, the amount of combustion in the gasifier can be minimized, and the calorific value of the fuel gas can be maximized. This enhances both the performance and burnability of the fuel.

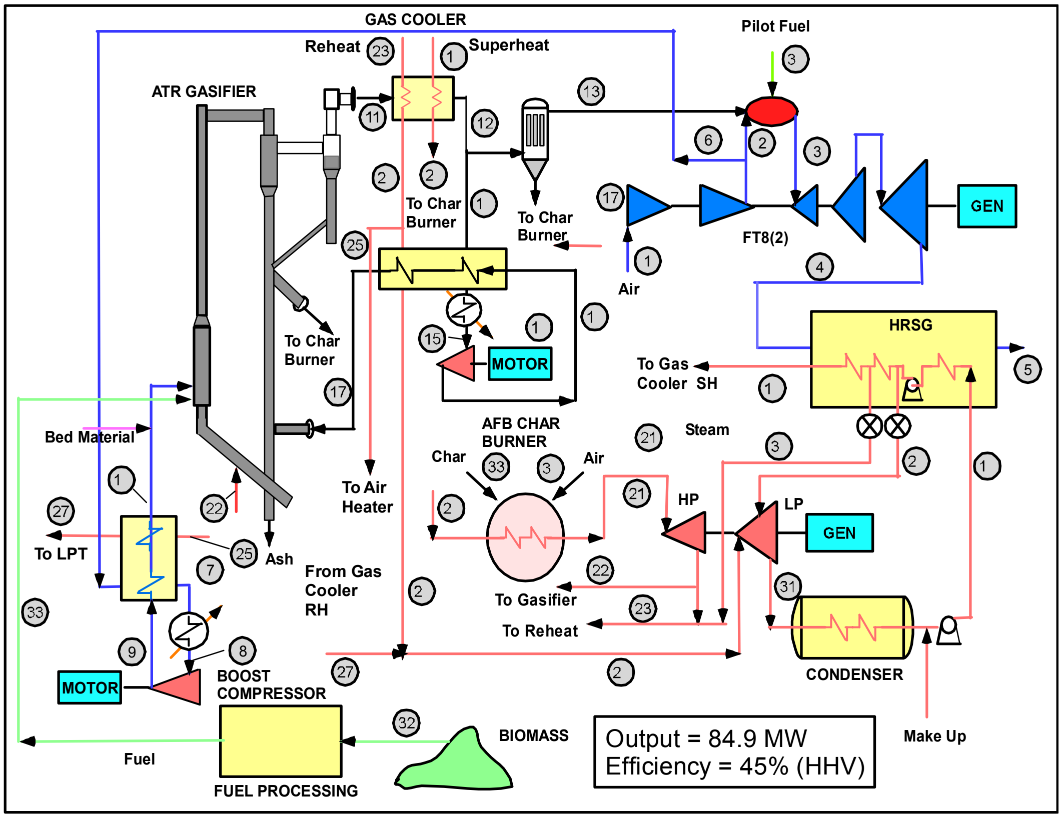

Figure 5: Flow Sheet for Biomass Gasification Combined Cycle Power System

The fuel gas is sent to the gas turbine combustor where it is burned with the remaining compressor discharge air. After expansion, the combustion products exhaust to a heat recovery steam generator (HRSG) where feedwater is heated and boiling is initiated. From the HRSG, the steam is sent to the fuel gas cooler and superheated, then to the char burner where further superheating occurs. From the char burner, a fluidized-bed combustor, the steam flows to the HP steam turbine. After expansion, the steam plus a portion of the steam from the HRSG are returned to the reheater, which is located in the fuel gas stream. Here, it is very highly superheated. Also at this point, some of the steam is sent to the gasifier. Before going to the reheat (LP) turbine, the steam is used to heat the recycle gas and the gasifier air. To maximize system efficiency, another steam slipstream is throttled and sent to a third pressure level in the steam turbine.

Alternative Power System

The power system described above requires nearly 900 TPD of RDF. For those areas where RDF availability might be more limited, there is an alternative. A single-engine plant would require approximately half the RDF input. Adding a steam turbine bottoming system could enhance its power output, albeit a costly addition at this scale. A second approach would use the GT exhaust to raise steam for injection into the GT. Funding limitations precluded a complete evaluation of this latter configuration, but preliminary estimates indicate potential for over 30 MW and efficiencies approaching 38%, see Figure 6.

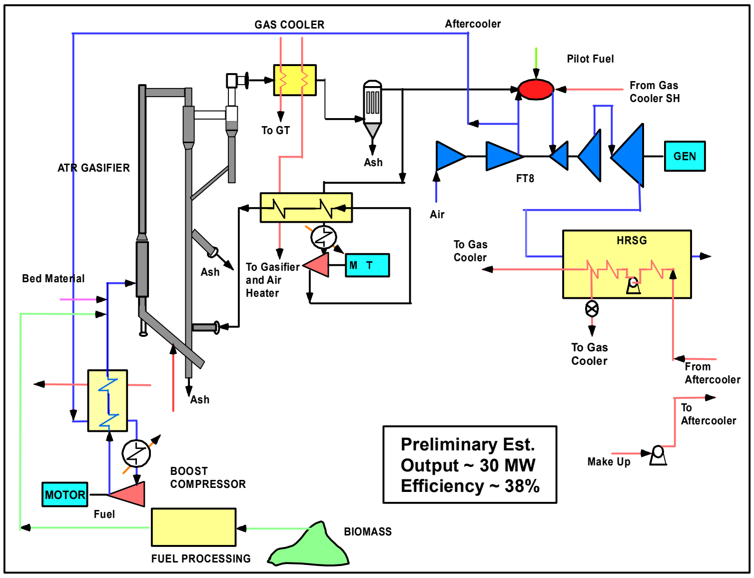

Figure 6: Flow Sheet for Biomass Gasification Steam Injected Gas Turbine Power System

A more futuristic biomass gasification hybrid power system is shown in Figure 7. Here, a solid oxide fuel cell (SOFC) is the major power system component. A gas turbine uses the exhaust gases from the pressurized SOFC as its working fluid and provides additional power and increases the system efficiency. A steam bottoming system adds more power and also increases the efficiency. While funding limited the consideration given to this system, prior studies of other gasification (coal) hybrid system indicate that such a system could approach 60% overall efficiency.

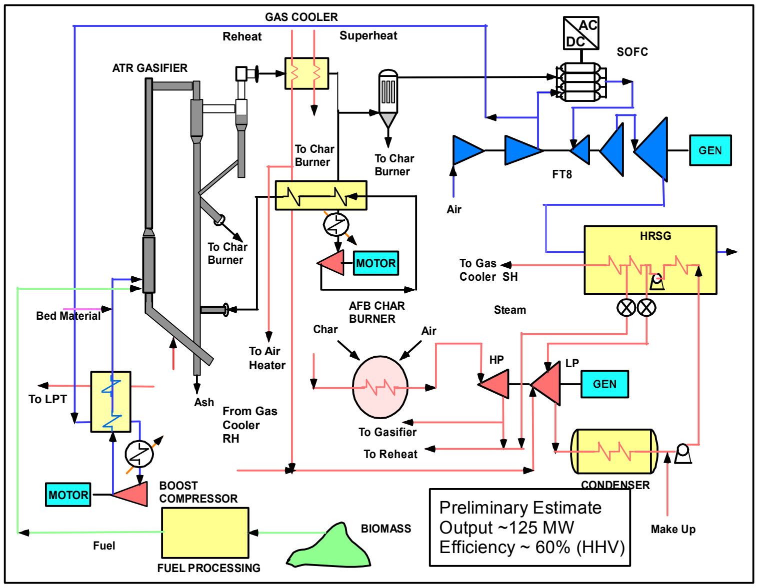

Figure 7: Flow Sheet for Biomass Gasification Hybrid SOFC/Gas Turbine Power System

An additional configuration that used an oxygen-blown version of the ATG was also investigated. This configuration was investigated to ascertain the effect on system performance that would occur were the gas turbine combustor to require a higher calorific value syn-gas. In this configuration, an air separation plant supplied 95% pure O_2 to the ATG. Additional steam was required to moderate the gasification reactions and recycle was increased to maintain fluidization. The overall effect was a 4% reduction in system efficiency to 42.9%. Most of this difference was due to the energy required to produce the oxygen. There was also a small decrease in steam turbine power. Since the preliminary combustion analysis indicated that the system would work over a wide range of conditions with the lower heating value syngas, no further work was done on this system.

Power System Technology Requirements

The power system has been chosen from commercially available equipment. The gas turbine is widely used, and the steam turbine is similar to those that can be specified and purchased from a number of vendors. This approach was chosen to reduce the risk associated with commercialization of the biomass gasification technology. The only component on the gas turbine that would not be bill of material is the combustor and its associated fuel control. While there are gas turbines that currently burn low-calorific-value fuel gas, they are generally large-scale heavy frame machines with very large combustors. Even with this advantage, significant R&D (research and development) took place to arrive at suitable low-emission combustors. Similar R&D would be required for the more compact combustors typical of aeroderivative gas turbines.

COMBUSTION MODELING

Introduction

Several computational simulations were made with various low-heating value fuel blends to qualitatively assess combustion characteristics. Both "wet" and "dry" fuel blends were evaluated in order to predict combustor performance compared to operation on natural gas. The fully detailed GRI-3.0 kinetic mechanism and the perfectly stirred reactor (PSR) code of the Chemkin-II software package [3] were utilized to determine each fuel mixture's PSR extinction time as a function of initial temperature and overall equivalence ratio (phi). The fuel blends (mole basis) are shown in Table 4.

Table 4: Fuel Blends Used in PSR Calculations

	Wet Mix	**Dry Mix**	**Natural Gas**
CO	9.91	11.26	
CO_2	14.69	16.69	0.81
H_2	17.13	19.47	
H_2O	12	0	
N_2	41.82	47.52	1.6
CH_4	2.78	3.16	94.69
C_2H_6	0.17	0.19	2.58
C_2H_4	0.28	0.32	
C_3H_8	0	0	0.29
N_3	0.1	0.11	
Ar	1.04	1.18	
O_2	0	0	0.02
(sum-moles)	99.92	99.9	99.99

Simulation and Estimation of Combustor Extinction Times

To first determine the overall equivalence ratio operating range, equilibrium calculations for the various blends were performed at 20 atm and 644 K (i.e., typical combustor entrance conditions during turbine operation). The threshold exit temperature was 1480 K (the minimum temperature necessary to provide enough power in the turbine). Additionally, the equilibrium calculations were performed with steam added to the fuel and air, to determine its operational effects if various alternate cycles (e.g., HAT or Cheng cycle) were considered. Shown in Figure 8 is a graph of exit temperature as a function of equivalence ratio and added steam. As expected, the wet gasifier blend with an additional 20% (mass) steam has the smallest range of equivalence ratios (~0.5–1.6), and the dry blend with no added steam has the highest operational equivalence ratio range (~0.4–2.1). These calculations were used to determine the applicable equivalence ratio range necessary for the PSR extinction calculations. (Essentially, the determination of the extinction time, for a given set of conditions, is made by continually reducing the reactant residence time in the reactor until no reaction occurs and the system "blows out." The residence time at which this occurs is the extinction time.) Plots of inverse extinction times for the various fuel mixtures (for inlet conditions of 20 atm and 644 K) are shown in Figure 9. Also shown is a similar profile for natural gas.

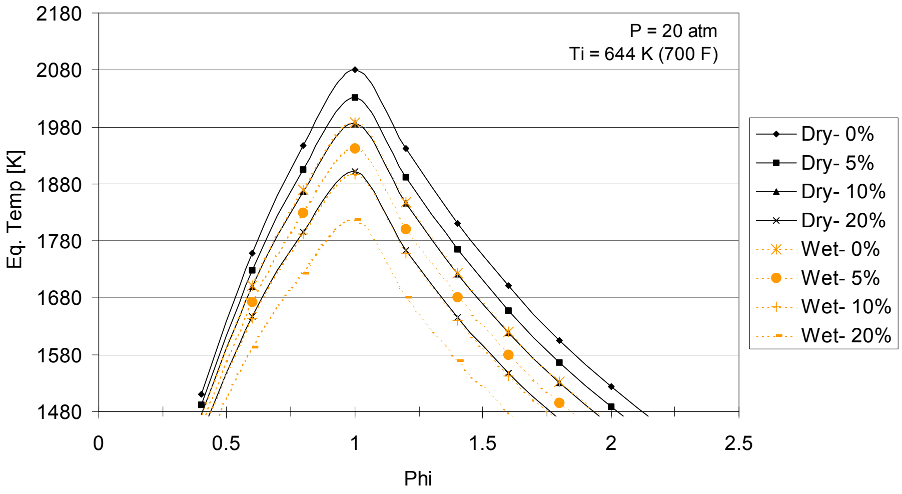

Figure 8: Equilibrium Temperatures vs. Fuel Composition and Equivalence Ratio

As shown in Figure 9, the dry gasifier fuel blend profile is within a factor of two of the corresponding natural gas profile. In contrast, the profile associated with the wet gasifier fuel blend is approximately 15–20 times lower than that of natural gas. These results suggest that minimal hardware modification may be necessary in converting the turbine from natural gas to a wide range of dry blend operation, while more extensive modification may be necessary for wet blend operation. Also shown in the figure is the reasonable result that the manner in which steam is added to the combustor (i.e., either injected into the combustor during dry blend operation or supplied via wet blend consumption) is immaterial to overall combustor extinction times.

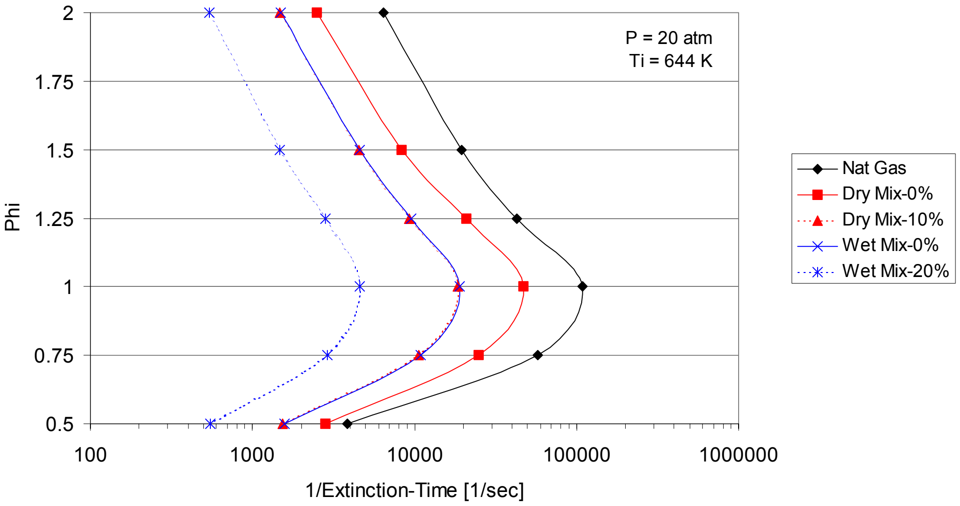

Figure 9: Inverse Extinction Time vs. Fuel Composition and Equivalence Ratio

BIOMASS FEEDSTOCK EVALUATION

Introduction

Biomass already accounts for 3.3% of the energy consumption in the US (biomass sources include wood, wood waste, peat, wood sludge, municipal solid waste, agricultural waste, straw, tires, landfill gases, fish oils, and/or other waste) [4]. Recent actions such as the Kyoto Protocol and implementation of a 20% renewable portfolio standard will drive increased biomass utilization. Many studies [4,5,6,7] estimate future biofeedstock availability and power generation capability will increase by 50% before 2020. However, increased biomass availability and utilization depend on feedstock price. Estimates are that feedstock cost will need to approach $5/MBtu in order to increase biomass availability by 50% by 2020 [6]. This high cost of obtaining biofeedstock will be a considerable impediment to wide-scale implementation of biomass-based power generation given the much lower fuel costs for coal or natural gas-fired plants which are likely for the next 20 yrs.

