Commercial Gasifier for IGCC Applications Study Report

By
John E. Notestein

With Contributions From
Larry A. Bissett
Larry D. Strickland
John S. Halow
Julianne S. Klara
Hoanh Pham
John M. Rockey
H. Mitch Spengler

U.S. Department of Energy
Office of Fossil Energy
Morgantown Energy Technology Center
P.O. Box 880
Morgantown, West Virginia 26507-0880

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Executive Summary

This was a scoping-level study to identify and characterize the design features of fixed-bed gasifiers appearing most important for a gasifier that was to be (1) potentially commercially attractive, and (2) specifically intended for use in integrated coal gasification/combined-cycle (IGCC) applications. It also performed comparative analyses on the impact or value of these design features and on performance characteristics options of the whole IGCC system since cost, efficiency, environmental traits, and operability -- on a system basis -- are what is really important. The study also reviewed and evaluated existing gasifier designs, produced a conceptual-level gasifier design, and generated a moderately advanced system configuration that was utilized as the reference framework for the comparative analyses. In addition, technical issues and knowledge gaps were defined.

Three overall goals were to define system (and gasifier) configurations that (1) were exceptionally clean from an emissions point of view, (2) were well below the capital cost of conventional coal plants with flue gas scrubbing, and (3) offered projected cost-of-electricity (COE) values that were at least highly competitive with alternative technologies -- including natural gas-fired combined cycles. These goals were accomplished. Three major variations on a reference Simplified IGCC system configuration were identified that offered respective sulfur oxides (SO_x) and nitrogen dioxide (NO_2) reductions of about 98 and 60%, and projected N^{th} plant Total-Process-Capital costs in the range of \$480 to 520 per kilowatt electric (kWe). (This cost corresponds to a Total Capital Requirement in the vicinity of \$700/kWe, using representative assumptions for the "soft" costs.) Both one-stage and two-stage, dry-bottom, air-blown, fixed-bed gasifiers were found to provide attractive bases for these types of systems. Similarly, a one-stage, slagging-bottom, air-blown, fixed-bed gasifier was found to offer lower environmental burdens for potentially the same (or perhaps, lower) costs. Two-stage gasifiers that recycle pyrolysis gas to a deep-bed location were found to be unattractive.

One significance of these parameters is that many utility plants/systems can be upgraded or expanded to achieve growth in electricity output of as much as 20 times current levels for no net increase in absolute emissions from levels that would correspond to current power generation rates while emitting SO_x at 1.2 lb per million Btu (MBtu) of fuel fed to the plant. In reality, this relationship should be recognized as either (1) a *large* potential capability for growth in generating capacity with a simultaneously *large* reduction in emissions burdens, or (2) an *immense* reduction in emissions burdens coupled with a small potential for growth in generation capacity. This occurs because most problem plants/systems currently emit SO_x at rates well in excess of 1.2 lb/MBtu of fuel fed. Further, the associated costs are more attractive than any competing generation technology, with the current exception of natural-gas-fired combined cycles, and the long-term availability and pricing of natural gas relative to coal casts doubt on the long-term competitiveness of natural-gas-based systems.

In general terms, design features and options were examined that offered routes to utilize a greater fraction of the coal mined, increase the energy output rates from the gasifier, lessen emissions from the overall system, and reduce costs for the overall system. A significant number of gasifier design features were identified that had the potential to improve the gasifier (and system) performance and operability, but in almost all cases, a technically adequate basis with which to quantify cost reductions associated with these improvements was lacking. As a result, the gasifier costing was done on a very conservative basis and largely reflects Lurgi costs. Many of the cost variations, consequently, reflect variations in the approach taken to system integration. However, it should be recognized that another by-product of this is that it would be expected that system costs will be further reduced when enough data become available to quantify the effects of these improved gasifier features.

The design philosophy was to emphasize technology that could be reasonably expected to be implementable, at least in a basic form, within the next 10 years, in order to be capable of making a national contribution as the need for new generation capacity rises. Further, the nature of the technology should be such that it would be anticipated to evolve into still more attractive technology or systems as a result of progressive design maturation activities over subsequent years. An example of this would be the evolution of a desulfurization sorbent into a desulfurization/tar-cracking media, thence to a desulfurization/tar- and ammonia-cracking media, and finally to a combined desulfurization/tar- and ammonia-cracking/granular-bed filter media. The potential commercial attractiveness was also judged to be enhanced to the degree that component *standardization* and *modularity* could be enhanced. In particular, the intent was to define a universal gasifier that, with minor pre-engineered hardware options, could function well on all U.S. coals and could service a major portion of the perceived new power generation applications. One embodiment of this was the selection of a desired maximum vessel diameter of 12 feet with a diameter limit of 14 feet. This size provides a capability for over-the-road trucking and forms a basis for maximizing the fraction of the plant that can be shop-fabricated as standard components or modules. Considerable system cost reductions have been found to be achievable when standardized components are purchased and major subsystem modules are essentially only connected (as opposed to constructed) at the plant site. The maximum power output from a single gasifier unit train resulting from this design philosophy was estimated to fall in the 70- to 150-MWe range, depending mostly on the specific gasifier configuration and gasifier pressure level. The resulting commercial attractiveness at this unit scale virtually ensured merit for larger-scale systems.

The study was entitled "Commercial Gasifier for IGCC Applications," or CGIA. In several areas, this study raised more questions and issues than it provided answers; however, that was logical considering the scoping nature of the study, and the identification of unknowns is a useful outgrowth. The CGIA study report consists of seven Chapters. Chapter 1 provides the relevant background and defines the study in more specific terms. Chapter 2 defines the design/functional requirements, describes the issues and options, and sets the study Ground Rules. Chapter 3 describes and evaluates relevant existing fixed-bed gasifier designs. Chapter 4 describes and evaluates major design features and options applicable to the fixed-bed gasifier itself. Chapter 5 provides a conceptual design for a fixed-bed "Commercial

Gasifier for IGCC Applications," the reference IGCC system configuration used in the comparative analyses, and descriptions of the major non-gasifier components and system-level options. The methodology and results of the comparative analyses are described in Chapter 6, and Chapter 7 encapsulates the conclusions of the CGIA study. Appendix A includes the assumptions for each of the 12 cases, and Appendix B presents capital cost details as spreadsheets for the individual cases.

Chapter 1

Introduction

1.1 BACKGROUND

One of this nation's inherent strengths is the existence of relatively low-cost and plentiful electric power. However, there are factors that should be considered when contemplating future trends in the cost and availability of electricity. The nation's existing electric capacity (including all modes of generation) was installed at an average capital cost, in as-spent dollars, of slightly less than \$340/kWe (U.S. Department of Energy 1987) - all contemporary types of generating plant significantly exceed that number, often by factors of four or more. The ages of existing plants and the projected growth in electric power demand combine to indicate a need for a significant increase in new generation capacity, particularly in the early years of the next century - less than 15 years away. Concerns about the environmental burdens associated with electric generation (e.g., acid rain and greenhouse effects) effectively dictate a need for all new plants to be very clean relative to past performances. National and local socioeconomic factors make it highly desirable that new sources of electricity utilize U.S. industries and provide a capability to utilize indigenous U.S. resources (e.g., high sulfur coals). Lastly, it typically takes more than 10 years for a technology that has proven itself workable and apparently attractive in a laboratory setting to find its way into a *first* commercial plant.

While there are a number of promising emerging electricity generation technologies, few are both (1) exceptionally clean, and (2) low in cost. One of these emerging technologies that shows very attractive potential is the integrated gasification combined-cycle- (IGCC) based system, a schematic of which is shown in Figure 1. Numerous studies have been done

Figure 1. Schematic of a Generic IGCC System

that predicted great reductions in environmental burdens to be achievable at costs that are competitive with conventional coal-fired boilers through the use of the IGCC approach. Two such reports were presented at recent conferences sponsored by the Electric Power Research Institute (1988, 1990). At least four studies have been published (Notestein 1989; Pitrolo and Bechtel 1988; Corman 1986; Bajura et al. 1986) that show promise of achieving both significant reductions in air emissions and capital costs of under \$1000/kWe through the use of a *simplified* IGCC approach. The simplified IGCC system seeks (1) to maintain nearly constant pressure and temperature levels through the fuel gas preparation portion of the system as a means of minimizing the number of operational units within the overall system and reducing losses, and (2) to size the components such that they are consistent with both shop fabrication (as opposed to field) and contemporary classes of gas turbine (GT) prime movers. These characteristics should provide

- Minimum component cost,
- Minimum construction time (and cost),
- Minimum plant "footprint,"
- Maximum system efficiency, and
- Maximum system reliability.

The key operational characteristics of a simplified IGCC system are (1) it utilizes air derived from the GT compressor to provide air blast to the coal gasifier (through a small boost compressor), (2) the product gas is cleaned and then combusted in the GT without significant cooling or depressurization, and (3) no wastewater streams are produced. The resulting systems are more tightly integrated than most other IGCC system approaches, which has both positive and negative aspects, but they retain essentially all the virtues associated with the IGCC concept (e.g., low fuel cost, extremely low emissions, and high efficiency).

While the simplified form of the IGCC system does not generally achieve the ultra-low emission levels associated with low-temperature, liquid-based, fuel gas scrubbing systems, the resulting emission reductions are much more cost effective and can approach the ultra-low levels. This relationship and its significance to the electric power industry may be understood by examining Figures 2, 3, and 4. These figures are based on the Environmental Protection Agency (EPA) list of 107 power plants that have been targeted for emission reductions, and on President Bush's proposed Clean Air Act emission limitations. These 107 plants are considered to represent a capacity of 56,273 MWe. (This value is 65% of the totaled "nameplate ratings" in 1985, and thus, it is also roughly equivalent to the average summed output of the 107 plants.) Under Phase II of the proposed Clean Air Act, these 107 plants will have to be modified in some form or utilize different fuels, so as to emit sulfur dioxide (SO_2) at a rate of no more than 1.2 lb of SO_2 /MBtu of fuel consumed.

Considering various plant modification scenarios for these 107 plants, Figure 2 shows the total level of power production that could be obtained from these 107 sites while releasing no more than the Phase II, 1.2 lb/MBtu, SO_2 emissions (assuming the currently utilized coals

M90000561

Figure 2. Future Generating Capacity Under the Clean Air Act

M90001888

**Figure 3. Capital Cost Versus SO₂ Removal
(3% Sulfur Coal)**

M90001888

**Figure 4. COE Versus SO₂ Removal
(3% Sulfur Coal)**

continue to be utilized). The first bar shows that if all 107 plants were retrofitted with flue gas desulfurization (FGD) systems (i.e., scrubbers) that were 90% effective (which would also have the effect of increasing the average heat rate from the existing 9,850 Btu/kilowatt hour [kWh] to 10,155 Btu/kWh), plant usage would have to be limited to an average aggregate output of 50,110 MWe in order to maintain total sulfur emissions within the Clean Air Act Phase II limit. The second bar of Figure 2 provides an example of what might be achieved if all 107 plants were retrofitted with a pressurized fluidized-bed combustion (PFBC) process, based on a 92% effective SO₂ removal and a 9,350 Btu/kWh heating rate. The bar shows that the average power generated could be increased to 67,007 MWe for the same Phase II SO₂ emissions. The third bar represents the same analysis but utilizes a simplified IGCC system with a 95% effective SO₂ removal system (which would be representative of a lower limit for hot gas desulfurization) and an 8,110 Btu/kWh heating rate. This bar shows that the total generating rate could be raised to 123,600 MWe. The fourth bar considers the same simplified IGCC system but with a 99.5% effective removal system, which would correspond to an advanced hot-gas desulfurization system, and shows that the total generating capacity can be raised to more than 22 times the baseline rate, or 1,236,041 MWe. The last bar considers a mature cool-water IGCC plant, which could be representative of the ultra-low emission levels attainable with low-temperature, liquid-based, fuel-gas scrubbing systems. This bar is based on a 99.9% effective SO₂ removal system and a heat rate of 9,280 Btu/kWh, and illustrates that an increase of slightly more than two orders of magnitude in generating capacity can be provided without increasing Phase II SO₂ emission levels. It is worth noting that while upgrading plant efficiency does help increase the amount of power that may be produced while still remaining compliant with emission limits, increased sulfur removal effectiveness is much more influential, and the combination of increased sulfur removal effectiveness and increased generating efficiency is far more significant (than improved efficiency only) in terms of providing a capability for system expansion without increased SO₂ burdens.

Figures 3 and 4 provide typical corresponding capital cost and cost of electricity (COE) values for these systems. A 90% sulfur removal level is representative of a conventional pulverized-coal boiler with flue-gas desulfurization (PC-FGD) and appears at the left edge of the figures. Projected costs and performances for several conventional IGCC systems, as well as the anticipated range for simplified IGCC systems, are also shown. While all the IGCC system performances are impressive, it appears the potential cost penalty associated with the conventional IGCC approach is at least 25% and could be as much as 100% of a simplified IGCC system.

Over the past 10 years or so, several systems-level comparisons have led to the conclusion that using fixed-bed gasifiers with hot gas cleaning results in economic parameters comparable to (and often better than) using other types of gasifier in a conventional IGCC application (Chandra et al. 1978; Hackworth and Mann 1989; Electric Power Research Institute 1983, 1986). Furthermore, there are a number of operational realities (e.g., product gas temperature and dust loading) that appear to make fixed-bed gasifiers more readily integrated with low-complexity fuel-gas cleanup subsystems and the types of contemporary gas turbines of interest to electricity producers. While not yet established as state of the art, the needed

techniques and devices for particulate and sulfur removal at conditions consistent with fixed-bed gasifier raw-gas outlet streams are well along in their development process and should be well established in the next 5 to 10 years. Comparable technology for higher temperature gasifiers is much less well developed. Currently, there are a number of approaches for NO_x control, ranging from the state-of-the-art water injection or selective catalytic reduction (SCR), through various approaches for "gas reburning" and "staged" combustion, to the developing techniques of direct ammonia destruction and catalytic chemical cracking. It would appear NO_x control will not be a significant problem. Thus, the fixed-bed gasifier potentially forms an excellent basis for a simplified IGCC system.

It also appears credible that a single gasifier of a size that could be fabricated in a factory and transported by truck could serve an application of 50 MWe to possibly 150 MWe and be reasonable in price. (For reference, the installed cost of a Mark IV Lurgi but without the direct cooler and its associated hardware is about \$200/kWe at the 50- to 70-MWe scale.) A "conventional wisdom" limitation that is related to air blowing of the Lurgi gasifier occurs when an excess of coal fines is blown out of the gasifier and causes problems in the direct cooler. If a fixed-bed gasifier or system could accommodate higher than conventional air blast rates, and if pressure levels higher than those required by a GT are utilized, then significant increases in system electrical output from one gasifier, up to perhaps 250 MWe, are credible. In the world of contemporary GTs, one should recognize that 250 MWe is represented by one General Electric (GE) Frame 7-F or Westinghouse 501-F plus a steam turbine. Furthermore, 150 MWe could be one of these GTs alone or a whole host of other combined-cycle options, and 50 MWe is probably the minimum power block worth considering for a generating capacity addition. It is quite possible that standardization of the gasifier hardware would minimize the unit cost of the gasifier, and various power levels could be accommodated with the same gasifier configuration using various pressure levels.

An approximation of what the simplified IGCC system capital cost should be may be developed as follows:

Gas/steam turbine power-cycle hardware	=	\$400/kWe
Gasifier subsystem producing raw gas	=	\$200/kWe
Hot-gas-cleanup subsystem allowance	=	\$200/kWe
Balance-of-plant allowance	=	<u>\$200/kWe</u>
Total	=	\$1000/kWe

While these cost values are admittedly very rough estimates (and the hot-gas-cleanup cost allowance is the least firm), there is historical evidence to say they are qualitatively reasonable for an IGCC system that has benefitted from a conscious effort to simplify the system and to modularize and standardize major subsystems within it.

1.2 THE STUDY

This was a scoping-level study to examine simplified IGCC systems and, particularly, fixed-bed gasifier designs that would be suited to those systems. Two overall goals were (1) to define one or more systems that were both exceptionally clean from an emissions point of view and also well below the cost of conventional coal plants with flue gas scrubbing, and then (2) to define a fixed-bed gasifier consistent with this type system(s). Other work, some of which is noted in the references previously cited, has shown that considerable system cost advantages can be accrued if standardized components can be purchased and major subsystem modules simply connected on site. To take advantage of this and to the degree possible, this study examined the feasibility of developing standardized components and modules for major subsystems. In particular, the intent was to define a universal gasifier that with minor pre-engineered hardware options, could function well on all U.S. coals and could service a major portion of the perceived new power generation applications. Such a gasifier or system was viewed as having excellent commercial potential. One measure of this potential is the realization that, if system capital costs in the vicinity of \$800/kWe can be achieved (which is not incredible), a simplified IGCC system will produce electricity at a COE that is below that of a natural-gas-fired, combined-cycle plant, even with the assumption that natural gas is priced at under \$3/MBtu (Pitrolo and Bechtel 1988), and it can also approach the natural-gas-fired system's low emissions level! This study was entitled "Commercial Gasifier for IGCC Applications" or CGIA.

The CGIA study was accomplished in three months and consisted of five overlapping activities: (1) Requirements Definition, (2) Definition of System and Major Component Configuration Options, (3) Systems Trade-Off Analyses, (4) Definition of a Conceptual Level Gasifier Design, and (5) Delineation of Issues/Critical Assumptions. These activities were initiated essentially in the sequence indicated, but once initiated, iterations occurred to the point that the activities overlapped nearly continuously during the study's execution. A discussion of the Requirements appears in Chapter 2 of this report; the other topics are covered in subsequent Chapters.

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Chapter 2

The System

2.1 SYSTEM DESCRIPTION

Figure 5 provides a block diagram of the major components and functions of both conventional and simplified IGCC systems. While both of these approaches accomplish all the functions of the generic IGCC system shown in Figure 1 on page 6, they do so in significantly different manners. The upper portion of Figure 5 represents an oxygen-blown, entrained-bed gasifier system and is typical of the low temperature, liquid-based, fuel gas scrubbing approaches when applied to a coal gasifier of virtually any type. Basically, these systems include a significant number of process operations and equipment items relating to heat exchange, gas resaturation, and wastewater handling -- all of which disappear with the simplified IGCC approach. Also, the oxygen plant of the conventional IGCC system is in essence replaced by a boost compressor so that the pressure in the gasification portion of the system is above the pressure requirements of the GT fuel-flow control-valve by the necessary margin. In the ideal simplified IGCC configuration, with the exception of the boost compressor, there are no designed changes in gas pressure or temperature as the gas flows through the system. Comparison of the two system diagrams provides an insight into the potential for cost reductions available with the simplified IGCC approach.

Further potential for cost reductions arise with the consideration of combining functions within one operational block or of designing a key component in one block such that another block in the system may be significantly simplified. An example of this is to design a fixed-bed gasifier to utilize run-of-mine (ROM) coal so that functions like coal grinding and screening and coal slurry preparation, interim storage, and slurry pumping can be eliminated from the coal preparation block.

The specific functions that must be provided by each plant section and typical approaches and options for doing so are described in the following paragraphs, in the context of a simplified IGCC system that is based on a fixed-bed gasifier.

2.1.1 Coal Handling

In the basic system, this plant section provides only coal receiving, pile storage, and reclaiming and conveying functions. The intent is to design the system, and the gasifier in particular, to be able to accommodate whatever amount of fines (defined as coal sized at less than 0.25 inches) is in the coal as it is supplied. This has the effect of lowering the fuel cost and increasing the overall coal-to-electricity conversion efficiency. An upgrading of this plant section to provide features to better accommodate fines may be the economic choice for some

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Figure 5. Comparison of Conventional and Simplified IGCC Systems

U.S. coals; it could also be dictated by local air quality regulations relating to fugitive emissions. The essential modification would be to add a screening unit to provide a fine coal stream that is separated from the main coal stream. This fine coal stream may then go to a separate portion of the plant, it may be routed to a separate feeding port of the gasifier, it may be agglomerated within the coal handling section, or it may simply be handled differently so as to reduce fugitive emissions or fire hazards (e.g., through the use of an enclosed storage silo).

2.1.2 Gasification

This plant section takes in coal, steam, and high-pressure air to produce raw fuel gas and ash. Air is supplied to the gasifier by withdrawing a fraction of the GT compressor discharge stream, which is essentially at GT combustor pressure (and at an elevated temperature), and raising its pressure to the gasifier's requirement. This section, consequently, also includes a boost compressor to provide a modest pressure rise (a compression ratio in the range of 1.3 to 2.0 will cover all eventualities) to allow for pressure drops that occur through the gasifier and the cleanup system and that are required for a control margin across the GT fuel-flow control-valve. The system is enhanced by minimization of this boost pressure ratio.

The downstream portion of the total system can be simplified to the degree the gasifier is designed to minimize the quantities of contaminants in the gas stream: particulates and dust, sulfur, and nitrogen-bearing compounds. Condensible hydrocarbons (i.e., tars) are the contaminants of most concern. Minimization of dust carry-over can be accomplished by not feeding fines to the region of the gasifier near the gas outlet port, by minimizing gas velocities in the vicinity of the outlet port, or both. While in-bed sulfur sorbents can be made to work to some degree in fixed-bed gasifiers, none (and no technique for their use) have been found to be effective from an economic standpoint, and their usage prompts a host of design and operational uncertainties (Bissett and Strickland 1989). Consequently, it is expected that all gaseous sulfur species will have to be dealt with downstream of the gasifier. Tars represent both potential operational problems and also potential sources of sulfur and nitrogen that, if unchecked, will contribute noticeably to air emissions. Nitrogen-bearing compounds (primarily ammonia) are of concern because of their propensity for forming NO_x during the fuel-gas combustion process. Most of the tars will chemically decompose (i.e., crack), given an adequate residence time at elevated temperature in the presence of a chemically active solid surface (e.g., hot char); and the products of decomposition are more easily dealt with, or even beneficial, in comparison to the parent tar. However, most tar- and nitrogen-bearing compounds are released from the coal as it warms shortly after entering the gasifier and, consequently, little tar cracking takes place deep within the gasifier itself. One approach to tar reduction is to cause the product gases to flow through hot regions of the coal bed for a sufficient residence time to provide an adequate amount of cracking; this is a basic premise behind the two-ported gasifier in which the top gas is recycled to the lower portion of the gasifier, and the raw product gas is extracted from a mid-level port. Tar can also be effectively dealt

with downstream of the gasifier, and the gaseous nitrogen species are typically controlled within the combustion process itself.

In a fixed-bed gasifier, the ash is generated in a hot, oxidizing zone that limits the possibility of environmentally significant or hazardous compounds existing in a leachable state within the ash. Further, the ash layer in a dry-bottom or non-slugging gasifier is cooled as a result of heating the rising air- and steam-blast gases; consequently, the ash is typically a "warm gravel" on exit from the gasifier and poses no special concerns. The slagging-bottom, fixed-bed gasifier is a special case, and it generates a glass-like, granulated ash that is very environmentally inert and benign. This option is discussed in more detail in subsequent chapters.

2.1.3 Gas Cleanup

The gas cleanup section takes in raw product gas from the gasifier and cleans it such that the resulting fuel gas complies with the needs of both the GT, initially, and the environment, ultimately. This section will also take in air for sorbent regeneration, and typically produces effluent solids (e.g., the dust catch from cyclones), flue gas streams that are virtually sulfur free, and possibly liquid elemental sulfur. As mentioned earlier, the cleanup process is designed to incur as few temperature and pressure changes as possible in order to maximize the overall system efficiency. This plant section overtly deals with the product-gas stream's contaminants: particulates and dust, sulfur-bearing compounds, and condensible hydrocarbons (or tar). In a conventional IGCC plant, this is usually the most expensive and extensive plant section by far, and while it is not as dominant in the simplified IGCC plant, it is still a very significant section.

Two of the contaminants, the dust and tars, are relatively easily dealt with. Unless incredibly high air- and steam-blast rates are utilized, fixed-bed gasifiers have lower rates of dust carry-over into the product gas, and the dust is at larger sizes than in other types of gasifiers. More to the point, an adequate level of dust removal is within the state of the art for hot, pressurized cyclones -- which are not expensive devices. The separated dust has the character of finely ground coal, which is slightly devolatilized (and dried) and very warm, and represents no problem when treated accordingly. It can be disposed of as a solid waste or, preferably, accommodated within the plant as a coal fines stream, as was discussed in the paragraphs above on coal handling. The key to handling the condensible hydrocarbon vapors (tar) is to keep the stream hot such that these vapors do not condense. Given that the stream is kept hot, for many U.S. coals and for existing emission standards, the presence of tars can effectively be ignored. However, for high volatile, high sulfur, eastern coals, the tar can contain enough sulfur to preclude achievement of really low SO_x emissions unless the tar is removed or destroyed. Similarly, there are nitrogenous species within the tar molecules that will result in NO_x emissions upon combustion. Fortunately, the potential emission contribution resulting from tar-borne nitrogen is not large since physical removal of the tar from the gas stream is currently the only means to realistically limit this component of the fuel-bound

nitrogen is passed to the combustor. This occurs because, while tar can be cracked to release gaseous sulfur and nitrogen species, the resulting ammonia and cyanide compounds (related to the tar's nitrogen content) will pass on through the system to the GT combustor while the hydrogen sulfide, carbonyl sulfide, etc. (from the tar's sulfur content), will be removed by the desulfurization units within this plant section. However, work on both ammonia cracking catalysts and modification of the GT combustion process for emissions reduction is showing considerable promise, and may cause this comment to be revised in the next few years.

In contrast, gas-phase sulfur compounds can be dealt with quite effectively. The technology base for sorbing sulfur compounds onto mixed metal oxides at product-gas stream conditions is well developed, and most of the experience base exists with a zinc oxide/iron oxide compound (zinc ferrite). In effect, zinc ferrite removes nearly all the gaseous sulfur species. The precise phenomena by which this occurs is not currently known. However, there is a possible model wherein the zinc ferrite removes hydrogen sulfide from the gas stream and then catalyzes the remaining species to react (shift) in an attempt to restore the relative hydrogen sulfide concentration in the gas stream while the hydrogen sulfide continues to be removed by the zinc ferrite. The overall removal process then becomes limited by the ability of the existing sulfur-bearing species to shift to hydrogen sulfide. The consequence of this removal is the formation of solid metal sulfides that are chemically bonded within the porous zinc-ferrite-sorbent particle. When the sorbent particle has become sufficiently loaded with sulfides to reduce its sorption effectiveness, the sorbent is isolated from the coal gas stream and is regenerated with an oxygen-bearing gas, typically air, to "burn off" the sulfur and restore its effectiveness as a sorbent. Zinc ferrite has a range of effective operational temperatures for sulfur sorption of 900 to 1200 °F and has demonstrated an ability to produce cleaned gas sulfur levels in the range of 10 to 20 parts per million by volume (ppmv) when employed in fixed-bed absorbers operating on actual coal gas; this level is also essentially independent of the input raw-gas sulfur level. There are other sorbent formulations that while less well developed, show considerable promise. Foremost among them is zinc titanate, which has shown an ability to reduce cleaned gas sulfur levels to the 1 to 10 ppmv range (in a laboratory setting using a synthetic coal gas), and is effective up to 1350 °F.

The sorbent regeneration step is essentially a combustion process, and the sorbent must be held below about 1500 °F (for zinc ferrite) to preclude a level of sintering that would reduce its effectiveness during subsequent sorption periods. The regeneration offgas is the means by which sulfur leaves the fuel stream of the overall system. This offgas stream contains hot SO_2 , and the regeneration step can be designed so the offgas is directly suitable for the production of sulfuric acid. The offgas can also be reacted with limestone to produce a disposable solid waste, or it can be reduced to elemental sulfur.

The alkali content of the fuel gas stream is not of environmental concern but is a traditional concern for the GT. One of the inherent characteristics of most fixed-bed gasifiers is that the product gas temperatures are low enough that the vast majority of the alkalis are condensed on the particulate exiting the gasifier, rather than being in the vapor phase.

Consequently, removal of the particulate also results in removal of the alkali, and virtual elimination of a potential problem for the GT.

2.1.4 Power Production

This section takes in the hot, pressurized, cleaned coal gas and burns it in the GT combustor. The hot GT exhaust gases are then passed through a heat recovery steam generator (HRSG) to raise steam for the rest of the system and to cool the exhaust to the point it becomes an acceptable flue gas. Product steam from the HRSG is primarily utilized by the gasifier and the steam cycle, if present as in a combined-cycle power system. In the combined-cycle configuration, electric power is generated from output shafts of both the GT and the steam turbine. The thermodynamic waste heat leaves the overall system as the sum of the energy in the flue gas exiting the HRSG and the steam cycle condenser heat rejection. A further discussion of combined-cycle power systems is beyond the scope of this report.

The fuel gas supplied to the GT will be at desulfurization sorbent-bed exit conditions and, for the base cases, essentially tar-free. The following would be typical characteristics:

Temperature:	= 1100 °F
Hydrogen:	= 18 vol %
Carbon monoxide:	= 9 vol %
Methane:	= 4.5 vol %
Other hydrocarbons:	= 1 vol %
Nitrogen plus argon:	= 36 vol %
Water vapor:	= 16 vol %
Carbon dioxide:	= 15 vol %
Ammonia, etc.:	= 4500 ppmv
Hydrogen sulfide, etc.:	= 15 ppmv
Particulate > 10 µm:	= essentially none
Particulate < 10 µm:	= < 70 ppmw
Lower heating value:	= 141 Btu/scf

These values will change with the coal utilized and with the operating conditions of the gasification process, but they may be considered representative.

The state of the art with respect to GT fuel-flow control valves is described as "being extensible to become capable of accommodating 1200 °F gas temperatures," and pending further developments, this value should be considered as representing a maximum for the supplied fuel gas temperature. Combustion tests at GT combustor conditions have shown this type of gas can be successfully burned with only minor modifications to the existing types of GT combustors. The key fuel gas parameters are the hydrogen content and the gas temperature. Hydrogen concentration in the fuel gas has a lot to do with its ignitibility characteristics, and while one could theoretically tolerate a lower level, 7 or 8% (by volume) probably

should be considered the functional minimum hydrogen concentration. Hot gases are much easier to burn than cool gases, and the "over 900 °F" fuel gas temperature is a definite assist. Carbon monoxide has a very high flame speed (comparable to hydrogen), which is an asset, but it has a slightly higher ignition limit at low concentrations and so is not all that effective as an ignition aid. Methane contributes a relatively high heating value, but otherwise is not a particularly strong participant in the combustion process.

As noted earlier, there will be reactive nitrogenous compounds (mostly ammonia) in the fuel gas, and their concentrations will correlate to the amount of nitrogen in the parent coal. For most U.S. coals, this level of fuel-bound nitrogen will cause NO_x emission limits to be exceeded if all of the fuel-bound nitrogen is converted to NO_x in the GT combustion process. As a consequence, it is expected that most of these systems will employ either gas turbines with some form of low NO_x combustor or will utilize an SCR process at an appropriate location within the GT exhaust gas stream. Combustion testing with high levels of fuel-bound nitrogen (e.g., simulated coal-gasifier product gas) with various types of staged GT combustors has produced very low levels of NO_x emissions, which is quite encouraging; however, more discussion of the gas turbine combustion process phenomena is beyond the scope of this report.

2.2 ISSUES, OPTIONS, AND CONCERNS

The highly integrated nature of the simplified IGCC system approach results in significant potential advantages, most of which are economic, and it introduces a host of new design options and operational nuances that must be understood to get the most out of the concept. This section discusses the primary design options and the associated technical issues and concerns.

Two typical characteristics of these options are (1) they affect more than one plant section, and (2) they can represent significant cost or operability impacts. Unfortunately, in some of the technical areas, a definitive discussion is not possible because achievable performances and cost data were not predictable with existing tools or within the available time and effort. Generally, these areas can be categorized as (1) design tradeoff areas that were not explored (e.g., the specific tailoring to optimize the system designs over the spectrum of U.S. coals), or (2) areas where the data to make definitive predictions were lacking (e.g., the cost penalty/advantage to utilizing a gasifier pressure that was significantly higher than what the GT requires). However, this section defines the key issues and concerns and should provide useful background to future workers in the field.

2.2.1 Coal Fines

The cost of feedstock coal normally is reduced if it can be utilized by the plant on a ROM basis. However, this typically entails accommodating a large fraction of fine coal that in turn may have a large fraction of the coal essentially as dust. The amount of fines is primarily a function of both the coal type and the mining technique; eastern U.S. coals and surface mining generally producing the lesser quantities of fines. The concern with fines stems primarily from the difficulty in utilizing its energy value in a fixed-bed process. To maximize the energy output from the gasifier, in effect one wants to maximize the amount of gas (i.e., velocity) leaving the top of the coal bed, which means that if fines are fed to the top of the bed with the rest of the coal, this operating mode also maximizes the amount of fines carried out by the product gas. Similarly, if the coal is particularly friable or thermally decrepitate, a significant amount of fines can be generated while passing through the feeding system or during the initial warming period in the bed, both of which exacerbate the "fines problem." The entrained fine coal represents a contamination of the raw product-gas stream and undergoes little more than drying and a slight amount of devolatilizing before it is removed, typically by cyclonic separators. Consequently, the byproduct dust production associated with a counterflow, fixed-bed gasifier is a function of (1) the basic fines content of the feed coal, (2) the coal type, (3) the gas velocities in the above-bed region of the gasifier, and (4) the gas pressure and temperature.

The simplified IGCC approach lacks downstream liquid process loops and the associated level of susceptibility to solids contamination problems, so whether dust production becomes a limit on gasifier operation in effect only depends on the economics related to fines disposition. Since the byproduct fines are very similar to the as-supplied coal fines, there is essentially one set of options for dealing with all coal fines:

- *Fines Separation:* Screen fines out of the coal as supplied to the plant to create a fines stream for off-site disposal, agglomeration, fines injection, or indirect or non-gasifier utilization (e.g., combustion to provide heat for coal drying or to raise steam).
- *Off-Site Disposal:* Coal that has undergone little or no chemical alteration can normally be disposed of by landfilling, e.g., returned to the mine from whence it came. This is usually an uneconomic option. In some special cases, the fines could be sold, but this is expected to be an unusual situation.
- *Fines Agglomeration:* Pelletize, briquette, or extrude the fines into lumps suitable for feeding to the gasifier. This adds costs to the Coal Handling Section of the plant, but is conceptually simple to implement and likely results in the highest coal-to-electricity conversion efficiency.
- *Fines Reinjection:* Inject the fines into the gasifier at a location deep enough in the bed that they can be expected to react rather than simply being blown out of the bed. The injection could occur at either the *gasification* or *combustion* level of the bed, but

injection into the gasification zone is far more practical (because of bed temperature levels and the probable lack of problems if a fines maldistribution occurs). Injection into the combustion zone has been proven viable by British Gas/Lurgi (BGL) for a slagging bottom design. The approach would appear to not be very costly, and it could approximate or equal the coal-to-electricity conversion efficiency of agglomeration.

- *Indirect Utilization:* The typical fixed-bed gasifier utilizes a combustion process occurring at the bottom to provide heat to drive the gasification reactions occurring above. Conceptually, this could be accomplished, at least in part, by a pressurized coal fines combustor providing hot products of combustion (POC) to the bottom of the gasifier. However, with this approach there are significant design issues and concerns relative to the resulting product-gas heating value and achievable temperature levels. The basic design issues include achieving a controllable, pressurized, external combustion process; providing adequate POC distribution across the gasifier diameter; and predicting the effect on gasifier operations if the external heat source is to be the major heat source for the gasifier (e.g., inertness of the bottom ash). This appears to be a relatively complex approach, which is probably best suited as a partial heat source to a slagging-bottom gasifier design (but it was not thoroughly evaluated). It also appears likely that some level of product gas dilution (by POC) will occur with this approach, which will both lower the product-gas heating value and raise the gas temperature.
- *Non-Gasifier Utilization:* Coal fines are a logical fuel for a pulverized coal boiler (assuming one exists on site). If sulfur and nitrogen levels are low enough, fines could also serve as a fuel for either coal drying or for a fired HRSG. One potentially interesting application is as a fuel for "reburning" within the HRSG with the intent of reducing NO_x emissions for the whole system (this is a totally untried concept at present). There are also possible uses for fines within the Gas Cleanup Section as either a carbon source in the production of elemental sulfur (e.g., RESOX) or as a heat source to drive calcium-based chemistry.

In summary, coal fines will exist within the system. They are likely to come in with the feed coal, and they certainly will be a byproduct of the process. The preferred fines disposition or application depends on the quantity to be accommodated and the consequent relative economics of the various alternatives. It appears prudent to design a gasifier to produce the smallest quantity of fines for a given gas output, and to have a means of accommodating reinjection of the fines in some form, if at all possible.

2.2.2 Boost-Pressure Level

The boost compressor pressurizes the GT's compressor discharge air to provide air to the gasifier at a pressure high enough to overcome the pressure losses through the gasifier and

cleanup subsystems so as to supply product gas at the required pressure to the inlet to the GT's fuel-flow control valve (which must be above actual combustor pressure levels by about 15%). It is likely that a boost-compressor inlet-to-outlet pressure ratio of 1.3 would be nominally correct. The GT's compressor discharge air is at temperature levels of between 600 and 1000 °F and pressure levels of 150 to 450 psig (depending on the specific type of GT in the system), and there is no thermodynamic incentive to deliver cooled air to the gasifier (because of the attendant loss of useful energy). Consequently, the boost compressor can be typified as a low pressure ratio, high mass flow device that operates at a high inlet pressure and has a high inlet temperature. This compressor is technologically closest to the types of compressors associated with catalytic cracking operations in the petrochemical industry; however, at the highest temperatures, it will barely be within the state of the art. The machine is not expected to be highly efficient (because of the high inlet temperature and the lack of incentive for intercooling), but this is not a major consequence because most of the inefficiency is likely to show up as increased energy supplied to the gasifier. The key question is, can the compressor be built for a reasonable cost?

The principal key to configuring a single gasifier's geometry to serve a wide range of power output levels is to utilize boost compressors of higher pressure ratio for systems that are to have higher power outputs. Since the gasification and cleanup portions of the system are costed and function primarily on the basis of actual cubic feet of gas throughput, the effect of a "higher than necessary" pressure level is to pack more Btus into each actual cubic foot of gas produced. (That is, power output is directly dependent on pressure but capital cost is roughly constant.) The increased pressure differential can be dissipated across the fuel-flow control valve or, if it is really large (e.g., more than 200 psi), it can be dissipated across an expansion turbine that would also contribute power to the system. This latter option rarely makes sense because the pressure levels must be very high to make the useable expansion ratios (which produce the useful output work) high enough to be worth the expenditure -- and expansion turbines are relatively expensive. Again, these turbines are technologically similar to units found in the catalytic cracking portions of many petrochemical plants. The questions of how much the operational bandwidth of an IGCC can be enlarged by over-boosting, what over-boosting costs in terms of overall system performance and economics, and at what point does an additional gasifier become cheaper than the cost of an expansion turbine plus the incremental portion of a boost compressor have not been explored.

2.2.3 Gasifier Control Functions and Instrumentation

To be commercially viable in today's world, an IGCC system should be capable of running "automatically," with a control system managing the specific operations based on appropriate external inputs and parameters that are sensed within the system. This is within the state of the art for all components of a simplified IGCC system except the gasifier. The gasifier has five essential functions that require a control methodology:

- *Ash Removal Rate:* This is modulated to maintain the combustion zone of the gasifier in a relatively fixed vertical location. With a dry-bottom gasifier, this is accomplished by controlling grate rotational speed, and with a slagging-bottom gasifier, it is accomplished by controlling the volume of liquid slag allowed to leave the gasifier. In the dry-bottom gasifier, two important functions of the ash bed are to serve as an insulating layer to protect the grate from excessive temperatures and to assist in distributing the rising reactant gases across the gasifier diameter. Both of these functions require the maintenance of a relatively constant thickness of ash below the combustion zone, and the key to this is a technique to sense the location of the combustion zone. While armored thermocouples both on the top of the grate and on the bottom arm of a deep-bed stirrer have provided enough of a combustion zone sensation to allow hundreds of hours of controlled operation of the fixed-bed gasifier at METC, it is doubtful these techniques have the inherent reliability necessary for a commercial application. In contrast, the location of the combustion zone in a slagging gasifier is readily established by the tuyere locations; however, management of the ash removal from a pressurized, slagging gasifier is an esoteric process. It is usually controlled by a combination of the designed tap-hole diameter and temperature, gas pressure balance between the slag quench hopper and gasifier, slag tapping frequency and period duration, and slag temperature. Achieving a controlled, reliable ash removal function is the most critical developmental area related to gasifier control!
- *Coal Feed Rate:* This is modulated to maintain a fixed top-of-bed level or location within the gasifier. Historically, executing this function has not been a problem if the bed height can be measured reliably. Gamma ray densitometry has been a workable technique and is likely to be so for a commercial-scale gasifier.
- *Bed Stirring Protocol:* Slow stirring of the upper portions of the bed maintains a relatively uniform bed gas permeability, and for caking coals, can also be designed to preclude the formation of significant agglomerates, gasifier interior bridging, or clinkers. The protocol is expected to be determined empirically and fixed for each coal or family of coals.
- *Air Blast Rate:* Basically, this is modulated to maintain a constant gasifier pressure. However, it is also a primary determinant of the exit gas temperature and heating value during steady-state operations; consequently, the air flow rate will be based on all three parameters. The specific relationships to be utilized in an automatic control system have yet to be defined; however, a simplification is that all the needed parameters are readily sensed.
- *Steam Flow Rate:* This is normally slaved to the air flow rate by a constant ratio. While it does enter into the gasification reactions, the major function of the steam flow is to moderate the combustion reactions. (In a dry-bottom gasifier, this moderation is both the means to prevent clinkering and to cool the grate.) As such, the

steam-to-air ratio is expected to be empirically derived and to be a function of the specific gasifier design and the coal utilized.

Three control functions in the above list have characteristics that make them potentially good topic areas where the application of control techniques based on the "neural network" subset of artificial intelligence will prove useful (in sequence of suitability): air blast, ash removal, and bed stirring.

2.2.4 Temperature of the Raw Fuel Gas

As mentioned earlier, fuel desulfurization by zinc ferrite or other high performance, mixed metal oxides will occur at temperatures above 900 °F, and the combination of a lack of an effective and economic desulfurization media and the very high-temperature fuel-flow control-valve technology will limit the cleaned fuel gas temperatures to levels below 1600 °F for likely the next 10 years. Currently, 1200 °F would have to be considered the maximum acceptable raw gas temperature if intentional cooling were not to be utilized. A single-stage fixed-bed gasifier operating on a bituminous or higher rank coal will produce a raw fuel gas whose temperature falls within the 900 to 1200 °F range. However, lower rank coals and very high moisture feedstocks in general tend to result in gasifier outlet temperatures that are well below these levels, sometimes as low as 400 °F. In addition, an air-blown slagging gasifier and some two-stage configurations are likely to have outlet temperatures well above 1200 °F. How energy exchange is handled to bring the fuel gas into consistency with the desulfurization and GT subsystems' requirements can be expected to have a significant effect on overall system costs and thermodynamic performance. For example, cooling the gas by generating a heated fluid stream at temperatures much below 950 °F is certainly not going to be thermodynamically attractive. Similarly, heat exchangers for operation in a high-pressure, dirty, and potentially corrosive atmosphere are not likely to be low in cost or particularly reliable.

Reducing excessive raw gas temperatures by incorporating a partial water quench is probably the most practical cooling technique. It requires relatively little water and is quite low in cost to implement. While a partial water quench will lower the gas heating value somewhat, the increased mass of steam in the fuel gas will enhance GT power output (and perhaps lower NO_x production). In addition, for zinc ferrite desulfurization technology, a steam content in the range of 15 to 20% is desired to chemically stabilize the zinc-ferrite iron fraction, and while water addition is normally not required to reach these levels when using fixed-bed gasifiers, the fixed-bed configurations with high output gas temperatures also tend to produce drier gas. It is also quite probable that ongoing sorbent-chemistry development work will reduce the need for a minimum steam content.

Raising low raw gas temperatures is only slightly more complex. The simplest means is to introduce a little air and let the resulting partial combustion achieve the desired temperatures. In this case, inadequate steam levels are never an issue because the lowered gas

temperatures are a direct result of high moisture feedstocks; however, the consequent reduction in fuel-gas heating value can be of concern. The more thermodynamically elegant technique for raising the raw gas temperature is to dry the feed coal, at least partially, before gasification. This does not require high quality heat, and the HRSG can readily serve as the heat source, typically through direct use of a fraction of the flue gas stream or by incorporating an air heater into the HRSG. Since the coals of concern for this case are typically low-rank western coals that are also typically quite low in cost, a desirable alternative may well be to dry the gasifier feed coal with a conventional fired heater using coal as fuel. As discussed above, this is also a possible use for an excess of coal fines. Unfortunately, the drying approaches are all likely to entail more capital cost than the partial combustion approach, but the specific impacts have not been evaluated.

2.2.5 Desulfurization-Bed Operating Mode

Desulfurization using a bed of mixed metal oxides is the key process that provides a basis for the immense potential of the simplified IGCC concept. There are three potentially credible bed operating modes for the sorption step of the process: fixed-bed, moving-bed, and fluidized-bed. Each mode has a set of advantages and disadvantages, and this sequence is in the approximate order of (1) increasing requirements for physical "hardiness" of the sorbent, (2) increasing gas throughput capability, and (3) decreasing desulfurization effectiveness.

The fixed-bed approach is the simplest and utilizes multiple vessels, only one (or some) of which is valved onto and desulfurizing the fuel gas stream at any one time. While the desulfurization step is in progress, at least one other vessel is being regenerated or is in a stand-by status for subsequent desulfurization usage. This approach utilizes a relatively large number of valves to effectively shift the gas stream between a minimum of two vessels containing a stationary sorbent. The actual sorption takes place in a narrow reaction zone, which progressively moves through the vessel toward the gas outlet end. As a consequence, outlet levels of gas-phase sulfur approach the theoretical minimum, and the fraction of the vessel volume being utilized for chemical reactions at any one moment is quite low.