Faced with a no-growth projection due to biofeedstock cost, the teams's analysis focuses on the northeast region of the US and, in particular, on the favorable economics associated with the use of municipal solid waste. This fuel source is indigenous, sustainable, typically greater than 70% biomass and can be obtained at zero to negative cost in sufficient quantity to fuel moderate-scale (under 100MW) power systems. Currently 102 WTE facilities are in operation in the US, however those plants only handle 14% of the available MSW [7]. Upon closer inspection this feedstock has several unique properties:

1. MSW grows with the population (and therefore with electrical demand)
2. MSW is located near the load source (existing T&D lines can be used)
3. MSW is an excellent gasifier material
    a. fairly homogenous once processed
    b. consistent heating value (HHV ~8000 Btu/lb, dry)
4. MSW usage mitigates major environmental problems
    a. Reduces trash volume by 90%
    b. Prevents release of CH_4 and CO_2 that would occur in landfills
    c. Supports recycling efforts
5. MSW is available at zero to negative cost (nobody wants it)
6. MSW is garbage in and clean power out

Team member CRRA processes 2200 tons/day of MSW into RDF for use in steam boilers at its Mid-Connecticut facility (and handles an additional 3400 tons/day of recycling and solid waste at three other facilities in Connecticut) [8]. For over a decade, CRRA has been able to process MSW to RDF while producing over 65 MW_e at its Mid-Connecticut facility [9]. As shown in Figure 10, RDF has tremendous potential as a biofeedstock for localized power plants.

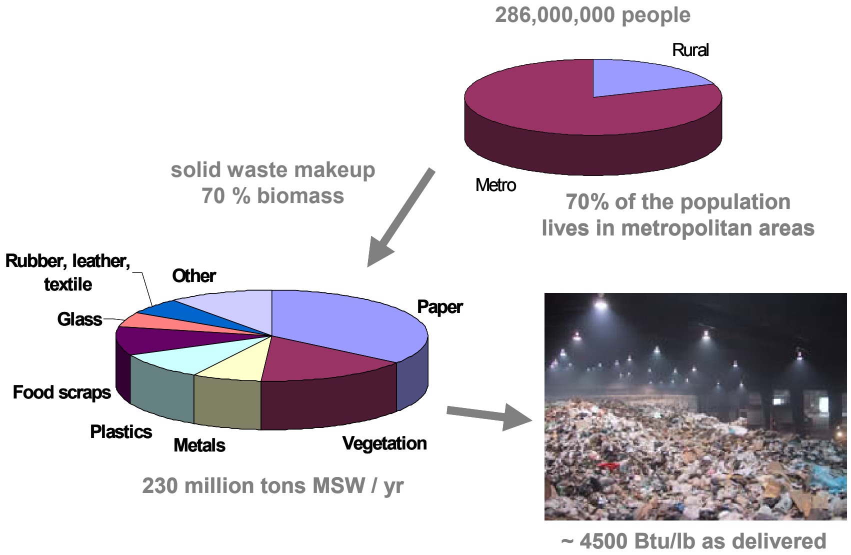

Figure 10: Availability and Quantity of MSW/RDF in the US

Wood waste is also available at negative cost. A recent CRRA report on waste wood availability in Connecticut [10] is summarized in Table 5. Although these estimates indicate ~$7x10^{13}$ Btu/yr[1] can be obtained at zero cost from wood in the Northeast, the actual amount of wood that can be obtained is likely less due to logistics and the fact that wood waste is, in some cases, desirable. Except for localized regions, wood is not sustainable at zero cost in sufficient quantity to be considered a primary biofeedstock for a moderate sized plant.

Table 5: Waste Wood Availability in Connecticut and Northeast

Waste Wood Category	**Description**	**Total Tons / yr (Connecticut)**	**EstimatedTons / yr, Zero Tip Fee (Connecticut)**
Forest Management Residues	Chips	210,000	50,000 to 75,000
Sawmill Residues	Chips, Bark, Sawdust, Scrap	300,000	Negligible
Pallet & Wood Product Manufacturers	Scraps, Sawdust	63,000	10,000
Tree/Utility Services	Chips, Round Wood	25,000 to 50,000	Negligible
C&D Wood	Mixed Wood	260,000	170,000
Land-Clearing Wood	Wood Chips	300,000	150,000 to 200,000
Totals		1,158,000 to 1,183,000	380,000 to 455,000

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[1] From reference [5] the available forest and mill waste in the Northeast totals ~13.7M tons/yr. Estimating 35% availability at zero tip fee and 7700 Btu/lb, ~$7x10^{13}$ Btu/yr is available from waste wood in Northeast

Municipal Solid Waste Availability

Data for the nationwide generation, recovery, and disposal of MSW were obtained from the U.S. Environmental Protection Agency [11] and *Biocycle* [12], an organics composting and recycling journal, and are presented for the year 2000 in Table 6.

Table 6: MSW Generation, Recovery, and Disposal Rates for the U.S. in 2000

	EPA		**Biocycle**	
	(Mton/yr)	**%**	**(Mton/yr)**	**%**
Generated	231.9	100	409.0	100
Recovered[1]	69.9	30.1	130.5	31.9
Incinerated	33.7	14.5	28.2	6.9
Landfilled	128.3	55.3	250.3	61.2

[1] Includes materials recycled and composted.

The variation in data can be attributed principally in the methods of data estimation. The EPA figures are generated using the material flows method, i.e., a mass balance approach that takes into account the quantities of physical goods (food, clothing, appliances, etc.) purchased. These purchased goods are the precursors of the generated waste. Corrections are made based on imports and exports and assumed life of a product. Data sources include industry and business (including their representative associations), other governmental agencies, and surveys performed by industry, government, or the press. MSW for EPA purposes includes ". . . those materials from municipal sources sent to municipal landfills." Construction and demolition (C&D) residue is not included in the MSW stream. Municipal sources are considered to include homes, institutions (schools, prisons), commercial (small business, offices, restaurants) and, to a limited extent, industry.

The *Biocycle* "State of Garbage" report [12], conducted yearly for the past 13 years, relies on questionnaires sent to solid waste management and recycling officials in all 50 states and the District of Columbia. Participation is high with all entities except Montana represented in the current survey. Data gleaned include MSW generation, recycling, incineration, and landfilling rates. Sources and types of waste counted as MSW are similar to the EPA approach with several notable inclusions in the *Biocycle* data: C&D debris (29 states), industrial waste (24 states), and agricultural waste (14 states). The contribution from each of these three categories to the total MSW generated is not ascertainable within the *Biocycle* data. However, using factors developed by Wiltsee [13], it is suggested here that C&D debris could contribute approximately 100 million of the 177 million tons/yr variation between EPA and *Biocycle* MSW generation values. This is based on a 0.076-ton/yr/person C&D wood generation value, an assumed wood content of 20 wt % in C&D debris, and a U.S. population of 275 million residents. As such, the *Biocycle* and EPA data appear to largely support each other. Between approaches, there is also reasonably good agreement concerning the quantity of MSW incinerated.

Using the more conservative EPA numbers for landfilled MSW, an average nationwide factor (0.467 ton/yr/person) was used to estimate the quantity of MSW available within 38 metropolitan areas of the United States with populations over 1 million people. It was assumed that MSW currently incinerated would not be available and only MSW going to landfill would be ascertainable as a gasification feedstock. The results are presented in Table 7. Further, by assuming a conservative heating value of 4500 Btu/lb for the MSW, the total megawatts (electrical) of generation capacity was estimated for each city based on an MSW gasification combined-cycle system operating at 45% efficiency.

The estimates show the available MSW to range from approximately half-million ton/yr (Hartford, Connecticut) to 7 million ton/yr (New York) with new electrical generation capacity ranging from 92 to 1200 MW, respectively. Consequently, a metropolitan area of 1 million people would provide sufficient MSW for a BIGCC system as proposed in this study. Further, the total capacity added nationwide would be 9925 MW based on MSW contributions from the approximately 45% of the U.S. population represented in Table 7.

Table 7: Ranking of Metropolitan United States by Population of Cities Larger than Hartford

City	State	Population [2] (millions)	MSW (tons/yr)	Total MWe (CC @ 45% eff.)
New York	NY	15.000	7,005,000	1201
Los Angeles	CA	13.000	6,071,000	1041
Chicago	IL	8.008	3,739,736	641
Philadelphia	PA	4.950	2,311,650	396
Dallas–Fort Worth	TX	4.910	2,292,970	393
Washington DC		4.740	2,213,580	379
Detroit	MI	4.475	2,089,825	358
San Francisco–Oakland	CA	4.035	1,884,345	323
Houston	TX	4.011	1,873,137	321
Atlanta	GA	3.857	1,801,219	309
Miami–Fort Lauderdale	FL	3.711	1,733,037	297
Boston	MA	3.297	1,539,699	264
Seattle-Tacoma	WA	3.260	1,522,420	261
Phoenix–Mesa	AZ	3.014	1,407,538	241
Minneapolis–St. Paul	MN	2.872	1,341,224	230
San Diego	CA	2.821	1,317,407	226
St. Louis	MO	2.569	1,199,723	206
Baltimore	MD	2.491	1,163,297	199
Pittsburgh	PA	2.331	1,088,577	187
Tampa–St. Petersburgh	FL	2.278	1,063,826	182
Cleveland	OH	2.221	1,037,207	178
Denver	CO	1.979	924,193	158
Portland–Vancouver	OR	1.846	862,082	148
Kansas City	MO	1.756	820,052	141
San Jose	CA	1.647	769,149	132
Cincinnati	OH	1.628	760,276	130
Sacramento	CA	1.585	740,195	127
San Antonio	TX	1.565	730,855	125
Norfolk–Virginia Beach	VA	1.563	729,921	125
Indianapolis	IN	1.537	717,779	123
Orlando	FL	1.535	716,845	123
Columbus	OH	1.489	695,363	119
Milwaukee	WI	1.462	682,754	117
Charlotte–Gastonia	NC	1.417	661,739	113
Las Vegas	NV	1.381	644,927	111
New Orleans	LA	1.305	609,435	104
Salt Lake–Ogden	UT	1.275	595,425	102
Hartford	CT	1.147	535,649	92
Total		**123.968**	**57,893,056**	**9925**

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[2] population of urbanized area

Refuse-derived Fuel Characterization

Physical Composition

An 18.2 kg (40 lb) sample of RDF, obtained from the CRRA Mid-Connecticut facility, was subjected to nondestructive physical analysis testing. Firstly, the bulk density was determined at several compaction levels, including as-received, loose, and spill. The as-received density was determined from the mass and volume of the RDF within the original hand-stuffed plastic package. A loose density was determined for the RDF after it was removed from the plastic package and then allowed to attain an expanded volume within a 0.2 m^3 (55 gal.) barrel. For the as-received and loose density determinations, the volumes were calculated from tape measurements of container height and diameter. A spill density was determined by pouring the RDF into an 0.0283 m^3 (1 ft^3) aluminum box typically used for measuring coal bulk density. The calculated values for bulk density kg/ m^3 (lb/ft^3) were:

As-received (compacted):	214.6 (13.4)
Loose (expanded):	124.9 (7.8)
Spill:	86.7 to 94.5 (5.6 to 5.9)

The RDF from the second determination of spill density was subjected to manual sorting to determine the primary constituents. The results of manually sorting the 2.32 kg (5.9lb) RDF sample are shown in Table 8. Almost 95 wt% of the RDF is combustible, with approximately 87 wt% of the RDF fraction comprising paper, paperboard, cardboard, and plastic film. Although no additional separation of this fraction was performed, visual analysis showed that the plastic film constituted a significant portion of the RDF by volume. Minor combustible fractions included wood and various forms/densities of plastic fragments (beverage containers and caps, toys, utensils). The principal noncombustible components were glass (2.0 wt%) and grit (1.8 wt%). Pictures of the RDF fractions are shown in Appendix B.

Table 8: Composition of CRRA RDF Based on Manual Sorting

Fraction	**wt%**
Paper, Paperboard, Cardboard, Plastic Film*	87.03
Wood	2.95
Glass	2.08
Plastic Beverage Container	2.01
Grit <10-mesh	1.76
Dense Plastic	1.65
Light Plastic	1.05
Aluminum	0.94
Ferrous	0.54

* Grocery and garbage bag-type plastic.

Chemical Composition

The combustible fractions (paper, paperboard, cardboard, plastic film, wood, grit, and beverage containers, dense, and light plastics) were combined and allowed to air-dry. The air-dried combustible RDF fraction was shredded using a laboratory-scale Nelmor knife shredder equipped with a 1/8-inch-diameter-round opening material retention screen. A representative sample of the shredded RDF was subjected to proximate, ultimate, heating value, and ash x-ray fluorescence analysis (XRFA). The results of this analysis (corrected to reflect the impact of the noncombustible fraction) are compared to those reported by CRRA in Table 9 indicating consistent values. Results of the ash XRFA are presented in Table 10; no attempt was made to correct this analysis for the noncombustible fraction.

The RDF is characterized by low sulfur content (0.35 wt%) and high volatile matter content (81 wt%), the latter value indicating the potential for high carbon conversion efficiency. The anticipated high reactivity of the RDF could allow gasification at low enough temperatures to minimize any impact from the high ash content (14 wt%) and high sodium content (12 wt% as oxide in ash). Variation in composition of the as-received RDF reported by CRRA is shown for a 1-year period in Table 11. Surprisingly, the variation is relatively small, resulting in a fuel Btu value that varies less than ±10% month-to-month over more than 4 years of data [14]. From the physical and chemical analysis, the homogeneity and viability of RDF as a biofeedstock is evident.

Table 9: Properties of RDF (sample from CRRA Mid-CT Facility)

	CRRA data (typical)		**UND/EERC Analysis**
	Average (dry basis)	**Standard Deviation**	**Average (dry basis)**
HHV (Btu/lb)	8191	347	8230
Volatile Matter (%)	72.41	1.46	81.2
Fixed Carbon (%)	11.18	0.89	4.8
Ash (%)	16.42	1.67	14.0
Carbon (%)	47.68	1.75	40.5
Oxygen (%)	28.91	1.73	39.1
Hydrogen (%)	5.79	0.49	5.6
Nitrogen (%)	0.42	0.06	0.43
Chlorine (%)	0.54	0.14	-
Sulfur (%)	0.26	0.05	0.35
Moisture (as-fired %)	21.54	2.33	30

Table 10: Ash XRFA for CRRA RDF (wt%)

	As Oxide	**Elemental**
Silicon	32.8	25.8
Aluminum	13.1	11.6
Iron	3.6	4.2
Titanium	2.8	2.8
Phosphorus	1.8	1.3
Calcium	26.5	31.8
Magnesium	2.6	2.7
Sodium	11.9	14.8
Potassium	2.1	2.9
Sulfur	3.0	2.0

Table 11: Composition Variation (as-received) of RDF (data from CRRA Mid-CT Facility)

Month (in Yr 2000)	Moisture (%)	Ash (%)	Sulfur (%)	Btu/lb
January	30.46	8.53	0.19	5052
February	30.47	7.94	0.19	5563
March	29.67	10.31	0.24	5634
April	30.26	9.22	0.16	5019
May	30.91	10.79	0.23	5246
June	30.84	8.98	0.17	5826
July	30.01	9.02	0.18	5086
August	34.82	9.06	0.17	5137
September	30.15	11.36	0.16	5307
October	29.87	8.96	0.22	4852
November	29.28	9.32	0.15	5533
December	31.80	6.16	0.12	5662
Average (over 55 mo.)	30.45	9.06	0.18	5585
Standard deviation (over 55 mo.)	2.50	1.14	0.03	382

RDF Preparation and Feed System

One of the major subsystems that will require further development is the one that prepares and feeds the RDF to the gasifier. Because of its importance, a significant portion of the study effort was expended to identify the components and potential vendors of equipment that would successfully handle RDF with all its variables. The fuel preparation and feed system design is based on the use of RDF from the CRRA Mid-Connecticut facility [8].

Estimation of RDF Processing Rate

To facilitate sizing and eventual capital cost estimation of the RDF processing and high pressure feeding equipment, the mass rate of RDF was estimated. For purposes of calculation, the following assumptions were made:

As-fired RDF heating value	6000 Btu/lb
Hot-gas efficiency	95%
Gas turbine output	50 MW
Simple cycle efficiency	38%

The calculated mass feed rate was rounded to 80,000 lb/hr (40 short tons/hr), and this value was used as the throughput of all major unit operations. At the time of mass rate estimation, no mass loss estimates were available for potential cleaning, size reduction, or drying stages.

Major Unit Operations

Based on visual inspection of the CRRA RDF and assumed feedstock properties necessary for feeding to and entrainment within the ATG, the following major unit operations were incorporated into the proposed feed system design:

1. Nonferrous removal – to principally remove and recover aluminum. Recovery of aluminum, which is present in high concentrations (1 wt%), would provide an added revenue stream as well as reduce any operational problems that could be associated with this low-melting-temperature material.

2. Gravity separation – to remove glass, ceramic, rock, and ferrous items that contribute to wear of downstream unit operations such as mills, rotating equipment, and high-pressure feeders. Further, glass is a low-melting material that could result in agglomeration of circulating bed material.

3. Size reduction – to reduce the material from a nominal size of minus 4 inch to minus 2 inch to improve utilization within rotating equipment such as screws and to improve entrainability within the ATG.

4. Thermal drying – to reduce moisture content of the RDF from a nominal 30 wt% to an as-fired value of 12 wt%. Reducing moisture content external to the ATG will result in an improvement in cycle efficiency.

5. High-pressure feeding – to move RDF at a controlled rate from ambient pressure to ATG pressure (estimated to be between 300 and 350 psig).

6. Other unit operations that will be required are conveyors for transporting RDF between major processing steps, buffering/metering bins, and a system for measuring mass flow rate or providing totalized mass.

Vendor Discussions

Vendors have been approached for each of the major unit operations. For each vendor, the following specifications were provided:

- RDF properties
  - RDF processing rate: 40 tons/hour
  - Primary constituents: paper, cardboard, plastic film (such as that from grocery store bags)
  - Minor constituents: glass (2%), wood (3%), dense plastic (3.5%), grit (2%), aluminum (1%), ferrous metal (1%)
  - Moisture content: 30wt%
  - Input size: minus 4 inch
  - Spill bulk density: 6 to 8 lb/ft

- Information desired
  - Cost per system (including controls)
  - Power requirements
  - Annual maintenance cost
  - Frequency of major repair
  - Turnaround time for major repair

In addition, drying system vendors required submission of a detailed questionnaire. The major unit operations and the respective vendors contacted were:

1. Nonferrous Removal
   - Eriez Magnetics – marketer of eddy current separation systems

2. Gravity Separation
   - General Kinematics – marketer of Gravity Destoner with Single Air Knife
   - Forsberg, Inc. – marketer of Float-Air Destoner and Air Classifer
   - Karl W. Schmidt & Associates, Inc. – marketer of Air Classifier Vacuum Separator

3. Size Reduction
   - American Pulverizer Company
   - Marathon Equipment Company
   - Williams Patent Crusher & Pulverizer Co.

4. Thermal Drying
   - Heyl & Patterson, Renneburg Division
   - Barr-Rosin, Inc.

5. High-Pressure Feeding
   - Metso Corporation – marketer of plug screw feeders (formerly Sunds Defibrator) for continuous thermochemical wood pulping
   - Stake Technology Ltd. – marketer of CO-AX Feeder for thermomechnical wood pulping
   - Fortum – technology rights holder for “Piston Feeder for Solid Fuels”; developed by company formerly known as Imatran Voima Oy

6. Buffer Metering Bin
   - Keith – walking-floor bin
   - HALLCO – walking-floor bin
   - SITA – Sverige AB-BRINI® metering bin

7. Conveyors
   - Williams Patent Crusher and Conveyor – mill-in feed conveyor

Estimates for capital and annual maintenance costs as well as power and other utility (e.g., Btus for fuel drying) are presented in Table 12.

Table 12: Cost and Utility Estimates for Major Unit Operations in RDF Feed System.

Unit Operation	Capital Cost ($1000s)	Maintenance Cost ($1000s)	Power hp (kW)	Btu/hr
Non-Ferrous Removal	300	10–20	(50)	NA*
Gravity Separation	92–300	4–5	106	NA
Size Reduction	782–1680	53–239	300–2400	NA
Thermal Drying	1480	74	350	30 MM
High-pressure Feeding	3760–6600	82–330	1000	NA
Conveyors/Metering Bins	1000-1200	4–5	100	

* Not applicable

The range of costs presented for size reduction reflect the differences in proposed systems for attaining the desired product consistency. This stems from a near nonexistent knowledge base within North America for reducing RDF to a size less than 10 cm (4 in). Both slow-speed shearing and high-speed hammer-milling have been specified; determination of the proper methodology will be required in subsequent phases of testing.

Similarly, the definition of a suitable method of “high-pressure feeding” will only be attained by subsequent testing. The EERC feels that feed systems that have good probability of success with the ATG are 1) those that form a pressure sealing plug to move the RDF across the pressure boundary and 2) piston systems that rely on ring seals on the piston to maintain a pressure seal. Other approaches that have been considered and the reasons for not pursuing these options include the following:

- Lock hoppers – Perform best with dry, free-flowing material that will not easily bridge across valves, chutes, transfer points, and other openings. Systems for biomass have had some success in alleviating this by using diverging lock hoppers with bodies the same size as lock valves. This however, results in very large systems for feeding sufficient quantities of the low-density, low-heating value biomass. Further, densification by cubing or pelleting, in the instance of RDF, may be required to assist product flow; this fuel form would be compatible with only fixed or bubbling-bed systems. Also, lock hoppers require large quantities of pressurization gas and sophisticated systems for lock-gas recovery.

- Rotary valves – Have limited pressure differential service (approximately 10 bar (150 psig)) and are prone to short-lived pressure sealing capability. Further, bridging of fuel is possible especially with stringy, sticky materials.

Cleaning RDF with Commercial Air Classifer

A sample of the CRRA RDF was brought to Forsberg, Inc., to evaluate the Float-Air Destoner for removing glass, aluminum, and other heavy material. After visual inspection of the RDF, Forsberg personnel determined that the Destoner is better suited for more granular, free-flowing material. Forsberg proposed testing the sample in an air-classifier that is typically used to clean agricultural products such as corn, sunflowers, beans, and wheat. The air-classifier appeared to be very suited for application with RDF, as the first test was successful at high separation of undesirables and high product recovery.

The yield of product was measured at 87.4 with 78 wt% of the noncombustibles (glass, ferrous, aluminum, etc.) and only 7.6 wt% of the combustible fraction passing with the rejects based on manual sorting of the product and reject fractions. After sorting, the combustible fraction was sorted and subjected to determination of moisture and ash. From these data, the ash reduction was estimated to be approximately 35 wt%.

Although marginally effective with aluminum (25 wt% removal), the air-classifier removed 80 wt% of metal and over 99 wt% of the glass as determined by hand-sorting of the product and reject fractions. Further, the air-classifier appeared to significantly reduce the quantity of loosely adhered grit (sand, fine glass). Digital photos of the air-classifier system and product and reject fractions from the testing are presented in Appendix B.

PROJECTED PLANT EMISSIONS AND CONTROL

Gasification is generally used to convert a "dirty" fuel resource into a form that can more easily be cleaned of the various contaminants. The ATG has several operating characteristics that reduce the cleanup task for the major contaminants. The emission of trace elements, however, remains as an area requiring additional R&D. This area is deemed an important enough consideration that a separate discussion of recent efforts has been included as Appendix C.

Particulate Emissions and Control

Particulate emissions will be controlled utilizing a hot-gas filter system with either ceramic or metallic candle filters. It is assumed that these filters will be backed up with safeguard devices that will prevent the release of catastrophic amounts of dust to the gas turbine and the environment in the event of candle filter failures. Both the EERC and the Southern Company Services (SCS) Power Systems Development Facility (PSDF) have demonstrated the successful operation of hot-gas filters operating in the temperature range from 370 to 540°C (700 to over 1000°F). Dust emissions have been less than 1 ppmw at face velocities around 1.2 to 1.37 m/min (4 to 4.5 ft/min). Vapor phase alkali have been shown to be below acceptable gas turbine inlet conditions when operating the filter system in this temperature range. The dust emissions from these systems are more than ten times less than those allowed by current NSPS standards. No further particulate cleanup is expected to be needed in this system.

Sulfur Dioxide Emissions

Current testing with the coal fuels on the ATG has shown that 90% or greater sulfur retention can be achieved in the circulating bed material by the addition of a calcium-based sorbent such as limestone or dolomite to the bed at a Ca/S ratio of 1.5 to 2.0 on a molar basis. Since the selected RDF fuel is low in sulfur (0.35 wt%), sulfur dioxide emissions would initially be relatively low. Assuming 85% sulfur removal in the bed material would further reduce the final sulfur dioxide emissions to less than 0.05 lb/MBtu. Addition of limestone to the bed material will also help reduce the potential for ash agglomeration problems from the high alkali in the ash by providing more surface area in the bed material for interaction with the alkali species. Sulfur dioxide emissions from the RDF-fueled ATG system are not expected to present a serious emissions issue. Further sulfur removal could be accomplished using the zinc-based hot-gas desulfurization sorbents currently being developed, however, this technology is not being proposed here, due to low projected emission levels already and to the higher removal costs associated with a smaller scale removal system.

Halide Emissions

The addition of limestone to the system will also have the effect of reducing the amount of chlorine and fluorine that will be emitted from the ATG system. The utilization of calcium-based sorbents either by injection into flue gas or as a packed bed in a fuel gas has been shown to remove up to approximately 90% of the HCl and up to 99% of the HF [15]. These calcium-based sorbents remove the halides by the formation of solid $CaCl_2$ and CaF_2 via the reaction:

$$CaO + 2HCl\ (\text{or } 2HF) \leftrightarrows CaCl_2\ (\text{or } CaF_2) + H_2O \qquad (1)$$

The EERC has successfully utilized CaO as a chlorine guard bed for the removal of chlorine from the thermal decomposition and pyrolysis of chlorine containing plastics. This testing removed greater than 90% of the chlorine in a fixed bed of CaO at temperatures between 450 to 650°C (840 to 1200°F) [16,17]. SRI International performed some research with low cost nahcolite which reduced the HCl to less than 1 ppm at temperatures from 400 to 650°C (840 to 1200°F). Other sodium and potassium-based sorbents such as a commercial Katalco Chloride Guard 59-3, and three natural carbonate minerals, shortite, dawsonitie, and nahcolite [18]. While all of these worked very well, only the nahcolite was felt to be economically feasible.

Nitrogen Oxide Emissions

Previous results have shown that hot or warm fuel gas mixtures are combustible in lean pre-mixed type burners typically found in gas turbines, however, the conversion of the fuel-bound nitrogen (mostly ammonia) can result in NO_x emissions significantly higher than desired. From combustion modeling performed on both air-blown and oxygen-blown gasifier operation on lignites, the NO_x formed via the thermal formation mechanism is very low at less than 25 ppmv at 15% O_2 (0.023 lb/MBtu of fuel gas). However the NO_x formed via the chemical conversion of nitrogen containing species such as ammonia and HCN is high at approximately 330 ppmv at 15% O_2 (2.06 lb/MBtu of fuel gas) for air-blown operation (NO_x formed for oxygen-blown gasification is even higher at approximately 475 ppmv at 15% O_2 (2.60 lb/MBtu of fuel gas)) [19]. These calculations assume that 100% of the ammonia is converted to NO_x in the dry low-NO_x (DLN) burner. Actual conversion rates, however, were expected to be between 20 to 50% for this type of combustor. Correction of these NO_x emission rates to this level of NH_3 conversion to NO_x would result in actual NO_x emissions of 0.64 to 1.05 lb/MBtu for air-blown operation (0.80 to 1.32 lb/MBtu for oxygen-blown operation), which would be too high under current regulations.

Utilization of cold gas cleanup with a water scrubber would alleviate the ammonia conversion issue since the water scrubbers can remove all of the ammonia. However, this could reduce the overall cycle efficiency and it would generate an additional waste water treatment problem.

Installing SCR technology in the heat recovery steam generator and dealing with the ammonium bisulfate deposition that will occur with the reaction from the ammonia slip and the sulfur dioxide in the flue gas is being considered as an option by SCS. Cost of electricity increases on the order of 0.15¢/kWh have been projected for the use of SCR on a 150 MW gas turbine power plant.

One technology to reduce NO_x emissions is the use of rich-quench-lean (RQL) type combustors which were being developed by various gas turbine manufacturers in the past to reduce the amount of ammonia converted during combustion. These RQL combustors have been shown to substantially reduce the amount of ammonia converted to NO_x. The Siemens-Westinghouse Multi-Annular Swirl Burner (MASB) tests performed at the University of Tennessee Space Institute in Tullahoma, TN reported the conversion of the fuel gas ammonia to NO_x at approximately 4% [20]. GE also reported on a small RQL combustor that was tested on hot syngas that achieved less than 500 ppmv (on a dry, 15% O_2 basis) which corresponded with 5% conversion of ammonia to NO_x [21]. However, these gas turbine manufacturers have not commercialized these combustors and are not currently pursuing the further development of these combustors.

Another way to control NO_x emissions is to remove the ammonia from the hot or warm fuel gas (without the need for cold-gas water scrubbers) by the use of ammonia decomposition catalysts such as nickel, ruthenium, or iron-based catalyst. This concept has been investigated by a number of other researchers [22, 23, 24, 25]. One study, conducted with nickel and ruthenium-based catalysts at temperatures between 700 and 900°C (1290 to 1650°F) [22], showed the greatest ammonia decomposition (>85%) at the highest temperature of 900°C (1650°F). An approach of 99% of chemical equilibrium was possible with a residence time > 1 second. The catalysts were deactivated over time in the presence of H_2S but could be easily regenerated. Another study described the development of a mixed metal oxide sorbent, a zinc-based material containing 5 wt% of Ni, Co, and Mo. It was capable of the simultaneous removal of NH_3 and H_2S at temperatures from 500 to 700°C (930 to 1290°F) [23, 24]. This sorbent/catalyst was shown to be effective at removing H_2S to less than 20 ppmv and at decomposing greater 90% of the NH_3 . Decomposition of the ammonia improved with increasing temperature and increasing pressure. Deactivation of the ammonia catalyst material was more rapid than the H_2S saturation of the Zn-based sorbent so the regeneration cycle time would be dictated by the ammonia decomposition kinetics; however, the HART-49 material seemed to regenerate very well over thirty cycles. The third study conducted ammonia decomposition tests in the presence of a RA-330 honeycomb catalysts in the temperature range of 650 to 700°C (1200-1300°F) and 870 to 980°C (1600-1800°F) and with a alumino-silicate catalyst in the 425 to 540°C (800 to 1000°F) temperature range [25]. Ammonia decomposition exceed 99% at temperatures above 1300°F, but in the presence of hydrogen the conversion drops to approximately 85% at 870°C (1600°F). A catalytic removal process was also discussed which involved the injection of some air or oxygen into a alumino-silicate honeycomb catalyst that could remove 80 to 95% of the ammonia at temperatures from 450 to 540°C (850 to 1000°F). It was unclear how the oxygen was reacting with the ammonia in the fuel gas mixture at these conditions.