The moving-bed approach effectively duplicates fixed-bed chemical phenomena, but does so by providing a solids bed that slowly moves downward and sequentially moves the sorbent through the countercurrent fuel and regeneration gas streams. The rate of sorbent flow is matched to the rate of sorbent reaction-zone progression such that the reaction zone remains at a spatially constant location within the vessel. This greatly reduces the number of valves required in the system, somewhat increases the ability to utilize reaction vessel volume efficiently, and considerably eases the materials selection process because hardware is not exposed to alternating reducing and oxidizing atmospheres. However, more care is required to provide pneumatic isolation between fuel and regeneration gas streams, and since these are typically gravity flow devices, some form of hot solids "lifting" technique is required for sorbent recycle. Desulfurization effectiveness is virtually as high as in the fixed-bed

approach, but the physical abuse of the sorbent (caused by its constant movement) and the potential for dust generation are markedly higher.

The last approach utilizes a fluidized sorbent bed to greatly increase the effectiveness of gas-solids contacting over the fixed- and moving-bed approaches. The macroscopic solids flow generally resembles the moving-bed approach. However, in this case, the bed would be lightly fluidized (a gas velocity only modestly above minimum fluidization) to minimize the physical abuse of the sorbent, but it would be fluidized well enough to provide a free flow of sorbent through and out of the desulfurization bed. While the increased effectiveness of gas-solids contacting can greatly reduce the desulfurization vessel size, the inherent back mixing results in less complete sorbent utilization and outlet gas sulfur levels that are a factor of perhaps 5 to 10 higher than occur with either fixed- or moving-bed desulfurization. The increased sulfur levels are still not high but represent a "lower than attainable" environmental performance. The key issue with this approach is the degree to which fluidization physically breaks up the sorbent, causing dust to enter the fuel gas stream and increasing the sorbent replacement rate and costs.

2.2.6 Desulfurization Sorbent Regeneration

The most important step in the hot gas desulfurization approach, and the one which really determines the value of its contribution to the overall simplified IGCC concept, is the regeneration step. In this step sulfided sorbent is oxidized to restore its desulfurization effectiveness and to cause the sulfur originally introduced with the feed coal to leave the system. Conceptually, all of the bed operating modes discussed above for desulfurization are useable for regeneration, and the comments are generally applicable. In theory, with the exception of the fixed-bed approach, the bed operational modes do not have to be the same for desulfurization and regeneration; however, practical considerations (such as the size of the sorbent particle desired for the absorption bed) make it probable that both absorption and regeneration beds will operate in the same mode. A major difference is that the regeneration step is significantly exothermic, and there is a maximum temperature (about 1500 °F for zinc ferrite) that is not difficult to reach and, if exceeded, can greatly degrade sorbent porosity or its physical integrity and, thus, its subsequent effectiveness for desulfurization. In addition, some sorbents (specifically, zinc ferrite) have a tendency to form sulfates during oxidative regeneration; if unchecked, this results in a large pulse of SO₂ being released into the fuel gas stream when the sorbent is initially placed back in the desulfurization mode. Probably more important for an IGCC application using zinc ferrite is the fact that the reaction of fuel gas with sulfated zinc ferrite is both very fast and exothermic. The result is that (1) large amounts of water vapor and CO₂ are formed, reflected in a transient but massive reduction in the fuel-gas heating value (and a possible flameout of the GT); and (2) the zinc ferrite decrepitates to the point of becoming unusable. The solution is to perform a last stage of sorbent regeneration in a reducing atmosphere (i.e., "reductive" regeneration, described below). The product of regeneration is a hot offgas stream, relatively rich in SO₂, which must be dealt with in an environmentally and economically acceptable manner. The degree of richness is the key

determinant in what technologies are applicable and how the system economics will be affected; as a generalization, the *richer* the stream, the *better* the economics.

As a consequence of all this, and recognizing that regeneration, like sorption, takes place in a narrow reaction zone in a fixed-bed reactor, it becomes obvious that the oxygen stream used for regeneration must be accompanied by an inert diluent or an aggressive approach to heat removal from the regeneration reaction zone. This is most easily accomplished through utilization of a fluidized-bed regeneration vessel (which could also readily accommodate internal heat removal surfaces); however, both moving-bed and fixed-bed regenerators (which require greater degrees of dilution) have been designed to function effectively. Typically, air is the oxygen-bearing stream, and steam or recycled regeneration offgas is utilized as the diluent. In a concept described by GE (Smith 1988), recycled SO_2 was utilized as a diluent to provide an offgas stream that was directly compatible with sulfuric acid synthesis. For sorbents susceptible to sulfate formation (like zinc ferrite), the final phase of regeneration should be done in a relatively sulfur-free, reducing gas atmosphere. Functionally, this is not a large gas requirement and cleaned fuel gas is entirely suitable, but the system must be designed to provide the gas. Fixed- and fluidized-bed regeneration approaches have, respectively, been found to be the most and least susceptible to operational problems caused by sulfate formation with zinc ferrite sorbents.

The most broadly applicable regeneration offgas disposition approaches are (a) reduction to elemental sulfur, or (b) reaction with calcium species to form gypsum; in either case, the end product can either be landfilled or sold. Two specific versions of these approaches were examined in this study and are discussed in following chapters. However, regeneration offgas disposition is a technically large subject, and a reasonable coverage of it was beyond the scope of this study.

2.2.7 Combined Function Devices and Techniques

As these systems and the supporting technologies develop, there is potential for further improvement of the system economics and performance (thermodynamic or environmental) by accomplishing more than one function in a single process unit. A good example in the relatively near term is the addition of a catalyst to the desulfurization sorbent to chemically crack tars and free the tar-bound sulfur for removal by the desulfurization sorbent. Given a physically hardy sorbent, the role of the desulfurization bed could be further extended to serve as a granular-bed particle filtration device. Use of the exothermic regeneration process as a heat source for application elsewhere in the system is another relatively near-term example. Techniques that allow the gasifier to process increased quantities of fines or provide for the total consumption of all fines and process dusts also have obvious potential.

Generally, application of these combined function approaches represents an advancement of the base technologies employed in the simplified IGCC concept. Consequently, they

are expected to appear in the maturation stages of this system and after successful experience is obtained with the base technologies.

2.3 GROUND RULES

Some ground rules had to be established to keep the scope of the study within manageable bounds, and this section presents and discusses the Ground Rules that evolved. The background and rationale for many of the specific Ground Rules were provided by the above discussions.

The basic philosophy and goal of the CGIA study was to explore the viability of a simplified IGCC approach in the context of its being able to make a meaningful contribution to the nation. The most credible route to this goal is to

- Provide the system configurations that minimize both the capital cost and the COE. It is expected these systems will need to be significantly lower in cost than other coal- or oil-based generating technologies and at least competitive with natural-gas-based systems.
- Provide system configurations with emissions that are comfortably below applicable air emission standards and that have the capability to be readily improved to levels significantly below existing standards.
- Provide system configurations that employ technology that is mature enough so "first plants" could be purchased sometime between 1995 and 2000.

As a result of this last step or characteristic, the technologies that will be considered must be currently existing or expected to be developed over the next few years, such that a banker could be successfully approached in the 1992 to 1998 time frame to participate in one of these projects. The list of characteristics also allows the possibility that a commercially desirable simplified IGCC concept could have a *plain vanilla*, near-term embodiment, which would form a basis for evolving the design by maturation stages into the ultimate commercially desirable configuration. Each maturation stage would represent a family of product improvements and an improvement in the overall economic and technical performance of the system. Given a positive outcome of the CGIA study and ensuing development and commercializing activities, it could reasonably be expected that significant numbers of simplified IGCC systems will find their way into the utility sector in the 2000 to 2020 time frame.

The specific Ground Rules and goals and some discussion of the basis for each is provided by the following.

1. The system capital cost and the COE will be minimized. A functional ceiling for capital cost is to be \$1,000/kWe, assuming the owner's "green-field, no-problem," site and an Nth plant design and construction process.
2. The system's maximum emissions of SO_x and NO_x will each be 0.1 lb/MBtu of coal input to the process. This represents a level of 97.8% SO_x removal and 95.6% NO_x removal for the reference coal. It is expected (but not known) to be near a break point where achieving an additional level of pollutant removal will be considerably more expensive per unit removed. These are to be total system air emissions, considering all source streams. As an example, this Ground Rule will have the effect of requiring tar cracking for essentially all U.S. coals. An alternative will also be examined wherein an emissions limit of 0.5 lb/MBtu for SO_x and NO_x will be utilized. It is expected that this level will be below the New Source Performance Standard (NSPS) limit (e.g., an NSPS limit value of 0.6 lb/MBtu of coal input to the process is most generally applicable), and will allow some simplification of the process (and associated cost reductions).
3. Technology levels to be utilized will be restrained to the point that only those technologies that could be credibly developed and commercialized by the 2000 to 2010 time frame will be considered. This has aspects of technology risk: what "entity" will logically be the technology developers, and what are the routes to commercialization? All of these must be considered.
4. The coal gasifier will be an air-blown, fixed-bed type. It can be dry-bottom or slagging, and can have novel features to reduce tar production, facilitate the use of fines, and reduce the ultimate SO_x and NO_x emissions. The gasifiers offered by Lurgi and BGL are considered as departure points. It is recognized that an early system configuration could utilize a more conventional configuration of the gasifier if the configuration could logically be extended to the more ultimate system, and the early system could provide data that would be representative of the ultimate system. The purpose of the study is to define the characteristics of this ultimate gasifier (and to understand the system into which it fits).
5. The system is to be designed to utilize all U.S. subbituminous through bituminous coals (and on an as-mined basis). It is understood that the performance, output, and operating conditions of a system will change as a function of the coal utilized; however, the configuration of that portion of the system involving the major cost pieces of hardware should be expected to remain identical. The intent is to have the gasifier itself evolve into one specific configuration with one operational pressure rating that in combination with various gas turbines, would provide the capability to cover the range of U.S. coals and all power levels. It may turn out there is a cost advantage to have two specific (but very similar) gasifier configurations (coal dependent) and two operational pressure level ratings (primarily power-level dependent).

6. A reference coal was selected to facilitate this study, and its properties are defined in Table 1. This coal is like a Pittsburgh seam coal but has been modified slightly to be generically representative of the bulk of coals consumed by utilities east of the Mississippi River. This coal will be difficult to run in a fixed-bed gasifier; thus, the resulting gasifier design should be capable of using virtually any U.S. coal. System performance levels are planned to be evaluated by considering variations from this reference coal.
7. The power subsystem is to utilize either a steam turbine (ST) and gas turbine combined cycle (STAG) or a steam-injected gas turbine (STIG) power block; the total output of one system module is to be 50 to 250 MWe. (The STIG cycle appears potentially applicable only at the nominal 50-MWe level.) Ideally, one module consists of one gasifier/GT/ST combination; however, if cost dictates, two gasifiers would be permitted for output levels over 150 MWe. The intercooled, steam-injected gas turbine (ISTIG) configuration is potentially attractive for this study, but its future development is uncertain; consequently, it was not considered. At a nominal 110 MWe, the ISTIG would compete with some good industrial GTs that currently do exist, and its projected price/performance ratio does not make it obviously superior. In addition, it could require a unique gasifier design.
8. A hot gas cleanup system will be utilized. Zinc ferrite desulfurization is the point of departure, but any technology consistent with the general philosophy of this study and Ground Rule 3 may be considered. The sorbent regeneration/sulfur disposition technique and its effect on total system emissions (and costs) is probably the principal issue. As in Ground Rule 5, the intent is that the system configuration would remain constant and the major components would be allowed only the minimum number of variations, e.g., sorbent-bed length or dust cyclone size.
9. Coal fines may be used directly (e.g., inclusion with coal fed to the gasifier), indirectly (e.g., fuel for subbituminous coal drying), or agglomerated (pelletized or briquetted and fed to the gasifier); however, their disposition must be accounted for in system costs and emissions.
10. Coal drying prior to gasification may be considered; however, the energy required for drying may be obtained only by combusting a fraction of the plant's input coal and using it as a heat source or by using heat derived from a process unit, e.g., the HRSG.
11. The operational pressure level at the outlet of the boost compressor will be 130% of the GT combustor pressure. This is to be utilized as gasifier and cleanup subsystem pressure drops plus control margin across the GT fuel-flow control valve. It is recognized that this boost-pressure level is well below what one GT manufacturer claims to need, but it is deemed feasible (and is above the requirements of at least one other major GT manufacturer).

Table 1. Reference Design Coal¹

Proximate Analysis, wt%, as received	
Volatile Matter	30.0
Fixed Carbon	52.0
Ash	15.0
Moisture	<u>3.0</u>
	100.0
Ultimate Analysis, wt%, as received	
Moisture	3.0
Carbon	68.6
Hydrogen	4.6
Nitrogen	1.2
Chlorine	0.1
Sulfur	2.8
Oxygen	4.7
Ash	<u>15.0</u>
	100.0
Sulfur Forms, wt%	
Pyritic	1.4
Organic	1.3
Sulfate	<u>0.1</u>
	2.8
Ash Fusion, °F	
Initial Deformation	2,200
Softening (H = W)	2,275
Fluid	2,400
Other Parameters	
HHV, Btu/lb, as received	12,500
Hardgrove Index	58
Free Swelling Index	8
Fines (< .25 in) Fraction, %	30

¹ Note: this is a synthetic coal but similar to an Upper Kittanning coal.

12. The gasifier will always produce some level of tar in a raw gas stream, and this must be dealt with. Increased pressure will lower the amount of tar, and a two-stage design will provide one of two streams that can be nearly free of tar, but the presence of a condensible hydrocarbon vapor in the raw gasifier output must be considered in all cases.
13. There will be a need for some form or level of tar removal from the product gas stream. For eastern coals, the presence of sulfur within the condensible hydrocarbon molecules (tar) is likely to contribute 0.20 to 0.25 lb SO₂/MBtu to the flue gas if the normal product gas is desulfurized without chemical cracking of the tar molecule (or physical removal of the tar), and if the gas is burned by the GT without further desulfurization steps. While this contribution to the SO₂ emission level is likely to be only 20 to 50% of the current EPA limit, it is also likely to result in system emissions that are significantly above what could be realized with a little extra expenditure to provide a tar destruction function coordinated with gas phase desulfurization. Chemical cracking of tar-like molecules has a large technology base in the petrochemical industry, and from a system perspective, it is viewed as having a far lower cost and performance penalty than physical removal of the tar. Consequently, the base-case systems in this study will utilize some form of tar cracking. Lower cost systems will also be evaluated and will have low emissions (below 0.5 lb/MBtu), but will only overtly destroy tar if there is essentially no extra cost.
14. The use of an in-bed sorbent for gas desulfurization within the gasifier will not be considered for this study. It is viewed as too ineffectual and likely to require post-gasifier ash treatment steps, and to be fraught with operational problems; the expectation is that it would result in a more costly system than would result from an external (to the gasifier) desulfurization approach.
15. If the raw coal gas exiting the gasifier is below 950 °F, it will be heated prior to desulfurization. This is primarily of concern when using western coals that have not been dried to some degree prior to gasification; typically, their raw product gas temperatures will be in the 400 to 700 °F range, which is too low for effective zinc ferrite (or similar) sorbent utilization. If this condition should exist, a fraction of the fuel gas will be burned to bring the bulk gas temperature up to the temperature level required for the desulfurization sorbent. See also Ground Rule 10.
16. A major revision of the GT combustor concept will not be required for these systems. Specific examples of this Ground Rule are if the manufacturer normally uses a *multiple can* combustor concept, this study will not assume the manufacturer would redesign to utilize a *silo* combustor; and similarly, use of a very advanced combustor (such as an off-base, staged, dirty fuel combustor), is outside the scope of this study. Modifications to a GT or a combustor design to the degree required to handle the low Btu character of the gas stream, its high temperature, or to utilize more than one fuel-gas input stream are considered legitimate, however. A subset to this Ground Rule is that

fuel gas supply temperatures up to 1600 °F are considered acceptable; it is recognized that temperature levels of 1200 °F and below are far easier to deal with and are to be considered as a goal.

17. The HRSG will be designed to not condense vapors out of the flue gas. It is recognized there is some system efficiency to be gained by operating in a condensing mode, but the gain is estimated to be outweighed by the increase in HRSG initial and maintenance costs.
18. The cost of a plant will be estimated as the summation of capital costs for major equipment plus an allowance for balance-of-plant (BOP), which generally covers necessary but non-process items. Additional funds for non-hardware line items (which are frequently included in plant cost estimates), such as land, allowance for funds during construction (AFDC), and contingency are not to be included in order to approximate the basic cost of the Nth plant, and in order to reflect the influences of gasifier/system design without obscuring the conclusions with factors that are site-specific or project-specific (and, for this study, would be assumptions in any event).
19. The BOP cost item will be a fixed 20% of capital cost used for all system configurations. Admittedly, this is not strictly correct (e.g., BOP is unlikely to change with the same scaling exponent as capital cost); however, the errors are not judged to be significant to the study results. Typical items expected to be included in BOP are
 - Plant support water supply and wastewater treatment systems,
 - Fire protection subsystems,
 - Auxiliary subsystems for plant startup,
 - Administrative building, parking lot, site security, etc., and
 - Overall plant control subsystem.

2.4 REFERENCES

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Chapter 3

Descriptions of Candidate Gasifiers

3.1 EXISTING GASIFIER DESIGNS

There are a number of existing fixed-bed gasifier designs or design concepts that potentially fit the simplified IGCC system concept and either are commercialized or are near enough to commercialization to be considered as a group with generally common characteristics. Those characteristics that are essential requirements for the simplified IGCC approach may be described as follows:

- At least a pilot-scale gasifier or test facility exists that can provide reputable information and confidence in gasifier performance predictions.
- There is reason to believe the gasifier, either as currently designed or with readily conceivable modifications, could operate
 - On essentially all types of United States (U.S.) coals,
 - With air blowing,
 - At a pressure of several hundred psig, and
 - Feeding a hot-gas-cleanup subsystem.
- The gasifier incorporated in a system can be expected to produce a cleaned gas that will be an adequate GT fuel. Adequate is defined as having a lower heating value of greater than 100 Btu per standard cubic foot (scf) with a minimum hydrogen concentration of 7%, a temperature of between 900 and 1600 °F, and no particles larger than 10 μ m in size.
- The gasifier can be expected to have a carbon utilization rate that exceeds 95%.
- The gasifier can be expected to have a large turn-down capability, e.g., the capability to operate satisfactorily at an output of less than 40% of full load if required. It is also expected to have a significant standby capability, e.g., to produce essentially no product gas for periods of as long as 24 hours and then to return to a full-load output state in less than 30 minutes.
- A potential gasifier vendor exists who either now has, or can reasonably develop, the capability to take an order and deliver a gasifier that will operate as expected.

One interesting commonality of the gasifiers discussed below is that while they have all the characteristics listed above, *no commercial-scale units have been sold that actually had the above list of attributes.*

A listing of the sponsoring organizations and the gasifiers in this grouping of potentially suitable designs is provided below. The list is ranked corresponding to the perceived nearness to commercial availability of a suitable gasifier.

- a. *Lurgi*: Pressurized, oxygen- and air-blown, commercial-scale units have been sold, and all have utilized a close-coupled, direct-quench, cold-gas-cleanup approach. Units of 12- and 15-foot diameters have been delivered; virtually all were oxygen-blown. Operability on U.S. coals is demonstrated.
- b. *British Gas/Lurgi (BGL)*: Pressurized, oxygen-blown, slagging, pilot units at up to 8-foot diameter have been operated successfully on most U.S. coals. An early unit was run successfully using highly preheated air in lieu of oxygen. No units have been sold.
- c. *Woodall-Duckham, Wellman Galusha, Riley*: There are several vendors offering generally similar, air- (or oxygen-) blown gasifiers that are capable of operating on non-caking or very mildly caking coals and are suited to producing a crude fuel gas for furnace and boiler applications. While Wellman Galusha has looked at pressurized designs, virtually all units sold in this subgrouping have been for operation at near atmospheric pressure. A few of the units sold have incorporated internal stirrers designed to allow use of coals with greater caking tendencies. Many of the units sold have had diameters of about 12 feet.
- d. *Kohlegas Nordrhein (KGN)*: This is a novel concept (Bauer et al. 1981) that utilizes internal recycling of the top gas, which is claimed to produce a tar-free product gas. The design is based on pressurized operation with oxygen or air blowing and the use of anthracite as a feedstock. The 6.8-foot diameter pilot unit, with a designed throughput of nominally 2 tons per hour (tph), is described as having operated successfully since early 1979. No recent information was found, and the status of this concept was not known at the time of the study.
- e. *Voest-Alpine*: This is a novel concept (Voest-Alpine Ag 1985), based on incinerator technology and having the capability to directly utilize fines. The design is based on operation with run-of-mine (ROM) coal at atmospheric pressure in an oxygen-blown and ash-slugging mode. It is claimed to produce a tar-free product gas and to be capable of utilizing coals with free-swelling indices up to 4. A pilot plant with a nominal throughput of 1 tph has run successfully.

The following provides a more substantive description of these gasifier designs and a summary listing of pros and cons for each.

3.1.1 Lurgi

The oxygen-blown, pressurized, dry-bottom, fixed-bed Lurgi gasifier is clearly the most mature fixed-bed gasifier in the world. Successful operation has been demonstrated at commercial-scale on coals that cover the range of U.S. coal characteristics. All units sold have utilized a gas cleanup approach that is initiated by the direct quench/liquid scrubbing of the raw output gas. The gasifiers are operable with either air or oxygen blowing; however, only five are known to have been put into commercial operation using air blowing and all five are in a combined-cycle system configuration at a Steag electric power plant in Lunen, West Germany. This system operated successfully for several years, but the gasification/combined-cycle portion of the power plant has been mothballed for more than 10 years - for reasons unrelated to gasifier performance. A diagram of the Lurgi gasifier, including the close-coupled scrubbing cooler (i.e., the direct-quench vessel), is provided as Figure 6.

The general characteristics of the Lurgi gasifier are reasonably well known and will not be repeated here. It is, however, worth noting that there has historically been a cited problem relative to the use of feedstock coal with a significant fines fraction (defined as that portion of the coal less than 0.25 inches in size). With the conventional Lurgi design, this concern is real and arises for two reasons: first, the liquid-based gas cleanup system is susceptible to reduced performance, plugging, etc., as a result of excessive dust and fines being carried over in the raw product gas and depositing in the scrubbing liquor; and second, the design of the top of the gasifier does little to reduce the propensity for solids carry-over.

Figure 6 should be regarded as geometrically representative rather than dimensionally accurate; however, some correct implications can be derived from this figure. The top of the coal bed is essentially at the elevation of the cup-shaped pan immediately above the stirrer blades. The blades turn through the upper portions of the coal bed, probably within a foot or so of the top of the bed (to maintain porosity of the devolatilization zone and break up any forming agglomerates). The raw gas outlet is near the top of the coal bed and represents a localized port in a region where a significant portion of the gasifier cross-section is unavailable for gas flow (because of blockage by the coal feed and distribution machinery). Consequently, over at least some portion of the bed surface, local gas velocities probably are actually accelerating as the raw gas leaves the coal bed, passes through the overbed volume, and reaches the gasifier outlet to the scrubbing cooler. As a result, any fines that are picked up by the gas leaving the bed are likely to remain entrained with the gas, at least until it enters the scrubbing cooler.

This problem has been addressed by Lurgi with two basic approaches: (1) make the coal sticky so individual fines are attached to larger nonentrainable lumps, and (2) cover the fines quickly so there is a more tortuous pathway to be followed before the average coal fine can exit the bed. An example of the first approach is indicated in Figure 6: recycle tar is ejected onto the coal in the distributor to serve as a dust suppressant. Similarly, tests have shown that a higher fraction of fines can be tolerated with "tarry" coals, such as Pittsburgh No. 8, which become sticky as they warm while in the top of the gasifier. One of the most

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Figure 6. The Lurgi Pressurized Gasifier

sophisticated embodiments of the second approach is found in Lurgi's U.S. Patent No. 4,405,340. When coal or any solid is dumped in a pile, it will assume a natural angle of repose, which is characteristic of the particular solid material, and it will also become segregated, with the preponderance of larger particles falling to the outside of the pile and the smaller, or fine, particles residing in the middle of the pile (i.e., the motions inherent in the piling process encourage the fines to be located near the centerline of the distributor and the larger particles near the periphery). This patent allows the size segregation to take place within the distributor and positions coal distributor-pan outlet chutes such that predominately fine coal is laid on top of the bed; this fine coal is immediately overlain by predominately coarse coal exiting from a second distributor discharge chute (as the distributor pan rotates). This is a relatively recent patent (1983), and the degree to which this design has proven to be beneficial in practice is not presently known.

The grate design of the Lurgi gasifier tends to emphasize controlled gas distribution and relies primarily on the amount of steam utilized and the chemistry of the coal char/ash to preclude excessive clinker formation. The grate surfaces are hard-faced, and this appears to mitigate wear; however, there are no features to overtly deal with clinkers. As alluded to above, if the bed-temperature distribution is as designed, the stirrer will reduce the formation of agglomerates in the upper portions of the bed (incipient clinkers), and the steam will suppress the lower bed temperatures enough to preclude the formation of significant clinkers. Under these conditions, there is no need to deal with clinkers; however, this grate design cannot easily accommodate untoward events.

This gasifier effectively only needs to be coupled to a hot-gas-cleanup subsystem to be consistent with the simplified IGCC concept. This is not viewed as a technical challenge, but could become a technology issue since Lurgi considers the scrubbing cooler an integral part of the gasifier, and scrubbing would obviously be inconsistent with a simplified IGCC system. Substituting a hot cyclone for a scrubbing cooler essentially removes all operability concerns related to an excess of fines in the coal feed. The remaining issue is primarily economic in nature, i.e., the *costs* of disposing of the dust caught by the cyclone. Table 2 provides a summary of pro and con statements relative to this gasifier design.

3.1.2 British Gas/Lurgi

British Gas plc and Lurgi GmbH combined to develop a version of the Lurgi gasifier that operates with a molten ash pool in the bottom of the gasifier (Thompson and Vierrath 1989a, 1989b; Thompson et al. 1989). This slagging fixed-bed gasifier design has been built and operated successfully as nominally 6- and 8-foot diameter test units at the British Gas research center in Westfield, Scotland. The gasifier design is considered only as an oxygen-blown unit by its sponsors, and good operational history and data exist on the test units using U.S. coals and dating back to 1975. The 8-foot unit is considered a commercial size and is nominally rated at a coal throughput of 500 tpd for 350 psig operation. It and larger designs,

Table 2. Dry-Bottom Lurgi Concept

PRO		CON	
1.	Huge general experience base.	1.	Little experience with air blowing and none with nonquenched operation.
2.	Capable of using all U.S. coals.		
3.	Can accept up to 35% of the feed as fines, but has a coal-specific throughput.	2.	Grate not tolerant of "rocks."
		3.	Top-bed stirring only.
4.	Commercial availability.	4.	Large steam usage.
		5.	Tars and fines in product gas.
		6.	Internal/central feed system raises over-bed gas velocities, and complicates stirrer design.

up to 15 feet in diameter, are currently offered for sale by British Gas/Lurgi (BGL); however, none have been sold to date.

The BGL gasifier is diagramed in Figure 7 and resembles the conventional Lurgi except at the bottom. Oxygen is injected using sidewall-mounted lances, or tuyeres, to provide the intense heating necessary to melt the coal ash and maintain an acceptably low slag viscosity. The tuyere locations and blast rates assure complete combustion of the descending coal char and essentially fix the physical location of the combustion process and molten slag pool. The higher temperatures created provide for increased char reaction rates in the lower portions of the gasifier and result in greatly increased coal throughput, relative to a dry-bottom design. The bed temperatures in the lower regions of the gasifier are high enough that lack of char reactivity has never been an issue or formed an operational limitation for this gasifier design. Coal throughput (per unit of gasifier cross-section area) in the BGL gasifier has been measured at about three times that of a dry-bottom Lurgi. The characteristics of the upper-bed region, raw product gas, etc., are very comparable to those in the dry-bottom Lurgi design, and the discussion of the preceding section applies.

Management and control of the molten slag is key to the successful operation of this design. As indicated in Figure 7, an open slag tap-hole exists in the hearth plate at the bottom of the gasifier. Slag is intermittently withdrawn from the gasifier, and a principal piece of the BGL technology is the use of slag and hearth-plate temperatures and differential pressure between the gasifier and the slag quench chamber to control the initiation and cessation of the slag tapping operation. This technique has been well enough developed that successful

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Figure 7. The BGL Gasifier

operation down to 30% of full output has been thoroughly demonstrated. Possibly more significant from a system operational standpoint is the capability for standby (negligible product-gas output) for periods as long as 48 hours has also been developed. This is accomplished by shutting off the tuyeres and letting the slag drain completely from the gasifier while keeping the bottom hot (but not molten), with the slag tap-burner located in the top of the slag quench chamber. To restart, oxygen is reintroduced through the tuyeres and the gasifier "lights off" immediately, producing useful output in a matter of a few minutes. Practically, the ability to maintain the standby mode likely exceeds any duration for which it would reasonably be needed. According to BGL, the system cannot reliably produce useable fuel at below 30% of full output.

Using tuyeres for injection of steam deep in the bed offers additional possibilities as a means for disposal of coal fines. BGL has, in fact, experimented with injecting fines through the tuyeres; with dry coal entrained in the steam stream and with a coal-water slurry (CWS) at 60% coal by weight. Both the entrained dry coal and CWS have worked well; however, the CWS approach is the more recent (it was only utilized with the 8-foot test gasifier) and appears to be preferred. When injecting fines as CWS, the steam flow is reduced to keep the moisture input approximately constant. BGL has injected as much as 30% of the total coal input to the gasifier as fines through the tuyeres and felt they could readily have fed more but saw no need to. There appears to have been no degradation of gas characteristics during periods of fines injection.

The slagging bottom provides some significant advantages with respect to ash handling and disposition. Within the system, the high temperatures at the bottom of the gasifier (nominally 3000 to 3200 °F) assure that all organic species are burned to extinction. The liquid ash is dropped into a pressurized water pool that both quenches the liquid to a solid state (without letting it form buildups on internal surfaces) and causes the solids to shatter (as opposed to existing in large solid chunks). A result is that the ash collects at the bottom of the slag quench chamber as very coarse sand-like, granular solids that flow relatively freely through the pressure letdown lockhoppers (i.e., slag crushers are not required anywhere in the system). A further consequence is that on exit from the gasifier subsystem, the slag is easily dewatered, with the result being a glassy frit. This product frit has been evaluated as "environmentally benign" by several organizations. It can be directly landfilled, and there are usages as abrasives and construction aggregates that, at some sites, could cause the frit to have commercial value. In addition, the wastewater from this portion of the process has also been evaluated as benign (Ebbins and Ruhl 1989; Bieshon, Hood, and Vierrath 1989).

An important point is that, while this gasifier is developed and proffered based on oxygen-blown operation, it has been successfully operated in an air-blown mode. While data have not been published by BGL, successful air-blown operation was described as requiring highly preheated blast air, approximately 1000 °F, and producing gas at about 120 to 130 Btu/scf and a few hundred degrees hotter than the nominal 1000 °F obtained with oxygen, with everything else being operationally normal. BGL has felt their potential applications were severely limited by a high nitrogen content and/or a low heating value in the product gas, and

consequently, regarded the air-blown product gas as essentially worthless and the experiments as programmatically unusable. It should be recognized that the blast air temperature requirement is very close to GT compressor discharge levels, and the heating value, especially for a highly preheated fuel gas, is quite likely to be consistent with the needs of any GT of interest. For the simplified IGCC concept, this gasifier would need to be mated to a hot gas cleanup subsystem and operated in an air-blown mode, both of which are quite credible.

Another potentially significant aspect of the BGL technology is the closely held nature of the "art" which supports this technology. Over approximately 15 years of development, BGL has evolved, but not publicized, specific operational techniques, empirical correlations, etc., as well as the specific design features of the gasifier. All of these are critical to the successful implementation of the BGL design. Since no gasifier has been sold, currently there is no basis to judge the effectiveness of the technology transfer to the customer - but this function is expected to be very important for this type of gasifier. Table 3 provides a summary of pro and con statements relative to this design.

Table 3. BGL Slagging Lurgi Concept

PRO	CON
1. A large general experience base exists.	1. Very little experience with air blowing and none with "non-quenched" gasification.
2. Appears capable of using all U.S. coals.	2. Control of slag tapping is key and is an "art."
3. Capable of using any credible amount of fines.	3. Top-bed stirring only.
4. Yields an extremely "benign" ash and associated wastewater.	4. Tars and fines in product gas.
5. Commercially available (but no units sold).	5. Internal/central feed system raises over-bed gas velocities and complicates stirrer design.

3.1.3 Woodall-Duckham

There are several vendors of coal gasifiers who have sold gas producers intended to provide an essentially ash-free, fuel gas typically as furnace or combustion fuel for process applications at atmospheric pressure. The market for these gasifiers primarily existed prior to 1940 and essentially disappeared with the advent of nationally distributed natural gas and rising concerns over potential environmental issues. Woodall-Duckham, Wellman Galusha, and Riley designs are representative of this group of gasifiers; however, a number of other vendors could be cited. While the number sold was not ascertained during this study, it

probably exceeded several hundred in the U.S. and considerably more worldwide; there are more than ten in the U.S. that are considered operational today. Generally, these were atmospheric pressure devices of considerable physical size and low throughput (10 feet in diameter and 3 to 4 tph would be representative) that required a great degree of operator involvement to run successfully. While several vendors have upgraded the technology somewhat (Wellman Galusha, Stoic, and Allis-Chalmers are the most recent examples), there has not been any significant market penetration by these types of gasifiers.

The Woodall-Duckham design is shown in Figure 8 and is one of the more novel examples of this class of gasifiers. It is a two-stage, atmospheric pressure, air-blown gasifier designed to run on noncaking (or at least very mildly caking) coals. The design has a drafted diameter (larger at the middle than the top and then narrowing slightly at the bottom) that slows the coal's velocity as it moves down through the upper portions of the bed, tailors the upward rising gas velocity as desired by the designer, and provides some room to accommodate a modest amount of coal swelling. The two-stage feature separates a fraction of the gas emanating from the gasification zone to provide a nearly condensibles-free, hot, relatively low heating value, mid-level gas. The top gas concentrates the condensibles from the pyrolysis/distillation processes occurring in the upper regions of the gasifier and has a significantly higher heating value as a result. This type of gasifier is particularly suited to an application where two gas streams with considerably different characters are desired, or to an application that seeks to produce a coal-derived condensate co-product while using the gas for other purposes.

The difficulty of upgrading these designs to operate at pressure levels suitable to firing GTs, learning how to operate them on caking coals, and developing at least some level of automated control system is judged to be a considerable, but not impossible, task. (The alternative of fuel gas compression is not credible because of the cleanliness requirements imposed by the compressor and the resulting costs, together with the direct costs of the compressor itself.) It should also be noted that there is some concern, common to gasifiers of this general configuration, that since a major portion of the product gas is withdrawn from the periphery of the vessel (which has, historically, been a region susceptible to gas bypassing phenomena or at least nonrepresentative reactions), there may be significant transient variations in the gas heating value. Table 4 provides a summary of pro and con statements relative to the Woodall-Duckham design.

3.1.4 Kohlegas Nordrhein

The KGN gasifier was developed in Germany; the design is depicted in Figure 9. The concept was developed to provide a means to produce tar-free coal gas as a route to simplifying the gas cleaning task and increasing overall system thermal efficiency. Early test units were operated in both the cyclic ("blow" alternating with "steam") and continuous-flow modes. The concept was upgraded to a 6.9-foot diameter pilot unit with a design rating of 2 tph; this unit was built in Huckelhoven, Germany, in 1978. The pilot unit is described as

Figure 8. The Woodall-Duckham Gasifier

Table 4. Woodall-Duckham Concept

PRO		CON	
1.	Commercial units exist.	1.	All experience is at atmospheric pressure with essentially noncaking coals.
2.	Two-stage gasifier including a mid-level outlet gas (hot and very low in condensibles).	2.	Major changes would be required for IGCC use.
3.	The drafted diameter provides a capability to more easily tailor the design to a specific coal type.	3.	Top gas contains tars.
		4.	Both top- and mid-level gas streams contain fines.

being capable of being operated at 7 atmospheres pressure in either air- or oxygen-blown modes. A first commercial plant was planned for implementation of this technology and was designated as "Riedelland" (Bauer et al. 1981). During this study, it was not determined whether this next step actually occurred or not.

Figure 9 shows the key feature of this concept, which is the vertical top-gas recycle pipe down the center of the reactor. Steam is utilized to educt the top gas and drive it to the grate level where it is released into the ash zone of the gasifier. The intent of this is to induce chemical cracking of all condensible hydrocarbons to occur (and it may also result in utilization of a significant fraction of the coal fines, but this is not a claimed virtue). The effective diameter of the upper bed is drafted, and the resulting annulus provides the outlet for a mid-level type of raw gas (and a region in which the product gas decelerates on its way to the outlet - providing an additional potential for fines separation); otherwise, the gasifier design is not particularly noteworthy.

The central gas recycle pipe is also a key potential operational weakness in the system. There are materials in the gas stream (dust and tars) that make it susceptible to blockage, and if the pipe should start to plug, the phenomena is likely to be self-aggravating because of induced high-temperature coking of the deposit and the resulting adjacent hot spots inside the pipe. While the presence of flowing steam should be a significant deterrent, if a blockage should develop, there is no easy way to clear it from the pipe without a complete shutdown. The gas recycle pipe, consequently, constitutes a single-point failure mechanism for the whole gasifier and system. The batch feeding system implied in Figure 9 is of specific concern because of the likelihood of its producing wide transient variations in the characteristics of the recycle gas (and perhaps also the product gas).

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Figure 9. The KGN Fixed-Bed Gasifier

Figure 9 depicts the recycle gas as being released within the grate, which would then distribute it across the gasifier diameter. It is unlikely the gas is actually handled in this manner as the recycle gas is hot and quite flammable, and combining it with air (or worse yet, oxygen) in this location is nearly certain to result in excessive temperatures within the grate. It is unclear in the reference; however, one "sector" of the grate may be designed to handle only recycle gas. In this case, combustion will occur immediately external to the grate with likely the same result. In addition, practical grate rotational speeds are slow enough that a design approach based on a fuel-gas-handling grate-sector is likely to produce enough asymmetry in ash zone temperatures that operational penalties will result; e.g., excessive use of steam to preclude local slagging. Some form of recycle gas distribution in the lower portions of the bed, other than what is implied in Figure 9, would seem to be preferred. A related concern exists relative to possible degradation of the product-gas heating value; the product gas is withdrawn only from the periphery of the vessel that historically has been a region susceptible to gas bypassing phenomena (or at least nonrepresentative reactions), which could easily either reduce the gas heating value or increase transient variations in it.

The cited reference is circa 1980 (Bauer et al. 1980), and various improvements have been reputed to be "in development" since then. However, no newer substitute references were found during the course of the study. Unfortunately, the technology has not been visibly obvious since 1980, and it is possible that development of the technology has been deferred. What is known from the referenced paper (at the time it was prepared) is that the pilot unit was operated only in the cyclic mode and only on anthracite but had accumulated a few thousand apparently successful operating hours. The maximum coal throughput achieved was 1.3 tph (versus the 2 tph design point). A large reduction in the hydrocarbon condensibles output with the product gas was noted, but since anthracite coal was utilized (and contained less than 8% total volatiles), this was not a particularly impressive claim. The design was not claimed to have the capability of accommodating caking coals, and the drafted diameter notwithstanding, the absence of a stirrer was consistent with that. Table 5 provides a summary of pro and con statements relative to the 1980-vintage KGN design.

3.1.5 Voest-Alpine

This gasifier is derived from incinerator technology and is shown as a conceptual pressurized design in Figure 10. The basic process was developed for the atmospheric pressure gasification of industrial and municipal wastes, which were often combined with coal to enrich the product gas. The overall process consisted of gasification, followed by cleaning of the product gases (which was judged to be lower in cost than cleaning the flue gases), followed by gas flaring or combustion for industrial or municipal heating. The gasifier was oxygen-blown and operated in a slagging mode, and as a result, the ash was environmentally benign and compact. A nominally 1 tph, atmospheric pressure, pilot unit was built in Linz, Austria. The gasifier relies on lump coal that is fed to the top of the gasifier, forming a fairly

Table 5. KGN Fixed-Bed Concept

PRO	CON
1. Internally recycles top gas to produce a nearly tar-free, hot, product gas.	1. No stirrer and no ability to accommodate caking coals.
2. The drafted diameter provides a capability to more easily tailor the design to a specific coal type.	2. High potential for recycle gas burning <u>at</u> the grate surface.
	3. Potential for carbon laydown in the recycle pipe or plugging, resulting in gasifier failure.
	4. Fines exist in the output gas.
	5. It appears major changes would be required for IGCC use, and very little information is available.

free-flowing char with a relatively predictable angle of repose. As implied in Figure 10, the angle of repose governs the shape of a "char bank" at the bottom of the gasifier, and this bank is fired by steam-oxygen tuyeres on the opposing wall. Hot gases from this combustion zone pass up through the bed to provide heat for the gasification reactions above. It is reported that nearly any conceivable amount of coal fines may be injected by the tuyeres and, consequently, this concept has the potential to utilize the *whole* coal supply directly with minimal preprocessing. A puddle of molten slag is maintained below the char bank and overflows essentially continuously through a tap hole into a water-quench bath.

As with all slagging designs, the ability to keep the slag tapping process functioning is critical. This design is based on a continuous tapping approach for a vessel in which the solids bed does not rest on the hearth plate and tap hole. As a consequence, the slag removal process/hardware design is comparatively simple with essentially the only requirement being that the slag tap-hole not be allowed to freeze shut. This design should be easier for high-ash feedstocks with their correspondingly larger slag pools, but there is no reason that with care this design approach cannot accommodate low-ash coals.

A claim is made that the gases passing up through the coal bed are so hot that no tars appear in the product gases. While the product gases are defined as being at 1400 to 1750 °F, this claim is not very credible for many U.S. coals, since fresh coal is fed to the top (i.e., gas outlet end) of the bed, and the gas temperatures are similar to several other gasifiers that were found to be tar generators, such as the Synthane gasifier.

Figure 10. The Voest-Alpine Gasification Reactor

Most of this development work was done in the middle 1980's, and while atmospheric pressure, combined coal-waste incineration systems appear to have become commercial, it is not known whether any actual coal gasifiers have been sold. Upgrading the existing incinerator-based design to operate at pressure and with air blowing apparently has not been done and this will be a significant effort, the "pressure" vessel indicated in Figure 10 notwithstanding. Table 6 provides a summary of pro and con statements relative to the Voest-Alpine gasifier design.

Table 6. Voest-Alpine Gasifier Concept

PRO	CON
1. Claim of no tars in product gas because of high outlet temperature (1400 to 1750 °F).	1. No experience with pressurized operation.
2. Capable of using any credible amount of fines.	2. Gasifier operation is dependent on a <i>flowable</i> char and a predictable angle of repose.
3. Yields an ash that is probably environmentally benign.	3. Inability to utilize coals with a free-swelling index (FSI) above 4. (The claimed ability to handle coals with FSI values up to 4 is highly questionable.)
	4. The reliability of pressurized slagging operation with low-ash feedstocks is unknown.
	5. Output gas contains fines.
	6. Major development would be required for IGCC use, and the availability of supporting information is sparse.

3.2 PROTOTYPE GASIFIER DESIGNS

There is a second grouping of gasifier designs that have characteristics similar to the foregoing but differ principally in that they are generally more suited to the simplified IGCC system approach, and they are generally at a less mature development stage. Frequently, it is also unclear who or what organization would be the technology vendor. A listing of the gasifier designs and concepts in this grouping, in order of their perceived nearness to potential commercialization, is provided.

- a. *METC*: A pressurized, air-blown, dry-bottom pilot-scale gasifier coupled to various hot gas cleaning test units. This is a 42-inch diameter test unit at the U.S. Department of Energy's Morgantown Energy Technology Center (METC) in Morgantown, West Virginia. This gasifier was built in 1967 and has been utilized (1) to develop fixed-bed gasification technology that is suitable to all U.S. coals, and (2) to investigate improved product-gas cleanup subsystems. A similar but slightly smaller unit exists at GE's Corporate Research and Development Center in Schenectady, New York.
- b. *BCURA*: This is a novel, supported fixed-bed, oxygen-blown, slagging, experimental gasifier. This nominally 1-foot diameter experimental unit was built by the British Coal Utilization Research Association (BCURA), Surrey, England, to test the concept of a "composite" gasifier design. It is capable of directly utilizing the coal fines and will output the coal ash as a glassy frit.

3.2.1 METC Gasifier Concept

This is a 42-inch diameter, dry-bottom gasifier with a nominal 300 psig and 1 tph rating. It was originally built in 1967 to develop the design features and technology required to gasify the range of U.S. coals using fixed-bed gasification. In the course of this work, it was found that three families of coals essentially set all the design requirements: (a) North Dakota lignite provided an extreme with respect to dust, friability, and low product-gas temperature; (b) Pittsburgh bituminous coal was the most difficult operationally because of its strongly caking nature combined with a large tar output (and a tar with a large fraction of its hydrocarbons having high boiling points); and (c) Pennsylvania anthracite resulted in the lowest throughput because of its low reactivity. The principal areas of investigation were coal feeding, bed stirring, grate design, and instrumentation. A single design was developed that could accommodate this range of coals well, and the resulting configuration is shown in Figure 11. Since approximately 1980, various product-gas cleanup approaches have also been investigated using this gasifier as a test gas source. An integrated process development unit for an advanced, staged, gas cleanup subsystem has been built and tested, and several individual hot gas cleaning devices have been tested using this facility (Morgantown Energy Technology Center 1981; Pater et al. 1982).

Two independent subsystems are utilized to provide a dual, through-the-sidewall approach for coal feeding. A pressurized lockhopper supplies coal to a variable speed, rotary feeder that dispenses coal to a screw feeder. The screw feeder runs at constant speed (45 rpm), which is fast enough that it is always somewhat empty, and more importantly, moves the coal into the volume of the gasifier before it can become hot enough to be sticky and initiate blockages in the feeding subsystem. The coal exits the feeder with enough velocity that a significant fraction reaches the gasifier centerline and obviates any real need for a bed-leveling function. The top-of-bed elevation is sensed (using a gamma ray attenuation-based approach), and the speed of the rotary feeder is varied to maintain the bed surface at the elevation desired.