Further work is needed to determine the best method for reducing the expected NO_x emissions from the conversion of the ammonia in the hot fuel gas.

Mercury and Trace Element Emissions

Gasification facilities pose a challenge for both measuring and controlling mercury and trace element emissions. This area is an important issue with all types of feedstocks, not only biomass, and is discussed in greater detail in Appendix C. The gas is under high pressure and may be at high temperature. In addition the flue gas being sampled contains high concentrations of reducing gases such as H_2, CO, and CH_4. Very little work has been done to evaluate traditional wet-chemistry methods (EPA Method 29, EPA Method 101A, and the Ontario Hydro mercury speciation method) for measuring mercury under gasification conditions. In addition, the potential long-term interferences that may be present when using mercury continuous emission monitors (CEM's) are not known for certain. The two most common analytical methods for measuring mercury in mercury CEM's are cold-vapor atomic adsorption spectroscopy (CVAAS) and atomic fluorescence spectroscopy (AFS). In both cases, some form of pretreatment system is necessary to provide a gas to the analyzers such that the mercury can be measured accurately.

Except for recently reported mercury removal results from the Eastman Chemical Co. plant in Kingsport, Tennessee, few mercury control capabilities for gasification technology have been reported to date. The Eastman plant utilized presulfided activated carbon beds at 62 bar (900 psig) and 30°C (80°F) to capture 90% to 95% of the mercury from a syngas mixture [26]. The process added about $10/kW of capital cost. However, the demonstration of acceptable mercury control performance at higher temperature syn-gas conditions is needed if gasification technology is going to achieve the higher efficiencies attractive to the power generation market. The process efficiency penalties associated with cold cleanup make it imperative that mercury control technologies be developed that would allow for mercury control at temperatures higher than the moisture dew point. This warm-gas cleanup would allow for mercury control while keeping the extra mass flow and the latent heat of vaporization associated with the gasifier steam and vaporized fuel moisture.

Recently, an EPA information collection request (ICR) for mercury required sampling of operating IGCC systems in the United States. Two plants were sampled with the following results: 1) The Polk Power Plant (Texaco gasifier in Florida) reported nearly all the mercury emissions from the plant to be in an elemental form and only had < 40% collection efficiency; 2) The Wabash River facility (Global Gasification Technology) showed almost all of the mercury emissions to be in the elemental form. The level of control was found to be about 50%, based on the level of mercury in the coal [27]. This indicates some capture is taking place in the gasification system; however, the exact location is not known. It is thought that it could be occurring in the amine-based cold-gas cleanup equipment utilized on both systems.

ECONOMIC ANALYSIS

Introduction

With few exceptions, mostly in the wood products and paper industries, the cost of generating power from biomass has not been competitive with purchased power from the grid. Various incentives, generally direct subsidization, have reduced the gap in cost of electricity (COE). The use of RDF in the Biomass Integrated Gasification Combined Cycle (BIGCC), however, can change this scenario. For example, the cost of generating electricity from conventional natural gas-fired combined cycle power systems, currently the most efficient type of power plant, varies over a range of approximately 2¢/kWh to over 10¢/kWh, depending on the size of the power plant and the cost of the natural gas fuel supplied to the power plant. Because of the low capital cost of combined cycles, the COE is most sensitive to the fuel cost. The cost of gas supplied to electric generators has varied widely in the past several years; e.g., in 2000, the national average was \$2.90/MMBtu while in 2001, the national average was \$7.21 with a high of \$16.91 (Virginia). Early in 2002, the average was \$3.39/MMBtu with a high of \$11.71/MMBtu [28].

The following sections describe the costs identified for the BIGCC and the resulting COE. While it must be noted that the economic analysis was only at a first-order granularity, the availability of RDF at zero, or even negative cost, makes this system very attractive. The costs are for a battery limits BIGCC plant using RDF as a feed and electricity as a product.

Capital Cost

The capital cost estimate for the BIGCC plant was prepared primarily by using published budget estimates from the literature and UTC's experience from recent coal-fired and natural gas-fired and gas turbine combined cycle power plant work. Capital costs for the subsystems were estimated using a combination of capacity-factored and equipment-based estimates. Capacity-factored estimates utilize the ratio of the capacity (flow rate, heat duty, etc.) of an existing piece of equipment to the new equipment multiplied by the cost of the existing equipment to estimate the cost of the new equipment. A scale-up factor particular to the equipment type was applied to the capacity ratio. The equipment-based estimates were determined from more detailed equipment design calculations based on the process conditions and results of the simulations. Where necessary, costs were adjusted for scope and capacity to match the BIGCC requirements and corrected to the year 2000 using equipment cost indices.

There are, however, two unique subsystems in BIGCC: the RDF processing and Advanced Transport Gasifier. Each of these was the focus of both a technical and cost investigation.

RDF Processing Costs

An estimate was made based on information from EERC as to the cost of an RDF processing facility that would produce enough RDF to supply an 85-MW BIGCC. That estimate is provided in Table 13. The cost ranges from approximately $74/kW to nearly $119/kW. The lower figure was used as it would be more representative of the nth plant. All of the BIGCC plant costs are for the nth plant; thus there are no special, one-time costs associated with engineering, constructing, and operating the first-of-a-kind facility.

Table 13: RDF Processing Equipment Cost Estimate

Process			
Non-Ferrous Material Removal	$300,000		
Gravity Separation	$92,000	to	$300,000
Size Reduction	$782,000	to	$1,680,000
Thermal Drying	$1,480,000		
High-Pressure Feeding	$3,760,000	to	$6,600,000
Miscellaneous	$1,000,000	to	$1,200,000
Total	**$7,414,000**		

Advanced Transport Gasifier Costs

The ATG has been described in an earlier section. One of the major advantages of this type of gasifier is that it can be built with essentially off-the-shelf material. Because of its relative simplicity, the ATG scales well with little loss of efficiency. For the nth plant BIGCC, a single gasifier and cleanup train has been included in the costing. Costing information for the gasification train components was based on information in [29, 30, 31, 32]. Costs were scaled using tables and curves from [32].

Heat transfer analysis was performed on the ATG configuration shown in Figure 11 to determine the refractory thickness requirements to maintain shell temperatures less than 120 C (250 F) for operation at 980 C (1800 F). Excursions to 1093 C (2000 F) were determined not to be a problem. Two layers of refractory are used, a hard-faced refractory on the inside for protection with a backup insulating layer of refractory providing the majority of insulation to maintain low surface temperatures. It is assumed the maximum operating pressure is 25 bar (350 psig). Based upon these operating conditions, using API 5L grade B carbon steel pipe as a material readily available, it was determined that 20 mm (3/4 inch) thick pipe is marginally adequate. For pricing purposes quotes were obtained for both 20 mm(3/4 inch) thick and 25mm (1 inch) thick pipe. Based upon vendor recommendation, while the 25mm(1 inch) thick pipe is slightly more expensive, it would be their preference for fabrication purposes. The flanges priced and required are 136 kg (300 pound) A105.

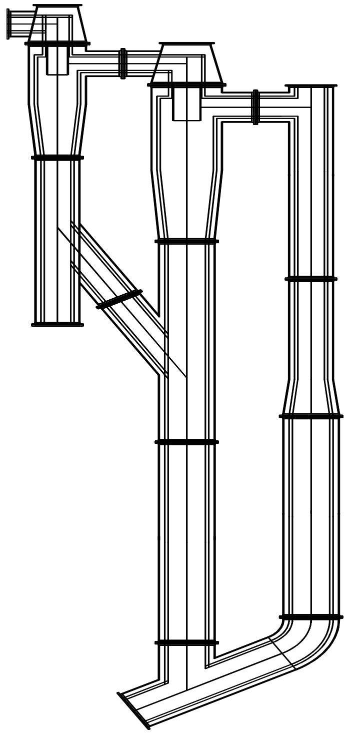

Figure 11: Cross-sectional Schematic of the ATG used to obtain Vendor Quotes

A quote for fabricating the pressure vessel according to the specifications above was obtained from GASPAR, Inc. in Canton, OH. The price quoted for a single train transport reactor designed to supply syngas to the FT-8 TwinPac™ was $305,000 for the 20 mm (3/4 inch) thick wall and $335,000 for the 25mm(1 inch) thick wall vessel. The inclusion of the necessary nozzles for air feed, biomass feed, and steam feed or recycle gas in the J-Leg is projected to cost less than $100,000. The cost of the completed gasifier vessel without refractory should cost less than $500,000. A quote for installing the refractory was obtained from Gagnon Refractory and Insulation Contractors in Hudson WI. Their quote for installing the refractory at their facility was for $513,000. Therefore, a complete gasifier ready for installation would cost approximately $1,000,000 excluding shipping from Canton, OH to Hudson, WI and from Hudson, WI to it final installation point. Sections of the gasifier were kept to 9.1 m (30 feet) or less in length and a maximum vessel diameter of 173 cm (68 inches), however, the maximum wet shipping weight of 36.3 tonnes (40 tons) probably would cause some issues with special shipping utilizing either railcars or special heavy weight load dispersing truck trailers.

A cost associated with the hot gas filter system for the Sierra Pacific Pinon Pine Project was obtained from Seimens/Westinghouse. This would be approximately the size needed for this project. This system cost was approximately $3,000,000 in 1997 dollars for the filter completed with all auxiliary components (such as the backpulse system) delivered on site [33]. This cost does not include on the installation costs. This vessel was approximately 3.05 m (10 feet) in diameter and approximately 15.2 m (50 feet) tall and contained 748 1.5 meter long ceramic candle filters in 16 clusters [34]. This system was designed to filter 136,116 kg/h (300,000 lb/h) of syngas at 543 C (1011 F) and 19 bar (260 psig).

Combined Cycle

The cost of the combined cycle system is based on information contained in [29, 30, 35]. Costs for additional compressor trains for the recycle and boost compressors are estimated from [36].

Balance of Plant

Within the balance of the plant are the water systems; civil, structural, and architectural components; piping; controls and instruments; and electrical systems. Using information developed as part of the High Performance Power Plant study [35], a value of 22% of the Process Plant Cost was used.

Engineering Services

The cost of home office and field engineering and fees are accounted for in this category. A value of 10.0% of the Process Plant Cost is typical.

Contingency

Depending on the level of engineering and technology maturity, a percentage contingency is added to account for costs, which are expected to be defined when more detailed engineering is accomplished. For the first four units, a process contingency of 5% is typical. However, for the nth plant, the process contingency can be assumed to be zero. Likewise for the nth plant, the overall project contingency can be assumed to be 7%.

Process Plant Cost Estimate

The estimate of process plant costs is presented in Table 14.

Table 14: Process Plant Cost Estimate for an 85MW BIGCC Plant

Plant Section Description	**Cost, $K**	**$/kW$_e$**
RDF Preparation	$7414	87
Transport Gasifier	$2800	33
Recycle Gas Compression	$4492	53
Air Boost Compressor	$2455	29
Gas Conditioning	$8350	98
CC GT&ST System	$51,600	607
Ash-Handling System	$1825	21
Char Combustor	$1577	19
	$80,513	947
Balance of Plant	$22,400	264
Process Plant Cost	**$102,913**	**1211**

Additional Capital Cost Items

Items shown in Table 15 are added to the Process Plant Cost Estimate to complete the capital cost.

Table 15: Total Capital Cost Calculations

Plant Section Description	Cost, $K	$/kWe
Engineering Fees	$10,291	121
Process Contingency	$0	0
Project Contingency	$7,204	85
	$17,495	204
Total Plant Cost[1]	**$119,238**	**1417**

[1] Overnight cost.

Annual Owning, Operating, and Maintaining Costs

Annual owning, operating, and maintaining costs for the BIGCC includes capital, fuel, maintenance, labor, and variable O&M (operating and maintenance) costs. The values used for these parameters are displayed in Table 16. The annual owning charges will vary according to the economic assumptions used and the financial structure adopted. The reviewed literature displayed values from approximately 10% to 18%; a range of 12% to 15% was chosen for this analysis. Plant capacity factor was selected to be 85%. The fuel cost used in the literature was also found to vary; the range varied from a revenue of $5.00/MMBtu to an expense of $2.00/MMBtu. CRRA receives revenues of $51/ton for the RDF feed material [9]. This equates to a fuel with a negative fuel cost of approximately $4.50/MMBtu.

The literature equated the annual maintenance cost to a percentage of the Total Plant Investment. The range for this parameter was between 2.0% and 2.2%. At least two references mentioned that there would need to be 14 people per shift; they did, however, differ on their hourly rate [30, 35]. Lastly, the variable cost is primarily comprised of consumables and varies according to the amount of fuel being used.

Table 16: Biomass-Fired System Operating Cost Parameters

Parameter		Low	High
Net Output	MW	85	85
Annual Cost Factor	%	12	15
Capacity Factor	%	85	85
Efficiency	%	45	45
Biomass Fuel Cost	$/MMBtu	(5.00)	2.00
Maintenance	% of total plant investment	2.0	2.2
Labor	Base $/hr	34	50
Personnel/Shift		14	14
Base Pay	$/hr	34	50
Variable O&M	$/MWhr/tons of fuel per year	4.10	4.10

Cost of Electricity and Comparisons

An upper-level economic analysis was performed for this preliminary investigation. The selected approach uses a levelized cost of electricity as a basis of comparison with more conventional power systems. Because absolute costs associated with the equipment, operation, and fueling of the power plant are 1) site specific and 2) not available within a level of confidence within ± 20% for many of the system components, it was decided to develop an envelope that would contain the anticipated range of power plant capital cost, operating and maintenance costs, and fuel costs.

As described above, capital costs were based upon various cost data appearing in the literature and extrapolations of costs for equipment given in a variety of reports. Cost for these components were scaled using standard scaling factors for the mechanical and chemical industries.

The resulting cost is approximately \$1417/kW as displayed in Table 15. This value agrees within 20% with an estimate of BIGCC costs reported in [6]. A range of O&M charges (excluding fuel) from 0.7¢ to 0.9¢/kWhr was identified as probable. A figure of 0.75¢/kWhr was chosen as a midrange value. The cost of fuel delivered to the gasifier site was also varied to include a range of values typical for MSW. As noted, CRRA has a fuel cost closer to the lower end of the fuel range picked for the analysis. An efficiency of 45% (HHV) was assumed for the COE calculations.

For comparative purposes, estimates were made of the COE for two natural gas-fired combined cycle plants. The values of capital cost, O&M costs, and efficiency for these combined cycles plants reflect current operating experience as reported in the literature [6]. For the 500-MW power plant, values of \$510/kW, 0.3¢/kWhr, and 56% (LHV) were chosen. For the 70-MW power plant, values of \$800/kW, 0.4¢/kWhr, and 49% (LHV) were chosen. The fuel cost range was based on the average by state of natural gas price delivered to utilities in the years 2000 and 2001.