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Figure 11. Sectional View of the Current METC Fixed-Bed Gasifier

A water-cooled, hydraulically driven, three-blade stirrer has been used to maintain bed porosity and provide a capability to utilize strongly caking coals. The design of the water-cooled stirrer shaft, support bearings, pressure seal, and hydraulic drive was based on providing a capability to sweep the entire bed (to essentially grate level). This range of motion is not necessary on a commercial gasifier but was useful for research purposes; similarly, various blade configurations were investigated and only the lower two blades appear to actually be necessary.

The grate is a three-tier, eccentric plate design that is hydraulically driven to rotate within a heavy steel cylindrical wear collar: a "Bosch ring." The top two plates are eccentric, and the bottom plate is concentric with the gasifier centerline. The eccentric plates in combination with the Bosch ring give the grate an ash crushing capability that, while not significantly utilized in normal operation, both provides an operational safety margin and allows the bottom of the gasifier to be run at slightly higher temperatures than would be desirable for a grate without some crushing capability. There is a "plow" on the grate that draws ash radially inward, and the ash exits downward through large central holes in the middle and bottom plates. The spacing between the three plates and the size of the ash exit hole in the bottom plate allow the grate to accommodate agglomerates of considerable size (e.g., 6 inches in diameter), which is a very forgiving feature.

In its final configuration, this gasifier has never been pushed to a throughput limit because the objectives of all recent testing have centered on downstream gas cleaning. What can be said is that the gasifier has run comfortably with a Montana rosebud subbituminous coal at an input rate of over 2 tph. Successful extended runs have been made using Pittsburgh No. 8 bituminous coal at slightly above 1 tph for periods of over 400 continuous hours. The coal feeding and ash removal subsystems can run at significantly higher levels. (For example, each of the two coal feeding units has been run at 2 tph.) Fines carry-over has never imposed a process limitation with this gasifier, even with North Dakota lignite; an economic limitation could exist, but it was not relevant to the tests and was not calculated. There appears to be no limitation for a combination of reasons. In all its incarnations, this gasifier has always utilized some form of a hot gas cyclone to remove the majority of solids/dust from the product gas stream, and these separated fines could not then degrade either the gasifier or downstream components. Further, the top of the coal bed is typically 8 to 12 inches below the coal feeder sidewall ports, the overbed region is only occupied by the stirrer shaft/bearing (constituting a minute fraction of the total volume), and the product gas outlet is in the vessel dome several feet above the feed ports and is large in diameter at the vessel wall. As a consequence, the gas velocities (and the quantity of entrained fines actually passing into the outlet port) are minimized, and those that do deposit at the entrance to the outlet port have a pathway to potentially fall back into the gasifier.

The most difficult part of this gasifier design is the provision of the deep-bed stirring function and the relating restrictions it places on means to accomplish other needed functions. As mentioned above, the stirrer depicted in Figure 11 has the ability to penetrate the entire bed, which will not be needed in a commercial unit; however, a commercial unit will be

physically larger with higher torsional loadings on the stirrer shaft and so many of the issues will remain. The principal issue is the bearing and pressure seal design for the erratically loaded, water-cooled shaft, which both slowly rotates and translates as it penetrates the gasifier dome. Relating to this is the need for the stirrer to predictably and repeatedly penetrate the needed zones within the gasifier (without damaging itself or the gasifier interior) and the need to know the location of the stirrer blades at any point in time. This last need becomes important when various approaches to instrumenting the gasifier are considered; the existence of the stirrer has both provided useful means to sense important in-bed phenomena and has also limited related design options, such as instrumentation utilizing penetrations through the sidewall. Instrumentation approaches for all the functions necessary to fully automate a commercial-scale gasifier have been proven on this test unit.

There is a similar gasifier at GE's Corporate Research and Development Center in Schenectady, New York. The principal differences are a single feeding subsystem, a 35-inch gasifier diameter, a refractory interior lining, and a somewhat higher pressure rating. This gasifier was built about 1978 to be the gas source for IGCC system types of tests, and the design was heavily influenced by experiences with the METC unit. Most of the comments above are also applicable to this unit. It is currently being incorporated into a configuration to allow system-level testing employing hot-gas-cleanup approaches.

A deterrent to commercialization of this concept is the lack of a "vendor" for the gasifier. There are no limitations relating to proprietary information; however, there also is not a currently significant market demand for this technology. Table 7 provides a summary of pro and con statements relative to the METC gasifier design concept.

3.2.2 BCURA Gasifier Concept

In the early 1980's, the BCURA (Corman 1986) conceived and built a small oxygen-blown, atmospheric pressure, experimental gasifier at the National Coal Board (NCB) Coal Utilization Research Laboratory (CURL). The concept then was to combine some of the virtues of fixed- and entrained-bed gasification into one unit, and it was dubbed a "composite" gasifier. The primary advantages were thought to be increased specific throughput, reduced steam requirement, total consumption of fines, and increased ability to utilize caking coals. This concept utilized a stationary, water-cooled, tubular "grate" to support the upper (fixed-bed) regions, which were heated by combustion product gases from the lower (entrained-bed) regions of the reactor. A schematic of the test unit is provided in Figure 12. Testing was limited in scope, and while reported as generally successful, it is not clear that the attributes of the concept were significantly better than those of other, more conventional, gasifier designs.

The operation of this approach is described in the cited reference and may be summarized as follows. Coal is separated into coarse and fine streams with the coarse coal being fed to the top of the gasifier and fines being fed through tuyeres to the below-grate region.

Table 7. METC Gasifier Concept

PRO	CON
<ol style="list-style-type: none"> 1. Highly flexible, reliable coal feeding subsystem: <ul style="list-style-type: none"> • Dual subsystems. • Large range of throughputs. • Minimal coal heating occurs. • External to the gasifier. 2. Capable of operating on all U.S. coals. 3. No operational limitation caused by fines (and features exist to minimize fines carry-over). 4. Capable of being completely automated in operation. 5. Grate can accommodate "large" ash agglomerates. 6. Hydraulic drives on stirrer and grate provide flexibility. 	<ol style="list-style-type: none"> 1. Design for deep-bed stirring is complex, particularly the stirrer shaft seal/bearing. 2. Tars and fines exist in product gas. 3. No obvious vendor.

The injected fines are burned with preheated air (or oxygen) in an entrained mode to provide the very hot combustion product gases that pass up through the water-cooled, tubular grate to provide heat sufficient to melt the coal ash and drive the rest of the gasification reactions occurring in the above-grate regions of the gasifier. The molten ash drips through the tubular grate and is collected in a slag pool at the bottom of the entrained section. From there, it continuously overflows into a quench bath and exits the system via lockhoppers as a glassy frit.

The general operating conditions utilized with the test unit are described as feeding 75% of the input coal (the ROM fraction greater than 0.25 inches in size) to the top, and feeding the fines to the entrained section (the fines were pulverized to 70% less than 200 mesh prior to feeding), while using 950 °F steam and preheated oxygen at nominally atmospheric pressure. Based on modifications of the test data, performance estimates were made for a 12.7-foot diameter commercial unit, using 950 °F steam but with 700 °F blast air and 30% of the coal fed as fines. The key results of these calculations are as follows (the cited

Figure 12. NCB/CURL Experimental Composite Slagging Gasifier

reference is not explicitly clear but the values appear to be for 20 atm gasifier pressure levels):

• Gas Constituent	<u>Volume %</u>
- Nitrogen	46.3
- Carbon Monoxide	27.2
- Hydrogen	12.3
- Water	7.6
- Carbon Dioxide	2.7
- Methane	2.4
- Hydrogen Sulfide	.6
- Argon	.6
- Ammonia	.3
- Carbonyl Sulfide plus "others"	Traces
• Raw Gas Heating Value (LHV)	151 Btu/scf
• Raw Gas Temperature	1,324 °F
• Raw Gas Output Productivity	3.61 lb gas/lb coal fed
• Coal Specific Throughput	875 lb coal/hr ft ² at 20 atm
• Steam-to-Coal Ratio	0.21 lb/lb
• Air-to-Coal Ratio	2.54 lb/lb
• Steam-to-Air Ratio	0.082 lb/lb

As with most designs based on slagging the coal ash, slag management is one of the most critical aspects of this design. In this case, the key is the ability to maintain the desired material flow rates through the tubular grate (liquid slag downward and gases upward); too fast a flow rate and solids enter the entrained region, too slow and gasifier output and gas quality are reduced. (A related issue is maintaining the desired solids flow rate in the upper regions of the gasifier.) In either case, gasifier operation could be expected to become unstable, and recovery would be difficult at best. During operation, it would be expected that liquid slag would freeze onto the water-cooled grate tubes and would reduce the effective tube-to-tube spacing. Presumably, this frozen slag layer would reach some equilibrium thickness. To operate as described, the upper bed must be supported by a combination of the slag-coated grate tubes and a meniscus of liquid slag in each tube-to-tube gap (which in turn is largely supported by the upward flowing gases). The downward slag-flow characteristics and the ability to retain solids above the grate (and perhaps the flow and lateral distribution of the upward moving gases) appear likely to be strong functions of the effective tube-to-tube gap and the slag viscosity - neither of which is readily controlled. From an operational standpoint, this is a very complex grate design. The gasifier's bottom design does separate the physical support of the bed and the slag-tapping functions, and there should be no problem relating to slag removal. (Keeping the slag tap hole unfrozen is the only real requirement.)

There is also a likelihood that macroscopic parameters such as variations in bed depth or gasifier pressure level will have a significant effect on material flow rates through the grate. In reality, this could be a virtue as these parameters are controllable and could offer a means to accommodate flow variations and keep the gasifier operation under control; admittedly, this type of dependency could complicate the gasifier's integration with the rest of the system. There are other concerns, primarily relating to the grate itself (such as the degree to which it cools the upward moving gases and corrosion/erosion in a very aggressive environment), but they are of comparatively lesser magnitude.

While the experience with this concept is quite limited, CURL reports they have been able to make it operate, which implies there may be more virtue to the concept than appears from this review. The degree to which the above potentially critical issues would actually manifest themselves in a commercial-size gasifier or the degree to which they would effect the integration of the total system is not known at this time. However, the above discussion points should be thoroughly understood before proceeding to system designs. CURL describes the gasifier as being potentially capable of being scaled-up and operating in an air-blown mode. Table 8 provides a summary of pro and con statements relative to the BCURA gasifier design concept.

Table 8. BCURA Composite Gasifier Concept

PRO	CON
<ol style="list-style-type: none"> 1. Likely to yield a benign ash. 2. Likely capable of accommodating a large fines input. 	<ol style="list-style-type: none"> 1. <u>Very</u> doubtful reliability of tubular grate design: <ul style="list-style-type: none"> • Slag/solids/gas flows. • Cooling of gas/slag. • Tube erosion/corrosion. 2. Only limited experience exists (small, oxygen-blown, CURL unit tested for a short time). 3. Tars and a level of fines are likely in the product gas. 4. Much development is required to commercialize this design.

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Chapter 4

Gasifier Characteristics: Key Features and Options

This chapter describes the characteristics of the most important design features and design options that are pertinent to the gasifier itself and relevant to the CGIA system concept. It should be noted that for what appear to be the most promising design approaches, none are currently developed at the scale and for the operating conditions required; however, all have a strong enough technology base to warrant the technical enthusiasm.

4.1 FUEL FEEDING SYSTEMS

The fuel feeding system for a fixed-bed gasifier utilizes fuel provided by the coal handling portion of the plant, and if applicable, fuels emanating from particular locations within the process portions of the plant, such as a dust removing cyclone on the product gas stream. The primary role of the fuel feeding system is to transfer coal, which is provided at ambient conditions, into the gasifier, which is at elevated pressure and temperature; the basic feeding mode is the deposition of lump coal onto the top of the gasifier's coal bed. However, the impact of fine coal must be considered, and "into-the-bed" feeding of multi-phase streams should also be considered, at least for the two-stage gasifiers that utilize recycle of the top gas to a location within the coal bed. The important operational features and requirements for the feeding system applicable to the CGIA concept may be summarized as follows:

- It shall deliver fuel to the gasifier on demand (as dictated by the control system) and be capable of operating at the required variations of feed rate (including a feed rate of zero).
- The same basic system configuration shall be capable of operating on the range of U.S. coals and across the range of applicable gasifier sizes. Coal feeding systems are essentially volumetric dispensers (functioning against a pressure differential), which must accommodate coals that vary considerably in energy content per unit volume, particle size and range, tendency to become sticky, friability, etc.
- It shall degrade the operation of the overall system to the minimum degree possible. The principal issues here are the possibility of product gas leakage, excessive power consumption (normally only a concern with extruder type systems), potential for fines generation within the feeding system, increased fines carry-over in the product gas stream, and the need for fines agglomeration (which would actually occur within the coal handling section of the plant).

All of these have aspects of reliability, maintainability, and cost that become significant design influences.

4.1.1 Top-Bed Feeding

For all fixed-bed gasifier concepts, the majority of fuel is fed to the top of the bed, and the design of the top-bed feeding system is extremely important to the overall gasifier performance. The reality of coal and coal feeding is that some level of coal fines will enter the gasifier, and some fraction of these will exit with the product gas. Excessive fines in the product gas stream potentially represent both a loss of gasifier feedstock and a gas contaminant that can cause operational burdens for downstream portions of the system. Typically, the last reactant solids encountered by the product gases prior to exiting the gasifier are the entering coal particles, and consequently, a major design issue is to find ways to minimize the entrainment of fine coal particles in the product gas stream. This is normally done by minimizing the fines fraction of the basic feed coal, treating the coal as "gently" as possible during its transit of the feeding system, minimizing gas velocities in the top of the gasifier, and trapping fine particles so they are less susceptible to entrainment -- and most typically by a combination of the above.

The fines fraction in the gasifier feed coal is a function of the coal supply and the coal handling section design, for example, whether screening and/or agglomeration capabilities are provided. However, this is not really an area benefitting from design niceties; it tends to become a pure economic tradeoff: the benefits of increased coal utilization versus the costs of waste fines disposal, fines agglomeration, or the effects of increased contaminant burdens. The technology of fines agglomeration is discussed in Chapter 5.

Generally speaking, the usage of belt or bucket coal conveying and pressure swinging lockhoppers appears to be the most practical gentle approach to handling the coal through the feeding system, and thus, it is a preferred approach to limiting fines generation within the feeding system itself. Techniques to be avoided or at least minimized are those that introduce "rubbing" and "impact" phenomena onto the coal particles. Similarly, significant heating of the coal within the feeding system can become a source of problems; many low-rank coals tend to become significantly more friable, and some high-rank coals become sticky with temperature increases as small as a few hundred degrees Fahrenheit. Increased friability primarily is reflected as an increased fines concentration in the product gas stream, and stickiness primarily is a source of operational problems within the feeding system itself (but it can reduce the amount of fines in the output gas stream).

A feed system with a single lockhopper positioned above the gasifier is probably the most commonly considered approach to integration with the gasifier vessel. The most sophisticated example of the top-located, single-lockhopper feed system is the Lurgi design, described in some detail in Chapter 3 (Section 3.1.1). This system has been employed on many commercial gasifier systems and has been modified and refined by Lurgi based on their experience. While it has been used on gasifier designs up to 15 feet in diameter, the design of this type of central feed system to achieve a really uniform coal feed distribution becomes an increasingly difficult problem for larger gasifiers. (Even distribution of the coal is important to avoid uneven gas flow, or channeling, through the gasifier and to ensure maximum

utilization of the gasifier volume.) Of likely more significance, the hardware internal to the gasifier, which is needed to ensure even coal distribution, results in restricting the product-gas flow area above the coal bed. The restricted flow area equates to higher than necessary gas velocity above the coal bed that in turn results in higher fines carry-over and lower realizable coal throughput per unit of gasifier cross-sectional area.

Two other potentially significant characteristics of the single-lockhopper approach are that it tends to reduce system overall efficiency and reliability. There are non-trivial gains to be made through savings of parasitic compression work requirements if lockhopper pressurization can be accomplished, at least partially, through the utilization of vent gases from another lockhopper that is being depressurized. In a system with multiple gasifiers (and, thus, multiple lockhoppers), this is easily accomplished, but in an installation with potentially only one or two gasifiers, this is not so readily accomplished.

In a similar manner, with a dual feeding system it is often possible to operate a gasifier at nearly full output with only one of two subsystems being operational, assuming each individual subsystem has the necessary throughput capability. The fuel feeding system is one of the more difficult parts of the system to make reliable, and this redundant capability can be very valuable in terms of improving plant availability and avoiding emergency shutdowns. This feature can provide a capability to operate the system on an interim basis (for periods of at least several days) until the problem can be remedied or the system can be brought to an orderly shutdown. Usually, there is not significant cost involved in oversizing each subsystem because much of the throughput increase can be obtained by operating the subsystem "faster" (in reality, cycling the lockhoppers more frequently) as opposed to using larger hardware. A byproduct of this is that the coal is handled more gently during normal (slower) operation. The drawback to dual subsystems is the doubling of the number of seals and other potential failure points, as well as the greater number of components involved. While these considerations intuitively mitigate against a larger number of feeding subsystems and while dual systems have been used with great success at METC, the specific reliability analyses to justify selection of any particular number have not been done.

Gasifiers that utilize swelling coals will require an in-bed stirrer that, practically speaking, must enter the bed from the top and be coaxial with the gasifier centerline. Top-bed feeding system designs having a significant amount of hardware located at the top center of the gasifier become seriously complicated when integration with the mounting of stirrer shafts and drive assemblies is considered. This is exacerbated if vertical travel of the stirrer is required, as may be the case for some strongly caking coals.

As a consequence of the above, some top-bed feeding system design guidelines may be defined:

- Maximize the free volume for gas flow above the coal bed. This can be accomplished by increasing gasifier diameter or by decreasing the volume consumed by internal components. Similarly, there is a virtue to increasing length (until the transport

disengagement height is reached) as it will lower the fines concentration in the product gas. However, there are obvious tradeoffs among increased gasifier performance, size, cost, and heat loss; these tradeoffs also must be considered.

- Minimize the temperature rise of the coal within the feeding system. This will help preclude other operational problems; that is, a 100 °F rise is not a problem, 500 °F is likely to be.
- Take extreme care to minimize the potential for any product-gas flow-back through the feeding system (i.e., leaks). These are an obvious nuisance and somewhat of a hazard but, of more real significance, they will invariably cause the feeding system to fail! The failure occurs as a result of a condensibles- and dust-bearing hot gas stream coming in contact with cooler surfaces and forming deposits on them, heating them to the point that something untoward happens, or both. Some typical examples of phenomena that can lead to failure include coal sticking to internal surfaces and restricting feeding system throughput, seals and bearings accumulating grit, seals and bearings overheating, and thermal decrepitation of the coal prior to entering the gasifier. A preventative measure is to provide a clean, cool gas purge through the feeding system but, while this might be useful for a research and development (R&D) oriented system, this is not very practical for a commercial system.
- Provide the greatest capability practically possible for maintenance of the feeding system. Solids handling systems operating at elevated pressure and temperature conditions have a difficult environment to contend with and have historically been problematical. Providing the capability to operate in a non-normal mode or to quickly effect repairs normally pays great dividends.

In the course of this study, a conceptual top-bed feeding system design was defined based upon the comments above and experience with the METC gasifier's feeding system. The METC gasifier, along with a brief discussion of its feeding system was described in Chapter 3 (Section 3.2.1). The conceptual feeding system design is shown in Figure 13. It was configured to be more amenable to the large diameter vessels typical of commercial gasifiers, and is a derivative of the dual through-the-sidewall feeding system successfully employed for years by METC on its 42-inch diameter gasifier.

As shown in Figure 13, this system incorporates a dual setup of a pressurized lockhopper releasing coal into a pocket wheel (rotary) feeder, which dispenses coal into a screw feeder for transfer to the top of the gasifier's coal bed. Lockhopper operation is not unique, but the existence of two parallel subsystems gives the opportunity to more efficiently provide lockhopper pressurization gas. It also greatly reduces the variations typically seen in gasifier output-gas properties whenever a feed system lockhopper is cycled, and the nearly continuous nature of the gasifier feeding process tends to smooth out operations in general. As is easily recognized from the figure, nearly all of the feeding system components are located outside of the gasifier vessel envelope. This feature maximizes the overbed free volume for gas flow,

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Figure 13. METC Coal-Feeder Concept

simplifies limitation of the coal temperature rise, and potentially facilitates maintenance activities.

In normal operation, both sides of this dual system are running continuously, but the lockhopper fill cycles are staggered such that one side of the system is always feeding the gasifier while the other side is having its lockhopper refilled. The gasifier's automatic control system senses the coal-bed level and controls the speed of the two rotary metering valves (which run continuously but at varied speeds -- including *stop*) such that the coal-bed level is maintained within an acceptable bandwidth. The rotary valves are pocket-wheel devices that sequentially present empty pocket cavities to the outlet side of the bottom lockhopper valve where each cavity is filled and then rotated to discharge coal onto a screw feeder. The design of the rotary feeder is simplified in that, other than the pocket wheel's drive shaft, it has no pressure sealing requirements -- the top lockhopper valve provides gasifier pressure isolation during the feeding cycle. The screw feeder is located in an inclined feed tube and is not used in the manner normal for a screw feeder. The primary role of the screw is to ensure the coal dispensed by the rotary feeder is moved into the gasifier before it can heat significantly. It also dictates the coal's *velocity* through the inclined feed tube and ensures the coal exits into the gasifier with a reasonably known velocity. In the course of doing this, it also mixes the coal to a relatively uniform size distribution (volumetrically) and removes any wall deposits as they initiate so as to keep the coal pathway open. The resulting design is a quite simple, loose fitting, constant pitch screw that is run at constant speed (45 rpm was used at METC) and normally operates in a *mostly empty* mode. Like the rotary feeder, its only pressure sealing requirement is for the drive shaft penetration, and the drive motor is mounted outside the pressure envelope and on the side of the screw housing (using a chain drive coupling) to facilitate both motor and screw maintenance. At the slow design speed, the screw does not significantly abrade the coal, and the coal's exit velocity, though non-trivial, is comparable to the free-fall velocity (given a clean, empty, inclined feed tube).

The coal exit velocity is selected to propel most of the coal across the radius of the gasifier to impact the coal mixer/distributor that is the only component of the feeding system located within the gasifier vessel. This component is located on the central stirrer shaft and rotates with it. Currently, it is not known whether the mixer/distributor needs to be rotated at a speed different than the stirrer and whether it must be held at a constant vertical position or can be allowed some degree of vertical movement. If the rotational speed is adequate for the mixer/distributor, and if the vertical motion required of the stirrer shaft is small (or zero), it may be mechanically fixed to the stirrer shaft which would simplify the design considerably. However, a faster rotational speed at a fixed vertical position for the mixer/distributor was assumed for the configuration shown in Figure 13. Similarly, the specific inclined feed tube angles (in both horizontal and vertical planes), coal exit velocities, and mixer/distributor blade shapes are not yet defined for use in a large diameter vessel.

The mixer/distributor is the only component of the feeding system that has not been extensively tested at METC. The basic principle of this mixer/distributor system is illustrated in Figure 14. With this concept, it is recognized that the size distribution of the coal exiting

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Figure 14. Basic Principle of the Rotating Coal Mixer/Distributor

the screw feeder will be in some mixed state because of the screw action, but is likely to become segregated as the stream arcs out over the bed. One can expect the fines to follow a shorter or lower trajectory than the lump coal, such that the fines will predominate in the lower portions of the stream. The concept shown here prevents this segregation from also appearing in the coal bed by utilizing deflectors (mounted on arms supported from the stirrer shaft) to deflect the coal stream in a downward direction. In this manner, the larger lump coal will be deflected downward through the fines, causing remixing. In addition, as the lump coal intercepts the fines, the fines will tend to collect or adhere to the coal lumps such that they will be carried onto the coal bed. The intent is that a combination of (1) the downward velocity imparted to the whole stream of coal particles, and (2) the co-deposition of large and small coal particles onto the same locations of the coal bed (without spatial segregation) will make entrainment of the fines by the rising product gases less likely.

In order to produce a nearly uniform distribution of coal over the top area of the coal bed, it is envisioned that a series of deflector sets will be required. A probable configuration of the mixer/distributor is shown in Figure 15 wherein a set of deflectors at a long radius and a set of deflectors at a shorter radius are utilized to distribute coal over different zones of the bed. The operation of this device can be understood by considering Figure 15 and assuming clockwise rotation of the deflectors. With the deflectors near the position shown, the long radius deflectors near the wall will intercept the stream and deflect coal in a downward and rotational direction. Thus, coal will be distributed near the wall in the quadrants located clockwise from the feed ports. Subsequently, the shorter radius deflector will deflect coal downward and around such that coal is distributed in the same quadrants but closer to the centerline of the gasifier. As the rotation continues, a period occurs wherein there are no deflectors in front of the inclined feed tube such that coal will be thrown out in line with the feed tube centerlines and essentially to the middle of the gasifier. The feed tube centerlines are oriented to be somewhat offset from each other and to not intersect the stirrer shaft centerline so as to effect this phase of coal distribution. Completion of one mixer/distributor rotation will then result in coal being distributed over all quadrants of the gasifier near its center and over the outer portions of the two quadrants clockwise from the feed ports. It is possible the peripheral portions of the two quadrants counterclockwise from the feed ports will receive an inadequate coal feed distribution, but this remains to be determined. If this is found to be the case, the addition of a bed leveler blade down from the long radius deflector support arm should smooth out this distribution. Alternatively, the use of a downward angled, vee-shaped, deflector surface to deflect coal in both the rotational and counter-rotational directions could also be a potential solution.

4.1.2 In-Bed Feeding

With some fixed-bed gasifier designs, it is necessary to feed a fuel stream, which is predominately gaseous but which contains a large fraction of condensible vapors and includes a significant fraction of fine coal, to a location deep within the coal bed itself. This is precisely the situation applicable to a two-stage, fixed-bed gasifier employing top-gas recycle. In

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Figure 15. Coal Mixer/Distributor Configuration

addition, this same capability is often desirable for other configurations as a technique for beneficially utilizing what is otherwise a difficult material to accommodate -- the really small coal fines. A good example of this situation is BGL's successful usage of the slagging gasifier tuyeres as an injection point for coal fines and other combustible wastes. In both cases, the intent is to place the fuel stream deep enough in the bed such that the indigenous temperatures are high enough and the injected fuels' residence times are long enough that the fuels are utilized in chemical reactions that complement the gasification process as a whole (usually by combustion to provide heat and reduce the quantity of char that is effectively burned to support the endothermic gasification reactions).

In order to accomplish this effectively, the key is to distribute the fuel more or less uniformly across the gasifier's diameter in the desired horizontal plane -- and this is difficult to do. Significantly nonuniform distributions, at best, result in gasifier inefficiencies, and at worst, lead to failures of a gasifier's internal components. The situation becomes more complex when the possibility of the fuel being a mixture of relatively fast burning gases and slow burning coal particles is considered (as in the case of top-gas recycle). It should also be remembered that practical hardware considerations effectively dictate that the horizontal plane for fuel injection will be physically fixed, and yet the physical location of any particular reacting zone within the gasifier (at any moment in time) is not well known, let alone well controlled. As a result, it is important to design the injection device to accommodate situations when the combustion zone is either higher or lower than the nominal position in the bed. This potential concern is greatly relieved in the slagging-bottom designs wherein the combustion zone is "stirred" by the injected gases to the point of being nearly homogeneous, and the temperatures are high enough that the combustion rates are very fast; in effect, the tuyeres effectively fix the physical location of the combustion zone.

Because of the temperatures involved and the need to distribute the fuel within the gasifier, the only practical means to carry the fuel into the gasifier is to entrain it in a gas stream (as opposed to some form of mechanical feeder). In all cases, the fuel to be injected is available at pressures below that of the lower portions of the gasifier bed, and consequently, the fuel stream must also be pressurized. Fortunately, both needs are satisfied by use of an eductor to propel the fuel into the gasifier. Eductor technology and pneumatic transport are reasonably well understood and will not be discussed in this report. Rather, the focus is on the key design issue: the development of a technique to penetrate the gasifier wall and solids bed at a non-slugging location and distribute the injected fuel relatively uniformly in a plane within the bed, without significantly disrupting the generally downward flow of solids in the bed.

Figure 16 depicts the key features of a conceptual design for a chevron injector unit, which appears to satisfy the above requirements and was evolved during this study. For purposes of discussion, it is assumed there is a general downward gravitational flow of solids as indicated in Section A-A of the figure and the injected fuel is a fluid-like mixture of noncondensable gases, vapors, and entrained coal fines. The chevron injector unit consists of the chevron shield, the injector pipe, and the reactor wall mountings.

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Figure 16. Chevron Injector Configuration

Operationally, the inverted V or chevron provides a mechanical shield for the injector pipe and generates an ullage volume around the injector pipe. This volume under the chevron acts as a channel to allow the injected fluid to flow longitudinally after exiting the injector pipe and promotes a reasonably uniform, ultimate distribution of the fluid into the moving bed of solids, at least in the vicinity of the chevron injector unit. For larger scale beds requiring more lateral uniformity as is likely to be the case with commercial gasifiers, multiple chevron injector units could be arrayed in a plane, e.g., as parallel units or even crossing in a "cruciform" configuration. The included angle of the chevron is chosen such that the solids flow downward smoothly without any tendency to be supported by the sides of the chevron. (That is, given that the chevron's centerline is vertical, the chevron's included angle should measure no more than 180° minus twice the angle of repose of the solids within the gasifier.) The shape of the chevron's apex is not particularly important and is primarily determined by structural and manufacturing considerations; both rounded and pointed shapes should work. The overall width of the chevron at its bottom represents a partial blockage of the general solids flow area within the gasifier, and this must be allowed for in the gasifier design. This is particularly important for designs utilizing an array of injector units.

The size of the injector pipe and its location relative to the chevron interior surface controls the major physical dimensions of the chevron, and thus is an important design aspect. The size of the pipe is determined by the amount of fluid to be handled, and thus is a relatively straightforward calculation. The pipe location is a less obvious determination and is a function of the angle of repose of the solids within the gasifier, the physical characteristics of the fluid being injected, and the design specifics of the injector pipe itself (e.g., perforation shapes and locations, pipe cross-sectional shape). The downward flowing solids within the gasifier will spill around the bottom edges of the chevron to refill the gasifier volume below it. The resulting shape of these lower surfaces of the ullage volume created by the chevron is, consequently, determined by the bed solids' angle of repose. The amount of void volume necessary around the injector pipe is determined by the injected fluid's ability to flow adequately in this voidage. This really needs to be determined empirically, but as a gross generalization, a reasonable minimum clearance-point dimension is likely to be in the range of 0.3 times the injector pipe diameter.

The injector pipe perforations must be designed (a) to not plug over the pipe's operational lifetime as a result of the injected fluid's characteristics and the general operating environment, and (b) to dispense the fluid into the ullage volume at a spatial frequency and pipe exit velocity that results in an adequately uniform ultimate distribution of the fluid into the bed of solids. Specific locations of the perforations are not otherwise critical, as long as the exterior (to the pipe) clearances discussed above have been provided. Some minimization of chevron width (and consequent reactor cross-sectional blockage) can be obtained by shaping the injector pipe cross-section so as to optimize the flow channel under the chevron, but this is expected to be a significant manufacturing complication for a small performance gain. The more general case, using a simple cylindrical pipe with axially distributed, "thru-drilled" holes to provide multiple sidewall perforations and having one sliding end support, is expected to be the most cost-effective approach and is shown in Figure 16.

It is expected that commercial gasifier applications of this concept will have dimensions large enough and operational temperatures disparate enough that relative thermal expansions could become a significant design issue when the need to mechanically support both the chevron and the injector pipe are considered. This is accommodated by structurally fixing the inlet ends of both the chevron and the injector pipe and providing a sliding support on the opposite ends, as shown in Figure 16. A "bayonet plug" type of support is shown for the injector pipe (which could also allow the pipe to be fabricated without an end closure), but conceptually, any type of sliding support would work.

The pressure seal to the gasifier volume and also to the injector pipe interior is achieved by use of a flange on the inlet end of the injector pipe with sealing surfaces on both faces. As shown in the figure, the injector pipe flange is trapped between the gasifier's wall-nozzle flange and the inlet fluid's supply-pipe flange and their associated pressure seals.

Maintainability is a key consideration for equipment of this type, and this concept's design features provide significant capabilities in this regard. First, there is to be an appreciable clearance between the inside diameter of the gasifier's wall nozzle and the outside diameter of the injector pipe (0.5 inches would be representative). Most importantly, this feature enhances the ability to easily withdraw the injector pipe, once the flange connections are removed. Specifically, the clearance provides (1) an ability to accommodate some level of thermal distortion in the pipe and/or some level of exterior deposit buildup; (2) a means to better thermally isolate the injector pipe from gasifier wall influences; and (3) with the injector pipe removed, the larger opening provides enhanced access to the underside of the chevron for possible scraping to remove deposits. Second, it should be recognized that the existence of the ullage volume under the chevron means that the vessel does not have to be drained of solids for removal of the injector pipe. (The system would have to be depressurized, however.) The injector pipe, while inherently physically hardy and expected to be reliable, is judged to be the only component in this concept that could need maintenance more frequently than the gasifier itself. Third and finally, the use of an open-ended injector pipe configuration with thru-drilled wall perforations (as shown in the figure) offers an additional benefit; with this pipe design, it would be extremely easy to ream out any deposits that did occur during the pipe's operational lifetime. (Removal of deposit buildup within the injector pipe is judged to be the most likely reason for any maintenance associated with the whole concept.)

4.2 INTERNAL STIRRER

A very large fraction of the U.S. coals of commercial interest have significant swelling/caking tendencies (typified by the coal having a free swelling index greater than 2.0). Various approaches have been postulated as means for utilization of swelling coals within a fixed-bed gasifier. The most frequently suggested are pre-oxidation of the coal, dilution with noncaking feedstocks (e.g., char), and gasification using grossly excess quantities of steam. However, the most proven and practical approach for a CGIA system appears to be the

incorporation of an internal stirrer. In many senses, the stirrer can be regarded as technically mature since it has been successfully utilized at 12-foot-gasifier-diameter scale with strongly caking coal in a Lurgi gasifier during tests at SASOL. However, as discussed in Section 4.1.1, it appears preferable to deviate from the Lurgi coal feeding design, which will also result in a deviation from the Lurgi stirrer support/drive design. In addition, there are broadened operational modes possible for the stirrer, which are discussed later in this Section (and in Section 4.9), and these potentially represent new stirrer technology. It should be remembered that for the fixed-bed gasification of strongly caking coals, the stirrer and its operational protocol are among the most critical aspects of the whole gasifier.

The basic role of the stirrer is to perturb the upper portions of the coal bed enough to (a) break up agglomerates that tend to form as the coal heats and becomes sticky (i.e., as the coal swells), but (b) not so much that the stirring action generates significant fines that leave the top of the bed or mix the bed to the degree the essentially "plug-flow reactor" character of the fixed-bed gasifier is destroyed. As the downward flowing coal continues to heat in the gasifier, at some point well below its ignition temperature, it loses its agglomeration tendencies, and if significant agglomerates have not been allowed to form by this point in the gasifier, they will not be formed subsequently. (Operation of the gasifier bottom at clinkering temperatures is an exception, and this will be discussed later.) More to the point, the maintenance of a relatively pebble-like nature of the upper bed assures a relatively uniform bed porosity, which translates into a really uniform upward gas flow, maximum coal throughput capability, and maximum useful gas output. The key, of course, is to stir in the right temperature zone of the bed and with an adequate, but not excessive, amount of "violence." This is complicated by the reality that in-bed temperature zones vary spatially with gasifier operating conditions (and are difficult to measure reliably). In addition, the temperature zone in which stirring should be accomplished and the degree of violence to be utilized are functions of the specific coal.

Ideally, stirring would only be done from the middle portions of the coal devolatilization zone to the upper portions of the gasification zone within the gasifier. As can be imagined, the mechanical loadings are formidable but, with some empirical design, a shape for a water-cooled stirring "blade" or arm can be developed that is structurally sound and adequately (but not excessively) perturbs the bed in the plane of rotation and for reasonable distances both above and below it. (While various shaped arms have been tried in the METC gasifier, a simple design based on round pipes was as effective as any and was the lowest in cost.) Further, it appears that two blades in the same plane are not significantly more effective than one, and as a consequence, a design with two opposed arms at vertically offset locations will effectively stir an appreciable vertical region of the bed. This appears to be the basis of the Lurgi stirring arm design, and it is the basis for the METC stirrer arm design, both of which were shown earlier. If the vertical height of the zone to be stirred exceeds the capability of two offset arms, either additional arms may be added (at the cost of increased driving torque requirements and bed heat loss), or the stirring arm assembly may be translated vertically as it rotates to helically sweep a larger bed volume (at the cost of considerably

increased complexity of the stirrer shaft/support/drive design). What is operationally important is that the bottom arm must adequately sweep the deepest necessary portion of the gasification zone when at the *maximum* gasifier coal throughput condition, and the top arm must sweep the uppermost necessary portion of the devolatilization zone when at the *minimum* gasifier coal throughput condition, and there must also be no unstirred region between the arms. If this can be accomplished with a particular stirrer arm configuration (and for a particular coal), the capability for vertical translation can be omitted from the design. What, if any, vertical motion capability exists in the standard Lurgi stirrer is not public knowledge, but a qualitative evaluation would lead one to believe that without it, the range of useable gasifier operating conditions would appear to be limited -- at least for many U.S. coals.

Functionally, it would appear that both rotational and vertical translation capability are required to accommodate at least the more difficult eastern U.S. coals. However, the subject of stirrer arm design and stirring protocol is one requiring empirical development and is currently unoptimized. Probably the most experience that is publicly available exists as a result of the work on the METC and GE gasifiers described in Section 3.2. Both of these gasifiers were intended for research purposes with multiple coal types, and both employed similar stirrer designs with rotational and translational capability -- and both were found to work well. It is probable that with some development, a stirrer configuration and rotational rate can be selected that for a particular coal or a range of coals will provide an adequate bandwidth of gasifier operating conditions without the use of vertical translation. This should result in a considerable gasifier-design simplification and cost reduction.

The stirrer used in the METC gasifier had the capability of sweeping essentially the entire bed (down to within a few inches of the grate). While this is well beyond what will be required for a commercial gasifier (which is normally expected to operate on a defined coal and thus a relatively narrow range of characteristics), this type of capability will be required as a tool to define a preferred stirrer configuration and stirring protocol for a range of coals or coal properties. The existence of this capability in the METC unit provided an opportunity to utilize the stirrer to gain further insights into gasifier operation through instrumentation installed on the stirrer itself. The two most useful examples were provided by thermocouples installed on the trailing edge of the lower stirrer arm and strain gages installed on the stirrer shaft.

The lower stirrer arm environment is "hostile" enough that the thermocouple sensing junction had to be contained in a short, armored protrusion in order to have an operating life-time that exceeded a few hundred hours. (The output leads were routed inside the stirrer assembly to reach the gasifier exterior.) In addition, the entire stirrer shaft/arm assembly incorporated a cooling water circuit that was required for structural integrity of the assembly but had the effect of heat-sinking the thermocouple's sensing junction. The result of this was that, while the bed temperature measurement was not accurate in an absolute sense, the shape of the bed's thermal profile was clearly discernable. Given knowledge of the physical position of the stirrer arm, locations of ash and combustion zones then became clear, and in addition, any gross lateral irregularities could be identified as well. The temperature profile was

established on every stirrer sweep cycle through the gasifier and thus was available in near real-time. As a consequence, this profile was found to provide a very good basis for controlling the gasifier grate speed (which, in turn, controls the vertical location of the ash and combustion zones within the gasifier). An example of these data is provided in Figure 17. The various in-bed reaction zones and the temperatures that would be representative are also indicated on the figure.

Comparison of the representative temperatures with the measured values makes the contribution of thermal damping/hysteresis in the thermocouple data obvious. However, the transition to a steeply rising temperature slope (and a peak temperature) in the combustion zone and the steep drop off as the stirrer penetrates into the ash layer over the grate is equally obvious. During the operational period in which these data were taken, the gasifier was utilizing a Pittsburgh seam coal and a stirrer sweep cycle that was 52 vertical inches, 15 minutes long (with the descending and ascending portions taking 10 and 5 minutes, respectively), and at a rotational speed of 1.0 rpm. (These values may be considered representative for coals with free swelling indices above 8.5, but it should be mentioned that rotational speeds in the range of 0.3 to 0.5 rpm have also worked well.) It has been found that more orderly data are obtained if the ascending portion of the sweep cycle is done relatively quickly (and the data ignored) so that more discernable data may be obtained on the descending portion of the cycle (and this is the only reason for the unequal durations within the sweep cycle). These data have implications pertinent to gasifier automation, which will be discussed in a subsequent section.

While this temperature profiling technique has worked very well as a basis for gasifier control, this particular approach to obtaining the data may not be workable in the context of commercial operation. There are several potential problems related to long-term cyclic passages between gasification and ash zones. Of primary concern is the erosion/corrosion environment, which is severe (for the lower portions of the stirrer assembly) because of mechanical abrasion effects combined with exposure to varying high-temperature atmospheres that are alternatively chemically reducing and oxidizing in character. Similarly, the need for really deep penetration of the gasifier solids bed lengthens the stirrer shaft and puts very significant structural requirements on the design of the stirrer shaft, support bearing, and seal to the pressure vessel, in particular. The usage of an over-bed lateral support spider for the shaft, as might be necessary with the mixer/distributor design discussed in Section 4.1, could considerably reduce this particular potential problem.

In another series of experiments, strain gages were emplaced on the length of exposed stirrer shaft internal to the gasifier to measure the resistance to stirrer rotation resulting from the solids bed. Measurements were made during various stirring protocols, and these measurements provided data that can be utilized to predict the forces on the stirrer shaft and arms for various scaled-up designs under similar operating conditions. Not surprisingly, it was noted that the required torque was reduced by anywhere from 30 to 50% when the stirrer was held in a fixed vertical position (i.e., the stirrer arms sweeping a constant plane) as opposed to operating with combined rotational and translational motions. However, it was surprising

Start Time of Plot 0: 0: 0 9/21/1983
 Stop Time of Plot 1: 0: 0 9/21/1983

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Figure 17. Thermal Profile as Determined From Stirrer Instrumentation

to find a near linear dependency of the torque requirement as a function of bed depth for periods wherein the stirring protocol utilized combined rotational and translational motions. The orderliness of this dependency potentially provides another route to the determination of bed depth and perhaps an ability to locate bed structural anomalies (e.g., voids or crusts). This is a possible basis for improved stirrer control logic and is discussed later in Section 4.9.3.

As Figure 13 showed, the more recent configurations of the METC gasifier utilized a hydraulic stirrer drive, as opposed to the earlier mechanical drive. The ability to apply a pure torque to the shaft without associated bending forces was an obvious and major improvement. However, there are two other significant advantages to a hydraulic drive design approach: it greatly facilitates the use of nearly any stirring protocol (e.g., differing upward and downward translational speeds), and it can readily provide a direct analog to stirrer torque in the form of the hydraulic pressure required to drive the rotational function.

An additional non-trivial factor requiring consideration when designing a stirrer shaft to operate with combined translational and rotational motions is the means by which the integrity of the gasifier's pressure seal is maintained. Typically, this is a packing gland seal at the point where the shaft penetrates the pressure vessel. Designing this for replacement or maintenance with the gasifier at operational conditions (but depressurized) is an obvious requirement. However, a more subtle (but very significant) influence on seal life is the hardness of the stirrer shaft along the portion of its length that is a sealing surface, and this should be recognized. Similarly, the degree to which the shaft support bearing, which is generally a combined assembly with the seal, allows the shaft to laterally "wander" in the vicinity of the seal is critical, but is typically manageable. Excessive freedom in this shaft bearing both makes the pressure sealing capability more problematical, and potentially allows the stirrer arm tips to rub the interior walls of the gasifier. What complicates this area is the fact that the sealing-surface length of the stirrer inherently is both cooled and exposed to the dusty and tarry atmosphere (and, potentially, even submerged in the coal bed itself) inside the gasifier for portions of its operating cycle. In effect, tar can glue dust or grit onto the sealing surface of the shaft during this cycle and result in its being dragged through the support bearing and the pressure seal, rapidly degrading both and possibly also scoring the stirrer shaft.

The most success in this area has resulted from application of the hardest possible surface coating on the sealing area of the shaft combined with a wiper ring on the pressure side of the bearing. This approach appears to work because the wiper ring reduces the amount of contamination entering the bearing/seal region, and the stirrer shaft is harder than either the grit or the bearing/seal materials. Consequently, because the grit crushes and the seal and bearing are maintainable, the assembly can function for several thousands of hours without maintenance. Unfortunately, several potentially attractive hardening processes have not been capable of being successfully applied simply because of the inherent geometry and dimensions of the stirrer shaft. The coatings that have worked best have been titanium-carbide- and chrome-iron-based, particularly one where the final step is a boron carbide "pack cementation hardening" process that converts the exterior 0.01- to 0.02-inch layer to titanium diboride --

which is exceedingly hard. (Similar hard facings were utilized on the stirrer arms, but the shaft sealing area turned out to be the more significant problem.) The definition of hard-coating processes applicable to parts having the size and dimensional tolerances of a stirrer shaft is an area warranting more work. An alternative solution to some of these design problems is to package the entire stirrer drive unit inside the pressure envelope; however, this represents significant costs related to both the pressure vessel itself and the increased complexity of maintenance operations.

4.3 GRATE DESIGN

The grate in a fixed-bed gasifier provides three important functions: it is the physical support for the mass of the bed, it volumetrically removes the lowest portions of the bed solids at a controlled rate, and it (usually) provides the initial distribution of the reactant gases across the diameter of the gasifier. If everything works as intended, the bottom of the bed is removed at the same rate the combustion zone is moving upward in the gasifier, causing the combustion zone to remain vertically fixed. (Meanwhile, the feeding system maintains the location of the top of the bed at a fixed location.) The volumetric solids-removal rate is determined by the grate's rotational rate. (Other means are possible but are felt to be less practical.) The operating environment generates a requirement for cooling of the grate in order to maintain the grate's mechanical and structural integrity without the use of exotic materials or designs. Typically, this is accomplished in two ways: (a) the steam and air used for gasification are introduced below the grate and pass up through it, thus cooling the grate; and (b) the combustion zone is maintained well enough above the grate that an appreciable layer of ash (1 foot in thickness would be representative) rests on the grate and insulates it from combustion zone temperatures. Obviously, the efficiency of the gasifier is enhanced if the reactant gases are distributed appropriately (which is nearly uniformly) across the gasifier cross-section and are supplied hot to the combustion zone. To this end, the ash layer also serves as a final reactant gas distributor and preheats the rising gases prior to their reaching the combustion zone.