The resulting comparison shown in Figure 12 COE estimates produced by both systems reveals that there is indeed a range of biomass/RDF fuel costs for the BIGCC that will produce electric energy costs that are competitive with natural gas-fired combined cycle systems in the United States. The biomass-based systems would be competitive even if the natural gas price were \$2.00/MMBtu (the approximate minimum national average for the years 2000 and 2001.

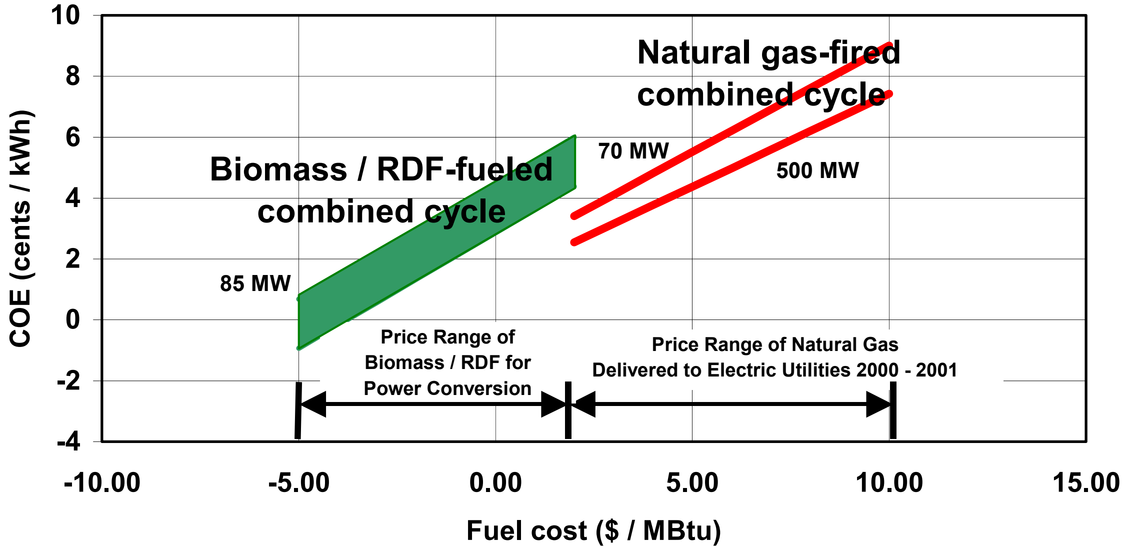

Figure 12: Impact of Fuel Cost on the Cost of Electricity

MARKET ANALYSIS

Introduction

The following section is a brief summary of the factors that could affect the future market for the technology described in the foregoing sections. It is by no means a definitive identification of the market that could develop for this system; that is beyond the scope of the current program. First, an overview of the potential world market is given and some projections for the market in Americas are identified. Then, the changes that are emerging as the power industry readies itself for the 21st century, and that could impact the implementation of the BIGCC technology, are identified.

World Market

According to the Integrated Waste Services Association [37], there are 102 waste-to-energy plants in the U.S. currently serving more than 37 million people in 31 states. These plants burn over 14% of the trash generated nationwide and generate more than 2,800 MW. A recent survey conducted in 70 cities nationwide [38] indicates that almost three-quarters of the Americans polled believe waste-to-energy plants are vital components for the nation's environmental and economic future. Those polled also believe that waste-to-energy programs mean cleaner disposal of trash, less need for landfill space and cost-effective, safe power generation. The U.S. Department of Energy has labeled waste-to-energy technology as a major part of a plan to reduce carbon dioxide emissions in the United States [39]. These are waste incineration plants that generate power by recovering the heat as steam for use in steam turbines. A newer technology, the gasification of biomass to form a clean synthesis gas (syn-gas) offers the opportunity to provide higher value products such as more efficient generation of electricity using gas turbines or production of chemicals and fuels in systems such as the BIGCC.

World wide, the projection [40] is that $11 billion of new gasification and pyrolysis plants are forecasted to be built between 2001 and 2010 specifically for energy production from waste. The increased interest in using biomass to generate renewable energy is the main factor behind this large market. Gasification and pyrolysis processes have distinct advantages over combustion, including potential for higher efficiency; lower emissions; lower costs for facilities; and smaller scale buildings and chimneys; see Table 17 from Ref. 41.

Many different companies, e.g., those listed in Table 18, are developing biomass gasification processes on a worldwide basis. Many of these are designed for specific applications. At this time, however, there are only a few commercial plants in operation around the world. Of the more than 100 facilities operating or ordered worldwide, many of the proprietary systems currently being promoted have only operated as small-scale pilot projects.

Table 17: Advantages of Biomass Gasification

Products/outputs	More flexibility in terms of outputs Can achieve higher levels of materials recycling vs. energy recovery
Image	Better potential public acceptance because of perception that new processes are 'greener', 'high-technology solutions'
Scale	Possibility of using at smaller scale than incineration
Energy efficiency and climate change mitigation	Can integrate with more efficient methods of energy recovery (gas engines, gas turbines, IGCC) as well as utilization (fuel cells, bio-oils)
Economics	Some processes, but not all, are lower cost than incineration Energy may command premium price (green tariff) May qualify for renewable energy grants
Visual impact	Many processes allow smaller chimneys (power gasification vs. heat gasification) Lower profile buildings
Residues	Possibility of reducing residues that require disposal Can lower costs

Table 18: Significant Gasification & Pyrolysis Suppliers For MSW Applications

Supplier	**Origin**	**Process type**	**Status**
Basse Sambre	France/ Belgium	Pyrolysis + gasification + combustion	Demonstrator
Compact Power	UK	Pyrolysis + gasification + combustion	Demonstrator being commissioned
Ebara	UK	Fluidized bed gasification + combustion + ash melting	Fully commercial (in Japan)
Brightstar Environmental	Australia/USA	Pyrolysis + gasification	Demonstrator being commissioned
Enerkem	Canada	Fluid bed gasification	Semi-commercial
GEM	UK	Thermal gasification	Pilot
IET TOPS	USA/UK	Gasification + combustion	Bench-scale/conceptual
Mitsui R21	Japan	Pyrolysis + combustion	Semi-commercial in Japan
Nexus	France	Pyrolysis	Pilot-scale
Nippon Steel	Japan	Gasification + combustion + melting	Fully commercial
Organic Power	Norway	Gasification + combustion	Semi-commercial
PKA	Germany	Pyrolysis + gasification	Fully commercial
Technip	France/ Germany	Thermal gasification	Semi-commercial
Thermoselect	Switzerland	Pyrolysis + gasification	Semi-commercial
Thide/Hitachi	France/Japan	Pyrolysis	Demonstrator in Japan
Von Roll RCP	Switzerland	Pyrolysis + combustion +melting	Demonstrator in Japan

A major factor that will drive the market for this technology is the realization by various governments that there is a need to accelerate growth in the renewable energy sector [42]. Many are offering a number of incentives to stimulate the industry. For example, in the U.K. a fund of £100 million (~$145 million) for capital grants for new projects is being offered and there is a regulatory system that provides a mechanism for premium pricing for green electricity. Policy makers in many other countries are considering similar incentives. Carbon taxes (on methane production from landfill) and landfill taxes are making landfill waste disposal more expensive. Thus, especially in Europe, biomass-to-energy is looking increasingly attractive from an economic perspective. Between 200 and 1000 new biomass gasification and pyrolysis plants could be built around the world over the next ten years. These facilities could reduce the need for fossil fuel power generation by processing 10 - 50 million metric tons (tonnes) of waste to energy. The majority of these plants would be in Europe with the U.S. as the next largest market. A projection of the U.S. market is shown in Figure 13 from Ref. [43].

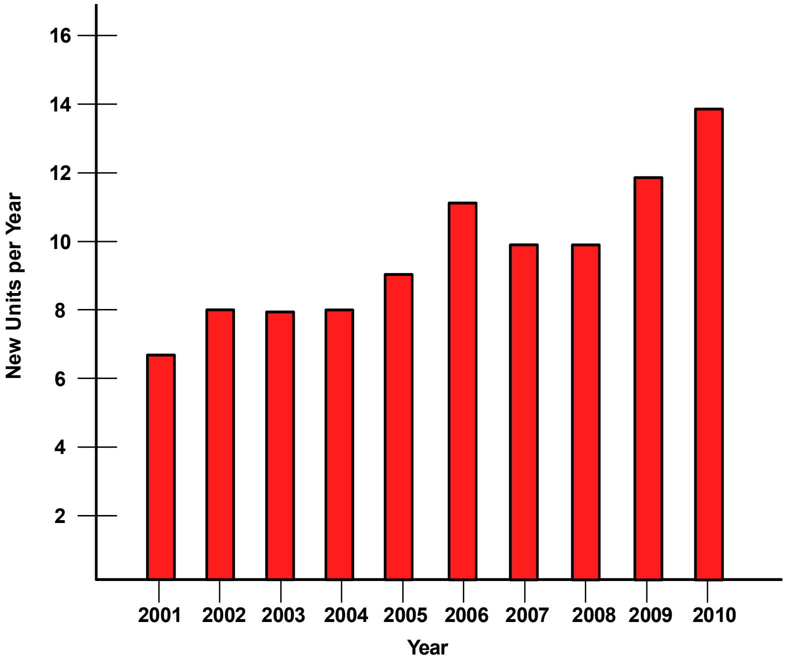

Figure 13: Projected market for Biomass Gasification in the U.S.

<u>New Factors Specific to the U.S. Market</u>

The above market projections do not reflect the advanced technology BIGCC plant described in the previous sections. Nor do they account for the recent changes in the U.S. power industry that will present new opportunities to meet an emerging segment of the electric power market. Changes in power plant ownership and operation brought about by the failed deregulation in several areas such as California and the deteriorating economic status of energy trading companies such as Enron will make small, fuel-flexible power systems especially attractive.

At this point, the electric power industry is undergoing significant changes. The Enron approach of owning little in the way of physical assets and concentrating on energy trading has completely unraveled. In the next several years, the surviving energy companies will move toward acquiring assets allowing them to move toward generation-rich and asset-rich portfolios. In the near term, however, they need a strong balance sheet, and they are selling or deferring assets [44], some on a discount basis. Other stronger companies are purchasing assets at bargain prices. Two types of companies are playing a big role in acquisitions. The first is foreign companies with great balance sheets that are cash-rich, asset-rich, and already have strong international presences. Two examples are Electricite de France (EDF) and Scottish Power. The latter, for example, purchased Sierra Pacific Power (an Oregon utility) in a surprise acquisition. The other major players are regulated utilities such as American Electric Power (AEP), Mid Atlantic, and Florida Power and Light (FP&L), which have moved slowly toward deregulation. California and those who quickly followed California's example were hurt by the California energy crisis. Those who moved more slowly have faired very well; Pacific Gas & Electric (PG&E) is in bankruptcy, while FP&L is quite financially strong.

The need to strengthen balance sheets is aimed at the ratings agencies so that they will give these companies a favorable bond rating (thereby allowing them to borrow money with an effective interest rate). This will lead to new entities as opposed to the existing companies such as those named above. For example, in the next several years, it is likely that there will be consolidation, primarily between the gas producers, pipeline owners, and electricity generators to form vertically integrated companies. This will open the market to new energy companies (as opposed to energy traders).

One of these new companies is Chevron-Texaco, which is getting into the generation business with a concept called municipalization. This approach could become an attractive one for the BIGCC. In this concept, the municipality would take over a regional (municipal) distribution network using the principle of eminent domain. Chevron would then furnish with generation assets, allowing municipalities to separate themselves from the burdens of the large utilities and regulatory agencies. Municipalities generally install smaller power plant size, (40–150 MW), which, not coincidentally, matches the MSW that would be generated in the same area.

Today there are very few (if any) companies interested (the exception may be industrial power plants (see the following)) in taking a technological risk. Those that will take such a risk will do it at a level that can be better absorbed: the 40–100-MW scale rather than at the 1-GW scale. Municipalities tend to be more of a risk taker and tend to be "green" (concerned with local emissions, etc). In the middle of the California energy crisis, two municipalities, Sacramento Municipal Utility District (SMUD) and the Los Angles Water and Power District used their facilities, some with newer combined cycle systems, to supply electricity to power-starved areas of CA. Thus the adoption of technology by municipalities is better because of both size and desire to absorb technological risk in return for "green" power. Also, municipalities, which have a broad tax base, are able to acquire bond funding at a good rate.

A second advantage of the fuel-flexible biomass plant is that today, all segments of industry are going after natural gas for power and process. Unfortunately, disruptions could occur because of exploration delays, pipeline failure, insufficient storage, and vulnerability to terrorist attacks, etc. Some large energy companies, such as Calpine, are including large-scale gas storage in its portfolio. The country is rapidly approaching a shortage of natural gas infrastructure and possible market manipulations (e.g., Ref. 45). The infrastructure is becoming a serious consideration especially since approximately 40,000 miles of additional pipeline is needed by 2015. Some 12,000 miles of offshore pipeline are also under construction. This construction is hobbled by the fact that lenders won't readily OK projects without regulatory permission, a long and arduous task in many cases. Thus, if a technology can be demonstrated that "insulates" a plant from fuel price fluctuations (by fuel flexibility), it will become a superior technology, banks will support it, and competitors may be forced to adopt a similar technology or have the banks refuse to accept the risk of fuel price variability.

Even though very few companies are now willing to absorb technological risk, the good news is that BIGCC plant sizes are of interest to both early adopters and green municipalities. Plants must be introduced at small scale and then increased in size in a stair-step manner as technologies become proven (as in the example of the gas turbine). This is done by using a development path that demonstrates the technology in stages, documented in both categories of 1) maturity of component technology and 2) integration risk. The use of proven components can then be demonstrated to have very little technological risk, thereby limiting the risk to integration issues and new technology components.

According to the Integrated Solid Waste Association [37] there are Industrial Power Producers (IPP's) that are converting various waste fuels into electricity. Currently, there are IPP's that combust wood waste, anthracite culm, tires, landfill gas, some in combination with natural gas. In addition to electricity production, the IPP installations are generally Combined Heat and Power (CHP) plants that also produce steam, which is sold to diverse markets such as a state hospital, a prison, sawmills, a particleboard plant and a pulp/paper complex. While the current IPP market is small, future applications are being identified. As described previously, for example, the paper industry has a need for both power and steam. In response to this need, the Gas Technology Institute has issued a request for programs that would develop and demonstrate a biomass gasification-based power generation system for the forest products industry [46]. While the proposed program does not specifically consider the ATG-type gasifier, interest is in systems up to 20 MW or more that would gasify wood and paper-making by-products and supply power and steam to the paper mill. This size could be attractive, especially if then ATG scales as currently thought and mass production techniques are applicable, as there are hundred's of potential users.

Because of its high efficiency and low emissions, the BIGCC can also be applied to a growing market niche, the repowering of older facilities. This is a very prevalent practice today using gas turbines with heat recovery boilers since it is cost effective with a short pay back time. These plants are also often near urban areas, making them amenable both to concentrating MSW and the municipalization concept mentioned above.

CONCLUSIONS

A multidisciplined team led by the United Technologies Research Center (UTRC) and consisting of Pratt & Whitney Power Systems (PWPS), the University of North Dakota Energy & Environmental Research Center (EERC), kraftWork Systems, Inc. (kWS), and the Connecticut Resource Recovery Authority (CRRA) has evaluated a variety of gasified biomass fuels, integrated into advanced gas turbine-based power systems.