Of the three functions, the maintenance of a controlled solids removal rate is the one that presents the most challenge by far. There are two basic grate design approaches: (a) design the grate to cause the ash to tumble generally downward along the surface of the grate to a withdrawal port, or (b) incorporate an auger function into the grate to withdraw a theoretically constant volume of ash for each rotation. The first approach normally requires less grate drive power and allows the grate to be designed as a very effective reactant gas distributor, but it requires critical control of the combustion zone (ash) temperature, consistent coal ash composition, or both. The second approach readily allows incorporation of an ash crushing function into the grate design and is, consequently, relatively forgiving of coal ash and combustion zone anomalies, but the design generally requires more drive power, is harder to cool using the reactant gases, and is a relatively poor gas distributor. The Lurgi and METC gasifier designs shown in Chapter 3 are examples of the first and second approaches, respectively.

The first approach typically results in a grate with a conical or pyramidal shape. If the upper portion of the gasifier is functioning properly, the coal char entering the combustion region will be in a pebble-like state and will have no agglomeration tendencies. The key to the operation of the combustion zone is to suppress combustion temperatures (typically by the use of steam) enough that major agglomeration or clinkering of the coal ash does not occur. Since suppression of combustion zone temperatures has the effect of also suppressing gasification rates (as well as causing the gasifier to require more steam than is necessary for gasification), there is incentive to operate a gasifier with a pyramidal grate at combustion zone temperatures that *just slightly* melt the ash to form hard ash nuggets as they leave the combustion zone. A key property of the nuggets is that they have a very shallow angle of repose in a bed. As a consequence, the slight agitation provided by the grate's rotation causes the nuggets to "tumble" down the sides of the pyramid and be discharged through a narrow gap between the grate's skirt and the gasifier interior wall.

Typically, the conical surface is perforated at frequent intervals, which allows a fairly precise radial distribution of reactant gases into the ash layer to be achieved and also enhances the ability of the ash to tumble down the conical surface. The pyramidal design is susceptible to problems caused by the formation of oversize nuggets resulting from local or global combustion-zone temperature excursions, significant variations in coal ash characteristics, or both. It would be possible to add an abrasion function to the grate to grind down oversize nuggets (at a slight increase in drive power requirement), but it is not known if this has been successfully tried. (A design to accomplish a grinding function with a pyramidal grate has been generated, but was omitted from this report to preserve the patent application process.) It is not credible to accommodate clinkers with a conventional pyramidal grate design. The conventional design could also have a problem with excessively low temperatures in the bottom of the gasifier for some operating conditions with particular coals. There are coals that decrepitate to a near powder state with heat, burn to produce a powdery ash if the temperatures are low enough, or both. A powder will have a larger angle of repose and, of more significance, if a powder were to reach the grate region, it could pack or bridge over a narrow discharge opening, which would greatly reduce the ash discharge rate and could easily shut down the gasifier. In this case, increased combustion zone temperatures would be required to provide for slight melting of the ash and subsequent agglomeration into nuggets.

The second approach is normally referred to as a "crusher" grate, though "scooper" would be more appropriate. There are several means to achieve what is effectively the scooping-out of a volume of ash for discharge through the grate itself, combined with a crushing function for any agglomerates too big to pass through the grate. A typical embodiment of this approach involves the use of multiple parallel horizontal plates, several of which are eccentric to the gasifier centerline (the axis of rotation) and carry plows to control the radial flow of ash; the assembly is rotated within a strong wear collar, or "Bosch ring." A three-plate configuration was utilized in the METC gasifier as shown in Figure 18. In operation, the grate rotates clockwise (as viewed from the top), and the top two plates are eccentric to the gasifier centerline. Ash is scooped from the gasifier wall region by the plow located on the middle plate; the ash is forced radially inward to pass over and under the middle plate

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Figure 18. Configuration of the METC Gasifier Grate

and fall through large central holes (in the middle and bottom plates) into the ash hopper below. The grate's top and middle plates cover these central holes with enough overhang to preclude ash from free-falling into the ash hopper below, and a deflector plow causes ash resting on the top plate to be deflected and fall to a lower level where it is picked up by the middle plate's plow. While some ash falls past the periphery of the bottom plate into the ash hopper, the vast majority exits the grate region through the center hole in the bottom plate. Any agglomerates too big to pass through the gaps between the plates tumble past the plow and are crushed as the gap between the eccentrically mounted plates and the surrounding Bosch ring goes to near zero clearance.

The principal virtue of this grate design is its ability to effectively control the flow of nearly any type of ash, from powder to small clinkers. The grate is quite forgiving of off-design operational modes in portions of the gasifier above the grate, and the size of clinker that can be accommodated is determined by the plate-to-plate spacing (6 inches in the case of the METC gasifier), the holes in the plates being larger than the plate-to-plate spacing, and the *aggressiveness* of the crushing function. In reality, it is not practical to make this design an extremely robust clinker-crusher, but the capability that can be provided is appreciable and can compensate for many untoward events which would otherwise shut down the system. As can be envisioned by looking at the figure, this grate is at best a poorly designed baffle-plate reactant gas distributor. Consequently, the ash layer must be relied upon to distribute the gases across the gasifier diameter to a greater degree than occurs with the conical grate. As with the conical grate, the ash layer provides thermal insulation for grate protection. As a result, a several-inch deeper ash layer would be anticipated with a crusher grate.

Both mechanical and hydraulic grate drives are useable for either grate design. However, the operability bandwidth of a gasifier with a crusher grate is further enhanced if the grate drive also has a capability to back up. It is possible to stall the grate if a really significant clinker is encountered (the assembly should be designed to *stall* rather than *break*), and it has been found that a capability to rotate the grate backwards, even as little as 10° of rotation, and then go forward often will allow a significant clinker to be crushed. It has not been apparent that more than 180° of reverse rotation was ever beneficial. A hydraulic grate-drive design would make this sophistication easier to incorporate.

At one time, experiments were done using armored thermocouples mounted on the top plate deflector plow to measure temperature of the ash descending on the grate. Like the stirrer tests discussed in Section 4.2, the objective was to determine whether a temperature rise could be sensed that would indicate approach of the combustion zone, and thus form the basis for grate-speed control logic. The tests were stopped for other reasons before any conclusion could be drawn, but a number of problems were found and some insights are possible. While the problems of extracting a useable signal from this location are formidable, the major problem was achieving a reasonable thermocouple lifetime in this operating environment, primarily because of a combination of the mechanical loadings and abrasion on the thermocouple sheath itself. There was an appreciable level of variation associated with the readings that were obtained, and it was not resolved whether this "noise" was a real phenomena or not.

The primary credible explanations included upward flowing reactant gas intermittently blanketing the thermocouple, large hot (or cold) ash chunks intermittently contacting the thermocouple, and unreliability of the signal transfer technique (which was based on enclosed slip-rings) from the rotating grate and within the ash hopper environment. As mentioned above, it is not expected that a stirrer-mounted temperature-measurement device will prove practical for grate speed control; however, the value to the system of a workable combustion zone sensor would be very high, and the grate mounted approach was not proven infeasible. These tests were done more than 10 years ago, and the results must be tempered by the gasifier's operating conditions and the experiences of the period. At the point the tests terminated, the next approach would have been to mount a thermocouple mast on the grate's rotational centerline (as opposed to on the deflector plow) as a means to minimize the imposed lateral loads. It is not known whether this approach will work in a gasifier functioning in an otherwise proper manner, but it would appear to be potentially feasible.

4.4 GASIFIER WALL DESIGN

Design of the gasifier wall is essentially a selection between a refractory-lined and an actively cooled wall design for the gasifier vessel itself. The gasifier pressure-shell design is simplified considerably by lowering its operational temperature, and the two conventional ways of accomplishing this are (a) to line the inside with an insulating, high-temperature refractory; or (b) to provide a jacket through which cooling water is circulated. Combinations of the two approaches are also credible, wherein some zones of the gasifier would utilize a water-cooled wall and others would be refractory lined. The preferred selection would be design specific.

The refractory lining approach results in a hot wall surface existing next to the solids bed, usually lower heat losses from the gasifier, and usually a thicker wall. As a consequence, the refractory-lined gasifier should have a higher efficiency, but it will normally also have a lesser amount of product gas output for a given exterior vessel diameter. Unfortunately, the experience with refractory-lined, dry-bottom, gasifier designs has not been very good. While, theoretically, this should be an excellent design approach, the insulating qualities are typically so good and the ability to effect temperature control of the combustion zone is so poor that a significantly increased tendency toward clinkering has been reported by nearly every user of this approach. (Improved gasifier-control capability could also make a significant contribution in this area.) For the slagging-bottom gasifier, the use of a refractory lining is almost the mandatory approach due to its insulating characteristics and also its general "inertness." An important consideration for refractory linings is their ability to withstand the abrasion inherent in the solids motion in what is really a "moving-bed" reactor, and similarly, the solvent/corrosive characteristics of the reacting atmospheres. This is a particular concern with slagging designs wherein the molten slag is known to be especially aggressive.

In contrast, the water-wall approach results in a cold wall surface existing next to the bed, usually the lowest pressure vessel temperature, and normally the more difficult wall

penetration design. The relatively cold wall will chill the adjacent portions of the solids bed and will result in noticeable wall effects on the exterior 9 to 18 inches of the bed, which can influence gasifier operation. Sidewall penetrations have the obvious requirement to maintain structural and hydraulic integrity, but the need to consider the cooling effects on the media that is passing through the wall can be more subtle and requires care in the penetration design. A good example is a gas inlet penetration to accept recycled top gas to a location within the gasifier's solids bed. Top gas inherently contains condensible tars; if these are cooled and condensed in the wall penetration area as a result of cooling jacket effects, the gas flow can be partially obstructed, which then allows more cooling to occur, resulting in more blockage, etc. The effect can be the formation of a flow blockage with a self-aggravating characteristic. The water-cooled approach is often preferred when the gasifier operates in a setting where there is a premium on steam raising capability, since the water jacket can be an effective boiler feedwater heater. This is not likely to be the case in a simplified IGCC application as these systems will virtually always include an HRSG to raise steam from the GT exhaust, and the feedwater heating potential provided by the gasifier is expected to be inconsequential.

The use of a steam-generating cooling jacket is in many ways a compromise between these two design approaches. In this approach, appropriately pressurized (and, perhaps, pre-heated) water is fed to an annular space in a dual-walled (metallic) gasifier. The annulus is designed to serve as a steaming surface, and the cooling jacket generates saturated steam at essentially the input-water pressure level. The effect of this is to raise the temperature of the metal wall facing the bed by several hundreds of degrees Fahrenheit (over that of a water-wall design), and to also provide an extremely high heat-transfer capacity. This reduces the heat losses (and the wall effects) incurred with a water-wall design but retains the capability to remove large heat fluxes when conditions require it. Both interior and exterior walls of the jacket effectively are designed as pressure vessels, but the interior wall (facing the coal bed) can be designed to operate at a near zero pressure differential, if desired. The exterior of the jacket must, of course, be packaged as a boiler would be (with appropriate insulation, for example). The conventional Lurgi gasifier vessel is based on this approach.

Probably the best approach (pending the development of improved control technology), is to utilize a steam generating jacket in the lower half of the gasifier and a refractory wall in the upper half. However, this design area was not addressed during the study. Similarly, the potential of a cost advantage of one approach over the other was not investigated, but no cost advantages were obvious.

4.5 SLAGGING-BOTTOM DESIGNS

A fixed-bed gasifier that is designed to operate with the combustion zone's global temperature at a level significantly above the melting temperature of the coal ash is referred to as a slagging-bottom (fixed-bed) gasifier. There are two primary incentives for this approach: (a) the increased temperatures in the lower portion of the gasifier increase the

gasification rates in the rest of the vessel (e.g., a specific coal throughput increase of about a factor of 3 is regarded as typical), and (b) the characteristics of the effluent ash itself make it somewhat preferable to ash from a dry-bottom gasifier (i.e., a gasifier with a grate). Specifically, the ash will be a hard glassy frit of lesser volume that is somewhat more environmentally benign and has some potential commercial value. Also important, but of generally lesser significance for the simplified IGCC application addressed by this study, the slagging gasifier requires less steam (because of the absence of grate cooling requirements), and because the volume of gases passing through the bed is lowered, there is less fine particle carry-over in the product gas per unit of energy produced. The disincentives to this approach are the increased design and operational complexities and the relatively small experience base, all of which are really one issue -- how well can the actual costs for a system based on a slagging-bottom design be predicted?

In theory, a slagging bottom could be incorporated into nearly any fixed-bed gasifier design; however, the only significant experience is with a modified, single-stage, Lurgi design. By far, the world's largest experience base with this gasifier design approach exists within the BGL organization and is embodied in the several BGL research/demonstration gasifiers (the general characteristics of which are discussed in Section 3.1), and few of the specifics are known outside of BGL. It should be noted, however, that most of what is potentially difficult about a slagging-bottom gasifier relates to dealing with liquid slag at nominally 3000 °F, and this is quite similar to basic steel making technology -- for which there is a large U.S. experience base.

The key to any slagging process is the management and control of the slag flow. For fixed-bed gasifiers, there are two approaches to the basic concept: continuous removal of the liquid slag, and intermittent removal or tapping of the slag. There was a relatively small, slagging gasifier built and operated by the U.S. Department of Energy (DOE) in Grand Forks, North Dakota, which was based on the continuous slag removal approach (Ellman et al. 1974; Willson et al. 1981; Hauserman and Willson 1983). The BGL gasifiers are based on the intermittent slag-tapping approach. The comparisons cited herein are largely intuitive because the small size and limited testing of the DOE unit and the paucity of specifics from BGL preclude anything more definitive. However, useful insights may be derived from understanding the pertinent basic characteristics. Probably the four most important factors to remember are

- The molten slag viscosity is *strongly* dependent upon slag temperature in the 2600 to 3400 °F temperature range, and the dependency is also a function of the specific coal and the amount and type of flux added (actually, a function of ash chemistry). Also, there is a hysteresis effect, that is, the viscosities (at the same temperature point) of the same slag after cooling and reheating are different.
- The generation of high temperatures in the combustion zone has major effects on the reactant gas requirements, specifically, flow rates and input temperature levels. These high temperatures also have a significant effect on the product gas characteristics,

particularly output temperature level, moisture fraction, and heating value (i.e., the hydrogen- and carbon-monoxide-constituent fractions).

- The density of the molten slag exceeds that of the solids in the bed, that is, the solids can float in the slag pool.
- The slag chemistry tends to have corrosive, erosive, and solvent characteristics that make it aggressive to many of the construction materials commonly used in high-temperature/high-pressure equipment.

Generally, the higher the combustion zone temperature, the more difficult and demanding the gasifier design and operation become. As an example, volatile ash species can migrate upward to condense on the cooler char and form sizeable agglomerates/crusts/bridges in the bed above the combustion zone. In contrast however, at higher combustion zone temperatures, the slag becomes more liquid-like, less variable, and has more predictable flow characteristics.

4.5.1 Design Considerations

In both approaches to slagging gasifier design, the physical support for the bed is provided by the bottom "head" of the vessel. Normally, the head is refractory lined and has one tap hole through which the slag drains. The art is to retain the solids within the gasifier and yet allow the liquid slag to drain through the tap hole at the desired rate. The reactant gases are introduced very low in the bed using sidewall mounted, multiple lances or tuyeres; to simplify achievement of slagging temperatures, pure oxygen has typically been the oxidant but air (preheated to approximately 1000 °F) is also reported to be useable. A small amount of moisture, typically steam, is added as a reactant for gasification and to a degree (and for the oxygen-blown designs in particular) also as a means to limit combustion zone temperatures. The reduced steam requirement in a dry-bottom design initially appears quite beneficial, but it must be considered on a system basis; potential downstream needs for product gas conditioning may obviate the benefit. (Either zinc ferrite desulfurization or GT fuel-flow control equipment could generate additional steam requirements.) The specific nature of the gas flow from the tuyeres appears critical to gasifier operation. It is known that seemingly small changes in tuyere positioning in the wall, specific tuyere dimensions, tuyere operational protocol, and tuyere gas-flow rates can have massive effects on the reliability of slag management and, thus, the general operability of the gasifier. What is not known is, are the tuyere parameters well enough understood that a slagging-bottom gasifier can be currently designed that will utilize an arbitrarily selected U.S. coal, be of commercial scale, and function successfully in a commercial setting? BGL states they are able to do this, but they are alone in this claim and since no units have been sold to date, this ability is unfortunately unproven.

For efficient operation of the gasifier, it appears the oxidant must penetrate to near the gasifier centerline, and the char/oxidant contacting must be reasonably areally uniform so that essentially complete burnout of the organic material occurs. However, it also appears the

fluid dynamic forces exerted by the tuyere jets on the solid/liquid mass in the combustion zone (and in the lower portions of the gasification zone) must be adequate to control the local solids flows as well as the slag formation in this region. Both DOE/Grand Forks and BGL experienced tendencies toward the formation of agglomerate "structures" within a functioning gasifier that were significant enough to impede operations and were greatly influenced by small changes to the tuyeres. As a consequence, the degree of solid/liquid mixing induced by the jets appears to be critical, and the actual physical and dynamic nature of the bottom portion of the bed is not entirely obvious.

The slagging bottom with a continuous slag-tapping approach, as utilized in the DOE/Grand Forks gasifier, is perhaps the easier to visualize. In this design, the combustion zone bed consists of physical pieces of hot char that are vigorously blown with the oxygen/steam tuyere jets. The jets nearly "wash" the bottom surface of the vessel and cause the locally resident char to burn at high enough temperature that the associated ash continuously melts on the individual char pieces, runs down to the bottom of the vessel, and runs out the tap hole. There is essentially no pool of liquid slag resident in the gasifier. Wall cooling effects cause a slowly moving crust of semi-frozen slag to accumulate (to a dynamically stable thickness) on the bottom and lower portions of the walls. The crust is continuously accumulating material, burning, melting, and descending to provide additional slag flow to the tap hole. The tuyere jets physically perturb the general solids bed somewhat and probably locally fluidize it, but not macroscopically. In addition, either the jets are strong enough (and pointed in the right directions) or the combustion rates are high enough that essentially no small pieces of char slip out through the tap hole. As mentioned earlier, the DOE/Grand Forks gasifier was small (nominally 16 inches in diameter at the tuyeres) with a correspondingly small slag flow (the tap hole was 1 inch in diameter). It is not known how this design approach would have fared at a commercial scale of perhaps 12 feet in diameter and with a tap hole of 6 to 12 inches in diameter. Certainly, the jet penetration and oxidant distribution into the solids bed and the preclusion of solids exiting through the tap hole could all become potential issues.

The intermittent slag-tapping approach has been followed by BGL, and the following is thought to be at least qualitatively correct but is not based on authoritative references. The BGL tuyere locations appear to be high enough that a pool of liquid slag of appreciable depth (but probably not more than 24 inches) can exist in the gasifier. The solids bed would float on the pool were it not for the weight of the solids above forcing at least some of the char pieces to be submerged in the liquid slag; nonetheless, many pieces (particularly the smaller ones) do float at or near the surface. A solids gradient would be expected in the pool, and the bottom of the pool is probably all liquid ash. The available descriptions imply that the tuyere jets are more energetic than those utilized in the DOE/Grand Forks gasifier. The indications are that the jets in effect stir this liquid pool vigorously and that the volume and velocity of gas flow, combined with vaporized ash species and near-molten char, promote the existence of pneumatically supported voids in the vicinity of the tuyeres, and possibly across the diameter of the gasifier. The bed is not macroscopically fluidized, but local solids circulation cells likely exist at or near the surface of the pool. As in the description above, a

dimensionally stable, continuously accreting and descending, slag crust likely exists on the walls and bottom because of heat losses. One significant difference is that it appears a slag bridge is allowed to form in or over the single tap hole. It is unclear what the physical state of the bridge is, but it is unlikely that it is very solid or BGL would never have been able to achieve the long and reliable operational periods they have been noted for.

The tapping operation has been described as initiated by a step reduction of the pressure in the slag quench chamber (which is nominally equal to the gasifier outlet pressure plus the pressure head through the total height of the gasifier), and this would place something approximating the gasifier's operating differential pressure-head as a differential pressure across the bridge. Assuming the character of the bridge is somewhere between a very viscous meniscus and a semi-solid crust, and assuming the tap hole is at least several inches in diameter, it would be credible that this pressure differential could break the bridge and allow the slag to drain. The restoration of the bridge is a less obvious process. It is described as being recreatable simply by raising the slag quench-hopper pressure. This would imply there is enough slag flowing through the hole, and that it is viscous enough that a pressure pulse to an operational level will burp the descending slag into the bridged position and hold it there long enough for the bridge to develop the necessary structural integrity. A credible alternative is that once tapping is initiated, the gasifier drains to an essentially dry state, the quench-hopper pressure is raised, and the newly produced slag is allowed to progressively freeze from the edges of the tap hole until the opening is bridged over with a semi-solid crust. It is also known that BGL utilizes a burner that is pointed at the underside of the tap hole. The burner's intermittent operation could greatly assist in control of slag tapping (e.g., "lit" to initiate tapping and "off" to stop). Either methodology requires quite precise knowledge and management of the slag's viscosity.

The method for precluding solids leakage out the tap hole is also not obvious with the intermittent tapping approach; it may be simply that a slag pool of sufficient depth is constantly maintained such that solids either float or dissolve and, thus, never reach the tap hole. There is an inherent virtue in the intermittent tapping approach, particularly if a liquid ash pool is constantly present, because the molten ash is a liquid with high thermal conductivity (essentially, a liquid metal in character). This pool will tend to make the bottom of the gasifier nearly isothermal and would serve as a ready ignition source for the descending char; as a consequence, the actual radial penetration and pointing vectors of the tuyeres are likely not nearly as critical with this approach -- assuming the slag pool is reasonably stirred by the tuyere jets.

A distinct virtue to the existence of the tuyeres is their role in controlling the physical location of the combustion zone, and consequently, the greater likelihood of predicting phenomena occurring at higher elevations and of generally controlling these aspects of gasifier operation. Similarly, the tuyeres offer potential as injectors for materials that could either be considered as fuels or materials that could be disposed of as a result of coming into contact with the molten slag. The best example is coal fines: BGL reports great success with tuyere injection of up to 30% of the coal feed as fines (in the form of coal-water slurry), and claims

to have not been limited at that point. At commercial scale, either of the slag tapping approaches will require a multiplicity of individual tuyeres, and the associated hot-gas plumbing that will be required will be a formidable design challenge.

If the key to a slagging process is the control of slag flow, the key to controlling slag flow is the design of the hearth plate area (and its ability to control and utilize the slag's viscosity). This area is at the bottom of the gasifier and contains at least the tap hole itself (the most critical single feature) and, normally, a "tap-hole burner;" however, "slag breakers" or similar devices to prevent formation of flow blockages are sometimes also present in a design. The hearth plate is typically a refractory structure with a round tap hole in the center and has provisions for a carefully designed amount of active internal or back face cooling. Normally, it is also designed to be an easily replaced insert in the bottom of the gasifier as it operates in an extremely hostile environment, and replacement during maintenance periods is common. The principal hearth-plate design requirements are to resist the corrosion and erosion effects of the molten slag, and to maintain the desired surface temperatures so as to maintain the ability for the slag to drain through the tap hole at the desired rate. The main thermal concern is the possibility of slag unintentionally freezing in the tap hole and precipitating a blockage. A similar concern is the possibility that slag could freeze on the top surfaces and form an undesired slag dam, bridge, or other "uncontrolled" obstructive shape. Usually, the consequences of excessive temperature are a wash out or a structural failure of the hearth plate itself. There is normally a tap-hole burner located below the hearth plate that focuses a flame on the slag tap-hole exit. The environment below the tap hole is much cooler, typically the pressurized, water-vapor-filled, ullage over a quench water bath. This cooler environment can cause the slag to form stalactites and, generally, to not flow freely (drop) from the tap hole exit into the quench bath.

The tap-hole burner's role is to provide any extra heat that may be required to keep the slag flowing freely in this exit area. It normally burns continuously (gas fueled), and the POC vent up through the gasifier -- they are a small fraction of the gases passing through the gasifier and do not impact the product-gas characteristics significantly. In theory, this is all that is needed in the hearth plate area; however, several designs have incorporated a slag breaker as a kind of insurance policy. The slag breaker is typically a metal bar that intermittently sweeps across the region just downstream of the tap-hole opening. Its purpose is to break off any forming stalactites and allow them to fall into the slag hopper. While the stalactites would be expected to be somewhat plastic in nature, the forces the slag breaker-bar might have to exert could be significant; as a consequence, the bar itself is typically water-cooled and is swept through the environment of the tap-hole exit/tap-hole burner only intermittently. If the bar is too close to the tap hole (and slag temperatures are too high), there is the possibility of slag freezing onto the bar; if the bar is too far from the tap hole, there is the possibility it will not be effective at assisting the tap hole to remain open. There is an additional concern in the event a stalactite is firmly adherent to the hearth plate: the shock imparted by the breaker bar to the stalactite may structurally fail a part of the hearth plate. The maintenance of the correct temperatures at the tap hole is absolutely critical to the operation of a slagging gasifier!

4.5.2 Slagging-Gasifier Mathematical Model

The ability to mathematically model at least the macroscopic phenomena within a slagging-bottom, fixed-bed gasifier is valuable as a tool to understand and, hopefully, estimate the important influences on overall gasifier performance. There are no really mature and robust mathematical models that allow adequate simulation of a dry-bottom, fixed-bed gasifier, and models for an air-blown, slagging-bottom design are even less mature. However, a methodology was developed in the course of this study that was felt to provide an adequate (for this study) approximation of the performance of an air-blown, slagging-bottom gasifier. The following describes the model.

The materials available were (1) mathematical models and data from air- and oxygen-blown, dry-bottom gasifiers; (2) published descriptions of oxygen-blown, slagging-bottom gasifier performance data and predictions from BGL; and (3) verbal descriptions from BGL personnel that, if the air was preheated enough, the gasifier would operate in a normal fashion but would produce a somewhat higher than normal output gas temperature with a lowered heating value (primarily reflecting the dilution by air-borne nitrogen). Three different gasifier analyses were conducted before selecting the one to be used in the system simulations. Each analytical approach predicted a product-gas temperature higher than would occur with oxygen-blown operation, which seemed at least qualitatively correct. The first approach was based on satisfying the energy requirements in the slagging zone of the gasifier. The key assumption was that switching from oxygen to air blowing would not change the configuration or the performance of the slag zone. In essence, this assumes that, if enough additional energy is provided to heat the inert nitrogen (entering with the oxygen) to the slag-zone temperature, the same slagging conditions will be achieved. The outputs from this calculation procedure were an estimate of the required air mass-flow and air preheat. The second approach used these estimates (700 °F air preheat and an air-to-coal ratio of 4 lb/lb) as inputs to an existing, but somewhat simplistic, Advanced System for Process Engineering (ASPEN) computer model for oxygen-blown, slagging-gasifier simulations (Stearns Catalytic Corporation 1984). (This computer model was compared with BGL slagging gasifier data and was shown to be in reasonable agreement [Stefano 1985].) This model was modified to use air rather than oxygen as the oxidant; however, the model predicted what were felt to be unrealistically high combustion zone temperatures (e.g., 3800 °F). It was concluded that either the estimated air preheat or flow requirements obtained from the first approach were too high, or this relatively simplistic ASPEN model for an oxygen-blown, slagging gasifier was not suitable for predicting air-blown operation without further modifications. Without experimental air-blown slagging-gasifier data, neither approach could be validated. To resolve these problems, a third approach was initiated and came to be adopted as the final gasifier simulation model.

The analysis procedure for the third approach was initiated in the combustion zone at the bottom of the gasifier and was based on oxygen-blown slagging-gasifier operation. Previous analyses and data from oxygen-blown units led to the use of assumptions that 70% of the carbon input with the coal would remain to be consumed in the combustion zone, the carbon/

ash (i.e., char) would be at 2000 °F on entry to the combustion zone, and that 0.3 lb of steam/lb of coal would be a reasonable operational ratio. Given these assumptions and the knowledge that compressed air will exit from a GT at above 700 °F, chemical equilibrium calculations were performed on a mixture of ash, carbon, steam, and air, for various levels of air preheat temperature, air-to-coal ratios, and combustion-zone heat loss. Figure 19 shows the results of these calculations for 700 and 1000 °F air preheat and heat losses of zero and 2%. (As a point of comparison, previous calculations had shown 4% represented a reasonable value for heat loss from the entire gasifier.) While the specifics of Figure 19 may not precisely represent a slagging gasifier, the trends are likely to be close to correct. This figure clearly shows the strong influences that air preheat temperature and air-to-coal ratio have on the combustion zone (i.e., slag) temperature. Observations from various slagging units had shown that 3000 to 3200 °F was an operationally workable range for slag temperature. Consequently, the starting points for analysis of the air-blown case were a slag zone temperature of at least 3100 °F, a steam-to-coal ratio of 0.3 lb/lb, a gasifier inlet air temperature of 700 °F, and an adiabatic combustion zone (all of which correspond to the middle curve of Figure 19). This resulted in an estimated blast-air requirement of 3.3 lb of air/lb of coal and a combustion zone temperature of 3150 °F. This third approach was checked by using the same procedure for an air-blown, dry-bottom gasifier. The results for the amount of carbon consumed and zone exit gas temperature agreed fairly well with calculated results from prior dry-bottom gasifier simulations and were felt to confirm this portion of the approach.

Predicting gasification and devolatilization zone phenomena probably has more uncertainty than any other aspect of the third approach to developing a slagging-gasifier mathematical model because of all of the assumptions that had to be made. The basic assumptions were

- Char resulting from devolatilization was composed only of carbon and ash.
- The tar yield was 4 lb/100 lb of coal, based on BGL data for an oxygen-blown slagging gasifier (as opposed to a tar yield of 13 lb as estimated for a dry-bottom gasifier).
- For the fraction of tar that cracks (or 9 lb/100 lb of coal), 46% of the carbon ended up in the gas phase (based on tar-cracking data from Battelle Pacific Northwest Laboratories).
- Gas-phase carbon from tar cracking ultimately ended up as carbon monoxide because of the much higher gas temperatures in the upper portions of an air-blown, slagging gasifier.
- Methane, ethane, and ethylene were formed only in the devolatilization zone.

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Figure 19. Effect of Air-to-Coal Ratio on Combustion Zone Temperature

- The volatile hydrocarbon gas yields were reduced proportionally to the tar yield (i.e., 4/13th of the yields estimated for a dry-bottom gasifier) because of the much higher gas temperatures.
- The carbon dioxide yield from devolatilization was the same for slagging- and dry-bottom gasifiers.
- The hydrogen, steam, and carbon monoxide yields from devolatilization were established from a material balance.

These assumptions allowed estimation of the amount of volatile carbon released in the devolatilization zone. The amount of fixed carbon that gets converted in the gasification zone could then be estimated by subtracting the volatile carbon and the carbon converted in the combustion zone from the total carbon. The gasification zone was then simulated using an equilibrium analysis that combined the gas flow, composition, and temperature from the combustion zone with the fixed carbon and ash. This provided the gas flow, composition, and temperature at the top of the gasification zone.

The gasification-zone exit gas was then considered to pass through a devolatilization zone wherein it was adjusted for temperature as a result of the coal's drying and heating (from its ambient temperature at the gasifier input), an assumed heat loss (4% of the coal's energy), and the release of volatiles. At the devolatilization-zone exit point, the gas temperature is the product-gas outlet temperature and was determined to be 1800 °F. This is 900 °F higher than the exit temperature reported by BGL for their oxygen-blown slagging gasifier using Pittsburgh No. 8 coal. The higher temperature is a direct consequence of the approximately 50% higher equivalent oxygen-to-coal ratio (0.77 lb/lb in the air-blown case compared to 0.5 for the oxygen-blown case), the 500 °F higher preheat temperature on the oxidant feed (700 °F as compared to nominally 200 °F), and the much greater sensible heat flow up through the air-blown slagging gasifier (as a result of the hot nitrogen flow). In addition, prior analyses had shown that the product-gas temperature correlated very strongly to slag temperature; the 3100 °F slag temperature used in these calculations could be slightly too high, which would noticeably elevate the product-gas temperature. As with dry-bottom units, the actual exit-gas temperature will also be highly sensitive to the coal type and its moisture content. The final product-gas composition was then obtained by combining the gases from the gasification zone with the volatile products released in this (devolatilization) zone and imposing a water-gas shift equilibrium condition at 1800 °F. Finally, elemental balances were made as a check on the overall material balance, and an overall energy balance was computed to check the validity of the assumed carbon flow into the combustion zone. Since the overall energy balance closed to within 1.3%, it was concluded that this solution was close enough, and the carbon flow to the combustion zone was not iterated.

This third approach was validated by simulating an oxygen-blown, slagging gasifier. The predicted gas composition agreed closely with that reported by BGL. In addition, the predicted product-gas temperature was only about 40 °F lower than the reported temperature,

and the overall energy balance closed to within about 2%. Since the agreement with reported oxygen-blown results was good, the decision was made to accept this third approach to analyzing an air-blown, slagging gasifier.

This calculational procedure was used as a basis for a larger scale mathematical simulation to predict the overall performance of a slagging gasifier in a system configuration; the results (System Case 7) are described later in this report.

4.6 TWO-STAGE DESIGNS

In a two-stage, fixed-bed gasifier, a portion of the product gas has characteristics representative of conditions at an in-bed location, and the residual portion of the product gas has top-gas characteristics. The Woodall-Duckham gasifier described in Section 3.1 is an example of this design. The incentive for this approach is that a single coal gasifier can produce two product gas streams with considerably different characteristics, and it can increase the ability to utilize caking coals. In comparison to a single-stage gasifier, the top gas is typically enriched in hydrocarbon condensibles, has a higher heating value, and is somewhat cooler. The side gas is typically withdrawn from a location within the gasification zone and can be nearly free of condensibles, has a lower heating value, and is hundreds of degrees Fahrenheit hotter than the top gas (e.g., 800 °F would be a representative temperature differential). The specific gas streams' characteristics are dependent upon the level in the bed at which the side gas is withdrawn and the relative volume flows of side gas and top gas.

The sum of the two streams' energy flows is essentially identical to the energy output from a comparably sized, single-stage gasifier. However, each gas stream's individual volumetric flows are considerably less than the volumetric output of a comparable single-stage gasifier, and the cleanup requirements for each stream are distinctly different: the top gas has a relatively high concentration of sulfur and nitrogen species and is complicated by the presence of condensible hydrocarbons but with a modest particulate loading; the side gas has a significant particulate loading but is low in condensibles and sulfur and nitrogen species. As a result, each stream's complement of cleanup equipment can be tailored to the stream's particular characteristics, and since the cost of gas cleanup equipment is greatly influenced by the actual volume of gas to be cleaned, the total system's cost may be reduced for some applications.

The two-stage gasifier was originally developed to expand both the types of coals and the markets that could be served by a fixed-bed gasifier without a stirrer. These designs typically provided comparatively low gas flows, long residence times, and cool temperatures through the gasifier's upper regions, which resulted in both a capability to accommodate higher swelling index coals (a capability to utilize coal with an index as high as 2.5 was often claimed) and a more manageable and desirable tar by-product (with a correspondingly lower tendency to deposit heavy pitch within the gasifier). The tars were lower in particulate content and viscosity than from a single-stage unit (because of reduced cracking within the

single-stage gasifier), which made them both more easily dealt with on-site and more marketable. The side gas would often be passed through a cyclone for particulate removal, the top gas passed through an electrostatic precipitator for tar removal, and the streams would then be recombined to provide a relatively clean, warm, fuel gas for combustion applications. Most of the commercial, two-stage gasifiers were atmospheric pressure units and were designed so that approximately 50% of the gas exited the top, and the top-gas and side-gas temperatures were about 250 and 1200 °F, respectively. In sharp contrast, for the pressurized, two-stage gasifiers considered in the CGIA study, the lowest top-gas temperature considered was 800 °F. The reasons for this temperature being considerably higher than what was practiced in the atmospheric-pressure units are as follows:

- Higher operating pressures lead to higher partial pressures of the various tar components, and thus there is a larger quantity of these components available to condense (deposit) and cause operational problems. Thus, compared to atmospheric units, pressurized units must operate with higher exit gas temperatures to avoid tar deposition within the gasifier.
- Tar, which is a complex mixture of literally hundreds of identified compounds, has a boiling point (vapor pressure) curve that spans many hundreds of degrees. Thus, the heavier tar fractions can condense at rather elevated temperatures (probably heterogeneously on entrained coal fines). Operationally, it becomes a question of how much tar can be permitted to condense before problems develop, and the amount of material condensed is reduced by increased gas temperatures.
- The published top-gas and side-gas temperature ranges for the high-pressure, two-stage, Lurgi Ruhr 100 gasifier are 750 to 1100 °F and 1100 to 1500 °F, respectively.

In addition, it has been found that the METC, single-stage, fixed-bed gasifier exit-gas temperature must be kept higher than approximately 900 °F (when operating with bituminous coals) to avoid significant tar deposition in the pipe from the top of the gasifier through the cyclone to the humidifier. Since tar composition depends on the conditions in the upper part of the gasifier bed, it is difficult to predict what the tar properties will be in a two-stage unit operating at higher pressures than the METC gasifier. However, based on the experiences with atmospheric pressure units, the tar becomes more manageable in a two-stage gasifier, and this translates into a lowered acceptable minimum top-gas temperature. This will probably also be the tendency in two-stage gasifiers at higher pressures, thus, the minimum top gas temperature was lowered somewhat from the METC experience.

The two-stage gasifier will also have a particular advantage in settings where in essence there are two differing applications for the coal gas. The side gas is potentially well suited to heat exchange applications or as a preheated, lean fuel-gas for combustion applications that can tolerate a small amount of sulfur in the fuel. The top gas is a reasonable source for the production of coal tars and is also a very rich fuel gas that can be utilized directly in any combustion process having an inherent emissions reduction characteristic.

(Some basic metallurgical processes -- and the prototype system considered in this study -- have this feature.) The attractiveness of this approach for a simplified IGCC application is a function of how cleverly these differing product-gas stream characteristics can be integrated to achieve an improvement in overall system cost and performance.

4.6.1 Design Considerations

As mentioned above, the volume of the side gas and the location in the bed from which the side gas is withdrawn are key determinants of both gas streams' characteristics. In the general case, the intent is to obtain about half of the total product-gas volume as side gas taken from the gasification zone. This gas will be expected to have a heating value of somewhat under 130 Btu/scf, a temperature approaching 1500 °F, and will contain about 5% of the total hydrocarbon condensibles output from the gasifier. The removal of the side gas obviously lessens the gas flow and gas velocity up through the rest of the gasification zone and all of the devolatilization zone (for a constant diameter gasifier). This slows the heating rate of the coal in the upper portions of the gasifier relative to a single-stage unit, and this tends to require a slight lengthening of the gasifier and allows the gas to cool more by the time it reaches the top of the coal bed. Since release of coal volatiles is a function of the coal temperature level (which is essentially the same for both single- and two-stage gasifiers) and since there is a lesser volume of gas passing through the upper portions of the two-stage gasifier, the result is a significantly increased fraction of hydrocarbon condensibles in the top gas. This is reflected as a significant increase in heating value of the top gas, and levels of 225 Btu/scf would be considered representative.

The reduced gas velocity at the top surface of the coal bed also has the effect of lessening the propensity for fines carry-over in the top-gas output stream. The formation and release of approximately half of the contaminant sulfur species (depending on the pyritic and organic sulfur fractions in the coal) and at least one-third of the nitrogen species occur in the very top of the gasification zone and higher regions in the gasifier. As a result, if the top-gas stream output flow is relatively low, it can be comparatively rich in environmentally significant contaminant precursors. As the volume fraction of side gas is increased (from 50% of the total output gas flow), its heating value is increased and the top-gas temperature is decreased (and the gasifier design length is increased). As the side gas withdrawal point is moved higher in the bed at a constant top-gas-to-side-gas ratio, the side gas temperature is reduced, its heating value and hydrocarbon condensibles content are both increased while the top gas temperature is reduced, and its hydrocarbon condensibles fraction is decreased (and the gasifier design length is also increased, but only slightly).

A prototype system configuration was considered in the CGIA study with a novel PFBC that utilized a limestone bed and was "fueled" by streams that were environmentally difficult to accommodate. (The PFBC unit is discussed in Chapter 5.) During this study, it was found that overall system performance was enhanced if the fraction of energy sent to this PFBC was reduced relative to the fraction of energy utilized directly as gas turbine fuel. As

a result, the design of the two-stage gasifier became a compromise between minimizing the quantity of top-gas flow (thus, minimizing energy to the PFBC) and maximizing the quantity of volatiles in the top gas (thus, reducing system NO_x and SO_x emissions). This had the result of shifting the preferred fraction of top gas to be only about 30% of the total output gas flow.

Another major consideration is the technique used to separate the side gas from the general gas flow through the solids bed. The Woodall-Duckham design was based on the conventional approach of utilizing a gasifier diameter increase and an internal skirt to provide a void volume into which gas could flow without the danger of significant solids flow into the gas passageway. The bottom of the skirt presumably defines the elevation within the gasifier from which side gas is withdrawn. While this approach has been reliably used by several designers, it has the characteristic of withdrawing a gas that is representative of the gas in the vicinity of the vessel wall -- not necessarily gas representative of the cross-sectional plane in the gasifier defined by the bottom edge of the skirt. Flow anomalies appear to be most common along the wall of a gasifier, and cooling effects are almost solely confined to the bed regions adjacent to the wall. As a consequence, it is unlikely the side gas with this design is representative of the cross-section of the gasifier, and of more significance, the characteristics of the withdrawn gas are likely to be much more transiently variable than the characteristics of the gas as a whole in the cross-sectional plane. The cooling effect on the gas is exacerbated by the need to pass the gas vertically upward for some length (to preclude solids entry into the gas passageway) prior to collecting the gas at an exit nozzle through the gasifier wall. While this design has been reported as mechanically reliable, it should be noted that the radial gap (skirt exterior to gasifier interior) for typical gas flows is small, and any solids that do find their way into this gap may very well stay there. An alternative is to use a chevron injector unit (described in Section 4.1.2) operating in the withdrawal mode. An additional advantage of this approach over the radial-gap design is the relatively large volume (and the associated larger minimum gap dimension) over the region of the bed from which gas is being withdrawn. This would be expected to provide a form of settling chamber and to reduce the population of fines that would otherwise be found in the side-gas stream.

A basic assumption frequently made when contemplating a staged gasifier is that the gasifier is operating as it is supposed to, and that the various reaction zones are in their proper spatial locations. Unfortunately, history has shown this is not always the case, and while off-design operation can be inconvenient with a single-stage gasifier, it can be a major problem with a staged gasifier. The worst problems can normally be expected to arise when the side-gas characteristics approach those normally associated with the top gas (because the combustion zone is too deep in the gasifier or has become too thin, or the top-gas-to-side-gas ratio has been shifted), and the downstream components are not equipped to deal with the levels of condensibles, the sulfur and nitrogen contents, and/or the lower temperature and higher heating value. However, problems can also arise when the top gas characteristics approach levels more typical of the side gas (because the combustion zone is too high in the gasifier or has become too thick); however, this situation is likely to be less severe. When the reaction zones are spatially mislocated, the *unusual* characteristics of the less affected gas

stream become exacerbated; i.e., with a low combustion zone, the top gas becomes even cooler and with a high combustion zone, the side gas heating value drops even further. Similar phenomena can occur with a disruption of the top-gas-to-side-gas output-flow ratio but, since these flows are controlled external to the gasifier, this condition should be easier to preclude. It is worth mentioning that a significant period of time (e.g., 1 hour) at off-design gas flows can easily mislocate the reaction zones. A way around this potential problem is to design both streams' cleanup systems to be able to accommodate a degree of off-design gas characteristics, but this obviates an advantage of this concept. For this concept, a sophisticated and reliable gasifier control system is vitally important.

4.6.2 Two-Stage Gasifier Mathematical Model

As with the slagging-bottom gasifier model, the existing mathematical models were not adequate to represent the thermal and chemical phenomena occurring in the vessel. A particular complication was the inability to mathematically remove an arbitrary fraction of the reactant gas flow at an arbitrary vertical elevation (i.e., the side-gas collection level). Again, the resulting model was not as elegant as it could have been, but was felt to be good enough for the purposes of this study.

The analysis was done in two steps: calculations were first done to develop the general mass flow and thermal trends and relationships, and then a particular top-gas-to-side-gas split was chosen for more detailed analysis. This second analysis generated the model that was utilized in the larger-scale systems simulations. A constant top-gas outlet temperature of 800 °F was assumed to limit the calculational options and to simplify the analysis process. While other top-gas outlet temperatures are certainly possible, this value is about the minimum that will cause essentially all tars to be in the vapor state (and thus to pass out of the gasifier with the top gas), and it maximizes the flow rate of the side gas. The 800 °F top-gas temperature is also consistent with the levels actually experienced in at least two units of this type when gasifying a good coal. A summary of the steps in the trends analysis is as follows:

- Coal at ambient temperature is considered to enter the top of the gasifier and be heated by only the top gas while descending through the devolatilization zone to a temperature plane -- which would correspond to the physical elevation at which the side gas was removed. Coal temperature levels of 750 and 1100 °F (corresponding to coal devolatilization levels of 50 and 90%) were considered for this plane to investigate two reasonable levels of devolatilization.
- The net energy requirements to heat and devolatilize the coal from its ambient temperature input condition to the temperature-plane level (either 750 or 1100 °F) were calculated, assuming the top gas was exiting at 800 °F and using established procedures to determine the total energy required by this devolatilization zone (Institute of Gas Technology 1978).

- For a Pittsburgh No. 8 coal, the air-to-coal ratio (or blast ratio) that would be utilized was estimated to be bracketed by the values of 2.12 and 2.41 (lb of air/lb of coal) as representative low- and high-blast ratios. The gas composition and characteristics at the two designated coal temperature-plane conditions were estimated for both low and high air-blast ratios and for side-gas temperature levels of 1200, 1400, and 1600 °F by subtracting published pyrolysis yields observed at these temperatures from the product-(top) gas component yields actually obtained in the METC and GE gasifiers.
- The top-gas flow that would satisfy the devolatilization-zone energy requirement and the corresponding side-gas flow (which is removed from the gasifier at the coal-bed temperature plane) were then calculated.
- The energy flows attributable to the top-gas and side-gas streams were calculated from the gas characteristics and the corresponding flow rates.

Figures 20, 21, and 22 provide the energy flow fractions, mass flow fractions, and the heating values resulting from these calculations for the two gas streams at the low-blast ratio. This analysis was used to estimate all the significant gasifier output species for a larger-scale simulation of a system employing a two-stage gasifier.

Refinements were then made to the trends analysis. These refinements consisted of revising the coal to be the reference coal for this study (which was only slightly different than the Pittsburgh No. 8 coal utilized above) and calculating more precise gas characteristics. An air-blast ratio of 2.12 lb of air/lb of coal (low-blast ratio) and a steam-to-air ratio of 0.4 lb of steam/lb of air were selected to enable the direct use of existing METC test data (gasifier Run No. 106), because the associated parasitic energy consumptions (air compression work and steam flows) would be lower and would result in a more efficient overall system, and because Lurgi (Corman 1986) felt they could achieve successful operation at these air and steam flows (which are lower than conventional Lurgi values). A limestone PFBC to consume environmentally noxious fuels was incorporated into the system configurations being considered; this was found to result in a cost/performance disincentive when a large volume of top gas was produced. Given this consideration and the 2.12 blast ratio, the conditions selected for the detailed analysis were a top-gas flow that was 32% of the total output-gas flow (but containing 64% of the total weight of volatiles output) with the side gas at 1600 °F. While a higher side gas temperature would allow a still lower top-gas flow, this value was at the maximum gas temperature used in the trends calculations and at the maximum to be considered based on the Ground Rules. (While 1600 °F was based on gas usage as a GT fuel, it still represents a reasonable limit based on gas handling difficulties.) The additional steps involved in the detailed analysis are summarized in the following.

- The weight of fines entrained in the top gas was considered to be 5% of the fed coal weight (which is probably high), and the fines were assumed to devolatilize completely to a char consisting of carbon and ash.

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Figure 20. Gas Energy Flow Fractions

Figure 21. Mass Flow Fractions

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Figure 22. Product-Gas Heating Values

- The amount of bottom ash was calculated assuming all carbon was reacted in the gasifier, i.e., the solids at the bottom of the gasifier consisted only of ash.
- The weight of the tar/oil yield was considered to be 13% of the fed coal weight. This was lowered from the 15% tar and oil yield measured with the METC gasifier at this blast condition to account for the CGIA gasifier being nominally 100 psi higher in pressure than the METC gasifier at the time the test data were taken. The yield difference (2%) was assumed to result from tar cracking to a mixture of methane, carbon monoxide, and carbon dioxide with the same total energy content. (Note: For the still higher pressure system cases using a STIG GT cycle, the tar/oil yield was lowered to 8%.)
- The required product-gas yield for overall material balance closure was calculated by subtracting the output tar and oil, coal dust, and bottom ash weights from the total weight of the input streams.
- The coal-borne input sulfur was assumed to all be converted to H_2S and the coal-borne input nitrogen was assumed to be 90% converted to ammonia with the residual appearing as nitrogen gas; all of these exited in the gasifier's output-gas streams.
- Gasification product gases and volatile species were split between the top and side gas in accordance with the trends analysis results.
- Using the heating values and estimated yields for the individual species, the energy flows of the top- and side-gas streams were calculated and checked against the results of the trends analysis. Similarly, an elemental balance was performed to assure consistency between input and output material flows.

This resulted in the two-stage gasifier representation utilized in the systems analysis (described subsequently as System Cases 1 and 2).

4.7 GAS-RECYCLE DESIGNS

A fixed-bed gasifier employing gas recycle is a derivative of a two-stage gasifier design approach. The gas-recycle design routes the top gas to a location deep within the gasifier's solids bed, and the gasifier's product gas is output from a single gas port at a location similar to the side-gas port of a two-stage design. The KGN fixed-bed gasifier described in Section 3.1 is an example of this design concept. The incentive for doing this is to eliminate (or at least greatly reduce) the process complexities related to dealing with a condensibles-laden top gas. Using the top-gas recycle approach, it is possible to achieve a fixed-bed product-gas stream with many characteristics similar to those of the product gases from fluidized-bed or even entrained-bed gasifiers.

In this concept, the top gas is injected into a hot region deep within the gasifier bed, and the associated tars and oils and coal fines initially pass upward within the solids bed along with the rest of the reactant gas flow. The fines become trapped by the mass of solids in the bed and are reacted along with the char as it moves down through the bed. The condensable hydrocarbon vapors are heated and in contact with active char surfaces, and given the proper injection point into the gasifier, the vapor can chemically crack to lighter hydrocarbons and elemental carbon (which is subsequently reacted as the char moves lower in the bed). The light hydrocarbon gases may then either further crack or remain as a portion of the upward gas flow within the bed, depending on the temperature, pressure, and residence time conditions within the bed. In addition, if the conditions are conducive, a significant fraction of the ammonia may be cracked to produce environmentally benign nitrogen gas. A majority fraction of the upward gas flow is extracted from the gasifier at an elevation in the bed below the level at which the coal is evolving problem volatiles. The resulting product gas characteristics consequently resemble those of the side gas of a two-stage gasifier, discussed in the foregoing Section, but the product gas will have a somewhat lower heating value as a result of the gasifier's inherently higher steam and air consumption.

4.7.1 Design Considerations

There are two basic approaches to this concept: injection of the top gas really deep in the bed so that, in essence, it serves solely as a fuel for the combustion zone and thus results in providing energy to drive the gasification reactions occurring in the bed above; or injection of the top gas into the gasification zone at a point wherein the temperature, pressure, and residence time conditions promote sufficient cracking that the output product gas is very low in condensibles. The first approach requires lengthening of the gasifier because a greater fraction of the coal's fixed carbon is being gasified (since this fraction does not need to be combusted to provide heat), and this will raise residence time requirements for the gasification zone, i.e., length. The KGN fixed-bed gasifier (described in Section 3.1) is an example of the first approach. The second approach is more technically elegant but is also difficult to achieve in an operational gasifier. The second approach also requires lengthening of the upper portions of the gasifier. (Which approach requires the greater degree of lengthening is not currently known.) The increased length requirement occurs because the residence times required for relatively complete tar and oil cracking are appreciable and because the residence time required for gasification is significantly increased (because of the lowered gasification zone temperatures resulting from the injection of relatively cooler, recycled top gas). In addition, with this approach, it is critical to maintain the various reacting regions within the operating gasifier at constant (and known) spatial locations. As a consequence, top-gas reinjection into or near the combustion zone of the gasifier is considered the most workable gas recycle approach; however, it normally entails the solution of a formidable technical problem, that of gas injector design.

Two other key considerations are the means by which the top gas is pressurized so that it can be transported into the lower regions of the gasifier and the means by which it is

distributed across the diameter to react in an essentially uniform manner with the solids inventory of the gasifier. The propensity for blockages and deposits to be created in piping as a result of the amount of fines and condensible tars and oils typically found in fixed-bed gasifier top-gas has been discussed earlier in this report and will not be repeated here. However, it is worth noting that the top-gas stream from a gasifier employing gas recycle is richer in condensibles than the product gas from a single-stage, fixed-bed unit, although it is not as rich as the top gas from a two-stage unit (without recycle). The use of steam as an educting media to pressurize and propel the top gas into the gasifier appears to be a logical design approach. The only potential concerns are (1) that the steam be hot enough that it not produce noticeable tar and oil condensation in the gas transport piping, and (2) that the quantity of steam utilized (and the location and manner of injection into the gasifier) be as consistent as possible with the reactions taking place within the gasifier (e.g., the combustion reactions should not be quenched). To elaborate further on this latter point, it should be remembered that the steam requirement for a dry-bottom gasifier is normally dictated by clinker prevention considerations and grate cooling, and these needs normally exceed gasification reactant requirements. When the realities of in-bed distribution of the recycle gas are considered, the steam used in the top-gas eductor is not very effective at mitigating clinker formation in the combustion zone; consequently, the eductor steam flow is mostly in excess of the gasifier's real needs (and it is totally in excess in the case of injection into the gasification zone).

It is a relatively straightforward design problem to achieve reasonably uniform distribution across the diametrical cross-section of the combustion zone of a slagging gasifier utilizing tuyeres for gas injection. However, accomplishing a reasonable distribution into the combustion zone of a commercial-scale, dry-bottom gasifier is another matter entirely! The major technical issue with designing a top-gas distributor in this application is achieving compatibility with the temperature level in this region (which typically exceeds 2000 °F in normal operation) and the mechanical loads imposed by the solids bed. While the referenced description was not entirely clear, it appeared the KGN gasifier utilized the grate as the top-gas distributor. (The relevant aspects of doing this were discussed in Section 3.1.) While an aggressively cooled distributor design is credible, it must be remembered that many of the top-gas constituents are very flammable and would be expected to burn essentially upon injection into the bed. As a result, when using an aggressively cooled distributor, a non-trivial fraction of the chemical energy released from the top gas is likely to go into the cooling system -- and to not provide energy to the gasification reactions.

The technique by which a portion of the gas flowing up through the gasifier may be collected from a plane within the solids bed and output as the product-gas stream has the same generic considerations as were discussed in the previous section for the side-gas stream of the two-stage gasifier. Two differences could be significant, however. First, in this case, all of the product gas exits from the side-gas outlet, and thus, gas velocities leaving the solids bed will be higher than occur in a two-stage gasifier (for gasifiers with the same coal throughput), and particulate entrainment could be increased. Second, it is likely the outlet port can be higher in the bed than the side-gas port of a two-stage gasifier, since for the same concentration of volatiles in the output gas stream, the total mass of volatiles that can be

tolerated will be higher. This would correspond to withdrawing gas from cooler coal that should be less friable (and perhaps even somewhat sticky) and consequently could release fewer fines into the product-gas stream. Whether the fines fraction of the product-gas stream is actually increased or decreased is not currently obvious.

4.7.2 Gas-Recycle Mathematical Model

The fixed-bed gasifier utilizing top-gas recycle is a variation on the two-stage gasifier design, and the mathematical model was derived from the analysis done to develop the two-stage gasifier model. The model is complicated by the single product-gas stream being a function of the characteristics of the top-gas stream that is recycled to the lower portions of the gasifier, and as a consequence, the analysis process is inherently iterative. An outline summary of the mathematical model is provided below.

- The two-stage gasifier model discussed in the preceding Section was utilized as a starting point. The top-gas temperature was held at 800 °F, the product-gas temperature was held at 1600 °F, and the product-gas outlet port was located at the horizontal plane in the bed at which the descending (and heating) coal had reached 750 °F.
- The top gas was considered to be recycled to a location somewhat above the combustion zone. This was done because preliminary versions of this model consistently predicted that slagging temperature levels would result if the injection point were in the combustion zone of a dry-bottom gasifier.
- The eductor steam requirements were computed using Zenz's relationships (n.d.), based on the amount of top gas to be transported to the lower portions of the gasifier. Initial assumptions were needed for the first iteration; the key assumptions were the gasifier's air-to-coal ratio was to be 2.12 lb of air/lb of coal, the split fraction for the gas within the bed was to be the same as that for the two-stage gasifier without recycle (i.e., 0.2864 for the 32/68 top-to-side-gas output ratio), and the recycle gas was to have the characteristics calculated for the top gas of a two-stage gasifier. (Note: these parameters will actually be the results of the iteration process.)
- The eductor steam and top-gas species flows were added to the gases leaving the combustion zone to obtain the reactant gases existing in the gasification zone. The gasification-zone outlet-gas compositions and flows were then calculated from previously derived gasification gas yields, but were adjusted to account for steam reforming of the hydrocarbon condensibles carried in with the top gas and those released below the side-gas withdrawal plane. (As an example, for the majority of U.S. coals, approximately 50% of the coal's original volatile matter is released below the plane at which the coal has reached 750 °F.) These compositions then became the gasifier's output product-gas compositions.

- The fraction of the gasification-zone outlet-gas flow that became gasifier product gas was determined, based on achieving an energy balance in the devolatilization zone above the plane from which the product gas was withdrawn. This then allowed recomputation of top-gas compositions and flow rates and allowed the iteration process to continue to stability.
- When iteration of the gas composition and flow calculations stabilized, the oxygen and overall energy balances were computed and the air-to-coal ratio was adjusted to satisfy the balances (while assuming the steam-to-air ratio -- as fed to the bottom of the gasifier -- remained fixed). This affected the energy balances internal to the gasifier and consequently redefined the recycle gas flow; as a result, the internal gas composition and flow computation iteration loop was restarted at this point.
- When all recomputations stabilized, the overall elemental and energy balances were checked for closure.

A (stabilized) mathematical characterization of a fixed-bed gasifier using top-gas recycle was utilized in the systems analysis activities (System Case 6), which is described in a later chapter.

4.8 OTHER GASIFIER FEATURES

A number of design features are not of major importance but are worth understanding and do offer possible routes to increasing the performance of fixed-bed gasifiers. The principal candidates for this list are an internal cyclone, a "drafted" diameter, and a gasifier dome outlet.

A key component in the simplified IGCC system approach is the hot cyclone that removes the vast majority of particulate contamination in the product gas while allowing any hydrocarbon condensibles to pass through. This component eliminates (or at least greatly reduces) a host of potential complications caused by the co-existence of tar, oil, and particulate in the equipment downstream of the gasifier. There are two primary needs to be satisfied for the successful use of a hot cyclone: it must be maintained at a temperature (and pressure) approximating the gasifier outlet conditions, and there must be a beneficial use for the particulate that is collected (which closely resembles coal dust). Both of these needs can be satisfied by locating the cyclone within the gasifier; however, the associated design problems are not trivial.

Conceptually, an internal cyclone would be located in the ullage volume at the top of the gasifier and would employ a dip-leg that routes the removed particulate to a location within the solids bed. The first location ensures that the cyclone would be at gasifier outlet conditions without the need for features such as a pressure shell or heat tracing. If the dip-leg outlet location is deep enough, the captured particulate becomes part of the reacting solids

bed without a need for additional processing -- rather than a waste material. The most complex issues with this design approach relate to the design of the dip-leg. For gasifier designs intended to accommodate coals that have even slight caking and swelling tendencies, an in-bed stirrer will be required, and it is doubtful (but not known) that the dip-leg outlet could be high enough in the bed to always be above the stirrer and yet be deep enough to cause at least the majority of the deposited particulate to be trapped by the coal in the upper regions of the bed. Routing the dip-leg around the stirred volume is possible, e.g., via a pathway within a thick refractory wall, but this adds questions relating to the nonuniformity of the eventual particulate distribution within the solids bed and to the somewhat increased gasifier complexity. Other solutions are possible but typically are exceedingly complex, e.g., incorporation into the stirrer shaft. The other critical need for successful functioning of the dip-leg is that it must always maintain enough of a particle inventory (i.e., pneumatic pressure drop) to prevent significant gas flow up the dip-leg, and this can translate into excessive length or complex valving and plumbing requirements. The internal cyclone approach also can impose a general thermodynamic performance and cost penalty in that it can add significant length to the gasifier vessel. Whether the gasifier penalty becomes a system penalty (or benefit) is a function of the design specifics.

The diameter of a gasifier is considered to be drafted when it changes with gasifier length, as with a tapered vessel. A vessel with an abrupt change or step in diameter may also be considered to be drafted. The solids in a fixed-bed gasifier are heated and reacted as a result of the effectiveness of gas-solids contacting and the residence time at appropriate conditions. A drafted diameter is a means to tailor the gas velocity (i.e., gas-solids contacting) and the solids velocity (i.e., residence time) in the various regions of the gasifier in order to maximize gasifier performance. Some designs employ both increasing and decreasing drafts; an example is the Woodall-Duckham design discussed in Section 3.1. Drafting also can assist in mitigating the consequences of potentially deleterious phenomena, such as coal swelling. In summary, drafting a gasifier's diameter is a means to *tune* the gasifier to achieve improved performance, but usually only for a narrowed range of feed coals.

The successful use of drafting in a gasifier design process requires an excellent understanding of the processes and phenomena occurring within the gasifier (including the coal's characteristics). While these are certainly generically understood, the ability to model and predict them is currently imperfect, at best. Drafting the gasifier's diameter can impose additional requirements on other components of the gasifier that can be significant; the best example is the stirrer (specifically, stirring arm length and extent of stirrer vertical travel).

The placement of the gasifier's outlet-gas port usually receives minor consideration in designing a fixed-bed gasifier, but it can have a significant effect on the product-gas particulate loading. To utilize a coal with even a small caking or swelling tendency, a pressurized fixed-bed gasifier will inherently have a somewhat empty ullage volume over the coal bed inside the gasifier. This volume will exist as a result of several factors: a curved pressure vessel head, a coal feeding system that places coal on top of the bed, an in-bed stirrer with constant diameter arms, and possibly the nature of the penetrations for the stirrer and feeding

system. There is obvious incentive to minimize this volume since it increases vessel heat losses and civil engineering complexities; however, a significant volume will still be there. It is also well known that the fines concentration in the gas leaving the top of the coal bed decreases with increasing height (assuming no flow blockages are introduced to accelerate the gas) until a distance known as the transport disengagement height or TDH is reached. For realistic conditions, lengthening the gasifier ullage volume to reach the TDH is not practical since this corresponds to a very significant increase. However, the rate of fines concentration reduction also decreases with increased height over the bed, and as a consequence, it does make sense to utilize this phenomena to reduce the fines content in the product gas to the greatest extent practical. Placing the outlet port as high in the dome as possible is generally a good idea. Further improvements may be achieved by placing a baffle at the entrance to the port such that there is no line-of-sight path from the coal-bed surface to the port opening; however, care must be taken to not induce significant gas acceleration. It should also be possible for any particulate clump to fall back out of the port without becoming trapped by the baffle. The baffle is not an area that has been well researched, however.

4.9 CONTROL SUBSYSTEM

In a very real sense, the control subsystem brings together all the individual design decisions relative to the gasifier; its function is to orchestrate the various components and features so they function as an integrated whole. The requirements and a description of the major gasifier control functions were provided in Chapter 2, Section 2.2.3, and will not be repeated here (but the reader will find it useful background for the following discussion). Similarly, most of the component-specific instrumentation discussions were provided within the previous portions of this Chapter. The following discussions are limited to the gasifier itself and assume the gasifier is supplying an essentially constant pressure fuel-gas line from which the GT draws fuel as required to generate the desired system power output. Higher level controls that manage the overall system and controls for other major subsystems are, consequently, not included in the discussions.

Few technical subjects have been laden with as much esoteric *lore* as the subject of controlling the operation of a fixed-bed gasifier. A fixed-bed gasifier is normally regarded as easy to control, and the state of the bed is well known. While these statements are true, they are unfortunately really only applicable to a gasifier at atmospheric pressure supplying a steady-state demand that also is not particularly sensitive to variations in gas stream characteristics. (In addition, these statements are usually based on operating practices that are temporarily unacceptable in the U.S.) When a fixed-bed gasifier is operating at an appreciable pressure and when there is a requirement to load-follow while satisfying the operational requirements of a contemporary IGCC system (with regard to acceptable operating costs, emissions, Occupational Safety and Health Administration [(OSHA)] practices, efficiency, and availability), the performance requirements for the control system strain the existing knowledge base. It appears probable that an advanced control system would allow improved performances to be obtained from conventional fixed-bed gasifiers, and advanced control systems will probably be required to realize the potential of some of the advanced design features described in this report.

4.9.1 Classification of Operational Phenomena

An important characteristic of a fixed-bed gasifier is that there are operationally significant phenomena occurring simultaneously but in vastly different relevant time frames. In addition, while most of the phenomena are continuous, some significant ones are intermittent in nature. Since the gasifier's role in life is to supply fuel gas on demand to the rest of a system, a useful background reference point can be provided by defining and classifying the phenomena that affect the characteristics of the gasifier's product gas (including the basic capability to continue to produce gas) and the time frames relevant to the phenomena.

Short Relevant Times -- Fractions of seconds to tens of seconds:

- Volume of gas output -- This parameter directly and very quickly responds to the input reactant gas-feed rates. Strictly speaking, it is also dependent on the volume of the bed that constitutes the devolatilization zone, but this is normally nearly a constant and so slowly changing in any event that it can be ignored for this parameter.
- Slag flow -- Changes in slag flow do not directly affect the output gas characteristics, but they do affect the spatial distributions of the various reaction zones and hence even the ability to continue to produce gas. While there are changes in slag characteristics that happen over periods of many minutes, the important changes happen in small numbers of seconds and basically relate to the freezing and thawing of slag flow obstructions. The most important slag property is its viscosity, and this is a very strong function of slag temperature.

Medium Relevant Times -- Fractions of minutes to tens of minutes:

- Product gas temperature -- A high-temperature condition, in theory, is a result of the combustion zone becoming too high in the bed (zone migration is a slow moving phenomena), which reduces the volumes available for the endothermic upper-bed reactions (given a fixed location for the top of the coal bed). The outlet temperature can readily be reduced in medium time frames by increasing the solids throughput as a means of artificially shifting all of the reaction zones deeper into the gasifier. This also has the effect of thickening the devolatilization zone. Unfortunately, the same effect is operationally produced by the formation of gas channels in the upper portions of the bed wherein combustion zone gases bypass the upper bed and raise the product-gas temperature (and reduce the heating value). In this case, the stirrer is utilized to destroy the channels. These conditions are practically indistinguishable (without very elaborate instrumentation), and the best approach is to maintain a stirrer protocol that precludes the formation of non-uniform bed porosities such that this possibility can be disregarded. A low product-gas temperature is the result of a low combustion zone location and is only slowly respondent (by reducing solids throughput). Changing the steam flow to the gasifier (with no other changes) is influential in the time frame of minutes; however, this has significant ramifications on reactant zone phenomena throughout the gasifier and must be done very carefully.

- Product gas chemistry -- Assuming the feedstock coal is unchanged, this is primarily a function of the input steam and air flow rates (individually), which is to say the volumes of the bed that constitute the gasification and devolatilization zones and the chemical constituency of the in-situ gas. While the bed solids and zone locations will change slowly if left to their own accord, again, artificial changes can be relatively quickly effected by more or less solids throughput rates.

Long Relevant Times -- Fractions of hours to tens of hours:

- Bed solids temperatures -- The masses are so large and the gasifier heat losses are low enough that significant changes in the in-situ solids' temperatures are very slow to occur. When a gasifier is put into a standby condition, infrequent pulses of air (e.g., for 15 minutes every 4 hours) are all that is required to keep the bed in a condition wherein the gasifier is capable of a rapid restart. The need for the pulse and the pulse duration are usually determined by measuring the temperature of the outlet gas or, more correctly, the temperature of the outlet gas piping, since there is little to no gas flow. The standby condition can be sustained essentially forever with the inclusion of very minor coal additions or ash withdrawals -- which are even less frequent. An exception to the above statements is represented by the molten slag in the immediate vicinity of the slag tap hole in a slagging design. While, realistically, the temperature of a pool of slag will change very slowly, what is important is the temperature of the relatively small quantity of slag at or in the tap hole, and the fact that the slag viscosity is an exceptionally strong function of temperature. The changes important to slag flow are small enough and can happen quickly enough that slag flow phenomena are felt to be governed by short-term effects. It should be noted that a slagging gasifier in standby mode would normally be drained of molten slag, i.e., dry.
- Reaction zone locations -- The ash, combustion, gasification, and devolatilization zones exist generically, but are not precisely defined as the regions from bottom to top within a fixed-bed gasifier. Each zone is characterized by the predominant chemical reactions that are ongoing, and these reactions in turn define heat release and consumption phenomena, which then define temperature levels within the zone. The vertical extent of each zone (i.e., volume) defines the reactants' residence times in each zone and is critical to both the characteristics of the product gas and to the continued operability of the gasifier. The extent of each zone is defined by the balance between the chemical reaction rates in the various zones and the solids throughput rate. The reaction rates are determined predominately by the temperatures and velocities of the rising gas (which in turn strongly influence the in-situ solids temperatures) and the fractions of water, oxygen, carbon dioxide, carbon monoxide, and hydrogen in the rising gas. The solids throughput is defined by matched coal-addition and ash-withdrawal rates. Consequently, the extent of each zone in the gasifier is primarily dependent upon what is occurring in the zone below it, and changes can take a few hours to work their way through an operating gasifier.

- Bed solids composition -- The solids residence time within a gasifier is a function of several parameters, but operational pressure is one of the major influences. For a dry-bottom, fixed-bed gasifier at 300 psig, a typical solids residence time would be about 2 hours, and as inferred above, the time for reaction zones to change in response to a new feedstock is appreciable. As a result, if an overt change is made to the solids fed to the gasifier, it takes about 8 hours for the effects of the change to equilibrate out. Fortunately, in the real world it is rare that a major change is made in the feedstock's characteristics, and while variabilities in real coals exist, they are small enough and the gasifier is forgiving enough that they can be ignored. Following a major change, the gas characteristics will stabilize first (a medium-time phenomena), and eventually the ash characteristics will stabilize (which is what this discussion is essentially concerned with). Consequently, this characteristic is typically only a consideration during the first 24 hours following a cold startup. Another major influence on solids residence time is the bottom bed temperature (because of its influence on gasification reaction rates, i.e., coal throughput). As a result of this and because of the aggressive character of the combustion process in a slagging gasifier (which virtually destroys all organic material), it has been argued that the ash characteristics of a comparable slagging gasifier would stabilize much more quickly (under 1 hour); however, this was not verified during the study.

Intermittent Phenomena:

- Lockhopper effects -- These are virtually only noticeable as a result of operation of the lockhoppers on the coal feeding side of the gasifier (as opposed to the ash withdrawal function) and occur after the lockhopper has been recharged and repressurized and pneumatically reconnected to the gasifier. The specific effects depend on the design of the coal feeding system, but they result from the possible introduction of cool gas (with a different constituency) into the product gas at the top of the gasifier and the introduction of cool coal (and perhaps an increment of fines) into the top region of the gasifier. The magnitude of the effect depends on the amount by which the repressurized lockhopper exceeds the gasifier pressure and by the degree to which fresh coal resides in a region of the gasifier accessible to the product gas as it exits the gasifier. The gas effect is typically mitigated by using a cleaned product-gas stream for lockhopper pressurization and a control system that pressurizes to only slightly above gasifier pressure. It should be noted that the cleaned product-gas stream in a simplified IGCC configuration does not meet the inlet requirements of state-of-the-art gas compressors; as a result, a tradeoff will need to be made between the cost of generating enough cool and perhaps more stringently cleaned product gas for lockhopper pressurization and the performance penalty for use of clean inert gas. (This was not evaluated during the study, but it is guessed that generating pressurized inert gas will be preferred.) The coal effect is mostly dependent upon how much of the fresh coal enters the physical volume of the gasifier. Generally, lockhoppers are sized such that recharging takes place on 15- to 30-minute frequencies and the effect equilibrates in a minute or two, so it is not a major event -- but the control system must not over-react

to it. Two possible extremes for the susceptibility of gasifier designs to these phenomena are represented by the Woodall-Duckham and METC units depicted in Figures 8 and 11. In the Woodall-Duckham design, discrete and relatively large batches of fresh coal are intermittently dropped onto the top of the coal bed adjacent to the top-gas outlet. In the METC design, the coal does not reside within the gasifier at the end of the lockhopper cycle, and the dual feeding system provides a nearly continuous stream of coal to the gasifier; as a result, discernable lockhopper effects will be almost zero.

- Increased fines content in the product gas -- This is primarily a function of the feed-stock coal, the abuse rendered on the coal by the feeding system, and the design specifics of the lockhoppering process. However, these should be essentially constant parameters for a given system or design. The next most influential effect should be the combination of the reactant-gas feed rates and the stirrer -- whenever it gets too close to the surface of the coal bed. As the stirring arms approach the top of the bed, they can cause an increased level of fines to be entrained in the product gas, both by stirring additional fines up to the surface and by generating new fines at the surface. The stirrer can approach the bed surface either because the stroke of the intended vertical traverse brings a stirrer arm to the surface (about once every 15 minutes) or because the bed surface is sinking to the plane being stirred by the uppermost arm (as the result of too rapid an ash withdrawal rate). The velocity of the gas leaving the surface of the bed is what entrains the fines, and this is primarily only a function of the air- and steam-feed rates. A change in either of these rates produces both an immediate effect (gross gas flow) and a long-term effect (the volumes of gas resulting from gasification and devolatilization). This was considered an intermittent phenomena since the stirrer motions are slow enough that a stirrer arm could only be near the bed surface (or the gas exit port) intermittently, and since the coal bed-level control should be able to restore a proper bed level that is too low within several minutes.
- Slag removal -- For intermittently tapped, slagging gasifiers, slag removal could be likened to abruptly moving the combustion zone downward within the gasifier by a distance corresponding to the volume of slag withdrawn. In reality, it seems unlikely that much physical movement of the upper portions of the bed occurs because of the bed's reputed semi-crustal nature and because of the large measure of support that is reputed to come from the reactant gas flows. However, there is likely to be a short-term change in gas flows and a reduction in the heat content of the bottom of the bed as a result of the tapping operations. Tapping is typically considered to occur several times an hour.
- Clinkering -- This is an unintended but significant phenomenon that is associated with dry-bottom gasifiers. It has no immediate effect on gas quality, but it can reduce or eliminate the gasifier's ability to discharge ash. This then allows the reaction zones to move upward in the gasifier and eventually destroys the gas heating value. As mentioned earlier, there is incentive in terms of gas output to operate with the bottom of

the bed as hot as possible, which places the combustion zone in a near-clinkering condition. The degree to which clinkers can be dealt with depends upon the grate design and the control system logic. The degree to which clinkers can be prevented (with a maximum gas output) depends upon the ability to (1) sense a temperature representative of the bottom of the bed, and (2) have control system logic designed to operate on it.

In summary, a successful control system must respond to stimuli to control short-term phenomena and to monitor additional stimuli so as to manage the medium-term phenomena but in the context of maintaining acceptable long-term conditions. The control system's responses must be conditioned based on a blend of medium- and long-term trends. This is a nearly ideal application of a control methodology based upon the "neural network" subset of the artificial intelligence field, but of course, this will require considerable successful development.

4.9.2 Sensed Parameters

The necessary functions to be performed by the control system were described in Chapter 2 (Section 2.2.3). The two major problem areas with developing an advanced control system are the determination of the methodology and logic to be embodied in the control system (the principal relationships and phenomena relating to this area were discussed immediately above) and the development of reliable and reasonably accurate sensations on which to base the control system's responses. There are five types of measurements that are critical to operation of a fixed-bed gasifier. In order of criticality these are (1) temperature levels in the lower bed regions, (2) product-gas temperature, (3) input flows and temperatures of the reactant gases, (4) physical location of the coal-bed surface, and (5) product-gas heating value. These parameters do not need to be measured directly (and several are nearly impossible to measure directly), but some sensation must be observable that is at least analogous to the needed parameter. In addition, there are four other types of measurements that are not needed for control but are very useful to have, particularly for system technical- and environmental-performance evaluation purposes: coal properties and input feed rate, product-gas output flow rate, ash output rate, and fines output rate (and size distribution, if possible) from the hot cyclone.

Observing temperature levels in the lower portions of the gasifier provides what is likely the most feasible means to locate the position and extent (and intensity) of the combustion zone. As inferred many times earlier, the characteristics of the combustion zone are extremely important to the current and continued operations of the gasifier. These measurements are complicated considerably by the combination of a massive moving solids bed, the possible wall effects (general cooling and local anomalies caused by gas and solids contacting), the possibility of an intermittent physical presence of the stirrer, and the generally high-temperature levels (within a pressurized environment). What has been found to work acceptably well is to measure the temperature of a "body" that is contacting an appreciable fraction of the bed solids in a single plane, is relatively isothermal (massive and metallic), and is not

in the combustion zone itself. Temperature gradients in the solids bed are fairly steep as the combustion zone is approached from either above or below, and consequently the approach of the combustion zone to the "body" (i.e., the location at which temperature is being measured) is readily sensed. However, without care, this approach provides only half of the information needed because when the combustion zone is not close, there is virtually no way to tell how far away it is with a single measurement location. The desired technique is to have an instrumented body located where the expected measurement will be up the thermal gradient slope by a discernable amount. This is depicted qualitatively in Figure 23.

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Figure 23. Thermal Profile in a Dry-Bottom Fixed-Bed Gasifier

This approach, particularly with a single sensing location, becomes a bit more complicated as the combustion zone moves away from the nominal location, changes in thickness because of other than a normal full-output operating mode, or both. However, these conditions need to be accommodable by the control system methodology. Another complication can result from a radially asymmetric combustion zone. (Unfortunately, the asymmetry will usually also be changing with time.) This can occur as a result of (1) nonuniform gas flow through the upper portions of the bed (i.e., channeling), which is eventually reflected as a nonuniform combustion zone; or (2) severe maldistribution of the reactant gases by the grate. The best prevention is to use an adequate bed stirring protocol and a reasonable grate design (and speed) coupled with an adequate ash zone depth so the reactant gases are dispersed across the diameter.

Of likely more direct significance is the need to design the sensing location to be able to withstand temperatures several hundreds of degrees above the nominal level as these conditions can and will logically occur during the gasifier's lifetime. For management of the gasification process, only an above-the-combustion-zone measurement is really necessary; however, to control grate speed, some representation of a below-the-combustion-zone temperature is highly desirable. This sensation has been attempted several ways (even to using the visual appearance of the ash), and none have proven really successful in the context of a commercial-scale application. The problem is theoretically simplified for a slagging gasifier since the liquid slag is, by definition, at the bottom of the gasifier. However, the more promising approach to slagging gasifier technology entails intermittent slag tapping, and this requires currently undefined slag measurements to control the tapping process. Fortunately, this area is a developed part of the BGL technology, so a potential source of this information may exist for purchasers of BGL technology. Possible techniques for making these temperature measurements in the lower portions of the gasifier include instrumentation on the stirrer, on a skirt in the solids bed (e.g., the underside of a chevron injector skirt, or an internal skirt as used in a two-stage gasifier), on a "mast" supported from the grate (or the stirrer), in retractable probes through the sidewall, or by gas samples extracted through the wall (e.g., using a suction pyrometer).

The use of in-situ gas chemistry to locate the combustion zone has been tried and is based on the reduction of oxygen and the production of carbon dioxide as reactant gases pass upward through the gasifier. In theory, this should be a very reliable technique; however, the difficulty of obtaining a representative gas sample and of maintaining the integrity of the sampling subsystem were found to be insurmountable on a practical basis.

The product-gas temperature is actually not needed for control of the gasifier -- given that the top portions of the combustion zone are reliably located. However, it is a potential substitute control parameter during abnormal operational periods, and it becomes very valuable if gasifier operations become truly abnormal with the combustion zone too high in the bed. In addition, it is likely this parameter will be needed for the control of downstream portions of the total system (e.g., a zinc-ferrite desulfurization subsystem). This temperature can readily be obtained from a simple gas-stream probe; however, the measurement needs to be truly reliable. To this end, it appears useful to have a second, essentially duplicate, measurement taken from a sensing body with a large surface area that is wetted by the outlet-gas stream and has more thermal inertia than the probe (e.g., a metallic plate on the outlet-gas pipe's interior wall). While the stream probe is the better operational measurement, it is inherently susceptible to error or drift resulting from the buildup of carbonaceous deposits, and the second measurement should be located so as to have a different tendency (lower is preferred) for deposit buildup relative to the probe. Operationally, both will normally provide the same temperature reading (but with less transient variation from the second measurement), and as deposits occur, both temperature readings will fall but at a slower rate from the second measurement. Observing the difference between the two readings yields clues to several phenomena, but the important one for this discussion is the amount of deposition and the need for probe cleaning. (Cleaning can be done on-line with a steam jet, which can be designed to

adequately clean both measurement locations.) The product-gas temperature measurement is perhaps most relevant during standby periods (even though essentially no gas is flowing) as a means to control the frequency and durations of the air pulses required to keep the gasifier capable of rapid restarting.

Measuring the temperatures and flow rates of the steam and air fed to the gasifier is well within the state of the art and should pose no particular problems. It must be recognized that these two flow streams are the primary control over the extent of the various reaction zones in the gasifier and have a major influence on whether the zones are moving or remaining spatially fixed. The two streams produce somewhat opposing effects; increasing air generally heats the gasifier, increasing steam cools it.

The location of the physical surface of the coal bed is the parameter that controls the coal feeding subsystem, and maintenance of a constant bed depth is important to the constancy of the product-gas characteristics. It is also a measurement that is much more complex to reliably accomplish in a high-pressure gasifier. Various approaches to this measurement have been utilized (e.g., plumb bobs, electrical capacitance/resistance probes, and nuclear gages); however, the most reliable technique utilized nuclear gages to measure the attenuation of a beam of nuclear radiation by the coal bed. Typically, this is densitometry and is based on gamma rays from an externally mounted, cobalt-60 source being sensed by an array of ionization chambers located on the far side of the vessel. However, other variations on the general technique are potentially practical. Cobalt-60 sources are commercially available, relatively low in cost, and emit gamma rays at 1.17 and 1.33 Mev (a level that is energetic enough to penetrate the appreciable mass of the gasifier and yet be discernably absorbed and scattered by the presence of coal in the beam path). This technique has been used on the METC and GE gasifiers, and it becomes more viable as the equivalent attenuation resulting from the presence or absence of coal in the beam path becomes large relative to the attenuation resulting from the gasifier's structure. (That is, densitometry should work better on larger diameter gasifiers.) A complicating factor is that the quantity of cobalt-60 required on a large gasifier is increased at least linearly with diameter. While this is not a particularly significant cost factor, it does become an operational consideration.

Large flux, cobalt-60 sources can become problematical because a significant fraction of the attenuated radiation is really scattered (in nearly all directions), and the associated radiation shielding and worker proximity restrictions become impractical or at least difficult to live with. A specific analysis would have to be done, but it is estimated that the amount of cobalt-60 required for a source-to-detector (gasifier chord) path length of 14 to 18 feet is about the limit that could be accommodated without significant OSHA restrictions (or inconvenient amounts of shielding on the gasifier exterior). It appears the commercial size gasifier can utilize this technique without any real operational problems. However, an alternative approach based on stirrer torque requirements was investigated during the course of this study and is presented in the next Section.

Nuclear radiation-based techniques, while more operationally complex than most others, also offer additional possibilities. Foremost among them is the possibility of locating radiation flux detectors lower along the gasifier (which could also be irradiated by the same sources that irradiate the bed-level detectors) to sense void formation (a possible parameter to control depth of stirrer travel). Similarly, it may be feasible to detect the location of the ash zone (because of an absence of carbon); however, this would probably require use of a neutron source (rather than a gamma ray source), and this also has operational implications. Nuclear radiation-based approaches for detection of reaction regimes higher in the gasifier have been postulated but are not felt to be workable. Void detection has been tested with limited success in the METC gasifier (and no other experience with this approach is currently known). No applications other than level detection have been reliable.

The heating value of the raw product gas is not really needed for gasifier control, assuming the measurements discussed above are available. However, it is possible to operate a fixed-bed gasifier with only knowledge of the product-gas heating value, product-gas temperature, and bed surface location plus a visual impression of the ash's appearance -- all of which are readily obtained. In a sense, product-gas heating value is a redundant control parameter; however, changes in the heating value often provide the first indications that something is changing inside the gasifier. In addition, this parameter is necessary if a performance analysis of the gasifier itself is to be performed, and this is often desired. There are basically two ways to obtain the heating value: quantify the major gas constituents and calculate the heating value, or measure it directly. Direct measurement is usually a calorimetric technique that has the advantage of being essentially a continuously available, real-time value, and it can measure nearly the whole gas stream's heating value (which may or may not be an advantage). The heating value attributable to nearly everything in the product stream is obtainable because the calorimeter needs only enough gas sample conditioning to allow the burner to remain operational for reasonable time periods. The conditioning is potentially only particulate removal. The disadvantages are that calorimeters can be very troublesome to calibrate and maintain, and there are no insights as to what characteristics of the gas stream are influential in producing an observed heating value.

Quantifying the gas stream constituents is generally done with on-line mass spectrometers, gas chromatographs, or both. Then a computer algorithm is utilized to calculate the heating value from the species concentrations. The advantages to this approach are that it provides precise insights into the gas characteristics of the product stream, and the equipment is state of the art. The disadvantages are that the water, condensable hydrocarbons, and dust must all be removed prior to gas quantitation (and thus, much less than the whole stream is characterized); the equipment is expensive; and the numerical result is a discontinuous value whose frequency of availability is dependent upon the cycle times of the instrumentation and the computer programs. (Once every 12 minutes would not be atypical for a chromatography-based system.) The difficulty in correcting back to a whole stream heating value is significant, but fortunately, this is not really necessary for control purposes. However, it does become necessary when evaluating gasifier performance.

4.9.3 Use of Stirrer Torque for Determining Bed Depth

The two problems with using nuclear gages for bed level determination are that they do not really provide a precise determination of bed level, and the associated administrative and operational overhead requirements are non-trivial. While the administrative and operational factors can be routinely dealt with and the real risks attendant with nuclear gages are minuscule, these do need to be understood when basing a design on the use of nuclear radiation. It should also be remembered that the skills required to survey, document, and maintain this class of equipment are unique and are likely to only be needed on an IGCC site because of the existence of this particular instrumentation. The lack of a sharp discrimination of the bed surface occurs mostly because a significant fraction of the radiation is scattered and because the surface of the bed characteristically has structures in the beam path that are irregularly or intermittently absorbing (and scattering) the gamma radiation. The effect of scattering is that detectors, which in reality are shadowed by the coal bed but are close to the surface, will receive extra photons that were scattered from both the coal and the metallic portions of the gasifier; consequently, these photons register just like the photons that did not strike the coal bed at all (thus, producing a falsely shallow-bed bias to the data). The irregularity occurs both because stirrer (or bed leveling) arms swing in and out of the beam path and because the coal bed surface can have hills and valleys that are significant in relation to a desired resolution scale of the bed surface of perhaps 4 to 6 inches. The determination of the surface is also degraded, but much less significantly, by the continuous decay of the radioactive source strength and because the path lengths to each detector are not identical. (More distant detectors receive less intense direct and scattered radiation.) Natural isotopic sources are by far the most economic to use, and cobalt-60, with its slightly over 5 year half-life, decays slowly enough that recalibration of the nuclear gages on annual or biannual frequencies has been found to be perfectly adequate. While nuclear gage technology represents what is felt to be the current best approach to this measurement, an alternative option was felt to be desirable.

In the course of this study, the possibility of using the resistance to stirring (which is offered by the mass of the bed) as an analog of bed depth was investigated. This was found to be potentially feasible but to also require a control system embodying a considerable level of intelligence -- well beyond overt signal-conditioning technology.

This concept is based on two effects that in combination (or also individually) can cause the resistance to stirring experienced by the stirrer to be correlatable with coal-bed depth. First, with a vertically traversing, multiple arm stirrer, there are periods when the top-most arm normally clears the top of the bed (and provides a bed leveling function), and the resistance resulting from, for example, two stirring arms being within the coal bed ought to be noticeably lower than the resistance with three arms in the bed. Second, as a submerged arm moves through the bed of solids, it forces solids out of its path, and the ability of the solids to physically move (at least upward) ought to be a function of the mass of bed overhead (i.e., the depth of the stirring arm in the solids bed). Fortunately, both of these phenomena are complementary (more stirring resistance corresponds to greater depth) and both could be

utilized with either vertically stationary or traversing stirrers. (Admittedly, the ability to hold a constant bed-surface location is likely to be very poor using only the first phenomena with a vertically stationary stirrer.) Conceptually, only a stirring resistance measurement and knowledge of the stirrer's physical location would be required to infer the location of the coal-bed surface.

While this seems quite credible, it must be remembered the solids bed is a very heterogeneous mass (in any one plane, let alone vertically), and the solids in the various reaction zones (which can also move spatially) would be expected to have differing flow characteristics because of their differing physical characteristics. Added to this is the expectation that forming agglomerates would be intermittently encountered (which is why a stirrer is utilized in the first place), and these would pose a temporarily increased resistance (and a false deep-bed indication). While the concept seemed viable, the amount of noise expected in the measurements is expected to be formidable.

As mentioned earlier, the stirrer shaft in the METC gasifier was instrumented with strain gages on a portion of the length that was always within the gasifier but was above the stirrer arms. (Figure 11 on page 53 shows the general configuration.) In this location, the strain gages measured only the torque corresponding to the stirring resistance imposed on the three arms by the solids bed (the resistance offered by the shaft support bearing and the pressure seal were reflected only on the shaft length outside of the gasifier). These data were taken in 1983 for other purposes and were of tertiary priority in the test series at the time. As a result, only a limited amount of data was found to be reducible for this evaluation; specifically, data from several hours of a stable period of Run No. 101, while feeding a Pittsburgh seam coal (Arkwright) and operating with two vertically traversing stirrer protocols, using a three-arm stirrer configuration. A plot of the stirrer vertical position during the hour that was analyzed in detail is provided in Figure 24. The two protocols differed only in vertical stroke lengths (approximately 3 and 4 feet) while maintaining a constant 1-rpm rotational rate. Note that the stirrer cycles were about 12 to 14 minutes in duration with the up-stroke being twice the down-stroke speed; consequently, the stirred helical path was quite different during the two portions of a given cycle. The stirrer height above the grate shown on the ordinate provides the location of the lower surface of the lowest stirrer arm as a function of time through the period.

A plot based on data from the five detectors used in the nuclear gage subsystem is provided in Figure 25. Figure 25a shows the distribution of detector data at the moment when the stirrer was at its maximum vertical position for the five sequential cycles analyzed, and Figure 25b shows the same type data when the stirrer was at its deepest location (for four cycles). It should be noted that high voltage readings from the data system correspond to low irradiation levels at the detector. From these plots it is quite noticeable that (1) the coal-bed surface is not blatantly obvious, (2) the surface appears to have been sinking slightly prior to 4:00 a.m., and (3) the surface was then stabilized until 4:30 a.m. when the plot was stopped. It is fairly obvious the detectors at the 47- and 56-inch locations must have been in the coal bed since the detector outputs were quite high and similar. The 86-inch detector is located so

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Figure 24. Vertical-Traverse History of the METC Gasifier Stirrer

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Figure 25. Typical Output From the Nuclear Gage Bed-Level Detection Subsystem

high that it is, by definition, out of the coal bed (barring major problems), and so intuition and the use of slope intersections would allow a manual prediction of a bed surface location between about 59 and 68 inches above the grate. It is believed the feeding system set point was 65 inches for this period (however, it may not have stabilized prior to the hour analyzed), and it appears the control system algorithm, which is based on nuclear gage calibration data, calculated a rational bed surface location and held it with apparently minor variations, especially after 4:00 a.m. The variation shown in Figure 25a when the stirrer is at its topmost position could also have been a result of the top arm being in the beam path by a varying amount. (The top arm's rotational location -- in or out of the beam path -- is not known.) However, Figure 25b shows similar data when the stirrer is at its deepest position and the top arm is well submerged in the coal bed. One conclusion from this is that the relatively massive top stirring arm's intermittent presence did not significantly degrade the nuclear gage subsystem's performance. Figure 25c provides the same type data for three successive data scans 30 seconds apart just past the first bottom of a cycle. Since the stirrer was well out of the beam path, this should have been a period of excellent repeatability (even though the bed's top surface appears to have been sinking slightly) -- and it was.