> **The team has concluded that a biomass integrated gasification combined-cycle (BIGCC) plant with an overall integrated system efficiency of 45% (HHV) at emission levels of less than half of New Source Performance Standards (NSPS) is technically and economically feasible.**

The higher process efficiency in itself reduces consumption of premium fuels currently used for power generation including those from foreign sources. In addition, the advanced gasification process can be used to generate fuels and chemicals, such as low-cost hydrogen and syngas for chemical synthesis, as well as baseload power. The conceptual design of the plant consists of an air-blown circulating fluidized-bed Advanced Transport Gasifier and a PWPS FT8 TwinPac™ aeroderivative gas turbine operated in combined cycle to produce ~80 MW_e. This system uses advanced technology commercial products in combination with components in advanced development or demonstration stages, thereby maximizing the opportunity for early implementation. The biofueled power system was found to have a levelized cost of electricity competitive with other new power system alternatives including larger scale natural gas combined cycles. The key elements are:

- An Advanced Transport Gasifier (ATG) circulating fluid-bed gasifier having wide fuel flexibility and high gasification efficiency
- An FT8 TwinPac™ -based combined cycle of approximately 80 MW_e
- Sustainable biomass primary fuel source at low cost and potentially widespread availability – refuse-derived fuel (RDF)
- An overall integrated system that exceeds the U.S. Department of Energy (DOE) goal of 40% (HHV) efficiency at emission levels well below the DOE suggested limits
- An advanced biofueled power system whose levelized cost of electricity can be competitive with other new power system alternatives

RECOMMENDATIONS

1. Perform detailed engineering and cost analysis.

The conceptual analysis reported herein describes a biomass integrated gasification plant that has the potential to be economically feasible. The technical challenges are low risk. A detailed A&E engineering and economic analysis is likely to show installation costs that are even lower than $1400/kW. Given reliable cost and schedule information, government agencies, municipalities, and investors can assess and implement the technology in the near-term.

2. Demonstrate Operation of the Advanced Transport Gasifier with RDF.

Processing and pressurized feeding is often viewed as a showstopper for gasification of MSW / RDF. A multi-hundred-hour test in the EERC pilot-scale ATG would provide a database to evaluate risk of scale-up.

3. Perform R&D to further assess removal of nitrogen compounds and elimination of trace metals (mercury) at warm gas temperatures.

Current warm gas (>350 C) cleanup doesn't remove significant amounts of ammonia / nitrogen compounds or of trace metals, thus R&D aimed at reduction of emissions of these important species is needed.

4. Fund development programs for biogas capable gas turbines

Currently and into the foreseeable future, advanced gas turbines are the most efficient and least polluting means of reliable power production. Commercial-scale operation using biofeedstock requires development of a combustor for 20-30 MW gas turbines capable of burning low calorific value gas.

5. Identify potential WTE demonstration sites for BIGCC.

Use of MSW as a biofeedstock is the most favorable economic scenario for larger-scale (>20 MW) biomass power generation at this time. Siting of a demonstration-scale project at an existing WTE facility would greatly reduce project costs. Currently, however, most WTE facilities are comfortable with mass burn. A study is required to identify potential sites and the incentives needed to place a demonstration project at these sites.

6. Implement a BIGCC demonstration project now in order to exploit current energy needs and MSW disposal requirements

A timely demonstration at reasonable scale, greater than 5 MW (e), of a RDF/ATG / gas turbine / HRSG would provide significant insight into the operation of this concept and give confidence for investors in a future commercial-scale venture. The National benefit of biomass utilization promoted by a single demonstration facility cannot be over-estimated.

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APPENDIX A

Review of Current Gasifier Technology

The Texaco Gasifier System

The Texaco design Figure A 1 is an oxygen-blown, single-stage, entrained-flow gasifier that uses a pulverized coal slurry feed containing approximately 25% pasting water. Raw product gas leaving the gasifier at 1371°C (2500°F) is cooled using either a fuel heat recovery system or is water-quenched followed by partial heat recovery. Capital costs for a 250-MW$_e$ Texaco IGCC plant operating on U.S. bituminous coal and using a cold-acid gas removal process are estimated to be $1446/kW with fuel heat recovery and $1300/kW with a water quench [1]. Plant efficiency is reduced when the quench option is used. Texaco gasifiers have been demonstrated to operate successfully on a wide range of carbonaceous fuels, including bituminous and subbituminous coal, heavy oil and refinery residues, petroleum coke, mixed-plastic wastes, and sewage sludge.

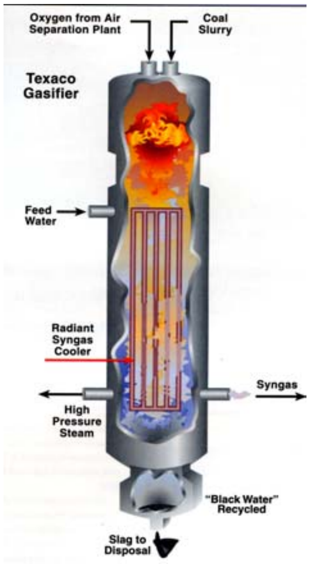

Figure A 1: Texaco Gasifier System

In the Texaco oxygen-blown pressurized entrained-flow gasifier, shown in Figure A 1, coal slurry feed and oxygen are combined in burners that are oriented downward from the top of the gasifier. Operating conditions are typically at or above 27.5 bar (400 psig) and 1371°C (2500°F) with a 2-second residence time. The Texaco design is based on an earlier commercial process for gasifying petroleum resid, and it is the most mature and commercially accepted design of this type. The Texaco design will gasify any carbonaceous material that can be fed into the gasifier, but economical operations on high-moisture, low-rank coals or biomass would require hydrothermal pretreatment to provide the high dry solids content in the slurry and low oxygen feed rate required for efficient operation. Commercial applications of the Texaco design for coal gasification have been demonstrated in the Tampa Electric IGCC project which gasifies

2200 tons/day of coal to generate 250 MW with a 36.5% efficiency (HHV) [2]. A typical fuel gas composition of the cleaned fuel gas is shown in Table A1. The Tampa Electric plant injects nitrogen from the ASU to lower the heating value of the syngas to approximately 120 Btu/scf in order to reduce the amount of thermal NO_x formed in the gas turbine combustor. Several other units are used in chemical manufacturing and petrochemical refining processes throughout the world. The high moisture content of biomass has precluded its use in Texaco gasifiers owing to the substantial increase in oxygen required to evaporate excess coal moisture in the slurry feed. The oxygen requirement increases substantially as the moisture of the feed slurry increases. Pre-drying the biomass feed would be far more economical than increasing oxygen to the gasifier if moisture were not reabsorbed when the biomass feedstock is slurried. For this reason, conventional gas-drying methods are not of practical use in this application. Texaco has not pursued a dry-feed system for its gasifier.

Table A 1: Typical Fuel Gas Compositions of Selected Gasifier Technology (% by vol)

	Texaco TECO	E-Gas Wabash	Shell Demkolec	Prenflo Puertollano	KRW Piñon Pine	ATG EERC/ PSDF	GTI Renugas Calla	FERCO Silvagas Vermont	BGL SVZ
CO	42.7	45.3	63.4	60.5	23.9	11.3	15.3	40.4	59
H_2	38.3	34.4	28.4	22.1	14.5	19.5	14.8	5.1	27
CO_2	14.4	15.8	1.5	3.9	5.5	16.7	12.3		3
CH_4	0.1	1.9	0.4	0.0	1.4	3.2	2.6	11.8	7
H_2O	0.3				5.5	0.0	13.9	42.6	
N_2	3.3	1.9	6.2	12.5	48.6	47.5	40.5		4
Ar	0.9	0.6		1.0	0.6	1.2			
H_2S	200 ppm	68 ppm	20 ppm	6 ppm	20 ppm	.18	190 ppm		1 ppm
COS	10 ppm			6 ppm	0 ppm				
NH_3	0 ppm	0 ppm	0 ppm	0 ppm	200 ppm	0.11	2000 ppm		0 ppm

The Global Energy E-Gas Gasifier System

The E-Gas (Destec) gasifier system (Figure A 2) is described in the EPRI Coal Gasification Guidebook [3]. The design is a two-stage, pulverized coal slurry feed, entrained-flow gasifier. The slurry is prepared with a minimum of pasting water and heated to provide some dewatering of coal particles and reduction in slurry viscosity. About three-fourths of the slurry feed is fed to the first stage operating at 1316° to 1427°C (2400° to 2600°F). The remaining slurry is injected into the second stage, from which the raw product gas exits at 1038°C (1900°F). The E-Gas IGCC system used for the Wabash River Clean Coal Demonstration Project [4] matches the gasifier with a heat recovery gas cooler, particulate removal, cold-gas sulfur removal, and fuel gas reheat. This project has successfully operated since 1996 on 2450 tons/day of an Indiana bituminous coal in Terre Haute, Indiana, and is currently gasifying 2000 tons/day of petroleum coke to generate 262 MW net to the grid at an overall thermal efficiency of 39.7%.

Figure A2 shows the gasifier configuration for the Wabash River Repowering project. Table A1 shows the fuel gas composition obtained from the Wabash facility. The Wabash River plant remoisturizes and injects steam to again lower the heating value of the syngas entering the gas turbine to approximately 120 Btu/scf.

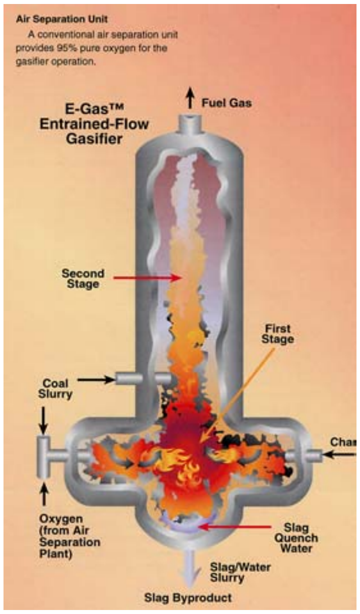

Figure A 2: E-Gas IGCC System

The two-stage design of the E-Gas gasifier reduces heat recovery requirements and increases flexibility for using higher-moisture fuels as evidenced by operation of the Plaquemine, Louisiana, gasification facility on U.S. Wyoming subbituminous coal. However, Global Energy has also projected [3] a large increase in oxygen demand for a Texas lignite containing 35% moisture as compared to a low-moisture Appalachian bituminous coal. A dry-feed system for this gasifier has also not been pursued. The hot product gas is passed through a hot-gas filter to separate unreacted char particles for recycle. The benefit of this design is that it increases the overall conversion efficiency of coal to gas while reducing the need for heat recovery from the product hot gas. Carbon conversion efficiency has been greater that 99%. The E-Gas gasifier, because of its two-stage design, may provide increased flexibility in the use of high-moisture fuels without hydrothermal treatment.

The Shell Gasifier System

The Shell gasification and cold-gas-cleaning system (Figure A 3) consists of a dry-feed, oxygen-blown, entrained-flow gasifier followed by a syngas cooler, wet scrubber, and cold-acid gas absorption unit (the Sulfinol process was used at Shell's Deer Park, Texas, demonstration plant). This configuration was demonstrated on Texas lignite at a capacity of 400 short tons/day where it achieved a carbon conversion of 99.77%, a cold-gas efficiency of 80.3% (HHV), and a total energy recovery (gas plus steam) of 95.7% [5]. Sulfur control levels of 99% can be achieved with cold-gas absorption, with conversion to salable sulfur in a Claus-type unit. Shell's estimate of capital cost for a 2 × 400-MW$_e$ IGCC plant is \$1500 to \$1600/kW; projected efficiencies are 42% to 46% (LHV) depending on coal quality and turbine efficiency. Hot-gas cleanup is estimated to improve efficiency by 1.3 percentage points, with no estimate given for the expected reduction in capital cost. A 250-MW_e IGCC plant based on Shell technology has been built and operated successfully by Demkolec in the Netherlands.

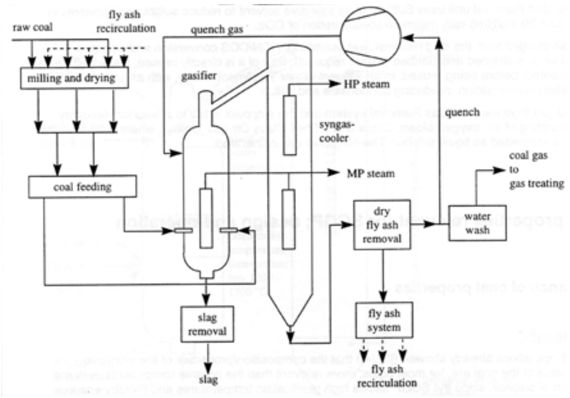

Figure A 3: Shell Gasifier

The Shell, pressurized, oxygen-blown entrained-flow gasifier is unique in that dry pulverized coal is pneumatically fed to gasifier burners in a dense fluidized phase transported in nitrogen. Oxygen is added at the burners, along with steam as required, to control the gasifier-operating temperature. The gasifier is designed to operate at up to 1649°C (3000°F) and 41.4 bar (600 psig). Carbon conversions of over 99% are achieved in a single pass. Development of the Shell process started in 1972 and has included work on a 400-ton/day demonstration plant near Houston, Texas. The process is reported to be insensitive to fuel properties and to be capable of more efficiently gasifying higher-moisture, higher volatile fuels. This technology is currently being commercially demonstrated at the Demkolec plant in Buggenum, Netherlands. In this demonstration plant, product gas is first cleaned of particulates and ammonia and is then sent to the Sulfinol-D acid gas recovery system operating at an inlet gas temperature of 32°C (90°F). The Shell gasifier system provides a higher overall efficiency of 43% in IGCC applications as compared to the slurry-fed entrained-flow gasifiers [6]. This gasifier utilizes remoisturization and nitrogen injection to reduce the fuel gas heating value to approximately 112 Btu/scf for NO_x control using a diffusion flame combustion concept [7]. Recently, the Demkolec plant has been successfully cofeeding a small amount (5–10 wt%) of a dried chicken litter biomass to its gasifier [8].

The Prenflo gasifier at Puertollano, Spain, is similar to the Shell gasifier in that a dry-feed system is utilized to cofeed a high ash Spanish coal along with petroleum coke to the gasifier to make 335 MW of electricity at approximately 45% thermal efficiency [9]. This system started up fairly recently and does not have the long operating history that the other entrained-flow gasifiers have. The projected Prenflo gas composition is shown in Table A1. This IGCC also uses nitrogen injection to lower the syngas heating value to approximately 115 Btu/scf for NO_x control purposes [7].

The principal advantage of the Shell and Prenflo gasifiers for biomass is the dry-feed system, which allows either as-received or thermally dried fuels to be fed to the gasifier without the issues associated with reabsorbing water in a slurry preparation step. The high gasifier exit temperature of 1371°C (2500°F) makes the technology less attractive for simplified IGCC designs based on hot-gas cleanup because of the substantial gas cooling required to match a hot metal oxide sulfur removal system. Plant availability remains somewhat of a concern for both plants.

British Gas/Lurgi

The Lurgi dry ash gasifier uses the temperature-moderating effect of a high ratio of steam to oxygen to maintain the gasification temperature at the bottom of the bed below the fusion temperature of the ash [10]. High reactivity coals are an ideal feedstock in that they are more reactive than high-rank coal at the relatively low operating temperature and are also noncaking. Lurgi gasifiers can be operated on either air or oxygen at pressures of up to 30 atmospheres. Dakota Gasification Corp, in Beulah, North Dakota, and SASOL in South Africa have been successfully operating these types of gasifiers for several decades. The typical dry/tar-free product gas leaving these gasifiers is shown in Table A1.