Using the stirrer's vertical position with time, the (nuclear-gage-controlled) bed surface position, and the corresponding torque data, it was possible to generate plots of stirrer torque as a function of the bottom arm's height above the grate. These data are shown in Figure 26 for successive up strokes and in Figure 27 for successive down strokes. Note that the torque measurements were more orderly on up strokes than down; whether this was because the stirrer was moving up or because of the higher vertical speed is not known. Significantly, these data are amazingly linear with a slope of about 50 ft-lb/in of bed depth, which is likely large enough to be discernable. Unfortunately, during the period when the data were taken, the stirrer was not raised to a point where only one arm was in the bed, but the regions on the plots where two or three arms must have been within the bed can be inferred from the stirrer's profile dimension, which is indicated on the figures. With a little imagination, it is possible to discern a steeper slope (lower torque requirement per unit depth) in the two-arm region, which is comfortingly logical; this is most noticeable on the Figure 26 up-stroke plots. The two- to three-arm transition period during which the topmost arm is presumed just sweeping at the bed surface is also the period that produced the most disorderly data, which is also logical. There is a possibility that routine transitioning of the uppermost arm in and out of the bed will be found inadvisable because it results in an increased release of fines into the product gas. (This is also of concern with respect to the usage of a bed-leveler arm.) If this is found to be the case, the general torque/depth slope appears to be of a significant enough magnitude to potentially be the sole parameter utilized for the bed depth computation. It should be remembered the arms were made from 4-inch pipe, and trying to discern bed depths to resolutions much smaller than 4 inches is likely to be futile.

Figures 26 and 27 imply that it should be possible to calculate bed depth based on stirrer torque measurements, and the resolution could be comparable to that obtainable from nuclear gages. Whether the resolution will be more or less than is obtainable through nuclear gaging remains to be seen. As with nuclear gaging, the viability of this approach is likely to improve with increased gasifier diameter because of the increased torque requirements. The torque data show a large degree of variability, which gives considerable pause for thought

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Figure 26. Stirrer Depth Versus Torque, Up Strokes

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Figure 27. Stirrer Depth Versus Torque, Down Strokes

when it is remembered that these data were taken from a steady-state operating period (except for the stroke length change -- which, interestingly, seemed to not effect the gasifier's operation). The use of a hydraulic stirrer drive could make this torque measurement trivial; it would be a direct analog of the hydraulic pressure, given a small and constant resistance attributable to the stirrer shaft-support bearings and pressure seal. What seems a certain requirement for this approach to be viable is a very smart control system. Elements of curve fitting, data trend/scatter analyses, and correlations with reaction zone movement will probably all need to be incorporated in any workable control methodology.

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Chapter 5

Prototype Simplified IGCC System

The objective of the CGIA study was to conceptually design and evaluate the potential of an advanced type of fixed-bed gasifier that would be attractive in commercial IGCC applications. However, this evaluation needed to be done in a system context to make sense, and to this end, a prototype system was configured to serve as a reference.

The basic requirements for the system were covered in the Ground Rules in Chapter 2. In essence, the system should have excellent efficiency, cost, and environmental performance characteristics. The system should be capable of being located essentially anywhere in the U.S., and any modifications required to accommodate local coals or site characteristics should be accomplishable by comparatively minor factory-installed options applied to the basic gasifier/system design. The philosophies of maximized factory fabrication with field assembly (versus field fabrication) and modularity based on an approximate 100- to 150-MWe unit size were also applied to the prototype system. The technology advancement level was allowed to stretch somewhat to include technology that could reasonably be expected to be proven feasible before 1995, with credible first units for major components and subsystems being operational before 2000.

The general configuration of the system and the major options that were considered in the CGIA study are depicted in Figure 28. The more intuitive material-flow streams are represented by solid lines connecting the various major operational units, and the credible options are indicated by dashed lines. This configuration meets the CGIA Ground Rules and incorporates some novel features that potentially make it very attractive; however, it was not optimized, and comparative analyses against other configurations were not done. The purpose of this system configuration was to provide a reference framework (with reasonably attractive technical and cost performance characteristics) that would allow comparative, system-level evaluations of fixed-bed gasifiers having significantly different design features, such as one- and two-stage gasifiers, with and without top-gas recycle, and slagging- and dry-ash-bottom designs.

As shown in Figure 28, the overall system utilizes ROM coal, but alternative approaches are considered for any fraction of the ROM coal that cannot be fed to the gasifier. However, the intent is to process the maximum coal fraction possible through the gasifier, based on the expectation that this will increase overall system efficiency and decrease both the emissions and the COE. Any significant gas stream loading of hydrocarbon condensibles (i.e., tars and oils) is to be eliminated prior to the fuel gas being fed to the GT. This is accomplished either by incorporating a cracking catalyst in the hot-gas desulfurization sorbent or by incorporating a relatively small but novel pressurized combustor (i.e., a PFBC) that uses a moderately fluidized bed of limestone and is fueled by particulate removed from the product gas or any other environmentally difficult stream (e.g., the top-gas stream from a two-stage

gasifier). The use of a hydrocarbon-cracking catalyst and the PFBC are small extensions of existing technology and were considered acceptable. The cracking catalyst is state of the art in the petrochemical industry, and the experience should be directly applicable to the CGIA system. Hot gas desulfurization is accomplished with a zinc ferrite subsystem using fluidized beds for both the sorption and regeneration steps. Regeneration offgas from the zinc ferrite subsystem is processed by either the limestone PFBC mentioned above, or a direct sulfur-recovery process (DSRP); in the former, the coal-borne sulfur exits the system as gypsum, and in the latter, it exits as elemental sulfur. A significant fraction of the GT compressor discharge air is utilized externally to the GT as blast air for the gasifier and also as the oxidant for the PFBC and zinc ferrite regeneration. The PFBC is operated with a minimum of 20% excess air to ensure the formation of both gypsum and nitrogen gas from the sulfur- and nitrogen-bearing species in the fuel streams supplied to the PFBC. The PFBC flue gas returns to the GT as highly preheated but somewhat oxygen deficient combustion air. Each of these major functional units, or system features, are discussed in the following sections.

5.1 PROTOTYPE GASIFIER

The CGIA study examined a number of gasifiers and gasifier design features, the more significant of which were described in the foregoing chapters. As mentioned in Chapter 4, the ability to analyze a fixed-bed gasifier in significant detail is currently limited, and the technical performance of one mode of fixed-bed gasification over another was not found to be clearly superior with the possible exception of recycling two-stage gasifiers, which were found to be inferior (based on the predictive tools at hand). However, the study evolved a preferred concept that could be representative of first units but, most importantly, contained the features necessary to provide the insights needed to select and support mature gasifier design configurations.

5.1.1 Pseudo-Drafted, Two-Stage Prototype Gasifier Concept

This concept is depicted in Figure 29 and is a pseudo-drafted gasifier incorporating an internal skirt. The design is oriented toward operation as a two-stage, fixed-bed gasifier; however, the major design features would be suitable to operation as either a one- or two-stage, fixed-bed gasifier and with either a dry bottom (as shown) or a slagging bottom.

Figure 29 shows a conceptual design; the expected virtues remain to be confirmed by testing. The general characteristics of two-stage, fixed-bed gasification were discussed in Section 4.6 of Chapter 4. The design features of this gasifier configuration, together with its incorporation into the system, are expected to result in very low emissions and very high performance. For example, the coal-feeding design approach and hot particulate-removal step should allow a higher gasifier air-blast rate (i.e., a correspondingly higher coal throughput,

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Figure 29. Pseudo-Drafted, Two-Stage Gasifier Concept

given a modest length increase) than a conventional single-stage, fixed-bed gasifier. It is estimated that specific coal throughputs exceeding 500 lb of coal/hr ft² should be achievable at 300 psig and 1000 lb of coal/hr ft² may be possible. On a vessel with a 12-ft internal diameter, this implies a system output capability of about 70 MWe to perhaps as much as 140 MWe per gasifier.

Many of the basic characteristics of this concept relate to the internal skirt. The skirt length selects a horizontal plane at which the coal bed diameter is allowed to increase and from which a mid-level gas may be extracted to provide the lower stage, or side-gas, product stream. The intent is that the narrower upper stage would retain the coal during the majority of its devolatilization and the period during which it would need to be stirred, and the smaller diameter would also increase the gas velocity through this portion of the reactor. The increase in gas velocity is important since the upper stage will have much less gas flow through it than would occur with a comparable throughput, single-stage gasifier, and hence, will have a lower coal heating rate. Diametral reduction to the point a normal gas velocity is restored in the upper portion of the gasifier is not practical, and the combination of less effective coal heating and increased lineal velocity of the coal will result in the upper stage being longer than a comparable single-stage unit. It is possible that a fraction of the coal swelling tendency will exist at the temperature corresponding to the point where the coal bed expands to fill the volume below the skirt. If so, this diametral expansion and the associated bed disruption will enhance the ability to accommodate strongly caking coals. The individual characteristics of various U.S. coals could be accommodated by selecting a particular skirt length and the operational parameters suited to that coal or family of coals. Figure 29 also shows a means to fine tune the skirt length during plant shakedown using jack screws. While this is a valuable feature for a first unit, it is unlikely to be needed on mature gasifier designs. (As shown, the jack screws would be contained under a pressure cap - necessitating depressurization of the gasifier to make an adjustment; however, other jack screw designs and other skirt length adjustment methods are certainly possible.) The intent is to not actively cool the skirt so that it can be instrumented to provide at least the upper portion of the longitudinal thermal profile that is so useful in controlling the gasifier's operation (and also to preclude reductions of gasifier efficiency). This may not be practical (and is at odds with existing experience), but in any event, the amount of cooling used will be minimized so that a temperature can be measured that is indicative of bed conditions at the lowest edge of the skirt.

The conceptual design of a new coal-feeding subsystem, described in Chapter 4, Section 4.1.1, was selected for the prototype gasifier. This design approach together with the use of a hot particulate-removal step are expected to allow a higher gasifier air-blast rate, and it is estimated that specific coal throughputs in excess of 500 lb of coal/hr ft² should be achievable, and 1000 lb of coal/hr ft² may be possible. Gamma ray densitometry will be used to determine the top-bed surface level and thus is the basis for control of the coal feeding subsystem. The grate design will utilize the METC crusher grate design described in Section 4.3 of Chapter 4, but with ash plows on both the bottom and middle plates of the grate as shown in Figure 29 and with a hydraulic drive capable of operating in the reverse direction for at least 30 degrees of grate rotation. The gasifier wall will utilize a steam generating cooling

jacket and will be designed to produce saturated steam at approximately the maximum gasifier operating pressure as a means to minimize cool wall effects.

The stirrer design will utilize a hydraulic drive and hardened surfaces on the stirring arms and on the section of the shaft engaging the pressure seals, and it will be based on the considerations discussed in Section 4.2 of Chapter 4. For first units, the stirrer will be designed for a limited amount of vertical travel (to no lower than the bottom edge of the skirt); this is expected to be a vertical stroke length of about 30 inches. The stirrer will utilize only one pair of arms located as shown in Figure 29. (A bed-leveling arm will not be utilized.) It is expected that the centerlines of the two arms will be vertically offset by about three arm-diameters from each other. It is probable that information gained from the first gasifiers will allow elimination of the stirrer's vertical traversing function in mature gasifier designs through optimized arm spacing, optimized stirring protocols, the addition of more stirring arms, or a combination of these. Even though the stirrer is not required to penetrate into the really high-temperature portions of the bed, the mechanical loads on arms stirring a bed potentially as much as 12 feet in diameter are formidable, and it is anticipated that aggressive water cooling will be required for long-term mechanical integrity.

A conceptual design for a stirring arm that should be adequate for this application is shown in Figure 30. This design is based on scaling-up the torque measurements taken from the METC 42-inch diameter gasifier operating on strongly caking coals to values that should be representative of the same conditions in a 12-foot diameter gasifier. The configuration is derived from one used successfully for years in the METC gasifier, and while not as elegant as configurations utilizing shaped blades (which have also been tested), it appears to be as functionally effective, much easier to fabricate, and is very effectively cooled. The shape for each arm is fabricated from two, 6-inch diameter, extra strong, carbon-steel pipe sections that are positioned side by side and welded together at their line of tangency. The cooling water supply flow is brought in vertically downward through a pipe located at the center of the stirrer shaft, and then is routed radially outward through the pipe, forming the leading edge of the stirrer arm. A cap at the outboard ends of the two pipes causes the cooling water to be returned to the stirrer drive-shaft through the pipe forming the trailing edge of the arm. The cooling water then returns up the annulus inside the stirrer drive-shaft (between the inside diameter of the stirrer shaft and the outside diameter of the cooling-water supply pipe). To make the top and bottom surfaces of the arm essentially flat and to increase the structural stiffness of the arm (without the cost and thermal stresses that would result from massive fillet welds), top and bottom valley-cover plates are welded over the top and bottom interstices between the two parallel pipes. Figure 30 also shows that a sizeable slot is cut in both of the pipes to make the void space behind the top and bottom valley covers become part of the trailing-edge and leading-edge water conduits. After all fabrication is completed, a hardened exterior surface is applied. As an example, a weld overlay of titanium carbide dispersed in a matrix of chrome iron (65% titanium carbide and 35% chrome iron by volume, resulting from Yeardley Engineering No. 1690), laid down to a nominally one-eighth inch in thickness and followed by a "pack cementation hardening process" has been used successfully at METC and appears to provide an operating lifetime of several years. This process diffuses boron

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Figure 30. Conceptual Design of the Stirrer Arm

carbide into the surface to convert the exterior of the weld overlay to a shell consisting of a mixture of titanium diboride and chromium boride with a nominal thickness of one-hundredth inch. The Knoop Hardness Number of titanium diboride exceeds 4000 and that of chromium boride is slightly over 2500, and this surface has proven very abrasion and corrosion resistant in the METC gasifier. It is possible that a less elaborate hardcoating process and scheduled stirrer replacement may be the preferred approach for a commercially mature gasifier, but this is not presently known.

As shown, the Figure 29 design would normally be thought of as a two-stage gasifier; however, the configuration is likely to also make sense for a one-stage design (without use of the side-gas port). In this case, the major functions of the skirt are to provide a means for deep-bed temperature measurement, and possibly to adjust the solids' velocities as they move down through the gasifier. It is possible the diametral expansion (and consequent bed disruption) that occurs at the bottom of the skirt will assist in accommodating caking coal, but this is not expected to be a highly significant feature. It is quite possible that design optimization studies will result in minimizing the gap between the skirt exterior and gasifier wall interior surfaces, in which case the preferred design will revert to a more conventional, non-skirted design. For a one-stage, non-skirted design, the principal attributes of Figure 29 are the coal-feeding system, stirrer, and grate designs (and instrumentation). For this configuration, obtaining deep-bed temperatures that would be representative enough to be useable for gasifier control will be difficult (without either stirrer- or grate-mounted sensors). In this case, a refractory-lined gasifier wall (or wall segment) should be considered as a means to obtain the information; however, it is recognized that this may entail development or operability risks.

The negative characteristics of the design shown in Figure 29, other than it has yet to be tested, are (1) the characteristics of the side gas may be influenced by wall effects and thus not be representative of the average gas properties at the plane formed by the bottom edge of the skirt, and (2) the vessel's volume is larger than that of a single-stage gasifier of comparable throughput. The use of a chevron injector unit, described in Section 4.1.2 of Chapter 4, for side gas withdrawal from the bed would eliminate the nonrepresentativeness of the side gas. However, this is not anticipated to be a major problem, particularly with a steam jacketed gasifier, and the skirt is so simple that it seemed like a good conceptual approach. It should also be remembered that there is a limitation on this design in that stirring below the plane defined by the lower edge of the skirt is not recommended. The principal reason is that this would be likely to greatly increase the amount of coal fines entrained by the side gas stream; however, the dimensional-control requirements also become more stringent to ensure the stirrer will be within the skirt's enclosure when it is elevated to the upper bed regions. A corollary to this limitation is that location of the skirt's lower edge above any of the stirring arms appears impractical, even for a stirrer that operates without vertical traversing. A stirrer arm that is long enough to approach the gasifier wall (below the skirt) will necessitate a significantly more complex and restrictive final fabrication, installation, or both. In addition, the likelihood is that there will be a minimum stirrer-arm-to-skirt-edge height dimension (or unstirred zone) that must be maintained if greatly increased quantities of fines in the side gas are to be precluded. Consequently, if stirring is required below

the skirt at all, it seems incredible that the gasifier would operate reliably with an unstirred zone between two stirred zones (since stirring will be required in the upper portions of the bed).

Table 9 provides a summary of the major pro and con characteristics of this concept, including facets of the discussions in Chapter 4.

Table 9. Pseudo-Drafted, Two-Stage Gasifier

PRO		CON	
1.	Environmentally significant species and tar are concentrated in the top gas.	1.	It is likely the gasifier can only be stirred within the skirt envelope (which may not be enough).
2.	Various coals are accommodated using various skirt lengths, grate and feeder speeds, and stirrer protocol.	2.	The gasifier is longer than a comparable one-stage unit.
3.	An advanced coal-feeding system should accommodate any coal, reduce fines loading in the top gas, and preclude the need for a bed-leveling stirrer arm.		
4.	The lowered gas velocity in the pyrolysis zone (relative to a one-stage unit) should slightly improve the ability to handle caking coals. However, this velocity is slightly elevated (relative to a constant diameter unit), which reduces the required increase in gasifier length.		
5.	The grate has some ash crushing capability and will tolerate sizeable agglomerates.		
6.	The skirt is a credible in-bed temperature-sensor location.		

5.1.2 Non-Selected Gasifier Concepts

There were three other gasifier concepts plus an adaptation of a feature from the recycle concept that had intriguing characteristics but were found to be either slightly attractive (the first concept discussed) or judged to be too impractical to pursue based upon currently existing knowledge bases and analysis tools (the latter concepts). As a group, they all have increased potential for the direct use of coal fines, and it is possible future analyses will identify virtues not found during this study. These concepts are described and documented herein to preserve them for possible future reconsideration.

Internally Skirted, Recycling, Gasifier Concept: This concept is shown in Figure 31; it features two fixed internal skirts and is intended for operation in a top-gas recycling mode. This concept came close to being selected as an alternative preferred configuration, but there were performance disincentives that precluded it. As mentioned previously, existing predictive tools and data also limited the depth of analysis that could be done on this concept, and a re-evaluation is probably warranted when better computer models or additional test data become available. This concept has much in common with the concept shown in Figure 29, and consequently, this discussion will focus on the differences without repeating the similarities.

The general characteristics of a fixed-bed gasifier that operates in the top-gas recycle mode were discussed in Section 4.7 of Chapter 4. Figure 31 shows that an additional port exists low in the side wall of the gasifier, and it is utilized for reinjection of the product gas emanating from the top of the gasifier. This type gasifier generally has only one product-gas port, and as shown in the figure, it is located just below the drying and pyrolysis zones at the top of the coal bed. The design of Figure 31 also utilizes internal skirts; however, in this concept the skirts are permanently fixed to the wall, and their primary function is to provide an ullage volume as a means to either collect or distribute gas in a relatively uniform manner from or to a plane within the gasifier.

Figure 31 implies that the top gas exits the dome of the gasifier through a pair of gas ports and is immediately propelled by steam eductors down a pair of well-insulated pipes to the gasifier ports corresponding to the lower skirt. The recycled material (mostly gas and condensible hydrocarbon vapors with some solids and a moderate amount of steam) is propelled into the ullage volume behind the lower skirt and wafts its way into the bed volume. At this horizontal plane, the bed's gasification reactions are nearly completed and the combustion reactions are just beginning. It is expected that most of the solids and some of the condensible hydrocarbon vapors adhere to or react essentially immediately with the indigenous hot solids and travel downward with the bed to add fuel to the combustion zone. Similarly, some of the more flammable gas species (e.g., hydrogen) are likely to combust at or near the reinjection plane. The majority of the gases and significant fractions of the other recycle stream constituents will exit the ullage volume and be carried up through the gasification zone with the bulk-gas flow. In the gasification region, a majority of the condensibles and a significant portion of the ammonia should chemically crack on the surfaces of the

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Figure 31. Internally-Skirted, Recycling Gasifier Concept

descending hot carbonaceous solids. At a plane near the top of the gasification zone or bottom of the pyrolysis zone, an upper skirt and its associated ullage volume are encountered. Gas collects in this volume and exits the single product-gas outlet port in the side of the gasifier as a hot fuel gas with lower hydrocarbon condensibles and ammonia loadings than in a single-stage gasifier.

The recycle-gas path design is one of the more critical aspects of a recycling gasifier. The top gas must be pressurized to propel it to its ultimate destination within the gasifier. It is laden with vapors having a wide range of condensation temperatures, some of which are virtually equal to typical gas stream temperatures. In addition, there is an appreciable fraction of fine particles that serve as condensation sites and form the basis for agglomerates with relatively hardy mechanical characteristics. Steam was selected as the driving media for the eductor both because it is effective and because it will be needed for the gasification process (and, thus, may not be a totally superfluous fluid). Steam at conditions approximating top-gas pressure and temperature is capable of precluding significant deposit formation and has also shown a non-trivial capability to remove hydrocarbon deposits from metallic surfaces. In addition, the generation of steam at conditions in the range of 500 psig and 950 to 1000 °F is well within the state of the art. However, producing steam at a temperature of 1100 °F (which is likely to approximate the requirements) will require either an auxiliary-fueled superheater or use of the PFBC for this purpose. Either approach does not represent a major cost or fuel consumption, but it must be considered.

Early in the study, an estimate of the eductor's steam requirements was made based on Dr. Zenz's correlations (n.d.) and the following assumptions:

- Gasifier air/coal mass feed ratio is 2.2,
- Gasifier steam/coal mass feed ratio is 1.0,
- Molecular weight of the top gas is 26,
- Top-gas temperature is 1000 °F,
- Top-gas pressure is 237 psia (This is 15 atm times a factor of 1.3 to allow for the pressure drop needed to get the fuel gas to the GT.), and
- The eductor operates at choked conditions using 1000 °F steam.

The correlations have approximately a 50% safety factor (relative to optimally designed eductors) and are claimed to fit high pressure data taken by both Westinghouse and Foster Wheeler. The results of the estimation are provided in Figure 32 as eductor performance curves as a function of the fraction of the gasifier's normal steam supply (i.e., required for a one-stage unit) that is supplied by the eductors. Since this steam is unavailable for moderating conditions in the grate region or the combustion zone (which are normally the major applications of the gasifier's steam demand), there is an obvious incentive to minimize eductor steam flow in order to minimize superfluous steam flow to the gasifier. While the gasifier conditions eventually selected for the systems analyses (presented in the following chapter) differ somewhat from the conditions pertinent to Figure 32, the indicated design point is representative. The design point would require 38% of the normal gasifier's steam

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Figure 32. Performance of Steam Eductor for Top-Gas Recycling Gasifier

flow to occur through the eductors and would recycle 20% of the gasifier's output mass flow with a 5 psi pressure rise (which should be more than sufficient to overcome the gasifier's internal pressure drop). The corresponding ratio of recycle-pipe-diameter to steam-nozzle-diameter is 5.2. Since these values were both conservative and within reasonable design bounds, and since further refinement would have required a specific gasifier design, it was judged safe to assume a credible design using steam eduction could be produced and to omit further analyses.

Two sets of recycle ports and piping were utilized in the concept to improve overall reliability and to make it easier to achieve a relatively uniform gas distribution at the reinjection plane. Similarly, a powered auger was located on a piping cross at each reinjection port. The reinjection port is the most critical location in the recycle pathway (and, with the top outlet port, is the most susceptible to plugging due to the influence of the cool pressure vessel wall), since the hot steam is expected to keep the vertical pipe clear to the point of the 90 degree bend in the gas flow. The use of a piping cross (instead of a tee) was chosen as a means to provide an easy access in the event through-reaming of the vertical pipe became a necessary maintenance operation (and to provide a volume for the possible accumulation of debris). Normally, the auger would be positioned back out of the gas flow, but the auger is capable of being driven forward to machine away any condensates or deposits that occur in this horizontal section, and of pushing any debris into the volume of the gasifier (behind the skirt). Since use of the auger is likely to temporarily restrict gas flow, the existence of "more than one" gas recycle pipe and asynchronous use of the augers essentially become requirements.

The use of a recycle subsystem also offers the possibility of entraining additional coal fines (e.g., those separated from the ROM coal supplied to the plant) and feeding them deep into the gasifier bed as a means to beneficially consume fines. Obviously this represents a complication to the recycle pathway design, and it is also likely to increase both the eductor steam demand and the gasifier air-blast requirement somewhat. However, it is probable this approach can be made to work. The injection of additional fines was not analyzed, and some testing would be required to determine the parameters useable for recycle-pathway design.

As with the preferred two-stage conceptual design, a principal problem with this concept is that it has never been built and tested. The primary concerns with the design relate to the expected need for more stringent operational control (to maintain reaction zone locations), the inherently low gas heating value, the design of the recycle pathways, the quantities of steam required, and the overall size of the unit. The need to keep the recycle pathways functioning is obvious, and the mechanisms that make them susceptible to blockage have mostly been discussed. One additional comment pertinent to the configuration in Figure 31 is that a build-up of coke-like deposits behind the lower skirt is possible, and there is no inherent mechanism for their removal. The recycled hydrocarbons and machined reinjection port deposits are swept into the lower skirt's ullage volume and are supposed to either be collected by the in-situ solids and proceed down into the combustion zone or to waft around the lower edge of the skirt and be swept upward with the general gas flow. The injected materials may

well impinge on the lower surface of the skirt, which is hot and quite capable of inducing chemical cracking and serving as a sticky substrate for deposit build-up. In the region immediately opposite the ports, the steam velocities may be sufficient to keep the deposits at acceptable levels, but it is unclear what will happen at locations circumferentially away from the reinjection ports. Whether this is a problem or not will have to be determined by testing.

Like the two-stage gasifier discussed in the preceding section, there is the question as to whether the skirts will distribute and recover representative flows with reasonable uniformity across the bed diameter (or whether it is necessary that they do so). Similarly, it is estimated that in-bed stirring will be required down to (but not below) the lower portions of the upper skirt in order to maintain solids flow through the skirt region. This results in a need for arms with differing radial lengths as shown in Figure 31, and will result in the possibility of a build-up on the converging slope of the upper skirt (since an elaborately shaped arm tip and excellent dimensional control of the stirrer motion would be required to preclude it). In addition, it is assumed that sufficient gasification reactions have taken place that the solids are free-flowing by the time they reach the vicinity of the lower skirt, and the recycled materials are assumed to not significantly alter this characteristic. While all of this seems acceptable and reasonable, tests with strongly caking coals will be required to determine the functionality of both internal skirts. As discussed above, it is likely the recycle concept will require more steam than a one-stage gasifier; however, it is also likely that testing and system optimization will allow the increase to be held below the 38% value cited above. Unfortunately, it is also estimated that the recycling gasifier will require a noticeably larger total volume per unit of coal throughput, and a length increase of 40 to 45% (relative to a single-stage unit) would be representative. Of likely more significance, the product gas, while predictably very hot, is quite low in heating value - less than 100 Btu/scf is possible - and this may have serious implications regarding use of the gas as a GT fuel. The heating value is lowered as a result of (1) dilution by superfluous eductor steam, (2) a requirement for additional air to produce the heat needed to raise the total mass of the recycle stream to the indigenous temperature, and (3) additional air required to supply heat to offset the endothermic cracking reactions (this should be a comparatively minor effect). The additional air allows essentially normal zonal temperature levels to be obtained, but it significantly adds diluent nitrogen and CO₂ to the product gas stream. Again, the reality of all this will have to be explored by more in-depth analyses and testing.

Table 10 provides a summary of the major pro and con characteristics of this concept, including facets of the discussions in Chapter 4. It is possible this concept may be better suited to a slagging-bottom gasifier configuration.

Other Non-Selected Concepts: None of the fixed-bed concepts discussed to this point really utilize coal fines well (without agglomeration); the possible exceptions to this are the slagging-bottom design (almost certainly) and the top-gas recycle design discussed above (perhaps). The following discussions were the results of an attempt to identify a gasifier configuration that had a high likelihood of being able to beneficially digest a large fraction of coal fines in the feed without creating a large particulate loading in the product gas. The

Table 10. Internally-Skirted, Recycling Gasifier

PRO	CON
1. The tar and oil and ammonia levels in the product gas should be reduced by internal cracking. The amount of reduction is currently unknown and is likely to be a strong function of the injection-port location.	1. Maintenance of spatially constant reaction zones within the gasifier is required for reliable operation (and is difficult to do).
2. Various coals are accommodated by the position of the upper skirt, feeder and grate speeds, and stirrer protocol.	2. Heating value of the hot product gas is very low compared to a one-stage unit.
3. An advanced coal feeding system should accommodate any coal, reduce fines loading in the top gas, and preclude the need for a bed-leveling stirrer arm.	3. Eductor steam must be very hot and is a significant fraction of the gasifier needs.
4. The grate has some ash crushing capability and will tolerate sizeable agglomerates.	4. The gasifier is significantly longer than a comparable one-stage unit.
5. Lowered gas velocities in the pyrolysis zone should slightly improve the ability to accommodate caking coals.	5. The design needs two stirrer arm lengths and <u>must</u> stir no lower than bottom of the upper skirt (which may not be sufficient).
6. External fines could be added to the recycle stream to beneficially dispose of coal fines.	6. A complex recycle pathway design exists. The main issues are the difficulty in distributing the recycle stream across the bed and the possible build-up of deposits in the recycle piping (and the consequent need for "thru-reaming" as a periodic maintenance operation).
7. The lower skirt is a potential fire-zone sensor location.	

resulting two gasifier concepts are shown in Figures 33 and 34 and share a novel feature: a co-current upper bed coupled with a countercurrent lower bed. The basic idea is to feed the total coal supply onto the top of the bed, but to have the hot-gas flow down through the upper bed so that all fines are carried into the bed and presumably bond with the lump coal. The single product-gas outlet would be at a mid-level position and probably somewhat below the point at which devolatilization has been completed. This would be expected to result in a

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Figure 33. Two Stage, Co-Current Flow Gasifier

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Figure 34. Overfired, Split-Feed Gasifier

product gas that was somewhat hotter and likely a bit lower in hydrocarbon volatiles and ammonia compared to a one-stage, normal, fixed-bed gasifier. The difference between Figure 33 and 34 is the means by which the hot gas is supplied to the upper bed.

The two stage, co-current gasifier concept of Figure 33 bears some similarity to the configuration shown in Figure 32, and discussions of the common characteristics will not be repeated. This concept would contain all of the combustion zone and some of the gasification zone in the lower, countercurrent stage. Hot gases, at probably 1800 °F (or above), accumulate in the ullage volume behind the skirt in the top of the lower stage and leave this stage by the side pipe. The gases are routed to the ullage volume over the upper coal bed and then pass co-currently downward through the upper coal bed. The downward flowing gas cools as it dries, devolatilizes, and partially gasifies the coal in the upper stage. The solids and gas eventually pass down through the exit plane of the skirt in the upper stage, at which point the gas turns and wafts up into the ullage volume behind the upper-stage skirt and out the gasifier's exit port. The partially gasified coal is now in an essentially non-reacting zone of the bed and should be free flowing. At this point, the coal/char flows onto the upper end of an auger, which feeds it into the higher pressure lower stage where gasification and combustion are completed.

The primary potential issues with this concept are (1) the ability to provide an auger that reliably feeds coal/char while maintaining a small pressure differential, e.g., a few psi (maintaining a small leakage gas flow is critical if the downward co-current gas flow is to be achieved in the upper stage); and (2) the ability to adequately gasify the coal in the upper stage. This latter issue is a summary of several separate concerns. For example, how hot must the lower stage exit gas be to be effective, how much gasification should occur in the upper stage, how long must the upper stage be (i.e., the slowest chemical reactions occur at the point the hot gas is at its coolest temperature), what is the likelihood that an alkali catalyst will be needed to obtain sufficient gasification in the upper bed, what is the probable pressure drop through the upper co-current bed, and what is the probability of low overall thermal efficiency because of the co-current gas and solids flow and the high susceptibility to performance degradation caused by heat losses? There are also secondary issues that are non-trivial: for example, what is the effect on the stirrer drive design (because of the presence of very hot gases in the bearing or seal region of the shaft and since it appears the portion of the bed that would need to be stirred will be lengthened), will the total gasifier height become oppressive (the interstage region will also add a length increment), how is the interstage feed auger designed to operate in the hot environment and controlled, will significant tar and oil cracking really occur in the upper stage, and does the lower stage need a bed-leveling function to effectively react the coal/char exiting the inter-stage auger? This formidable list of issues kept the concept from further consideration.

The over-fired, split-feed gasifier concept of Figure 34 bears some similarity to the configuration shown in Figure 31, and discussions of the common characteristics will not be repeated. This concept would split the combustion process, but again a significant portion of the gasification zone will be located in the lower, countercurrent portion of the reactor. In

this concept, lump and fine coal are separated prior to being fed to the gasifier; lump coal is fed normally and fine coal is entrained through a burner into the ullage volume at the top of the gasifier (in either a steam or air stream, or possibly as a coal-water slurry). A portion of the gasifier's air blast also enters through the burner, and the coal fines are burned in the ullage volume; the hot POC gases (and fine ash) and lump coal then move co-currently downward in the gasifier. As with the concept shown in Figure 33, the coal bed is dried and devolatilized, but in this case, is mostly gasified by the cooling, downward flowing gases, and some level of tar and oil and ammonia cracking is likely to occur. Eventually, the downward flow passes below the inter-stage skirt (which is the low-pressure point of the gasifier), and the gases waft into the ullage volume behind the skirt. The char continues to flow downward through the countercurrent combustion zone where it is converted to ash, and the rising gaseous POC are allowed to mix and react with the upper-stage product gases to form the gasifier's product gas, which is collected behind the inter-stage skirt and flows out the single product-gas exit port. An amount of air necessary to combust the char, and steam necessary to moderate ash clinkering tendencies and provide grate cooling, are provided to the bottom of the gasifier.

The major concern with this concept is, can the amount of heat necessary to adequately drive the upper stage reactions be practicably obtained by combustion of coal fines in the ullage volume at the top of the gasifier? As in the above concept, this is a summary of a host of sub-issues; for example, how much fines are needed (and will this usually result in having to grind lump coal to provide them), how hot must the overbed combustion gas be to be effective, what is the overall effect on the stirrer drive design, what is the impact on total steam and air requirements for the gasifier, how much gasification should occur in the lower stage, and how long must the upper stage be? There are again secondary issues that are non-trivial; for example, will the total gasifier height become oppressive, how is the gasifier controlled at part-load, will significant tar and oil cracking really occur in the upper stage, and what is the likelihood that an alkali catalyst will be needed to obtain sufficient gasification in the upper bed? This list of issues also kept this concept from further consideration.

As mentioned earlier in this report, the direct injection of coal fines to a location deep within the bed is a logical alternative route to fines utilization. The chevron injector unit discussed in Section 4.1 of Chapter 4 would be one type of device for accomplishing this; however, it may not handle a very high fines concentration well. A viable alternative design that should accommodate high fines concentrations is an adaptation of the lower skirt region from the recycling gasifier concept, shown in Figure 31. This adaptation is shown in Figure 35. Conceptually, this design feature could be applied to any type of fixed-bed gasifier; however, practically speaking, it is much better suited to a one- or two-stage, dry-bottom gasifier than a recycling gasifier or a slagging unit.

The design is based on an internal skirt being fixed to the wall of the gasifier at a location in the lower portions of the gasification zone. In operation, the fines would be entrained in a steam stream and transported in dense-phase mode into the gasifier, specifically into the ullage volume behind the fixed skirt. The fines would mostly associate with the in-situ solids and would be carried downward to provide fuel to the combustion zone; however, a small fraction may be carried upward with the bulk gas flow and be gasified in higher

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Figure 35. Direct Coal-Fines Injection Approach

regions of the reactor. In either case, the fines are beneficially consumed. The concerns described in the above discussions of recycling gasifier design, in-bed feeding systems, and coal fines in general are all relevant and will not be repeated here. Of most concern is whether the steam flow can be kept low enough that the gasifier operation and output will in fact be benefitted by this approach. It is likely the gasifier height will grow somewhat to accommodate a larger fraction of the carbon in the top-fed coal being gasified (slow reactions) where this fraction was formerly being combusted (fast reactions). It is also likely that multiple, circumferentially located, entrained coal injection nozzles will be preferred to make the distribution of the coal fines as uniform as easily possible within the bed; fortunately, uniform distribution is not likely to be a stringent requirement in this case.

5.2 FINES AGGLOMERATION

As discussed earlier, the reality of these IGCC systems is that coal fines will exist in significant quantities and that, to the degree these fines cannot be utilized like lump coal, they represent a burden upon the overall system. This burden is obviously both site-specific and coal-dependent and can approach zero; however, identification of a nationally universal means for coal fines utilization appeared desirable, if possible. A means of utilization would be judged advantageous only if its costs were less than the costs (i.e., burden) of utilizing (or disposing of) fines as "fines." After looking at several alternatives, a potentially low-cost and low-risk approach to fines utilization was found to be provided by agglomeration: specifically, briquetting; this is expected to be the preferred option for a significant number of plant sites. The information in this Section is provided to serve as a reference point for site-specific evaluations.

Agglomeration is the process of particle size enlargement in which small, fine particles (such as dusts or powders) are gathered into larger masses, clusters, pellets, extrudates, or briquettes, and normally using a binder material. There are three basic processes for agglomeration:

Pelletizing: The formation of pellets or balls by growth agitation and rolling of fines with moisture, which is generally done in rotating disc- and drum-type pelletizers. The process usually consists of green-pelletizing and drying of the pellets, which is typically followed by firing, calcining, or heat induration.

Briquetting: A pressure agglomeration process that forms, or pressure-molds, a shaped product called a briquette. In this country the usual shape resembles a pillow, but a large number of shapes are possible. The most common type of process equipment is described as a double-roll press briquetter.

Extruding: A pressure agglomeration process that forms a two-dimensional shape by forcing a plastic mass of fines through a die. Types of extruders include auger/screw, ring-die extruders, pellet presses, ram and piston extruders, and gear extruders.

Available studies and reports were reviewed (Davy McKee Corporation 1983; Holley 1986) and vendors were contacted, a number of whom provided valuable information and insights.^{1,2,3,4} This body of information formed the basis for this Section.

The intent of the agglomeration process in an IGCC system is to combine whatever fines exist at the site, either as a result of separation from the as-received ROM coal or as an output from a process step (e.g., cyclone dust), to form a coal-like feed stream to the gasifier. The result of the agglomeration is to be a lump product that is suitable for stockpiling and direct feed to a fixed-bed gasifier, preferably by simple addition to the coal feed stream. It is reputed that costs associated with conventional agglomeration processes have significant scale effects associated with them, and so a reasonably representative set of assumptions was necessary to arrive at credible cost values for an agglomeration process. The assumptions were

- Fines would be defined as all coal-like solids less than 0.25 inches in size.
- The minimum and maximum amount of fines requiring agglomeration would be 300 and 750 tpd, respectively.

The latter assumption was felt to be generally representative of the fines load that could correspond to a simplified IGCC system with a rated power level in the 100- to 250-MWe range and encompassing virtually any U.S. coal.

The initial reviews led to some conclusions regarding agglomeration in general, at least from a traditional point of view. As an initial comment, it was obvious that most people judged agglomerates to be good or bad based on comparisons to the performance of lump coal in a similar situation. However, some of these comparisons may be overly stringent in the context of a real IGCC application; particular examples of thought-to-be-valuable agglomerate characteristics that realistically are not vital in an IGCC application are as follows:

- Outdoor storability in uncovered piles for long time frames (e.g., 1 year) without significant decrepitation.

¹Rumley, Jamieson B. October 1989. Ferro-Tech, Michigan. Personal correspondence with H.M. Spengler, METC.

²Dalton, Barney. November 9, 1989. K.R. Komarck, Inc., Illinois. Personal correspondence with H.M. Spengler, METC.

³Hinkle, Robert G. November 15, 1989. MMC Mars Mineral, Pennsylvania. Personal correspondence with H.M. Spengler, METC.

⁴Engelleitner, William H. October 17 and November 14, 1989. Teledyne Readco, Pennsylvania. Personal correspondence with H.M. Spengler, METC.

- Mechanical integrity at both ambient and elevated temperature conditions (up to 1200 °F).
- Physical strength adequate for cross-country transportation and handling (values approximating solid coal were generally desired) with little to no degradation over long time frames (e.g., 1 year).
- Heating value to approximate lump coal and not degrade over long time frames (e.g., 1 year).

For the IGCC application, it should be remembered that the required rate of fines processing is in essence only a function of the quantity of lump coal gasified per day (i.e., the operating status of the system), and consequently, the amount of finished agglomerates required to be on hand only needs to be the quantity desired to provide surge capacity for the system. This is generally a small enough quantity that special storage conditions and short storage durations (and even special handling), if required, are not major problems or costs.

From a traditionalist's point of view, each coal is unique and should be processed in a particular way to make a good agglomerate. In addition, some coals are said to have good agglomerability and some have poor agglomerability. However, most approaches utilize a binder of some form that acts much like a glue and tends to obscure the coal-unique characteristics. Further, if maximum structural characteristics are not demanded, one type of binder can work reasonably well for a wide range of coals. An agglomerate generally will not be as strong as the parent lump coal, and the choice of binder is a major influence on the strength of the product agglomerate. The costs of the agglomerating process typically exceed 50% of the cost of the raw coal (on a per ton basis), and the binder is the most important single item in the cost. Binders that have been utilized have costs ranging from \$1.75 to over \$50.00 per ton of agglomerate produced; operation, maintenance, and labor are found to be relatively secondary cost items.

Binders are used in three basic physical states: as dry powders, as liquids, and as molten pitches. The pitches are quite effective but tend to both be expensive to purchase and also to require expensive equipment to manage the pitch temperature and the fines and pitch blending portion of the process. Fixed-bed gasification processes are unique in that the vast majority of them have a product stream rich in hydrocarbon volatiles that with careful condensation, could serve as a source of "free" pitch (conceptually, at least). Existing experience has defined 10 to 15% of the weight of the coal fines to be the weight of binder pitch required for good agglomerates. An analysis was done to determine what fraction of the fixed-bed gasifier product-gas stream would need to be cooled to provide an adequate amount of pitch using an approach such as that shown in Figure 36. A condensation temperature above the boiling point of water at the gas stream pressure was necessary to retain all of the inherent and added water in the gas stream, (in order to preclude additional system complications), and 425 °F was selected as a representative adequate temperature. For the study's reference coal, this would result in about half of the hydrocarbons present in the gas stream

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Figure 36. Concept for Process-Derived Agglomeration Pitch

being condensed. In summary, this approach turned out to be inadequate. While there is variation in the amount of hydrocarbon volatiles that will find their way into the gas stream (and two-stage gasifiers were obviously best), even if the whole stream were cooled, it would just barely provide an adequate pitch supply for the corresponding amount of fines assumed to be in the system. Since additional questions remained regarding the cost of pitch gathering and handling equipment and even the suitability of the pitch as a binder for its parent coal, the whole approach of using pitch, either purchased or process-derived, was dropped.

Dry powders mixed with water and liquids are widely used as agglomeration binders today: bentonite, corn starch, and bituminous coal dust are common powders, and molasses is a typical liquid. It would be nice if one low-cost binder would work well on all U.S. coals; unfortunately, that does not appear to be possible. However, it does appear possible that two classes of binders will be acceptably low in cost and capable of reasonably accommodating the vast majority of U.S. coals:

- 5% corn starch per unit weight of coal fines - This is a binder for lignites, sub-bituminous, and some bituminous coals. The typical cost is \$7.30 per ton of agglomerate (in 1989 dollars).
- 10% bentonite plus 10% high volatile bituminous coal dust per unit weight of coal fines - This is a binder for some sub-bituminous and most bituminous coals. The typical cost is \$5.80 to \$6.70 per ton of agglomerate (in 1989 dollars).

Binder costs can obviously be lowered to the degree that lesser quantities of binder can be utilized, e.g., only 4% corn starch (which frequently is adequate). It should also be mentioned that water alone is a reasonable binder for many bituminous coals; for those coals where water is an adequate binder, it will be the low-cost option.

The specific properties and features of the three agglomeration processes - extruding, pelletizing, and briquetting - were reviewed to identify characteristics relevant to the IGCC application. Extrusion (of a mixture of coal fines plus a binder) was deleted from further consideration because of problems associated with both the process and the product. It was found to be critical to both the process and the product that the feedstock mass be kept at high pressure for a 5- to 10-min retention time and within a relatively narrow range of physical properties (i.e., plasticity) until the actual extrusion occurs. The process results in two-dimensional-shaped extrudates of intermediate strength, which are cut to length for further handling and storage. Extrusion equipment is expensive per unit throughput compared to other options, and would be expected to require a relatively high level of maintenance. The extruding process appears best suited to materials with inherently plastic properties (e.g., clays, foods, and plastics).

Pelletizing was also dropped from consideration, primarily because of the large number of inherent process steps, but secondarily because the product pellet was felt to be structurally inferior to a briquette. Pelletizing requires a fine and uniformly sized feedstock with a

closely controlled moisture level for successful operation. Consequently, grinding the fines to a uniform size (typically minus 100 mesh) and additional drying will be required prior to the actual pelletizing process. In addition, the product requires drying, curing, or both following pelletizing, and depending upon the binder used, usually some form of post-pelletizing heat treatment (and associated cooling) is also required. The product pellets are of a uniform size; however, they are also of intermediate to low crush strength. While this is recognized as a potential handling and breakage problem (and is similar to extrudates), it has sometimes also resulted in an alternative problem -- (multi-pellet) agglomeration occurring in the storage bins. The operation of a pelletizing system requires a relatively high degree of operator involvement to monitor feedstock and product pellet size.