Some of the limitations of dry ash fixed-bed gasification are remedied in the British Gas/Lurgi slagging fixed-bed gasifier as shown in Figure A 4. The essential differences between the slagging designs and a dry ash Lurgi gasifier are in the substitution of a slagging hearth and taphole for the ash grate and the nearly fivefold reduction in steam flow per ton of coal, which allows the oxidation zone of the gasifier to reach temperatures above 1371°C (2500°F). This design has not been advocated for use with high volatile fuels by British Gas/Lurgi, even though extensive tests performed by DOE at the EERC have shown that the method is applicable to North Dakota lignite. Tests at the EERC on a 25-tpd oxygen-blown slagging gasifier operated on 25%–30% moisture lignite yielded 250 Btu/scf gas at 177°C (350°F) [11]. The British Gas/Lurgi gasifier has reportedly been tested in an air-blown mode operating on a 538°C (1000°F) air blast, reportedly producing 120–130 Btu/scf gas at an exit temperature higher than in a dry-ash design [12]. The advantages of the British Gas/Lurgi gasifier include the capability of using coal fines or some coal tars by injection into the high-temperature 1649°C (3000°F) reaction zone, a four- to sixfold reduction in steam consumption, improved system efficiency both at the gasifier and at the plant stack (by reducing the substantial energy loss due to water vapor leaving the stack), the generation of a environmentally benign vitrified slag, and a lower yield of ammonia (NH_3). Maintaining slag flow is a problem unique to the slagging design. The principal advantages of this type of gasifier are in the higher thermal efficiency and reduced volume of gas liquor that result from reduced steam flow. The high-temperature slagging design has been advocated for bituminous coals as a means of compensating for the lower reactivity of these coals at the relatively lower temperature of a dry-ash Lurgi design. The slagging design has been successfully tested on coals of all ranks by the British Gas Corporation (BGL [British Gas Lurgi]) in cooperation with the U.S. sponsors and is being offered by the BGL. Global Energy which has the license for U.S. commercial applications is currently developing the Kentucky Pioneer project which will use a BGL slagging gasifier to process a 50 wt% coal mixture to produce 540 MW RAP Pellet at a thermal efficiency of 47.8% [13]. Global Energy has stated that maximum RDF to coal ratio would probably be about 2 to 1 for this type of system because of issues with the limited amount of tar generated from the high volatile biomass which could be recycled back to the gasifier.

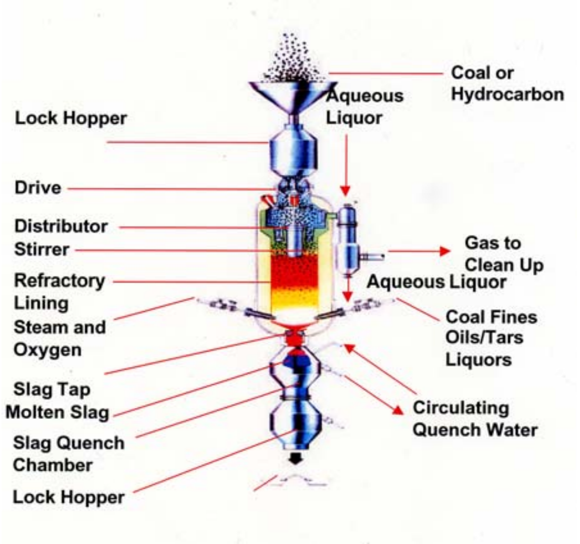

Figure A 4: British Gas Lurgi Fixed-Bed Gasification

High-Temperature Winkler (HTW)

Uhde GmbH together with Rheinbraun AG and Lurgi GmbH has developed the fluidized-bed HTW process Figure A 5 in Germany for both 1) a 10-bar oxygen-blown bubbling-bed design generating synthesis gas for methanol production from 720 tons/day of dried brown coal and 2) a 25-bar design using either air or steam/oxygen as the gasification agent for IGCC applications [14].

Dried fuel is fed through a lock hopper system to the lower fluidized zone of the gasifier operation at 850°C (1562°F). Gasification agent (air or steam/oxygen) is admitted both to the lower zone and to the above-bed freeboard to gasify entrained carbon at approximately 940°C (1724°F). Entrained solids leaving in the raw product gas are separated in a cyclone and fed by gravity back into the gasifier. Ash is withdrawn from the bottom of the gasifier and fed to a moving-bed cooler.

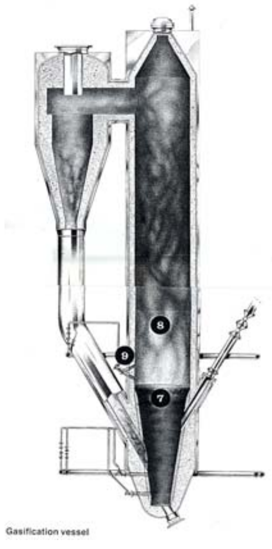

Figure A 5: HTW Gasifier

Coal is crushed to a 4-mm maximum grain size with fines retained in the coal feed. Beneficiation to reduce ash content is not recommended. Design measures are described for preventing ash and slag deposits in the gasifier system. High ash content is indicated to result in a lower cold-gas efficiency, estimated to be 68% to 70% for high-ash lignites versus 76% for German Rhenish brown coal. Carbon conversions have ranged from 91% to 96% for the lignite feedstocks. Rheinbraun has operated a HTW demonstration plant at Berrenrath since 1986 processing 600 tons/day of dried lignite to make 28 Mscf/day of synthesis gas for the production of 330 tons of methanol/day. This plant has a thermal input of approximately 140 MW$_{th}$ and has an internal diameter of 9 feet [14]. This demonstration plant has operated over 67,000 hours, including successful operation in 1997 on 1000 tons of RDF at 25 wt% and 50 wt% mixtures of lignite and dried pelletized (10 wt% moisture and ~12.5 to 19 mm(~0.5 to 0.75 inch) RDF fuel [15]. This gasifier showed no change in its operation and gas quality while on the lignite RDF mixture. The environmental performance was also monitored during these tests; most of the heavy metals (including Hg) and halogens such as chlorine reported to the filter dust. Over 99% of the amount of dioxins and furans fed in with the RDF were destroyed, with total emissions being 2% of the German regulatory limit of 0.1 ng/m^3. Furthermore, since the chlorine and fly ash are removed before the combustion process, the formation of the dioxin and furans during combustion is not possible since these necessary components are missing [15].

Operation on air from the gas turbine compressor as the gasifying agent as compared to oxygen or oxygen-enriched air is indicated by Uhde studies to have only a slight effect on capital investment, overall plant efficiencies, or emissions. A dry-gas heating value of 129 Btu/scf is indicated for air-blown gasification of Rhenish brown coal, compared to 271 Btu/scf for oxygen gasification.

Both cold-gas and hot-gas cleaning systems are described for the HTW gasifier. The cold-gas case includes 1) gas cooling to 260°C (500°F) in a high-pressure steam generator;

2) particulate filtration using ceramic candle filters; 3) wet scrubbing for removal of alkalies, chlorides, and other trace contaminants; 4) hydrolysis of COS and H_2S; 5) desulfurization using a selective process leaving CO_2 in the gas stream; and 6) final humidification and pretreating.

Hot-gas cleaning, which was previously discussed, is not commercially proven as described: 1) cooling to 650°C (1202°F), 2) particulate filtration, 3) treatment in a limestone/dolomite fixed bed for removal of alkalies and heavy metals, 4) metal oxide desulfurization, and 5) final particulate filtration. The retention of NH_3 in the treated gas (absent a water scrubber) is identified as an important problem affecting NO_x emissions.

The efficiency of a 310-MW IGCC (160-MW gas turbine and 150-MW steam turbine) is estimated to be 46% to 47% (LHV), without reference to a particular coal quality. Use of hot-gas cleaning is estimated to increase efficiency by about 2 percentage points. The estimated cost of the HTW gasification system for this plant is US$1129/kW, including coal drying, gasification, and cold-gas cleaning, but excluding the power systems.

Kellogg Gasification Processes:

Kellogg Brown & Root offers two gasification technologies: 1) the KRW process and 2) an advanced transport reactor process that is currently under testing at the EERC and at the Southern Company Services Power Systems Development Facility (PSDF) site in Wilsonville, Alabama.

Kellogg Rust Westinghouse (KRW) Gasifier

The KRW gasifier pneumatically feeds 1/4" × 0 coal through a central jet along with the combustion air or oxygen. Since this gasifier injects the oxidant with the fuel, significant quantities of the fuel's volatile matter are consumed to generate the process heat. The process has been operated in air and oxygen-blown modes on U.S. lignites from Texas and North Dakota at a 35-ton/day pilot plant in Waltz Mill, Pennsylvania [16].

The KRW fluidized-bed gasification system (Figure A 6) offers high thermal efficiency based on its low steam and oxygen consumption. The KRW system has been developed in both single-stage and two-stage configurations to match the processing characteristics of coals of all ranks from lignite to bituminous. The single-stage design is applicable to higher reactivity fuels, but less reactive fuels use a spouting-type bed wherein coal and recycled fines are pneumatically injected in a high-velocity flow of recycle gas through an axial feeding tube, where they are mixed with preheated steam and oxygen. The fuel undergoes rapid devolatilization and partial volatile combustion in the inlet jet zone. The gasifier shell is constructed in sections of successively larger diameter to accommodate combustion, gasification, and disengagement of char solids. The larger unreacted char particles fall back into an internal solids recirculating pattern within the gasifier to undergo further gasification. Fines entrained from the gasifier are separated in a cyclone and injected back into the gasifier with the coal feed. The gasifier is operated at temperatures of 815° to 1010°C (1500° to 1850°F) in an ash-agglomerating mode that causes dense ash to fall to the bottom of the gasifier to be removed through a rotary valve. Fuel feed is sized to minus 6.4 mm (1/4 inch) with in-mill drying to reduce surface moisture to less than 5% to accommodate pneumatic feeding. Fuel fines are not a problem, since they can be recycled until completely gasified. Carbon utilization in excess of 95% has been demonstrated for high-reactivity fuels.

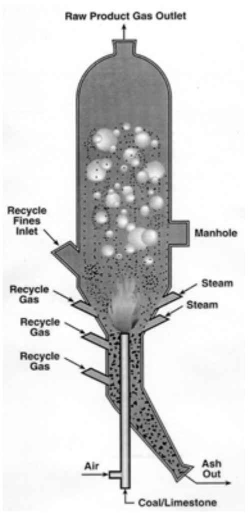

Figure A 6: KRW Gasifier

The Piñon Pine project utilizes the KRW gasifier to gasify 880 tons/day of Utah bituminous coal to produce 100 MW net of power with a thermal efficiency of 40.7% [17]. The internal diameter for this gasifier is 12 feet and operates with a velocity of 1.5 ft/sec in the freeboard. While the KRW gasification process has been well demonstrated in the 35-ton/day pilot plant, numerous start-up and operating issues have plagued the Piñon Pine Clean Coal Technology (CCT) demonstration. Because of operational problems with solids removal from the system and a desire by the owner Sierra Pacific Power Company to divest itself of its power generating facilities, the KRW gasifier at Piñon Pine has not operated for any extended period of time and has not supplied any syngas to the gas turbine. No testing of biomass in either scale gasifier has occurred. Successful demonstration of this technology at this larger scale should be accomplished before serious consideration of this gasification technology.

Advanced Transport Reactor

The transport reactor Figure A 7 design feeds a finely crushed (~ 1/16" × 0) coal into a high-velocity fluidized bed operating at a velocity of 20 to 40 ft/sec (6.1 to 12.2 m/sec). High carbon conversion is achieved by recycling a large flow of solids back into the reactor. The reactor operates at 815° to 1038°C (1700° to 1900°F) in gasification mode and at pressures up to 400 psig [18, 19]. The transport reactor concept was adapted from the proven design used for fluidized-bed catalytic cracking units in the petroleum industry. Development for coal conversion has been confirmed in a 2.4-ton/day pilot-scale unit at the EERC and on a 38-ton/day proof-of-concept unit at SCS's Wilsonville Power Systems Development Facility. The transport reactor design, owing to its use of finely crushed coal (fines are not a problem) and its simple design, may offer future advantages in designing IGCC systems at minimum cost. The separation of the fuel feed from the air or oxidant injection ports is an important part of this concept in that it allows the volatile matter in the fuel to be thermal-cracked and steam-reformed without being consumed by oxygen. The oxygen entering at the bottom of the mixing zone only partially oxidizes with the returning char from the standpipe, which is the hardest material to steam-gasify.

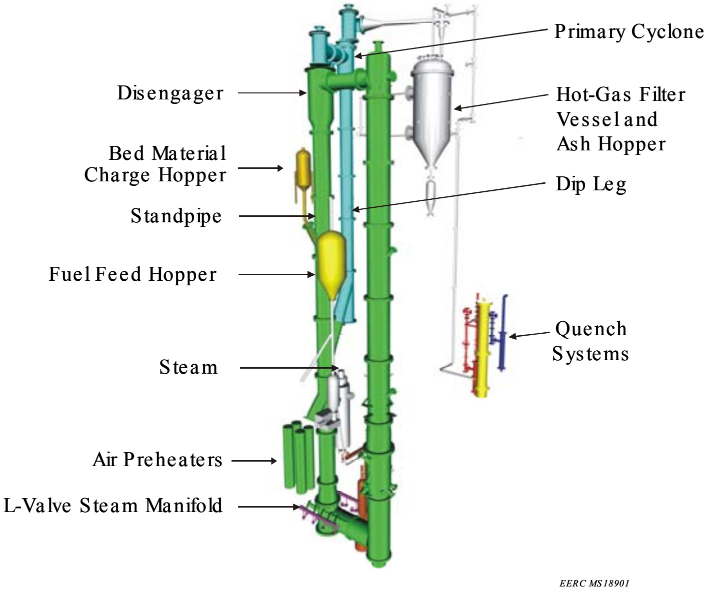

Figure A 7: Advance Transport Gasifier

The transport reactor is reported to be the highest throughput gasifier because of its high operating velocities. This high throughput allows a very small diameter gasifier to be utilized which reduces the capital cost of the gasification system as compared to other gasifier systems. Based on the required gas flow calculated by UTRC personnel, a transport reactor would need an internal mixing zone diameter of 45 inches and a riser diameter of 32 inches which should be considerably smaller than any other type of gasifier available and allow the use of off-the-shelf piping and flanges. This smaller diameter gasifier will result in reduced capital costs for the gasifier portion of the biomass gasification system. Carbon conversions from 85% to 98% have been demonstrated at both the EERC and the PSDF in Wilsonville, Alabama. No biomass gasification tests have been conducted in these two systems to date.

The GTI RenuGas™ Gasification Process

The U-Gas and RenuGas™ gasification process Figure A 8 was developed by Gas Technology Institute (GTI) formerly the Institute of Gas Technology (IGT) in the United States and licensed to Enviropower in Finland for commercialization in the European markets. Commercially, eight U-Gas gasifiers have been installed in China to supply fuel gas for coke overseas. The RenuGas process feeds dried and crushed coal (25% moisture sized to 1/4 in. × 0) through a lock hopper to a pressurized fluidized-bed (PFB) gasifier that incorporates a hot ash-agglomerating zone. Normal operating conditions are 760° to 980°C (1400° to 1800°F) and up to 30 atm [20]. Operation has been demonstrated with both oxygen and air. No steam is required for high-moisture fuels under air-blown operation. Ash agglomerates are discharged in an essentially nonleachable vitrified form through a countercurrent heat exchanger where they are cooled to 470°C (1200°F). The high-temperature agglomerating zone reportedly provides a higher carbon conversion (e.g., 95%) with lower external char recycle than does conventional fluidized-bed gasification [20].

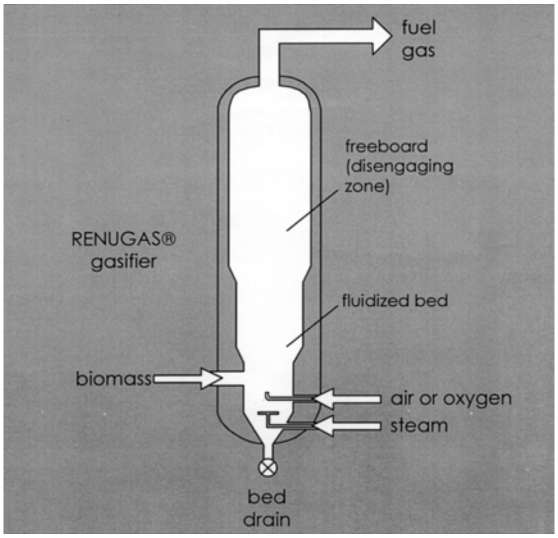

Figure A 8: U-Gas Gasifier

Lignite feed dried to 25% to 35% moisture can contain up to 10% fine coal (- 100 mesh or 149 µm). Hot-gas cleanup for particulates and sulfur have been investigated by IGT, including limestone addition for in-bed sulfur removal. Gasification of several biomass fuels have also been successfully conducted in the GTI gasifier. A 100-ton/day pressurized demonstration facility was built in Hawaii but was never successfully demonstrated primarily because of issues with feeding the sugar cane bagasse to the gasifier [21]. A larger 400-ton/day near-atmospheric-pressure gasifier is currently being designed for the Calla Energy site in Estill County in Kentucky. This gasification system feeds biomass sized to less than an inch through lock hoppers to a delivery screw feeder that inserts the biomass directly into the gasifier [20]; however, the fuel feed point is very close to the air injection ports.

The FERCO SilvaGas Process

Battelle has developed an indirect gasification process (Figure A 9) which utilizes two different circulating fluid-bed loops to pyrolyze and steam-gasify the injected biomass. The remaining char is captured in the cyclone of the first fluid bed and sent to the second circulating fluid bed where it is combusted with air to increase the bed material temperature and burn out the residual char. The heating of the bed material in the second fluid bed allows the endothermic pyrolysis and steam gasification reactions to continue in the first bed without having to add oxygen from an ASU or diluent the syngas with nitrogen by adding air to the gasifier. This allows a medium-Btu (300 to 400 Btu/scf) fuel gas to be generated. FERCO has bought the rights to the process and has renamed it the SilvaGas process. A 400-ton/day demonstration plant has been built at the McNeil Station in Vermont to utilize wood waste and chips to generate a fuel gas for combustion in an adjacent boiler [22]. While some very early work was done under pressure, this process has been developed and demonstrated as an atmospheric pressure gasification system.

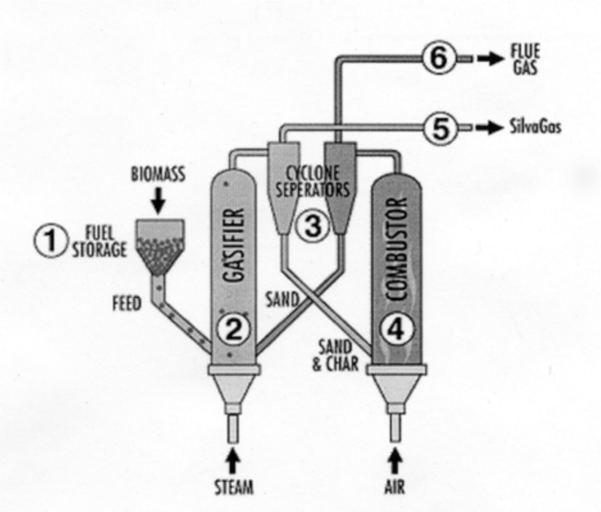

Figure A 9: FERCO – Silva Gasifier

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APPENDIX B

Images of Fractions from Hand-Sorting of RDF

Figure B1: Paper, cardboard, and plastic film fraction in RDF

Figure B2: Wood fraction in RDF

Figure B3: Glass fraction in RDF

Figure B4: Container plastic fraction in RDF

Figure B5: Dense plastic fraction in RDF

Figure B6: Metal fraction in RDF

Images of Air-Classifier and Fractions of RDF Recovered from Air-Classification

Figure B7: Forsberg air-classifier

Figure B8: Light (product) fraction from air-classification

Figure B9: Heavy (reject) fraction from air-classification

Figure B10: Combustibles in reject fraction from air-classification

Figure B11: Non-combustibles in reject fraction from air-classification

Figure B12: Non-combustibles in product fraction from air-classification

APPENDIX C

Trace Element Removal

The EERC transport reactor demonstration unit (TRDU) testing program has enabled a unique opportunity to examine the trace element partitioning within a pressurized circulating fluid-bed gasifier as well as at the inlet and outlet of a hot gas filter vessel (HGFV). Trace element sampling was conducted during a TRDU test run using a Wyodak subbituminous coal–dolomite blend as the feedstock [1,2].

The trace element composition of the gasification product gas was determined at the inlet to the HGFV using a modified EPA Method 29. The partitioning of trace elements between gas and solid phases was also evaluated based on the EPA Method 29 measurements. It is assumed in this evaluation that the sample filter and impingers of the modified EPA Method 29 train remove all the particle-associated trace elements and gaseous trace elements, respectively, from the particle-laden flue gas stream. Presented in Figure C1 is the trace element composition and partitioning results for this test. The trace element concentrations and gas–solid partitioning results determined at the inlet and outlet of the HGFV are presented in Figure C2. Trace element removal efficiencies for the HGFV are presented in Table C 1: Trace Element Removal Efficiencies of the HGFV. These results indicate, as expected, that the HGFV is effective in removing the nonvolatile trace elements (As, Pb, Cd, Cr, and Ni), but relatively ineffective in removing semivolatile Se and volatile Hg.

Table C 1: Trace Element Removal Efficiencies of the HGFV

Element	HGFV Removal Efficiency, %
Hg	2.6
Se	77.5
As	98.8
Pb	95.7
Cd	85.9
Cr	95.6
Ni	94.0

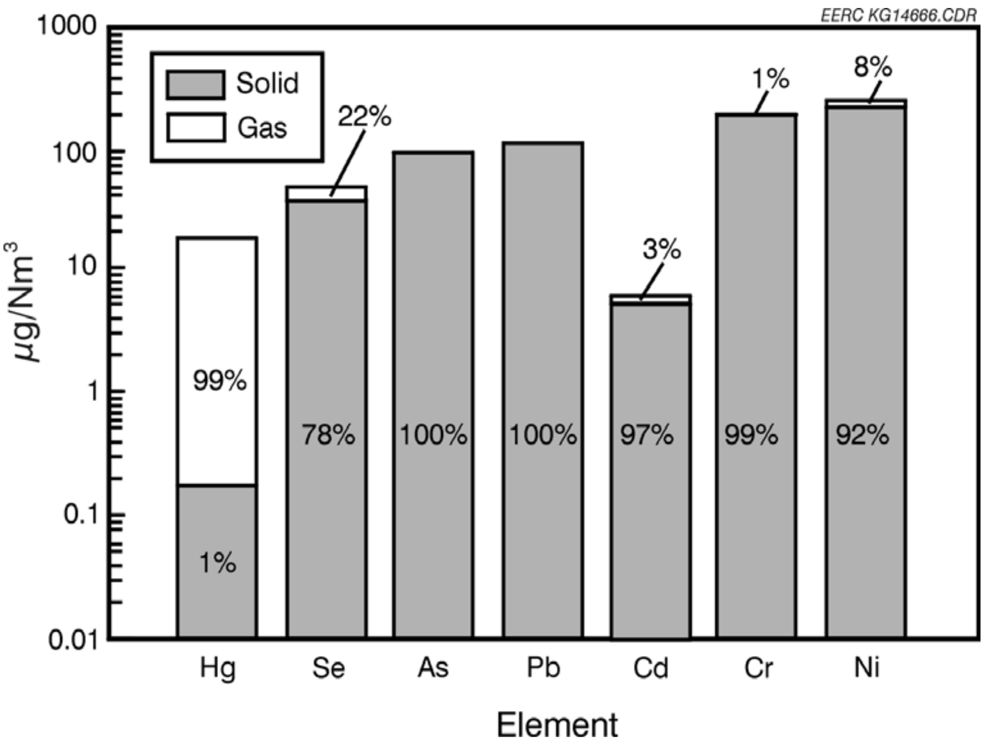

Figure C 1: Solid–gas partitioning of trace elements for Test P0501

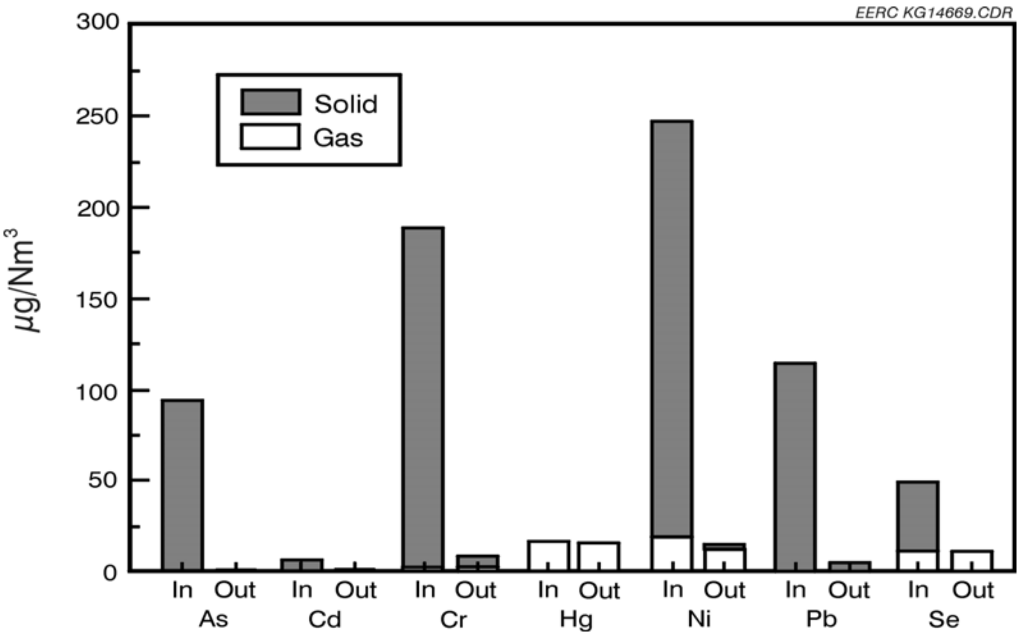

Figure C 2: Trace element concentrations and solid–gas partitioning measured at the inlet and outlet of the HGFV during Test P051.

In work conducted recently at the EERC, stabilities of various mercury compounds in a 2000-mL/min flow of inert (nitrogen) and hydrogen streams were evaluated by heating the compound isothermally in a tube at temperatures from ambient to 300°C (575°F) to decompose the compound and determining the concentration of mercury in the vapor stream using a mercury CEM [3]. Three gas compositions were investigated, 0%, 5%, and 25% hydrogen in nitrogen. The runs in 25% hydrogen concentration were conducted only for the sulfide, since it appeared to be more stable than most of the other compounds in 5% hydrogen. For this set of compounds, the effluent mercury was 100% in elemental form. These data show some degree of decomposition to elemental mercury for all the mercury compounds at 200°C (425°F) or even at 100°C (212°F) (not shown in table). This occurs in both N_2 and H_2–N_2 streams. The decomposition at 200°C (425°F) is relatively slow for most of the compounds, however. As expected, the decomposition rates increase greatly with temperature for all compounds.

With all compounds at all temperatures, the decomposition increased substantially with addition of H_2 to the stream, thus hydrogen in the gas stream reduces the mercuric compounds under these conditions. As expected the reduction is dependent on hydrogen concentration in the stream and goes off the scale of the CEM at 25% H_2 at temperatures above 250°C (530°F) It should be pointed out that 2000 $\mu g/m^3$ observed at the maximum scale limit represents only 0.004 mg/min, or 0.08 wt%/min, loss of the sample weight. Losses at high temperatures were measured independently in experiments using a larger amount of the mercury compounds and determining weight loss gravimetrically. These weight losses were recorded over a 10-min period.

Work in the last year under this project has shown that mercury metal compounds such as HgS and HgSe are relatively stable up to 300°C (635°F), even under reducing conditions. The melting temperature of HgSe is 770°C (1418°F), while HgS sublimes at 583°C (1081°F). If the mercury in the coal can be made to form HgS or HgSe, which are solids, the mercury could potentially be taken out with a hot-gas filter or some other solid particulate removal device. Condensation temperature for the high steam levels present in an oxygen-blown transport reactor would be approximately 200°C (425°F) [3].

The rate of decomposition of mercuric sulfate to elemental mercury is similar to that observed earlier for mercuric oxide. The rate increases dramatically at 275°C (525°F), especially when hydrogen is present. The other products of the decomposition of the sulfate are not known, but may be sulfuric and sulfurous acid. The decomposition of a set of mercuric nitrates in N_2 and in air was also investigated in another project. The set of mercuric nitrate compounds behave somewhat differently compared to the sulfate, in that mercuric nitrate is emitted to the gas phase in various proportions along with elemental, depending on the nature of the compound. But the rates are roughly similar. Mercuric chloride also devolatilized without decomposition to the elemental at these temperatures.

As indicated by the lower emission rates at 200°C (425°F) and 300°C (637°F), mercuric sulfide is more stable than the sulfate, especially in the 5% H_2 gas stream. The reaction is presumed to form H_2S in the hydrogen atmospheres, but elemental sulfur may also be evolved, especially at the lower H_2 concentration. The reason for the higher stability of the mercuric sulfide is that the sulfide is a very soft base and thus forms a strong bond with the soft acid Hg(II).

The red form of mercuric sulfide is somewhat more stable at lower temperatures, but becomes more reactive at higher temperatures. The black form is converted to the red at 300°C (637°F).

Lower decomposition rates were exhibited by the mercuric selenide. Thus the selenide is a little more stable than the sulfide under most temperature and hydrogen conditions. This stability can be attributed to the very soft basic character of the selenide ion.

Many of the elements of intermediate volatility will condense in the region of temperature in a gasifier where warm- or hot-gas cleanup takes place. Many of these elements form metallic and other reduced phases and accumulate in these regions of the system. These phases may have the potential to react with and form stable phases with mercury. The consistent characteristic is that they all appear to condense to form liquid or solid phases below 700 K (427°C, or 800°C). Thus, the operating window for potential mercury removal in a hot-gas filter system would be between 200°C (425°F) and 300°C (637°F). There is some potential that condensing tars could cause operation problems for the filter system; however, results from the HT Winkler gasification project in Berrenrath, Germany, have exhibited acceptable hot-gas filter operation at temperatures around 260°C (500°F). In addition, gasification tests completed on coal/RDF mixtures in the HTW plant at Berrenrath, have shown that 86 to 95% of the mercury seemed to be captured in the filter ash and all of the solid residues, respectively [4]. This high collection efficiency for mercury in the filter system is thought to be due to the formation of various solid mercury species such as mercury sulfide, or mercury chloride can be removed at these low filtration temperatures.

Work is continuing at the EERC and elsewhere on developing better mercury monitoring and control technology that will be effective at elevated temperatures. This is probably the area that will need the most research and development to generate an economic commercially available technology.

Appendix C - References

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