Briquetting was found to be the most promising agglomeration method. It combined the best characteristics of extruding and pelletizing and had essentially none of the problems. In particular, the product briquette has good strength, is of uniform size, and the production process is relatively low in cost. Pressure-roll briquetting can be made to work for all U.S. coals through modification of the binders and adjustment of the roll pressure. However, all U.S. coals are not low in cost to briquette, but most coals can be made to produce briquettes with acceptable strength and good storage characteristics for reasonable cost using one of the two binders described above. (A briquetting test is required to ascertain the capability for a particular feedstock.) Coal briquetting has been done since the early 1900's and is a proven agglomeration method for coal fines.

A conceptual-level design and costing exercise was done for a briquetting subsystem, based on the assumptions given above and information from equipment vendors. The general process-flow diagram for briquetting is provided in Figure 37. A configuration based on pressure-roll briquetting was selected as the most universal agglomeration method. It is a proven method, especially with bituminous coals. The desired capacity, to process 300 to 750 tpd of fines, translates to briquetting units with throughputs of somewhat less than 15 tons per hour (tph) to something over 30 tph. Standard size units are available with throughput ratings of 25 and 50 tph, with 25 tph being essentially the smallest commercial-size unit. The use of two parallel briquetting units was selected because the cost penalty was moderate and the resulting additional reliability and operating flexibility were very large. In addition, this configuration provides conservative costs for this approach to fines utilization, and thus if the costs are attractive even at the 300-tpd scale, briquetting could become the general solution to fines disposition.

The process flow diagram for the selected configuration is shown in Figure 38. This is in reality a 200- to 1200-tpd system and is based on using two, 25 tph, double-roll, pressure briquetters and associated feeding and storage equipment. The selected process flow diagram shown on Figure 38 eliminates the drying/curing unit of the typical briquetting process. This is possible as a result of the binder selection, since starch- and bentonite-based briquettes are expected to only require cooling following the briquetting process. The cooling function is to be incorporated into the product side-belt conveyor design. The capital cost is nominally \$2.5 million and is outlined in Table 11.

Figure 37. Process Diagram for a Typical Briquetting Facility

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Figure 38. Process-Flow Diagram for a 50-tph Briquetting Facility

Table 11. Capital Cost Estimate¹

ITEM DESCRIPTION	COST
Bin with two feeders	\$ 30,000
Feed side belt conveyor (2 required)	20,000
Pneumatically filled bin with two feeders	25,000
Model 24T79 turbulator (2 required)	100,000
Roll briquetter (2 required) ²	500,000
Vibrating screen/conveyor (2 required)	20,000
Recycle bucket conveyor (2 required)	30,000
Product side belt conveyor (2 required)	20,000
Electrical	200,000
Engineering and Construction Management	200,000
Installation	<u>750,000</u>
	Subtotal = 1,895,000
Contingency Allowance ($\approx 25\%$ of Subtotal)	<u>503,000</u>
	TOTAL = \$2,398,000

¹ Coal briquetting facility capacity is 50 tph.

² Briquetting unit costs are nominally \$250,000 for 25-tph and \$300,000 for 50-tph size units. Vendor-recommended facility configurations ranged from using one 50-tph unit to using three 25-tph units.

The production capability range of 200 to 1200 tpd for this system is based on operational modes ranging from using one 25-tph briquetter for one 8-hour shift to use of both 25-tph briquetters for three 8-hour shifts. Having parallel briquetters in the system allows for downtime and maintenance operations. If the facility is reduced in scope to include only one briquetter, the maximum capacity would be 600 tpd, and the capital cost of the facility can be reduced by \$405,000 of direct capital, plus the proportionate fractions of the last four line items in Table 11. If one 50-tph briquetter were utilized, the maximum capacity would remain at 1200 tpd and the direct cost savings would approximate \$305,000, but the redundancy feature would be lost. These operating modes and the capital cost/production capacity relationship are shown graphically in Figure 39.

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Figure 39. Briquetting Facility Capital Cost/Capacity Relationship

It is safe to assume the capital cost for a complete briquetting facility to be added to a fixed-bed IGCC module of 100- to 250-MWe scale will be in the range of \$2 to 3 million. This cost is a considerable variance from the Davy McKee report (1983), but the cost range of \$2 to 3 million (1989 dollars) has been verified by four vendors supplying independent quotes.

The operating cost of the facility is expected to be nearly \$10.00 per ton of briquettes produced, plus the cost of binder material. This value is verified both by the Davy McKee report (1983) as well as by the independent vendor quotes. A scale dependency associated with operating cost is defined in Figure 40. This effect could add nearly \$2.00 per ton to the operating cost of briquetting for very small facilities. The binder cost is primarily a function of the type of coal and the geographic location of the plant (i.e., local availability of the binder). Once a facility location and coal type selection is made, the binder or binders to be tested for suitability can be selected based at least partially on their potential to minimize operating cost. Initial consideration of using corn starch, bentonite, or high volatile bituminous coal dust as binder materials is highly recommended as these materials are widely available, suitable for a wide variety of U.S. coals, and are relatively low in cost. A delineation of the expected operating costs and binder cost options is provided in Table 12. It should also be mentioned that the use of a binder without high-temperature capability may be acceptable, in contrast to normal binders that seek to provide structural integrity at elevated temperatures. This possibility exists because when the briquette starts to warm up, the briquette has, by definition, been placed in the gasifier and is overlaid with additional coal; consequently, a significant level of breakage should be acceptable at this point. High volatile, bituminous, coal dust is a typical high-temperature binder material.

5.3 HOT FUEL-GAS CLEANUP

As described in the opening of this chapter, the fuel gas from the gasifier is routed through a particulate removal step followed by a combined tar cracking and sulfurous gas removal step to achieve a cleaned, hot, fuel gas suitable for combustion in the GT. Both steps are accomplished without fuel gas temperature or pressure changes, other than the nominal losses inherent in the cleanup and piping system.

The particulate loading in a fixed-bed gasifier's product-gas streams is typically lower and has a larger mean particle size than occurs with other forms of gasification. As mentioned earlier, the basic CGIA concept eliminates the major penalty associated with large quantities of fines carry-over, and the fines-disposition economic penalty is likely to be modest. As a consequence, there is a likelihood of increased gasifier air flows, which will increase particulate carry-over. However, it is expected that adequate levels of particulate removal from the product gas will be relatively easy to accomplish.

There are three basic technology options for particulate removal at fixed-bed gasifier product-stream conditions: cyclones, electrostatic precipitators, and barrier filters. Of these

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Figure 40. Briquetting Facility Operating Cost/Capacity Relationship

Table 12. Operating Cost Estimate¹

ITEM DESCRIPTION	COST PER TON
Binder cost ²	Not Included
Electricity	\$1.95
Operating Labor	\$1.95
Maintenance (Material and Labor)	<u>\$5.60</u>
	TOTAL = \$9.50
<u>BINDER COST OPTIONS</u>	
4% Corn starch	5.86
5% Corn starch	7.30
5% Bentonite	2.56
10% Bentonite	5.12
10% Bentonite plus 5% Pittsburgh No. 8 coal dust	5.88
10% Bentonite plus 10% Pittsburgh No. 8 coal dust	6.66
10% Bentonite plus 10% Illinois No. 6 coal dust	6.70

¹ Coal briquetting facility capacity is 50 tph.

² Adding the binder cost to \$9.50 is required to obtain the total operating cost per ton.

options, cyclones are currently state of the art and some electrostatic precipitator and barrier filter designs are nearly so. As a consequence, there is no need for any significant discussion of particulate removal technology in this report. What is important is to recognize that the performance of all of the downstream portions of the IGCC system is jeopardized if the vast majority of the entrained particulate is not removed from the gas stream as physically close to the gasifier outlet as practicable. Most of the reasons for desiring particulate removal are obvious, but a non-obvious reason is the basis for wanting the particulate removal device to be close to the gasifier. In the event there is local cooling of the gas stream (e.g., resulting from a temporary loss of exterior piping insulation), local condensation of the hydrocarbon vapors can occur. If there is a large population of particulate in the gas stream, this condensation can lead to aggregate-laden deposits within the process components and piping. The presence of internal particulate increases the physical strength of the deposits and makes them much more tenacious. As a gross simplification, the greater the internal surface area of the

process piping and equipment existing upstream of particulate removal, the more susceptible the system is to unplanned shutdowns.

Of somewhat more importance is the disposition of the captured particulate in a manner advantageous to the system once the particulate is removed from the fuel gas stream. Figure 28 (page 128) shows four fines disposition options: (1) disposal outside of the IGCC system, (2) recycling to a location deep within the gasifier, (3) transfer to an agglomeration facility, and (4) utilization in a novel PFBC. (The first three of these have been discussed in preceding Sections and option 4 will be discussed in the next Section.) The choice is essentially based only on relative economics, given that a gasifier capable of making option 2 workable has been selected. (Currently, this would be limited to BGL's slagging fixed-bed design.)

In the reference IGCC system shown in Figure 28, tar cracking and desulfurization are both accomplished in a single fluidized-bed unit using a zinc-ferrite-based sorbent. The discussion of this unit is the focus of this Section. It is pointed out that disposition of the regeneration gases is a subject of major importance to the overall system, and it is covered separately in Section 5.4.

5.3.1 Tar Cracking/Desulfurization Subsystem

The use of a fluidized bed was preferred for zinc-ferrite-regeneration purposes based on the perceived advantages that should result: (1) improved thermal control and the possibility of easily extracting heat for other uses from the regeneration step, (2) increased volumetric efficiency of the processing vessel and downstream equipment because of the capability for efficient gas/solids contacting and a consequently small volume regeneration-gas output-stream, and (3) the capability to regenerate zinc ferrite using air without forming much zinc sulfate within the sorbent particle. Consideration of the fluidization conditions (and the related sorbent particle sizes) appropriate to regeneration, and the ability to make the absorber vessel insensitive to modest levels of zinc sulfate in the regenerated sorbent, led to the conclusion that a fluidized-bed unit would also make the most sense for the tar cracking and sulfur absorbing step.

METC is currently addressing the design issues of fluidized-bed desulfurization for IGCC systems; this work is not complete. Therefore, the design concepts presented below must be considered preliminary in nature and may well change as experiments are completed. The major unknowns at this point include global reaction rates for absorption (whether controlled by bubble gas exchange with the emulsion phase, pore diffusion into the sorbent, or chemical kinetics), global reaction rates for regeneration, the effects of heat release on the sorbent during regeneration reactions, bubble size and related gas exchange parameters in a zinc ferrite system, alternative sorbent formulations, and finally, sorbent attrition rates. These questions are being addressed by specific portions of the METC program.

The approach taken in this design was to model the system after catalytic cracking, which has had a long and successful commercial history in the petroleum industry. A major difference between fluidized catalytic cracking (FCC) and the currently envisioned hot fuel-gas cleanup subsystem is in the operating pressure. The typical fluidized catalytic cracker operates at about 20 psig, compared to the 250 to 500 psia levels considered in the CGIA study. Specifically, the particle size of the sorbent chosen for this application was 70 μm , and an operating superficial velocity in the range of 1 to 3 ft/s was selected to match FCC practice. Besides being consistent with the commercial experience base, the small sorbent particle size also provided the best opportunity to achieve very high sulfur removal levels without staging the desulfurization bed. The actual (when developed) sorbent particle of this size may turn out to be too light in weight for the selected gas velocity range; in this case, the particle size will be increased slightly, or the superficial velocities will be limited to near 1 ft/s, or both. (However, it should be pointed out that small particles under 200 μm have been found to effectively clump in an operating fluidized-bed, which allows operation at much higher velocities with less carry-over than would theoretically be expected. This phenomena must be considered and evaluated when selecting the final sorbent particle size and the bed's operating gas velocity.) Laboratory data have shown that Y-zeolites, which are commercially available as fluidized-bed cracking catalysts, are capable of cracking and releasing the sulfur from most coal tars and oils. FCC catalysts are also highly attrition resistant and thus provide a good basis for the design of a commercial-grade tar cracking/sulfur absorbing media.

There are two viable approaches to the media. The first is to incorporate zinc ferrite into a Y-zeolite type of particle structure to achieve all the desired functions within each physical particle, and the second is to form zinc-ferrite particles with physical size and density characteristics that nearly duplicate that of the Y-zeolite particles. (Again, the clumping characteristics of small particles may well allow a bed of seemingly disparate size and density particles to fluidize as if it consisted of uniform media.) In the latter case, the desired functions are achieved with a physically well-mixed bed of similarly sized but discretely separate tar cracking and desulfurizing particles. With either approach, the tar cracking and desulfurization functions are combined in one vessel.

5.3.2 Two-Vessel Concept

Figure 41 illustrates this process configuration. The concept is based on a typical two-vessel FCC configuration employing sorbent standpipes with steam stripping and slide valves to ensure isolation of the reducing conditions in the desulfurization bed from the oxidizing conditions in the regenerator. This is conventional FCC practice, which has proven to be effective in commercial service.

The two vessels are positioned roughly side by side, one serving as a tar/oil cracker and sulfur absorber, and the second as the catalyst/sorbent regenerator. Regenerated catalyst/sorbent exits the regenerator standpipe, is contacted by fuel gas from the gasifier, and is

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Figure 41. Two-Vessel Fluid-Bed Absorber/Regenerator Concept

carried as a dilute phase suspension through the absorber/cracker grid into a bed of catalyst/sorbent media in the absorber tar-cracker vessel.

Fuel gas rises through the absorber bed in two ways. Under the conditions selected for this design, a relatively small amount of gas flows interstitially in the emulsion phase, while most of the gas flows through the bed as bubbles. The interstitial gas achieves intimate contact with the sorbent. The bubble gas, while in much less intimate contact with the sorbent, achieves increasing degrees of effective contact as a result of its continual exchange with the emulsion gas as the bubble rises through the bed. Providing a deeper bed of sorbent thus provides a greater degree of contacting and therefore more opportunity for tar cracking and sulfur removal. Tars and oils crack to deposit coke on the catalyst and to release sulfur-bearing gases, lighter hydrocarbons, and hydrogen. The sulfur-bearing gases react with the sorbent to form sulfides within the sorbent structure.

The sorbent level in the absorbing vessel bed is controlled by the system's total inventory of solids, the height of the standpipe (weir) within the absorber vessel, and the solids withdrawal rate (via a dense-phase standpipe). Solids flow down the standpipe is controlled by a slide valve located near the bottom of the pipe. Sufficient steam is injected along the standpipe to maintain a slightly fluidized condition and to ensure that fuel gases carried along with the flowing solids from the reactor are swept out (stripped) and returned to the absorber. The dense-phase fluidized state in the standpipe results in a pressure increase with length down the standpipe (caused by the static head of fluidized solids) such that the highest pressure occurs at the bottom of the standpipe, and it is designed to be sufficient to prevent incoming regeneration air from entering the tar cracking/sulfur absorbing vessel.

Solids exiting from the bottom of the absorber/cracker standpipe are picked up by transport air (or possibly steam) and carried as a dilute-phase suspension upward and into the regenerator vessel. Here air contacts the sorbent, burning off carbon and reacting with the metal sulfides to form sulfur dioxide. Gas contacting can be increased by increasing bed depth, but unlike the absorber, the regenerator can be operated with an excess of the fluidizing gas, air. Operation in the excess-air mode allows achievement of a higher level of regeneration, the utilization of shallower beds, or both. Commercial FCC experience has shown high levels of regeneration can be achieved with excess regeneration-air levels of only a few percent, and it is reasonable to expect that high levels of regeneration in a zinc-ferrite system can also be achieved even with very modest amounts of excess air.

The design of the solids flow path from the regenerator vessel mimics that of the absorber. Solids flow downward through a regenerator standpipe, which is slightly fluidized with stripping steam and controlled with a slide valve. The regenerated solid media is then picked up by the incoming fuel gas and is transported upward into the tar cracking/desulfurization vessel, thus completing the solids flow circuit.

As background for its possible selection, the quantity of steam required to transport the sulfided sorbent to the regeneration vessel was estimated. The amount of sorbent to be

transported (per unit time) is effectively only a function of the amount of coal gasified (specifically, the amount of sulfur sent to the absorber) and the sorbent utilization. The estimate was based on the following assumptions:

- Use of the reference coal containing 2.8% sulfur.
- All coal-borne sulfur finds its way to the absorber.
- Transport steam conditions of 1000 °F and 234 psia.
- A conveying velocity of 30 ft/s.
- The theoretical maximum sulfur loading at 35% of the sorbent weight.
- Zinc-ferrite sorbent with a particle density of 137 lb/ft³, a bulk density of 80 lb/ft³, and an average particle diameter of 200 μm.

(The assumed 200-μm sorbent diameter value exceeds the 70 μm size selected above but was utilized since it probably represents a maximum for the eventual sorbent size, and consequently, it results in a conservative steam estimate.) Using the above assumptions, the sorbent's specific circulation rate (i.e., the amount of sorbent per unit weight of coal gasified that must be circulated between the absorber and the regenerator) was calculated as a function of sorbent utilization; this is shown in Figure 42. The percent of sorbent utilization is the weight of sulfur actually absorbed divided by the weight of the maximum amount of sulfur that could theoretically be absorbed by the sorbent (expressed as a percent). The theoretical maximum is the amount of sulfur required to react all zinc and iron to their sulfide forms. As an example, a 95-tph gasifier feed-rate was considered (which would be representative of a 250- to 300-MWe plant module), and the required amount of conveying steam was calculated for choked flow conditions. These conveying steam-flow values were then multiplied by a factor of 5 to provide an operationally conservative design curve. The results of this example calculation are provided in Figure 43. A study of these two curves (with the knowledge that the gasifier will require about 160,000 lb of steam/hr for the conditions assumed) leads to the conclusion that there should be no problems with the steam-transport design approach, since (1) zinc ferrite utilization typically is in the vicinity of 40%, and (2) the conveying steam requirement (as illustrated in Figure 43) will be a minor fraction of the gasifier's demand and thus should not represent a significant system penalty.

A high particulate loading in the gas exiting either vessel (caused by possible dusting or decrepitation of the media) is a potential problem for the system, and particularly if it occurs in the cleaned fuel gas exiting the absorber (which would next pass to the GT). Because of the fluidized nature of both vessels, it will be difficult to prohibit very fine particulate from being suspended in the exit gas streams. The best approach to this is to provide a physically hardy, dustless media in the first place and to control the population of fine particles in the working media. As implied previously, the use of a modified Y-zeolite as a catalyst/sorbent particle should significantly reduce the dust generation potential. In addition, Y-zeolite offers a potentially increased temperature capability that would simplify design of the regeneration step and may also provide a basis for a sulfur sorbent with higher temperature capability than zinc ferrite. Control of fine particles within the system may be achieved by including a steam-operated, pneumatic, particle classifier at a convenient point in the solids

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Figure 42. Required Sorbent-Specific Circulation Rate

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Figure 43. Conveying Steam Requirement

flow circuit. (The bottom of the regenerator standpipe is a probable location.) This device could readily be designed to remove particles smaller than about 20 μm , thereby greatly limiting the particulate entrainment and also greatly easing the design of any cyclones or barrier filters (should they become necessary) on the exit gas streams.

It is also worth remembering that, assuming a physically hardy media is developed, the media is potentially the basis for an excellent combined function type of gas-processing unit. This unit would be designed primarily as a granular bed filter to remove solids, but enough gas residence time would be provided that both tar and oil cracking and sulfur absorption would be accomplished. Conceptually, it would take in raw gasifier product gas and output clean gas in one step. The particle classifier and media regeneration ancillary components of the unit would produce dust, heat, and sulfur dioxide gas as primary products. Whether or not this approach ultimately makes sense is obviously dependent on the existence of a suitable media; however, it is also dependent on whether there is a benefit to deleting one (or more) hot cyclones from the system. Since cyclones are not terribly expensive, there may not be much real system benefit to this approach, but it should be considered as media properties improve.

5.3.3 One-Vessel Concept

This is a variation of the two-vessel concept, and many of the same considerations are applicable.

One embodiment of this concept is shown in Figure 44. This concept is based on the fact that the reactions involved in catalyst/sorbent regeneration are fast and that a locally excess quantity of air is likely to be required around the particles to preclude their overheating. There is also the inherent need to transfer media from the bottom to the top of the absorber vessel, and there is a desire to minimize the volume of regeneration gas production. It is further recognized that there is some margin inherent in the absorbing and regenerating cycle since complete regeneration of the media is not really required if a higher media circulation rate can be sustained with an acceptable amount of physical degradation of the media (and parasitic effects, such as sorbent transport steam requirements, do not become excessive as a result of the lowered sorbent utilization).

In this concept, the regenerator vessel is replaced with a vertical, dilute phase, lift-pipe/regenerator. It is probable that the media exiting the absorber vessel will be hot enough to initiate combustion reactions; however, the use of preheated air (on the order of 700 to 800 °F, which is a level easily achieved with a GT-based power system) ensures that regeneration initiates promptly and proceeds smoothly. Representative lift-pipe lengths and media velocities are likely to be in the ranges of 45 to 90 ft and 15 to 30 ft/s, which would imply media residence times of 1.5 to 6 s in the lift-pipe/regenerator. Scoping tests at METC have shown that 85% regeneration of 30% sulfided zinc-ferrite sorbent (based upon the theoretical maximum) can be obtained in less than 2 s of entrained flow at temperatures of 1200 °F in

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Figure 44. One-Vessel Fluid-Bed Absorber/Regenerator Concept

several-hundred-percent excess air. This level of regeneration is quite adequate and implies that an increase in media circulation rate of less than 20% would make this concept viable (which should be achievable without markedly increased degradation). Estimates of the quantity of air required to convey the media have shown that less than stoichiometric quantities would be adequate to transport the media. Consequently, the lift-pipe air requirements will be determined by the thermal requirements of the media. If overt heat withdrawal is required to maintain an acceptable media temperature, the lift-pipe itself is a potentially adequate heat transfer surface (and can easily be cooled), and the addition of surface area internal to the lift-pipe should be quite feasible.

There is an obvious requirement to maintain separation of the regeneration air from the fuel gas. Isolation of the two gases is somewhat more difficult to accomplish in this concept than it was in the two-vessel approach; however, several means of achieving adequate isolation are well within the state of the art. Figure 44 illustrates the use of steam stripping, which is a typical practice in petrochemical plants and should be no problem to implement in an IGCC plant. In summary, this appears to be a very attractive concept.

5.3.4 Design of the Tar-Cracking/Sulfur-Absorber Vessel

The tar and oil cracking and desulfurization vessel for this system was conceptually designed, and was generically applicable to both concepts presented above.

The fuel-gas flow rate for the highest output IGCC configuration considered (250 MWe) was 1.5 million actual cubic feet per hour at 284 psia and 1100 °F. This gas rate can be accommodated by two absorbers each 12.7 ft in internal diameter operating at a superficial velocity of about 1.6 ft/s, or four absorbers of the same size operating at about 0.8 ft/s. Given that the eventually developed sorbent's physical size and density are found to be consistent with the desired fluidization conditions, some current uncertainties remain that will influence the choice between the nominally 1- and 2-ft/s superficial velocity: primary among these are sorbent durability, reactivity, and cost. The vessel design essentially becomes a tradeoff of superficial velocity against a combination of bed depth (i.e., sorbent reactivity) and sorbent attrition. Deeper beds can always be used to increase sulfur capture, but there is a practical limit set by vertical height considerations and to some extent by the pressure drop across the bed and distribution grid. Sorbent reactivity can also become an alternative to higher superficial velocities but only up to a limit set by the amount of gas-solids contacting occurring.

For this design study, sulfur removal was calculated using a "Davidson-type" bubble reactor model (McKetta 1985). In this model it is assumed that

- The reaction is first order in gas reactant concentration.
- No reaction takes place within the bubbles.

- Bubble gas is exchanged with emulsion gas by convection and diffusion through the "bubble wall."
- The bed is isothermal.
- Emulsion gas chemistry is uniform in a horizontal plane.
- Bubble gas is in plug flow.
- Emulsion gas may be in plug flow or back-mixed.

Essentially three unknown parameters are present in this model and these must be specified to calculate sulfur removal.

The first unknown parameter is the global reaction rate of the gas in contact with the solids in the emulsion phase. Louisiana State University (LSU) under contract to METC determined reaction rate data for extruded pellets prepared by commercial sorbent manufacturers. Similar studies have also been performed by AMAX, Research Triangle Institute (RTI), and others. The resultant data vary, which likely reflect varying degrees of reactivity in these relatively large extruded pellets. Until an actual material is prepared for fluidized beds in the correct size range and by a commercial process, a truly realistic global reaction rate will remain unknown. The LSU work found that over most of the reaction regime, the reaction was controlled by diffusion of reactants into the pores of the pellets. Assuming a similar pore size distribution in the comparatively finer sorbent being considered in this study, this reaction rate should be conservative since the envisioned particles will be much smaller and diffusion throughout the sorbent should be much more rapid. Within the Davidson model, pore diffusivity was the only resistance considered and it was related to the overall global reaction rate.

The second unknown parameter is gas diffusion through the bubble wall. This parameter was taken as zero (no diffusion), again a conservative assumption. The third unknown is the effective bubble size in the bed and it was assumed to be 2 inches in diameter. The basis for this assumption was (1) capacitance imaging measurements of fully developed bubble sizes made at METC using a fluidized bed of a 70- μm FCC catalyst, and (2) direct experience with the fluidization behavior of FCCs. (Note: bubble size will decrease with increased pressure, and bubble size is a critical parameter for gas-solids contacting, which decreases substantially with larger particles and, consequently, larger bubbles.) It is recognized that smaller bubbles are likely to be formed at or near the distributor, and that this will lead to increased gas-solids contacting and will reduce the actual bed-depth requirements. (This has been observed in commercial FCC units.) However, the consideration of a 2-inch diameter bubble represented a conservative assumption.

The model also includes bed depth at minimum fluidization velocity, which is really a measure of the solids inventory and superficial velocity, as well as the minimum fluidization velocity and the minimum fluidization voidage. These latter two parameters are determined by the sorbent and gas properties and are not independent. The model utilizes the input parameters to calculate the ratio of outlet concentration to inlet concentration of hydrogen sulfide; the percent sulfur captured is directly calculated from this ratio.

Figure 45 illustrates predicted absorber performance in an idealized case where there is no sulfur in the tar that is present with the fuel gas. The ratio of sulfur-outlet-to-inlet concentrations is plotted versus the expanded bed depth at superficial fuel gas velocities of 0.5, 1.0, 2.0, and 3.0 ft/s using a sorbent reactivity derived from the LSU results ($K_r = 1.172$). The expanded bed depth was independently calculated from the bubble diameter and the bed depth at minimum fluidization. The plot shows that sulfur capture efficiencies of 90% or better can be easily achieved in beds of under 15 feet in depth at velocities of 2 ft/s or less.

Figure 46 is a plot similar to Figure 45 except that the tar now contains sulfur. This plot assumes 5% of the input coal-borne sulfur (by weight) is found in the gas-borne tar and oil, and 90% of this tar and oil is cracked to release hydrogen sulfide, which may react with the sorbent. The figure shows that 90% overall sulfur capture is still attainable at a 2-ft/s superficial velocity with bed depths of less than 15 ft. The curves become asymptotic to a sulfur-outlet-to-inlet ratio of 0.005, which represents the uncaptureable sulfur in the uncracked 10% of the tar. Achieving a higher degree of capture, 99% for example, would require a lower bed design velocity, that is, 1 ft/s or less.

Figure 47 illustrates the performance that would be obtained with a reactivity nominally 200 times the value used for the two previous figures. This is an artificial condition and was computed to explore the case wherein sulfur removal would become limited by the rate at which gas is exchanged between the bubble and emulsion phases. For this calculation, the bubble size was held at 2 inches; however, in the event high sorbent reactivities are developed in the future, it must be remembered that bubble size will then become an important parameter (for a fluid-bed absorber approach). In this case 99% sulfur capture would be achievable in beds with design velocities of up to 3 ft/s.

Based on these results, bed designs with a 15-ft expanded bed depth and operating at fuel-gas superficial velocities of 1 to 2 ft/s should provide sulfur removal rates of 99% or higher. (To be conservative in the subsequently presented systems analyses and to allow for eventualities, a total sulfur loading of 100 ppm was assumed to exist in the cleaned fuel gas exiting the absorber. This would correspond to only a 97 to 98% sulfur capture level and is a level that should be improvable.) An additional 15 ft of freeboard is recommended to allow for disengagement of the larger entrained particles and to provide the vertical height that would facilitate using internal cyclones (i.e., for achieving pressure balance in the cyclone dip-legs) should they be needed for control of particulate loading in the cleaned fuel gas. A representative total active vessel size, excluding the vessel head and bottom, is a 12.7-ft inside diameter with a 30-ft length.

The commercial practice of FCC where heavy petroleum fractions are cracked suggests that coal tars can be effectively cracked in a fluidized bed; however, this has not been proven. Laboratory-scale studies of packed beds by Pacific Northwest Laboratories (PNL) (Baker and Mudge 1985) and Wen (1983) provide data showing that Y-zeolite, which is a commercial FCC catalyst, is effective at cracking coal tars. Figure 48 shows a plot of the fraction of coal tars that are cracked versus the total tars fed per unit weight of catalyst. The

Assumptions: 1. No tar sulfur & LSU reactivity
 2. Operating conditions
 - 70 μm particles
 - $K_r = 1.172$
 - no sulfur from tar

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Figure 45. Fluid-Bed Absorber Performance - No Tar

Assumptions: 1. Considers tar sulfur & "high" reactivity

2. Operating conditions

- 70 μ m particles

- $K_r = 1.72$

- 5% coal-borne sulfur is in tar

- 90% of tar is cracked

Figure 46. Fluid-Bed Absorber Performance - With Tar

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Assumptions: 1. Considers tar sulfur & "high" reactivity

2. Operating conditions

- 70 μ m particles

- $K_r = 200$

- 5% coal-borne sulfur is in tar

- 90% of tar is cracked

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Figure 47. Fluid-Bed Absorber Performance - With Tar and High Sorbent Reactivity

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Figure 48. Y-Zeolite Coal Tar-Cracking Activity

PNL data were taken using a packed bed of catalyst at 1070 °F with gas residence times of 1 to 4 s. The results were found to be insensitive to that residence time range, and consequently, 1 s was taken to be an acceptable design value. The tars were fed as vapors added to a gas mixture that simulated a Texaco gasifier's fuel-gas stream (which is not representative of a fixed-bed gasifier, but that characteristic is also not germane to this use of the test results). The PNL data show that over 80% of the tars were cracked initially, with a gradual decline to about 50% by the time 0.4 grams of tars/gram of catalyst had been fed to the bed. This decline is thought to reflect a deactivation of the catalyst by the deposition of coke from the tar cracking process. Later data from PNL, using packed beds of LZ-Y82 zeolite at 1022 °F and a gas residence time of 1 s, showed that a higher level of cracking (90%) could be achieved at low tar loadings. PNL determined that about 40 to 50% of the carbon present in the tars was deposited onto the catalyst as coke. PNL also found that approximately 95% of the sulfur in the tar was released. (The catalyst did not contain a sulfur sorbent.) The Wen data, also shown in Figure 48, was developed using the same kind of bed but at a lower temperature (840 °F) and tends to corroborate the PNL findings.

Figure 49 provides a plot of fuel-gas-bubble residence time in the catalyst/sorbent bed as a function of bed depth and superficial gas velocity. With a 15-ft deep bed at a superficial velocity of 2 ft/s, the gas bubble residence time will be over 4 s. While this does not correspond directly to the residence time in a packed bed of pellets, it does suggest that with the higher particle surface area resulting from the use of small particles (e.g., 70 to 200 μm in diameter), there should be sufficient time to achieve at least the level of cracking obtained in the PNL experiments. Commercial FCC experience supports the obtainment of high levels of cracking in short residence times. Many petroleum cracking units actually have little or no "dense" bed and accomplish the cracking reactions in the riser pipe going to the reactor (hence, the often-used term "riser-cracker").

5.3.5 Regenerator Vessel Design

The regenerator must provide a temperature-limited oxidizing environment in which (1) the coke deposited as a result of tar cracking can be burned from the catalyst/sorbent particles, and (2) the sulfides resulting from gas desulfurization can be converted to gas-phase sulfur dioxide. These processes need not be carried to total completion but must be carried to the point that the particles' reaction capacity is restored to levels that are adequate for the tar cracking/desulfurization vessel's requirements. For example, within limits, an incomplete regeneration can be offset by an increased solids circulation rate.

The air requirement for regeneration will be a strong function of the sulfur in the coal, the gasification tar yield, and to a lesser degree, to the fraction of carbon laydown resulting from tar cracking (i.e., the sorbent circulation rate and its carbon and sulfur content). Most coals considered for IGCC applications will have a sulfur content between 1 and 4%, and the expected tar and oil yields will fall into the range of 3 to 13% of the weight of coal gasified, with the percentage depending mostly on the coal's rank. Recognizing that the carbon

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Figure 49. Bubble Residence Time in a Fluid-Bed Absorber

laydown resulting from tar cracking has been reported at values ranging from 40 to just over 50%, it is credible to have requirements for stoichiometric regeneration air that are between about 0.25 and 1.25 lb of air/lb of coal gasified. It should be noted that the dominant influence is the amount of carbon on the sorbent, that is the tar and oil yield of the coal when gasified.

There is an incentive to minimize the air required for the regeneration step and, of course, regeneration vessel size. Unfortunately, both the sulfur and carbon combustion reactions involved in sorbent regeneration are strongly exothermic, and this combined with practical sorbent temperature limitations are likely to dictate a need for a significant amount of excess air. This is a near certainty for a riser-tube regeneration approach and could also be true with the use of a fluidized-bed regenerator. An appreciation of this potential problem can be obtained by examining Figure 50. This figure was developed assuming the adiabatic regeneration of zinc-ferrite and zinc-titanate sorbents without any deposited carbon being present. This is specifically applicable to regeneration of a non-tar-cracking sorbent in a riser tube regenerator, such as that discussed in Section 5.3.3. Limiting zinc ferrite to temperatures below 1500 °F will require on the order of 400 to 500% excess air. The temperature limitation for zinc titanate is not as well defined (and the lack of iron lowers the exothermicity of the reactions), but it appears likely that zinc titanate will require greater than 100% excess air. If carbon is present on the sorbent, the required air flow rate may be significantly higher; however, the percent of excess air required may not be significantly different.

This problem is most easily mitigated with the use of a fluidized-bed regenerator wherein the solids inventory could readily be made large with respect to the solids circulation rate; this would limit the effects of local media overheating (assuming acceptance of the costs associated with the increased vessel size). The use of aggressive heat removal from either fluidized-bed or riser-tube regenerators is, of course, the means to ultimately mitigate this potential problem; however, no design studies were done in this area. Without care, the magnitude of the resulting regeneration air requirement can easily become comparable to the quantity of air required by the gasifier (which will be on the order of 2.2 or 3.3 lb of air/lb of coal gasified, depending on whether the gasifier is a dry-ash or slagging design). In addition, a regeneration offgas with an inherently oxygen-rich nature will limit the technology choices for regeneration gas disposition. (However, it will have no significant impact on the reference system's PFBC approach.) In summary, this air requirement potentially represents a significant system-level performance penalty, and its quantification and design integration will require considerable care.

A regeneration bed depth of 15 ft with 15 ft of freeboard for a total of 30 feet of active vessel height should allow adequate gas solids contacting and the separation of the majority of the sorbent particles from the regeneration-gas exit stream. The superficial gas velocity in the bed will be a function of the developed media's properties and capabilities, but will almost certainly be under 2 ft/s. The diameter of the regeneration vessel will be determined based on the total gas flow that must be accommodated and the superficial velocity. Given the variations in coal properties and the uncertainties in stoichiometric air and excess

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Figure 50. Adiabatic Temperatures For Riser-Tube Sorbent Regeneration

air requirements, credible regeneration vessel diameters would be expected to fall in the range between 2.5 and 18 ft for the 50- to 250-MWe power ranges considered in the study. Based on a desire to simplify shipping and field construction activities, the worst case configuration would appear to require two parallel regeneration vessels of approximately 13-ft diameter.

5.4 APPROACHES FOR DISPOSING OF REGENERATION GAS

Cleaning the fixed-bed gasifier's product-gas stream to levels consistent with both GT and environmental requirements while maintaining modest amounts of thermal and pressure losses is not terribly difficult. However, one result of this design approach is a regeneration-offgas stream at physical conditions similar to those of the fuel-gas stream but containing sulfur dioxide (SO_2) plus a small amount of sulfur trioxide (SO_3) at levels that sum to nominally 3 to 8%, with the residual being a mixture of CO_2 (resulting from tar decomposition) and nitrogen (the majority component). In special configurations, the SO_2 plus SO_3 concentration could be as high as 13%, but this represents a practical maximum. The regeneration-gas stream represents essentially all the sulfur from the feed coal, but is in a gas phase and is concentrated from what would appear in a combustion process flue-gas stream. It obviously cannot be vented for environmental reasons, and its pressure and temperature characteristics provide incentive for a disposition technique that is energetically useful on a system basis. Accommodating this stream with conventional gas processing approaches is certainly possible, but a simple "disposal" approach to this stream can easily represent significant system performance and cost penalties.

In addition, there are other streams that potentially represent environmental burdens and/or have system-level energy or economic implications, the best examples of which are the coal-like dust from the hot particulate-removal unit and any coal fines that are not fed to the gasifier in some form. The possible incorporation of these streams into the regeneration-gas disposition approach is complementary to the IGCC system design concept and well worth consideration.

The philosophy of the study was to identify techniques that were both commercially attractive and also nearly nationally universal in terms of their potential applicability. As a result, the following effectively became Ground Rules:

- SO_2 reduction to elemental sulfur or gypsum (either of which can be physically disposed of or possibly sold).
- No overt steps to remove ammonia downstream of the gasifier and prior to the GT. (Low NO_x combustion techniques or post-combustion NO_x conversion, if needed, become part of the power producing system.)
- The avoidance of approaches that would be sensitive to the existence of local markets (e.g., the production of sulfuric acid for sale).

In addition, conventional gas processing approaches were really not given much consideration, since previous work showed them to be relatively costly most of the time.

As mentioned in the beginning of this Chapter, an approach that appeared potentially attractive for disposition of the regeneration gas stream was based on using a fluidized bed of hot limestone. This device was found to be capable of being an environmentally benign "trash burner" for the entire system, and as a result, it became the primary approach for the reference system configuration. Energy recovery was accomplished by the unit's ability to provide variable amounts of preheated combustion air to the GT and steam to the system in general. A much less mature but potentially very attractive alternative approach was found to be the DSRP. This process was found to be applicable to the regeneration gas stream, but only under the condition that the stream have a very low free oxygen content. The most credible "backup" to both of these approaches was found to be the Lo-Cat process offered by ARI Technologies that is applicable to the regeneration gas stream only, but is free of restrictions regarding oxygen content. Each of these approaches are discussed in the following Sections.

5.4.1 Limestone PFBC Description

In conventional PFBC, a substantial quantity of fuel is fed to a pressurized- and fluidized- (with air) bed of solids in which combustion occurs. Heat is then extracted from the flue gas, the fluidized bed itself, or both, and work is extracted from the flue gas, normally by an expansion turbine. Also, sulfur-bearing fuels are typically utilized, and calcium containing sorbents such as limestone are added to the bed and have been found to be very effective in lowering SO_2 levels in the flue-gas stream. The sulfur from the fuel becomes calcium-sulfur compounds in the ash (mostly calcium sulfate: gypsum), and the ash is regarded as environmentally benign. This technology, while not mature, is certainly well established.

The approach considered for the limestone PFBC unit of this system is similar but differs in that it is designed to be a minimum size or cost convertor of environmentally significant materials to benign ash or gas, and the heat or work output obtained is of secondary importance. The PFBC bed is limestone fed at a rate that provides the appropriate calcium to sulfur ratio considering the sum of all the feed streams, and the solids withdrawal rate is kept just high enough to maintain a constant bed level. However, since the total ash in the input streams is very low relative to a conventional fluidized-bed combustor, the vast majority of the bed's solids are calcium in some degree of sulfation. As a consequence, conceptually, the calcium-to-sulfur ratio in the bed is high, the solids throughput is low, and the quality of the output gypsum is high, compared to a conventional PFBC.

The primary sulfur source for the limestone PFBC is the regeneration-gas stream, which is in the range of 1000 to 1400 °F and at system operating pressure. An interesting feature of this bed is that its effectiveness is not dependent on calcination of the limestone

(that is, the direct reaction rates of sulfur species with calcium carbonate are essentially equal to those with calcium oxide) (Tulin and Ljungstrom 1989; Snow, Longwell, and Sarofim 1988). However, for the calcium sulfation processes to proceed at useful (and practical) rates, the bed temperature needs to be in the 1500 to 1750 °F range, and this requires that a small amount of "fuel" also be supplied to the bed. In all rational configurations of an IGCC system, particulate (i.e., coal dust) is removed from the gasifier's product-gas stream and must be accommodated by the overall IGCC system. This particulate is considered to be the primary fuel for the PFBC. While a minimum amount of fuel is required by the PFBC to reach temperatures appropriate to effective sulfur capture, functionally there is no maximum amount of fuel that can be tolerated, given additional heat removal capability within the PFBC. For example, fines from the coal handling area and top gas from a two-stage gasifier are both potentially credible as fuel to the PFBC. In addition, a useful characteristic of all PFBCs is that they are quite flexible regarding the fuel form - solids, liquids, and gases are all readily accommodated. The PFBC is operated with an appreciable amount of excess air, typically at least 20%, to ensure that complete chemical conversions occur within the PFBC. The sulfur and nitrogen species present in the fuel become inconsequential because of the PFBC's proven capability to capture sulfur and to release fuel-bound nitrogen as nitrogen gas. The flue gas from the PFBC is oxidizing in nature (a result of the high percentage of excess air) and is routed to the GT combustor to serve as preheated combustion air. While this stream is somewhat depleted in oxygen, it is quite hot relative to the GT compressor-discharge air, and so is useful to the GT's combustion process. However, some form of particulate removal from this stream prior to the GT will be required for compatibility with the GT, but this is nearly conventional technology.

There is a limit to the quantity of preheated combustion air that can be utilized by the GT, primarily because GT compressor-discharge air is needed to serve as a coolant within the GT, there are limitations on combustion product-gas temperatures at the turbine nozzle inlets, or both. Consequently, as the amount of fuel fed to the PFBC is increased, the practical ability to carry away the corresponding amount of heat as preheated GT combustion air can be readily exceeded. At this point, heat transfer surfaces must be added to the PFBC so that the additional energy can be recovered for utilization elsewhere in the system. Typically, this results in steam for use in either the gasification or steam power portions of the system or in preheated air for use in the (tar cracking and desulfurization) catalyst/sorbent regeneration step. While this approach to energy recovery can improve the economic performance of the system, it can reduce overall system efficiency compared to approaches that feed the energy content of the waste-fuel streams to the gasifier or the GT (because the energy contribution is limited to the steam side of the system).

The approach of utilizing a limestone PFBC makes the overall system quite flexible. The PFBC can accommodate the regeneration gas and whatever level of waste fuels are economically desired to produce a variable level of preheated combustion air and steam and benign wastes (gypsum and nitrogen gas). With reasonable care it should also be possible to design an increased load-following capability for the overall IGCC system by feeding more or less waste fuel to the PFBC. (The waste-fuel form has to be storable to accomplish this,

however.) In addition, the approach is not sensitive to the type of hot fuel-gas cleaning catalyst/sorbent that is utilized; the differences between zinc ferrite, zinc titanate, etc., should have a minor impact on the limestone PFBC design. The only restriction would occur if regeneration gas temperatures exceeded levels acceptable for the sulfation reactions (practically speaking, this would be about 1900 °F), which is extremely unlikely. This approach should enhance limestone utilization and lower the required calcium-to-sulfur ratios relative to a conventional PFBC, since the sulfur to be captured (in the regeneration gas) is already in the gas phase, is all introduced at the bottom of the bed, and is volumetrically concentrated.

5.4.2 Limestone PFBC Design

A conceptual design was developed for the limestone PFBC to better assess the viability of the approach and to provide input to the subsequent systems-level analyses. The various options considered for the basic reference system required the design of the PFBC to be capable of accommodating the regeneration gas plus a wide variation in fuel streams. (The specifics of the system configurations are covered in the next Chapter.) In summary, the limestone PFBC approach was found to be viable for all configurations for which it was considered. The following describes the methodology that was developed for sizing the PFBC.

Sizing the PFBC primarily consists of determining the vessel's active cross section and height. A schematic of the general geometry that was considered for the limestone PFBC unit itself and the configuration of this subsystem is shown in Figure 51. The cross-section area of the PFBC is determined from the volume flow and the density of the exiting POC gas stream (i.e., the preheated GT combustion air), the temperature and pressure within the PFBC, and the allowable superficial gas velocity in the fluidized solids bed. The height of the fluidized-bed is dictated by the gas and solid residence-time needed to provide adequate gas contacting and the freeboard space needed to disengage particles from the gas stream.

For these analyses, the PFBC was assumed to be operating at 1700 °F and at a pressure level of either 300 or 600 psia, depending on whether a STAG or a STIG gas turbine/generator set was employed. The PFBC flue-gas flow rate and density were obtained from preliminary versions of the ASPEN computer simulations for each system configuration examined, and were found to be configuration-unique. The acceptable operational superficial gas velocities were dictated by the size and density of the bed material, and were based on data from Grimethorpe Test Series 2.3 (NCB IEA Grimethorpe, Ltd. 1985). This test series used Ohio Plum Run dolomite, and the average bed-material particle sizes were found to be in the range of 0.8 to 1.1 mm. A representative size distribution of the bed material from these tests is shown in Figure 52. Figures 53 and 54 show minimum fluidization and terminal velocities for a range of bed-material particle diameters at pressures of 300 and 600 psia. A reasonable operating superficial gas velocity for the PFBC can be chosen by knowing the bed material sizes and referring to either Figure 53 or 54. The PFBC operating gas velocity must be above the minimum fluidization velocity, but not so high as to result in excessive carry-over of the bed material (i.e., below the terminal velocities of the majority of the bed's

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Figure 51. Schematic of the Limestone-Bed PFBC Subsystem

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Figure 52. Particle-Size Distribution of the Bed Material

Figure 53. PFBC Fluidization Velocities, 300 psia

Figure 54. PFBC Fluidization Velocities, 600 psia

solids). The reality is that a certain percentage of the small diameter material will be entrained in the gas stream from the fluid bed, and this must be accommodated; two stages of cyclones were depicted in Figure 51 for this purpose. The basic trade-off in all this is between the in-bed particle size and the bed's superficial gas velocity, and consequently, the vessel diameter and fuel input.

An example of this trade off is provided by Figure 55, which is a plot of the limestone PFBC bed diameter versus the fuel energy flow to the PFBC for Case 4a (defined in Chapter 6 as one-stage, dry-bottom gasifier/STAG power cycle configuration in which 30% of the plant's total coal input is considered to be fines that are fed directly to the PFBC together with the hot fuel-gas cyclone output dust, representing 5% of the gasifier's coal input). Figure 55 shows the PFBC bed diameters corresponding to fuel energy inputs of up to 600 MBtu/h with PFBC superficial velocities from 2.3 to 3.3 ft/s. Under the assumptions applicable to Case 4a, a coal-fines energy flow of 120 MBtu/h corresponds to a nominal 50-MWe plant output, and data points, consequently, are provided at nominal 50-MWe intervals. The parametric values of superficial gas velocity used in Figure 55 correspond to predicted bed-material carry-over rates of 5 to 20% of the bed (as noted on the figure), using the particle size distribution provided in Figure 52 and the corresponding minimum fluidization and terminal velocities of Figure 53. Figure 55 shows that for a 100-MWe plant, the bed particulate carry-over rates will vary from 5 to 20% as the vessel diameter varies from 13 to 11 ft. In a detailed design, a trade-off study would need to be done to determine the optimal vessel size, considering the economics of vessel size, particle carry-over rate, and the effect on downstream equipment. However, that activity was beyond the scope of this conceptual design.

The results of the sizing calculations done to support the systems analyses are provided in Table 13. The quoted ranges of values cover a bandwidth of particulate carry-over from 5 to 20% for a representative bed particle-size distribution, and thus they are likely to bound what would become the final design after a detailed optimization study. Multiple vessels of identical diameters are indicated for several of the system configurations. This approach was selected in the belief that a maximum diameter on the order of 15 feet would also be preferred for the PFBC. These results were normalized, based on the corresponding plant's net power output and using an average superficial velocity (2.8 ft/s for the STAG system and 2.0 ft/s for the STIG system). The resulting specific diameters for the various system configurations are provided in Figures 56 and 57 for STAG- and STIG-based system configurations. The effect on the required PFBC size (i.e., cost) of consuming more or less waste fuel can easily be seen by comparing specific diameter values. The Case 7 system configuration (a slagging gasifier burning only enough fuel to heat the PFBC to 1700 °F) clearly results in the smallest PFBC per unit of system power output.

Determination of the PFBC vessel height is primarily based on two factors: the depth of sorbent bed needed for adequate gas and solid contacting, and the height above the top of the bed needed to disengage particles from the flue-gas stream in order to limit carry-over. Design of the sorbent bed depth needed in the PFBC was also based on Grimethorpe

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Figure 55. PFBC Diameters for Case 4a

Table 13. PFBC Sizing Requirements

General Characteristics			STAG Systems	STIG Systems
- Operational pressure, psia			300	600
- Gas superficial velocity, ft/s			2.3 to 3.3	1.7 to 2.3
- Particulate carry-over rate, %			5 to 20	5 to 20
- Sorbent bed depth, ft			8 to 11.5	6 to 8
- Total active vessel height, ft			23 to 32	17 to 25

<i>System Configuration-Specific Parameters</i>				
Case No.	Plant Characteristics		Required Sizes of PFBC Units	
	Type	Size, MWe	No. of Units	Diameter, ft
1	STIG	60	1	9 to 10
1a	STIG	51	1	12.5 to 15
1b	STIG	77	1	9 to 10
2	STAG	341	3	11.4 to 13.5
3	STIG	55	1	3.5 to 4.5
4	STAG	250	1	10 to 12
4a	STAG	300	2	13 to 16
4b	STAG	252	1	10 to 12
6	STAG	284	1	8.5 to 10.5
7	STAG	253	1	6 to 7

information (NCB IEA Grimethorpe, Ltd. 1985). The gas residence time in the Grimethorpe tests was varied from 1 to 3.5 s and whenever the gas residence time exceeded 3.4 s with a calcium-to-sulfur feed ratio of 1.6 or greater, the in-bed sulfur capture exceeded 90%. (A few of these run periods showed sulfur retention values as high as 99%.) The Grimethorpe unit was fed solid coal, and in contrast, the preponderance of the sulfur fed to the PFBC will already be in the gas phase and will be introduced at the bottom of the bed; consequently, an increased sulfur capture effectiveness should be expected in the PFBC (with the same gross operational parameters). A gas residence time of 3.5 s was selected (since this was estimated to be enough to achieve a sulfur capture effectiveness well above 90%), and the corresponding sorbent bed depth required to achieve this residence time was calculated.

For fluidized-beds, the TDH is the height above the fluidized-bed at which the population of entrained particulate becomes essentially constant. In reality, the gas exit (or cyclone inlet) need not be located higher than the TDH, and considering the consequences of

Figure 56. STAG-Cycle PFBC Sizing Requirements

Figure 57. STIG-Cycle PFBC Sizing Requirements

unnecessary particle carry-over, the gas exit height is normally not located much below it. For simplicity, the bed's gas exit was taken to be at the TDH, and values for TDH were determined based on the empirical work of Zenz and Weil (1958). Their correlation provided TDH values that ranged from 15 to 21 ft for the 300-psia STAG configurations and from 10.8 to 16.7 ft for the 600-psia STIG-based systems.

For a fixed particle size distribution and a selected particulate carry-over rate, the PFBC bed cross-section area will vary, depending on

- The PFBC energy input (which also sizes the amount of GT compressor discharge air that is input to the PFBC),
- The zinc-ferrite regeneration-gas volume, and
- The operational pressure.

Similarly, the superficial gas velocity becomes only a function of the operating pressure, and given a constant gas residence time of 3.5 s, the sorbent-bed height also becomes only a function of the operating pressure. TDH varies with superficial gas velocity and vessel diameter; consequently, systems operated at 300 psia will have an active vessel height of between 23 and 32 ft and systems operated at 600 psia will have a vessel height of between 17 and 25 ft. For the systems analyses discussed in Chapter 6, the PFBC vessels were sized for each individual case as results of the preliminary ASPEN simulations became available.

5.4.3 Direct Sulfur Recovery Process

The DSRP has been developed by RTI under contract to U.S. Department of Energy. The DSRP is a high-temperature, high-pressure, process for the recovery of elemental sulfur from either oxidizing or reducing gases (Dorchak, Gangwal, and Harkins 1990). This process is much less mature than the limestone PFBC approach discussed above, but it promises much better performance in an IGCC application (and probably is suitable for other applications as well) and was felt worthy of evaluation. This process is currently being developed.

It is claimed that a two-stage DSRP is capable of recovering over 99% of the SO_2 in a regeneration-gas feed stream. In a conventional gas treatment approach, this level of performance would eliminate the need for a Claus plant and a tail-gas treatment step for the final offgas. The sulfur recovery of the two-stage DSRP is not only comparable to the performance of the so called "Superclaus" process recently developed in Germany ("Superclaus Increase Sulfur Recovery" 1988) but also has several advantages. The major advantage of the DSRP over the "Superclaus" (and all other existing elemental sulfur recovery processes) is the inherent match in operating conditions that provides an easy integration with the hot gas cleanup portions of the IGCC plant. It also performs well with a gas feed stream containing dilute sulfur levels (as low as 0.85% by volume) without a requirement for a sulfur pre-concentration stage, such as a Wellman-Lord process. The capital and operating costs for the two-stage DSRP also appear to be low compared to costs of the well-known elemental sulfur

recovery processes, such as Beavon/Stretford, Wellman-Lord/Resox, and Wellman-Lord/Augmented Claus. Unlike the limestone PFBC approach, the DSRP does not produce any solid waste. However, it also does not offer a means to directly utilize some of the waste-fuel streams within the plant, and its attractiveness is sensitive to the quantity of oxygen in the regeneration gas stream.

The two-stage DSRP was selected for this evaluation, and a schematic of how it might be integrated in an IGCC configuration is provided in Figure 58. The DSRP is shown in the right half of the figure, and it consists mainly of two catalytic reactors, two condensers, and a heat exchanger. The primary feed stream to the first-stage reactor consists of the SO₂-containing regeneration gas and a secondary reducing-gas stream (e.g., cleaned gasifier product gas). The secondary reducing-gas stream is best taken from the tar cracking and desulfurization unit and typically is approximately 3 to 5% of the volumetric flow of the coal gas, depending primarily on the oxygen content in the regeneration gas. This is sufficient to allow essentially complete conversion of the SO₂ in the primary gas stream to elemental sulfur. Sulfur conversions of over 92% have been obtained at a temperature of 1025 °F and a pressure of 270 psia in the first stage. The product offgas of the first stage is cooled to collect elemental sulfur and remove a small amount of water from the gas phase. The condenser effluent gas is then reheated in a heat exchanger and is fed to the second stage without additional reducing gas. The remaining SO₂ and reducing agents are continuously converted to elemental sulfur at a temperature of about 400 °F and a pressure of 270 psia in the second stage; over 98% conversion efficiency has been achieved in this stage. The final offgas is again cooled to around 300 °F to collect elemental sulfur and an additional small amount of water produced by the chemical reactions. The resultant sulfur should be of good enough quality to be entirely suitable for sale or disposal by landfilling, if required. The final effluent gas is at essentially GT-combustor pressure level, but is very cool and nearly inert. It is routed to the GT and utilized for internal cooling prior to being combined and expanded with the main GT POC gas stream. Many other gas disposition options are credible because of the general benignancy of the final effluent gas. However, this option was selected because of its probable enhancement of the overall system efficiency.

The chemistry of the sulfur conversion process involves several direct and intermediary reactions at or near the catalyst surface. These reactions may be summarized as follows:

The yield of elemental sulfur increases with increasing temperature between 750 °F (400 °C) and 1300 °F (700 °C) and with increasing pressure from atmospheric to 40 atm, at space velocities in the range of 4000 h⁻¹ to 5000 h⁻¹. Sulfur recovery levels of 92 to 95% have been

Assumptions:

1. Essentially all sulfur will be condensed as a liquid at 300°F.
2. DSRP is at system pressure (20-40 atm)

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Figure 58. Schematic of the Direct Sulfur Recovery Process

experimentally observed at these conditions in single-stage test units. The sulfur yield is further increased at lower space velocities; for example, at 2000 h^{-1} , over 99% sulfur recovery has been achieved in a single-stage test unit at a temperature as low as 480°F (248°C). The specific design and the resulting performance of the DSRP will primarily depend upon the gas compositions of the regeneration and reducing gas-feed streams and the catalytic-reactor operating conditions (i.e., temperature, pressure, space velocity, and stoichiometry).

The DSRP approach was incorporated in Cases 10 and 11 of the systems analysis activity. It was found to substantially lower sulfur emissions but to also slightly reduce overall system efficiency (because of the consumption of a fraction of the product-fuel gas). The need for reducing gas will have a strong negative influence on the desirability of the DSRP approach if the oxygen content of the regeneration gas stream cannot be made small (i.e., using near-stoichiometric regeneration). The technology is currently too immature for cost estimates to be meaningful; however, it was projected that capital costs would be very slightly above, and operating costs would be slightly below, those for a comparable IGCC configuration utilizing a limestone PFBC.

5.4.4 Lo-Cat Process

The viability of the reference IGCC system is absolutely dependent upon a means of disposing of the regeneration gas in an adequate manner. The limestone PFBC approach, which was selected as the baseline approach, is an unconventional application and an integration of reasonably proven technology. The DSRP approach, while quite promising, is only at the laboratory scale of development and could impose an unacceptable system-efficiency penalty. As a consequence, it was deemed advisable to identify and examine a fallback approach, or the alternative means appearing to be the most attractive of the conventional approaches to regeneration gas processing. The key to this approach turned out to be the Lo-Cat process marketed by ARI Technologies.

A survey of conventional approaches to regeneration gas treatment based on fixed-bed regeneration concluded the following three specific configurations had the highest potential merit (O'Hara, Chow, and Findley 1987):

- Wellman-Lord unit (or similar SO_2 -concentration step) followed by an Augmented Claus unit and an incineration unit. The offgas was to be recycled to the Wellman-Lord unit.
- Augmented Claus unit followed by a Beavon Sulfur Recovery Process (BSRP) unit to treat the Claus tail gas before venting it to the atmosphere.
- Wellman-Lord unit in place of the Augmented Claus Unit in the second configuration above.

All three configurations were approximately twice the capital cost and about triple the operating costs projected for the DSRP approach; however, they are all essentially guaranteed to work and all have successful commercial experience bases. These configurations also have the capability to accommodate the time-varying characteristics typical of the regeneration gas from a fixed-bed zinc-ferrite regenerator. Since the reference system configuration utilizes fluidized-bed regeneration with inherently near constant output-gas characteristics, it might be possible to simplify the above gas treatment configurations somewhat, but the cost margins are large enough that the DSRP approach is likely to still be favored. (However, this re-analysis was not performed.)

Re-examination of the three configurations in the context of treating gas from fluidized-bed regeneration led to a conclusion that the BSRP unit, by itself, is probably adequate. The BSRP combines a catalytic reactor unit for hydrogenation/hydrolysis with a conventional Stretford unit for elemental sulfur recovery. However, the capital and operating costs of the Stretford unit constitute more than 85% of the total cost of the BSRP. In addition, the Stretford process is known to have many inherent areas of difficulty such as purge disposal and sulfur product contamination. Because of this and based on past work, an alternative sulfur recovery process was investigated as a possible substitute for the Stretford unit. This alternative, the Lo-Cat process (designed and licensed by ARI Technologies, Inc.), was found to be a potentially attractive approach. A study (Pack 1984) showed that both capital and operating costs of the Lo-Cat process were significantly lower than the cost of the Stretford process. In addition, the inherent characteristics of the Lo-Cat process also eliminated most of the operational problems encountered with the Stretford process. A summary comparison of the gross characteristics of the two approaches is provided in Table 14.

The referenced study contained comparative cost estimates for a 25-MWe power station. The cited capital costs for the Stretford units considered ranged from \$6 to \$7.5 million and the capital cost for the Lo-Cat unit was \$3.5 million. Using conventional process cost-scaling factors, the corresponding fixed capital-investment cost for the regeneration-gas disposition portion of a 100-MWe IGCC system was estimated at \$74.50/kWe for a BSRP with a Stretford unit, and \$64.2/kWe for the same basic configuration with a Lo-Cat unit for sulfur recovery. These values compare to an estimate of \$38.81/kWe for the DSRP by RTI.

It is reported that 20 Lo-Cat units had been sold by ARI up to 1984; about half of these were in natural-gas purification applications, one was used to clean non-condensable gases in an oil-shale application, and the remaining units were sited in refinery and chemical plant applications. (Additional units have been sold since then but the specifics were not determined.) The operating experiences have been positive.

The result of this investigation is a conclusion that the use of the BSRP approach with the Lo-Cat unit as a replacement for the Stretford unit will provide an effective, essentially state-of-the-art, conventional gas treatment and disposition process for the regeneration gas. This system will capture in excess of 99% of the potential sulfurous air emissions. While it

Table 14. Comparison Between BSRP/Stretford and BSRP/Lo-Cat Subsystems

BSRP/Stretford	BSRP/Lo-Cat
Primary Chemistry:	
$\text{CO} + \text{H}_2\text{O} \rightarrow \text{CO}_2 + \text{H}_2$	$\text{CO} + \text{H}_2\text{O} \rightarrow \text{CO}_2 + \text{H}_2$
$3\text{H}_2 + \text{SO}_2 \rightarrow \text{H}_2\text{S} + 2\text{H}_2\text{O}$	$3\text{H}_2 + \text{SO}_2 \rightarrow \text{H}_2\text{S} + 2\text{H}_2\text{O}$
$\text{COS} + \text{H}_2\text{O} \rightarrow \text{H}_2\text{S} + \text{CO}_2$	$\text{COS} + \text{H}_2\text{OS} \rightarrow 2\text{S} + \text{CO}_2$
$\text{CS}_2 + 2\text{H}_2\text{O} \rightarrow 2\text{H}_2\text{S} + \text{CO}_2$	$\text{CS}_2 + 2\text{H}_2\text{O} \rightarrow 2\text{H}_2\text{S} + \text{CO}_2$
$\text{H}_2\text{S} + 1/2 \text{O}_2 \rightarrow \text{H}_2\text{O} + \text{S}$	$2\text{H}_2\text{S} + \text{O}_2 + 2\text{H}_2\text{O} \rightarrow 2\text{S} + 4\text{H}^+ + 4\text{OH}^-$
Pressure: Near atmospheric	Low pressure (elevated pressures are a future possibility)
Temperature: 500-700 °F for catalytic unit 85-115 °F for Stretford unit	500-700 °F for catalytic unit 100-130 °F for Lo-Cat unit
Operating Media: Syngas, catalysts	Syngas, catalysts, high pH solution (8-9.5)
Sulfur Conversion: > 99.5%	> 99%
Configuration: A catalytic reactor plus a Stretford unit	A catalytic reactor plus a Lo-Cat unit

is higher in cost than the DSRP or PFBC approaches, it is lower in cost than the traditional, conventional gas-treatment alternatives. In addition, the cost penalty was not felt to be intolerable compared to the DSRP approach (especially when remembering that the DSRP costs are based on immature technology), and the penalty is only slightly increased when compared to the limestone PFBC approach. Hence, the BSRP/Lo-Cat approach is considered a viable fallback technology for disposition of the zinc-ferrite regeneration gas.

5.5 POWER SYSTEM

The power system for the CGIA study initiates with the gas fuel-flow control valve to the GT and ends with the flue-gas stack and cooling tower discharging their respective portions of waste gases and heat from the power system (given a combined cycle configuration). The major components are the GT, the heat recovery steam generator (HRSG), the steam turbine (ST), and any emissions control equipment that may reside within the power system. Two types of power systems were considered, the more common stream and gas turbine combined cycle (STAG) system, and the stream injected gas turbine (STIG) system. The STIG system is a relatively newer approach; it deletes the ST and all related components

downstream of the HRSG and utilizes the HRSG primarily to raise steam for injection to the turbine section of the GT unit. This approach tends to be advantageous only when applied to a system configuration utilizing a high pressure ratio GT (one whose compressor discharge pressure is in the 30 to 35 atm range) and having appreciable other uses for the steam, such as the steam stream to the fixed-bed gasifier. The effect of the STIG approach is to produce much more power from the GT (and increase its efficiency) and to delete the cost of the ST portions of the normal STAG cycle. The contemporary GTs to which the STIG approach can be profitably applied are aircraft-derivative machines and, consequently, are limited to power levels under 75 MWe. At these lower power levels, the STIG approach results in more output power per unit area of plant, generally higher efficiency, and frequently a lower COE than would result from using a comparably sized STAG system (or any other contemporary form of power generation).

In general, the power system components are relatively mature technically, and this report does not contain an extensive discussion of turbine technology. However, there are two design issues that need to be understood: (1) how is a sizeable fraction of the compressor discharge air extracted from the machine and subsequently returned as low-Btu fuel gas and perhaps as preheated combustion air, and (2) how are the nitrogen-bearing species (typically ammonia) in the fuel accommodated such that very low levels of NO_x emissions are obtained for low cost and performance penalties? The reduction of potential NO_x emissions is essentially the only environmental issue to be addressed by this portion of the IGCC system; however, this is also the least mature aspect of the power system. There are other areas of concern, such as fuel-flow control-valve technology for fuel gas at temperatures of 1000 to 1200 °F (and above), and obtaining orderly and managed combustion of the low Btu fuel gas; however, these are regarded as areas of lesser concern.

5.5.1 Air Extraction/Gas Return

In the IGCC approach, typically 10 to 15% of the compressor discharge air is extracted to supply the gasifier with blast air and is then returned to the GT as fuel gas. In some of the more novel system configurations wherein extra fuel is consumed in a PFBC, as much as 40% of the compressor discharge is extracted for use external to the GT (e.g., Cases 2 and 4a in Chapter 6). The issues relating to the design of supply and return air ducting are relatively obvious; however, two related issues are more subtle and potentially very significant: (1) the GT in an IGCC system will be a modification of an existing natural gas- or oil-fired design (as opposed to a GT designed for IGCC use), and (2) a sizeable fraction of compressor discharge air is normally used within the GT for overt cooling or related functions (e.g., boundary layer control in combustors and transition pieces). The need for modifications to an existing design is not usually a significant cost factor (and, in fact, the modifications are usually minor in nature), but this characteristic limits the magnitude of changes that can be accomplished practically. The second issue recognizes that cooling functions within a GT are critical and to the degree the cooling ability is compromised, the GT performance and life are compromised. Other potential issues have been discussed in the past

(e.g., the mismatch of the mass flow through the turbine and compressor), and engineering solutions have been found to all of them. In summary, if the cost of the modifications to extract air and return gas are tolerable, and if enough high pressure air is left within the machine for cooling (or a substitute gas such as DSRP offgas is found to be feasible), then the particular candidate GT is applicable to the IGCC system approach.

There are two basic GT configurations in use today: the in-line combustor design, and the silo combustor design. In the former, the combustion process takes place in multiple cans (or an annulus) that are arrayed concentric with the GT centerline and closely fitted to the machine (i.e., the machine is shipped and installed as a one-unit structure). In the silo concept, the rotating machinery is located immediately adjacent to one or two silo-like combustors, there are appropriate interfaces for air flow to the combustor, and POC flow from the combustor. The GT is treated as a two- or three-unit subsystem with the rotating compressor/turbine being one unit and the silo combustor(s) being the other unit(s). The former design approach is typical of GE and Westinghouse GTs, and the latter approach is typical of GTs designed by Asea Brown Boveri and KWU. Technically, it appears more straightforward to integrate with the silo combustor designs, but either is useable. The system configurations wherein a large energy flow returns to the GT from the PFBC may, however, not be practical for the designs with in-line combustors. The decision on GT configuration is really based on cost and the related performance characteristics of the resulting IGCC system. A conclusion on which approach is best is not obvious at this point, and is best left to the GT suppliers in any event.

There is an additional opportunity offered in the event the DSRP can be incorporated in the IGCC system configuration. The DSRP approach results in a cool, high-pressure gas stream that is nearly chemically inert and could be an effective coolant gas for the GT. This is a potentially effective integration of gas flows throughout the system, but its viability depends on whether the DSRP is successfully developed and whether the oxygen content of the regeneration offgas (which is the DSRP feedstock) is low enough that the DSRP economics remain attractive. This approach is likely worth consideration for advanced IGCC system approaches.

5.5.2 Environmental Considerations

In the simplified IGCC concept, the hot-gas-cleanup portion of the IGCC system reduces sulfur species and particulate matter to levels well below both current and projected environmental limitations prior to the fuel gas reaching the GT; however, quantities of nitrogenous species are essentially unaltered (unless reduced within the gasifier itself). Research work on the catalytic reduction of ammonia and related compounds is currently ongoing, but in the context of the CGIA study, this should be looked at as a potential route to future system improvement. As a consequence, the subject of environmental controls within the power system primarily evolves to means for NO_x emission control, given a non-trivial concentration

of nitrogenous species in the fuel gas supplied to the GT; a representative concentration value would be 4500 parts per million by volume (ppmv).

Currently, acceptable NO_x emission limits are being tightened, and a trend has also developed toward project-specific limit setting that complicates the question of how low is low enough? There are two basic approaches to dealing with fuel-bound nitrogen (FBN) consistent with the CGIA Ground Rules: minor modifications to the combustion process, and flue-gas treatment downstream of the GT. One other approach, that of aggressively staging the combustion process, may soon be found to be effective (and thus will become applicable to these types of IGCC systems), but the hardware impact and development requirements were felt to be currently significant enough that physically staging the GT combustion process was not considered for this study. The minor modifications are typically the use of steam injection to the combustion process, or the use of special fuel injectors to create local reducing zones within the combusting gases followed by oxidizing regions. Both techniques have been applied to GT combustors burning fuels with varying amounts of FBN, and both have been found to show significant reduction of NO_x emissions from what would have otherwise been obtained. Thermal NO_x suppression is thought to result from limitations of both the peak combustion temperatures and the duration of the gases' exposure to those temperatures.

While the creation of local reducing zones is credited with suppression of thermal NO_x as a result of lowering peak-gas temperatures in the combustor, this approach is also thought to reduce the fraction of FBN that is ultimately converted to NO_x . It should be noted that fuel gas from a fixed-bed gasifier as operated in a simplified IGCC system will inherently contain a considerable steam fraction, and thus, a significant level of steam injection will occur without overt action.

There are two potentially applicable flue-gas treatment approaches: selective catalytic reduction (SCR) and reburning. SCR is a catalytic chemical reduction process that is state of the art. It utilizes ammonia, which is injected as a reactant into the flue-gas stream at a point where the gas is about 700 to 800 °F. The gas then passes through a catalyst section, and nitrogen gas and water vapor result. Implementation of this approach is by incorporating an "SCR bay" within the HRSG. Ammonia is a process consumable and the catalyst sections must be replaced on an every-few-years basis. The process has been successfully applied to both coal-fired boiler and GT flue-gas streams, and NO_x reductions of 90% have been achieved, but the process is regarded as expensive. Reburning is a process whereby a fuel can be injected into the flue-gas stream and burned to create a local reducing zone and temperatures in the vicinity of 1500 to 1700 °F. Reburning has only been applied to coal-fired boilers (typically within the boiler, either among or just ahead of the convection passes) and using a fuel that is free of FBN (e.g., natural gas). It has achieved 50 to 70% NO_x reductions for comparatively reasonable costs. In the last few years, data have been developed that show reburning could also be effective when the reburn fuel contains FBN, and one of the second-round Clean Coal project awards was to demonstrate reburning in a boiler using coal as the reburn fuel. This same approach may be potentially applicable to an IGCC system, and a reburn concept based on this approach is discussed in the following Section.

An additional potential environmental issue is the recently adopted Air Toxic Standards, which are part of the Clean Air Act. These standards pertain to the quantity of specifically defined elements that could be released from the feed coal and find their way through the process to be released as components of the flue gas emanating from the HRSG. An assessment identified (in order of significance) chlorine, lead, mercury, cadmium, and perhaps arsenic as the elements of most potential concern because of their concentrations in the raw product gas. However, it should be noted that mercury and cadmium are not found in significant quantities in the lower rank coals, and only by considering the high extremes of the concentrations reported for some Illinois coals, do these two elements become of potential concern. Similarly, arsenic levels vary widely with coal type and will only be a potential concern with some specific coals. The low product-gas temperature (relative to other processes for coal conversion) mitigates the quantities of these elements that are presented to the GT. A fraction of the volatilized elements will be sorbed by the dust particles and removed by the particle removal process, but some will pass on to the GT and will ultimately appear in the flue-gas. These elements are expected to exit in the form of trace levels of HCl and metal oxides (or chlorides).

A study by SRI International (Krishan et al. 1986) examined credible routes to chloride removal from coal gas streams. They conducted extensive experimental studies and found that a bed of Nahcolite was capable of lowering chloride levels to 1 ppm in coal gas at 1000 to 1200 °F. Conceptually, the bed would be placed upstream of the fuel control valve and would utilize disposable Nahcolite media. The projected capital and operating costs were low to the point of being negligible on a powerplant basis.

5.5.3 Reburning for NO_x Emission Reduction

A reburning system applied to a coal-fired boiler consists of three basic zones that occur sequentially in the gas path and are defined by air-to-fuel stoichiometric ratios SR_1 , SR_2 , and SR_3 :

- Primary zone: usually fuel-lean, an SR_1 of about 1.10 is typical.
- Reburn zone: located downstream of the radiant section of the boiler and must be fuel-rich, the optimum stoichiometric ratio SR_2 is in the range of 0.8 to 0.9.
- Burnout zone: overall fuel-lean, the typical SR_3 is about 1.25.

The general parameters that define the primary zone do not seem to have a tremendous effect on the overall reburn efficiency. One possible exception is the stoichiometry of the primary stage. Obviously, the larger the quantity of excess air in the first stage, the more reburn fuel will be required to attain the required SR_2 .

In the reburn zone, the best results seem to occur with an SR_2 of about 0.8 to 0.9, although a couple of investigators have gotten good results with SR_2 values as high as 0.95 and as low as 0.75. The amount of NO_x reduction is in the general range of 50 to 70%. One of the biggest problems with using coal as the reburn fuel instead of a clean hydrocarbon such as propane or natural gas is the potential of its own fuel-bound nitrogen contributing to the total NO_x emissions (Chen et al. 1983). This is particularly important when the NO_x emission level from the primary zone is relatively low (below 600 ppm). Under these conditions, the NO_x reduction percentage is significantly reduced (Greene et al. 1986). The total reburning-zone residence time could be as short as 0.4 s, but times in the range of 0.75 to 1.0 s are generally preferred. With regard to reburn-zone temperature, there appears currently to be no clear consensus as to what is the optimum. Most of the testing has been done with natural gas as the reburn fuel, and the data show NO_x reduction generally improves with higher reburn-zone temperatures, up to a limit of about 2700 °F, where thermal NO_x formation starts to become significant (Chen et al. 1988). There is evidence that use of pulverized coal as the reburn-zone fuel reduces the preferred reburn-zone temperature to the 1800 to 2200 °F range in conventional boilers. In tests with Beulah lignite as the reburn fuel, inlet NO_x levels of 630 ppm were reduced to below 100 ppm, and it was observed that the reductions were effective down to reburn-zone temperatures as low as 1700 °F. One credible explanation is that since this coal generates significant amounts of NH_3 under fuel-rich conditions, there is an in-situ "de- NO_x " process occurring that results in significantly lower NO_x emission values.

The burnout zone is fuel-lean, with typical SR_3 stoichiometry in the range of 1.2 to 1.25. The burnout air injection occurs at reburn-zone exit conditions, and the quantity of air injected is chosen to limit the temperature rise of the flue gas. Lower combustion temperatures prevent the formation of thermal NO_x and encourage the conversion of fixed nitrogen species to N_2 rather than NO. In addition, if significant quantities of ammonia are present, such as from the pyrolysis of Beulah lignite, the in-situ de- NO_x process can be expected to reduce NO_x levels even further. The sequence of reactions $NH_3 \rightarrow NH_2 + H$ and $NO + NH_2 \rightarrow N_2 + H_2O$ are important to this concept and have a fairly narrow temperature range (i.e., a few hundred degrees F) over which they occur at useful rates. At higher temperatures, ammonia converts to NO_x and at lower temperatures, sufficient ammonia decomposition does not occur, which limits the subsequent NO reduction. However, Chen and coworkers (1986) showed that at low concentrations of CO (i.e., in the range of 200 to 10,000 ppm) in the burnout zone, the decomposition of NH_3 is enhanced at temperatures as low as 1500 °F, and consequently, this condition can result in low NO_x emissions at low burnout-zone temperatures.

Conceptually, it is possible to add pulverized coal burners in the gas path ahead of a more-or-less conventional HRSG to form a reburn bay. The intent is to modify a normal fired-HRSG design to provide reburn and burnout zones so as to achieve significant flue-gas NO_x reduction. Gas residence times in typical HRSG designs are in the 0.4 to 2.0 s range, and CO concentrations in the inlet gas are expected to be quite low (10 ppm would be representative), so the chemistry is likely to be effective. It should be recognized that this

approach will require more fuel on a per-pound-of-flue-gas basis than is utilized in a boiler application, because of the lower temperature level and the larger oxygen concentration of the inlet flue-gas stream. In addition, in the IGCC application the fuel could introduce an unacceptable SO_2 emission. The incentive for this approach is that the fuel could be of little real value (e.g., coal fines), and its consumption would greatly reduce NO_x levels while providing a high quality energy stream that is directly utilized by the steam portion of the combined cycle. In summary, it appears the gas chemistry can be made to work; the important considerations are likely to be the additional amount of sulfur emissions, the possible impact on overall system efficiency, and the cost of this approach relative to the alternatives (e.g., fines briquetting and multi-swirl fuel injectors in GT combustors).

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Chapter 6

Comparative Analyses of Prototype Systems

This Chapter describes the results of the analyses done on the reference Prototype System and the various system-level options. This system and the various options were defined in Chapter 5 and are conceptually depicted in Figure 28 on page 128. The original intent of these analyses was to determine whether any particular gasifier configuration or design feature appeared exceptionally meritorious based on the resulting potential cost or performance advantage when incorporated into a Simplified IGCC system. The initial approach was to develop system-level performance and projected cost parameters for various gasifier options using a common reference frame in the form of a potentially desirable and only slightly advanced system configuration. The types of characterizations planned were based on the prototype gasifiers and included the following:

- One-stage prototype gasifier with and without a slagging bottom.
- Two-stage prototype gasifier with and without top-gas recycle.
- Impact of the DSRP versus the limestone PFBC.
- Performance with a conventional, air-blown, dry-bottom, Lurgi gasifier.
- Impact of briquetting for fines accommodation.
- Impact of overpressuring for increased gasifier throughput.
- Impact of coals other than the reference coal.

The planned systems analysis activities were far more time consuming than anticipated. Consequently, the first three items in the above list were completed as were portions of the fourth and fifth items, but the last two items were essentially undone.

The results presented in this Chapter are complete enough to provide a basis for valuable insights, but it is recognized they may stimulate more questions than provide answers. The thoughtful reader of this Chapter will also find a number of smaller technical areas that prompt unanswered questions, especially when the material presented in Chapters 4 and 5 is considered. The nature of the CGIA study was not to answer all relevant questions but to provide a status, to define and quantify some of the potentialities, and to point the way to productive future analysis and testing - and this it has accomplished.

6.1 ANALYTICAL PROCESS DESCRIPTION

The primary tool used in these analyses was the ASPEN computer program (Stearns Catalytic Corporation 1984), which provided a mathematical simulation of each selected total IGCC system configuration. The ASPEN simulation provided system-level performance parameters and the heat and material flow values for the integrated system configuration. The

resulting heat and material flow values were the basis for specific vessel or component sizing and costing. Capital costing was developed by "constructing" the Nth plant at a generic U.S. site. This approach was taken as a means to facilitate comparisons between system configurations without the potential confusion that can result from items such as differing levels of contingency cost allowance (i.e., current maturity levels of the involved technologies). See Ground Rules 18 and 19 in Chapter 2 for more discussion. This information was then utilized to estimate O&M costs and, from this, the COE.

Initially, 12 potentially desirable, specific system configurations were selected for analysis, and a set of general assumptions was established to define how the systems would be analyzed. The significant assumptions are delineated in Table 15. Of the 12 original systems, one was found to have major technical problem areas, and three others were deleted during the analytical process for lack of potential or because they were partially redundant. However, additional sub-options were computed for two of the configurations, coincidentally producing 12 sets of results. A summary description of the eight system configurations and the 12 total cases analyzed is provided in Table 16. The Case Numbers of the original 12 system configurations were retained to facilitate record keeping by the participants in the study, so the Case Number sequencing is non-monotonic. The abbreviated system descriptors in Table 16 define the important variations between the system configurations and the deviations from the general assumptions list given in Table 15.

Considerable effort was spent to make the analyzed configurations as technically credible as possible, but several will be recognized as suffering from a lack of substantive supporting data. The most uncertain areas of system-level significance relate to the computed temperatures and assumed dust content of the gasifier's output-gas streams for the more advanced gasifiers. Possibly the largest single technical gap is an estimate of the maximum throughput capability for the prototype gasifier configurations. Similarly, the estimates relating to stoichiometric zinc-ferrite regeneration and the assumed ability to extract and readmit substantial quantities of preheated combustion air to the GT will need to be validated. Nonetheless, the breadth of the spectrum of cases analyzed provides a basis for insights in many areas.

The detailed assumptions and departures from the general assumptions of Table 15 pertinent to each of the eight system configurations analyzed are provided in Appendix A. Because of the volume of detail, the presentation may be somewhat cryptic, but it should nonetheless be understandable to readers familiar with these types of analyses. In many instances, the information presented is redundant to the information in Tables 15 and 16, but this was done for the sake of completeness and was intentional. For reference, a diagram of each of the analyzed configurations is provided at the end of Chapter 6, and these diagrams illustrate the detailed results of the analyses.

Table 15. General Assumptions for the System Analyses

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1. The reference coal was utilized on a ROM basis, except for Cases 3, 4, 4a, 4b, and 10, which utilized a sized reference coal.
 2. The mathematical models representing the gasifier were as defined in Chapter 4. ASPEN utilized the results of those models (e.g., the gas outlet temperature[s], yields, and compositions) in the system representations. The exception was the one-stage, dry-bottom gasifier model, which was based almost exclusively on measured data from METC's gasifier.
 3. The mathematical models representing the tar cracking/desulfurization, catalyst/sorbent media regeneration, PFBC, and DSRP units were as described in Chapter 5. ASPEN utilized the results of those models (e.g., educting fluid requirements, heat release rates, gas compositions, sulfur capture) in the system representations. Some specific parameters are cited in the detailed assumptions in Appendix A.
 4. The fluid-bed tar-cracking/desulfurizer unit leaves 100 ppm of sulfur in the exit gas, and the ZnFe sorbent is 40% utilized. A tar-cracking catalyst is added to the ZnFe sulfur sorbent media for all systems wherein a product gas containing tar is routed to the desulfurization unit. The tar cracking catalyst is 100% effective.
 5. The fluid-bed tar-cracking/desulfurizer media (including ZnFe) is totally regenerated (sulfides decomposed and carbon burned off) in a fluidized-bed regenerator at 1400 °F with stoichiometric amounts of air.
 6. The fluid-bed tar-cracking/desulfurizer media lift pipe is driven by .085 lb of steam/lb of sorbent.
 7. The PFBC is operated with 20% excess air; solids are fed to the PFBC at a calcium-to-sulfur ratio of 1.5; the PFBC achieves 96% sulfur capture.
 8. All fuel-bound nitrogen entering the PFBC is completely reduced, irrespective of the parent fuel form, and exits as nitrogen gas in the PFBC flue-gas stream.
 9. The STAG power system model is based on an MS7000-F GT using a 2300 °F firing temperature, a 1450 psia/1000 °F/1000 °F ST, and a 284 °F stack-gas temperature. The STIG system model is based on an LM5000 using a 2200 °F firing temperature.
 10. Of the fuel-bound nitrogen delivered to the GT, 40% is converted to NO during combustion, and the NO₂ emission levels are calculated based on oxidation of the NO.
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Table 16. Summary Definitions of System Cases

Case No.	Gasifier/System Description
1	Two-stage, dry-bottom, 40-atm gasifier; PFBC fueled by top-gas plus 5% of the coal entering the gasifier (representing entrained dust carry-over); STIG power cycle
1a	Same as Case No. 1 but all GT combustion air is passed through the PFBC (resulting in 385% excess air) for preheating
1b	Same as Case No. 1 but a steam turbine is added, with steam supplied by the PFBC and HRSG
2	Two-stage, dry-bottom, 20-atm gasifier; PFBC fueled by the raw top-gas plus 5% of the coal entering the gasifier (dust carry-over); STAG power cycle
3	One-stage, dry-bottom, 40-atm gasifier; plant buys sized coal; PFBC fueled by 5% of the coal entering the gasifier (dust carry-over); STIG power cycle
4	One-stage, dry-bottom, 20-atm gasifier; plant buys sized coal; PFBC operating at a Ca/S ratio of 2, with 50% excess air, and is fueled by 5% of the coal entering the gasifier (dust carry-over); STAG power cycle
4a	Same as Case No. 4 but unsized coal is delivered to the plant and 30% of this, plus 5% of the gasifier feed (dust carry-over), fuels the PFBC; STAG power cycle
4b	Same as Case No. 4 but the PFBC is operated at a Ca/S ratio of 1.5 and with 20% excess air - for consistency with most of the other cases
6	Recycling, two-stage, dry-bottom, 20-atm gasifier; PFBC fueled by 5% of coal entering the gasifier (dust carry-over); STAG power cycle
7	Slagging, one-stage, 20-atm gasifier with all coal dust carry-over being fed back to the tuyeres; an adiabatic PFBC; STAG power cycle
10	Duplicate Case No. 4 but fired limestone PFBC unit is replaced with DSRP subsystem
11	Duplicate Case No. 6 but fired limestone PFBC unit is replaced with DSRP subsystem

6.2 DESCRIPTION OF COST ESTIMATING PROCESS

Once the selected system configuration's general integration, performance characteristics, and heat and material flows were developed using ASPEN, the information required to size key pieces of process equipment and to cost each plant section was available. This Section provides a description of the process utilized in computing the individual plant section costs. All costs were developed in first-quarter 1989 dollars.

As mentioned earlier, the costs were estimated for the Nth plant construction; this was accomplished by estimating learning-curve effects for the application of each plant section employing new technology and also for the overall construction activity. It is recognized that after a first-of-a-kind plant is built and has operated for a time, subsequent similar plants can be built at lower cost because of the experience gained. This happens for two reasons: (1) refinements in component or section design reduce the cost of the section, and (2) improved construction methodologies are developed (e.g., use of factory fabricated plant modules, pre-engineered scaffolding/forming systems, and better integrated construction sequences) that reduce both the construction time and cost. Consequently, the costs developed for early plants typically include a process-contingency factor to cover unexpected functional problems (i.e., costs), and larger costs for construction-related items. Process-contingency factors were estimated in this study for applicable plant sections. Unfortunately, little quantitative information is available on the subject of process design learning curve effects; however, one source⁵ stated that a reduction of 25 to 40% can be expected to occur over a period of about 10 years after a new technology is introduced. Similarly, few applicable studies have been done on construction-related learning curves. However, two examples are cited in an American Electric Power study (Guha and Singh 1989) wherein a 35% capital cost reduction was achieved in going from their first-of-a-kind supercritical steam plant to their first similar commercial plant, and a 15% reduction is anticipated in going from their currently under construction first-of-a-kind PFBC plant to their contemplated third plant (currently planned for 2006).

The methodology adopted for the CGIA study was to obtain the installed cost of each section employing new technology by estimating first-plant costs and then developing a section-specific reduction attributable to a normal learning curve. The specific reduction was based on a subjective evaluation of factors such as process complexity, likelihood of complementary R&D expenditures, and projected ease of scale-up, all of which affect the slope of the learning curve. The reduction percentage was limited to not being greater than the same section's first-plant process-contingency percentage. (It should be noted that whenever the percentages are the same value, this process results in a mature-plant section being slightly less costly than the basic un-escalated section cost.) This reduction was felt to not be unreasonable and to be reflective of both the effects of design improvements and the consequences of component standardization and reasonable quantity production rates. Once the subtotal of installed section costs was calculated, an additional 16% reduction (from the subtotal) was taken to reflect lowered construction costs resulting from both improved/mature construction practices and the erection of a plant whose section designs were optimized to minimize constructed plant costs. It should be recognized that a considerable amount of time and a significant number of similar (but not necessarily identical) simplified IGCC plants will have to be built to realize these cost reductions: 20 to 30 years and 10 plants are likely to be representative numbers.

⁵Gangwal, Santosh. January 25, 1990. Research Triangle Institute. Personal correspondence with S.C. Jain, METC.

The following define the bases on which individual section costs were developed. A detailed cost breakdown sheet for each system configuration is provided in Appendix B.

6.2.1 Coal Handling

The cost of the coal handling section was estimated directly from cost curves developed by KOH Systems Inc. (1989). These curves were applicable to direct material and labor only; consequently, the resulting costs were multiplied by a factor of 1.5 to account for other costs that make up the total installed cost (e.g., indirect costs, home office expenses, engineering). Plant-section-cost then became only a function of coal throughput, scaled in tons per day (tpd) of coal. No contingency factor was applicable for this section.

6.2.2 Limestone Handling

This is a standard, commercial type of single-train plant section. Its cost was estimated using cost data reported by GRI (Smith and Smelser 1987) and was scaled on the basis of limestone consumption in tpd. No contingency factor was applicable for this section.

6.2.3 Gasification Section

In generic terms, all the gasifiers considered in the CGIA study were of the same type: fixed-bed, air-blown, steam-jacketed, and operating at 285 psia pressure (for the STAG-based systems) on caking bituminous coal. While an effort was made to discriminate costs between the various specific gasifier design approaches, no attempt was made to quantify the potential cost impacts of the specific design features in the prototype gasifier designs.

The air-blown, pressurized, dry-bottom, one-stage gasifier is the most developed and was the basis from which the costs for this plant section were developed for all the system configurations using dry-bottom gasifiers. The reference costs for the single-stage units were based on Lurgi gasifier costs provided in a General Electric study (Corman 1986). It is assumed the costs of the peripheral equipment within this plant section (e.g., lockhoppers, ash transfer equipment) are similar. (It is recognized that the specifics of this equipment are likely to differ considerably from the Lurgi equipment complement; however, assuming similarity of the aggregate cost was felt to be a reasonable approach for this study.) No attempt was made to take credit for the increased specific throughput (tpd of coal per unit of gasifier cross-section area), which would be expected as a result of the prototype gasifier and hot-gas-cleanup design philosophies, and this results in a noticeable (but unquantified) degree of conservatism in the developed costs for this plant section. As a consequence, the gasifier throughput capacity and output mass flow of raw gas in the cited reference were applicable, since air and a caking coal were used, and since pressures were nearly the same. This was

the basis for estimating the number of gasifiers required, and the per-gasifier costs of the Lurgi subsystem (as estimated by Lurgi) were used directly. It should be noted that Lurgi estimates the cost of a gasifier operating at about 600 psia to be the same as the cost of the low pressure unit (Corman 1986). The specific reason for this is not known. No contingency factor was applicable to this section.

The two-stage feature is expected to add to the cost to the gasifier, and a multiplier of 1.1 was applied to the section cost (developed using costs for a one-stage gasifier) to cover this. Currently, the closest analog to a high-pressure, fixed-bed, dry-bottom, two-stage gasifier is the Ruhr 100. This design is currently in the later stage of full-scale development/early stage of demonstration. Consequently, a process contingency factor of 25% was judged to be appropriate. Recycling, two-stage gasifiers were estimated to cost the same and to have the same contingency factor on a per-unit-throughput basis as the non-recycling two-stage units (which are likely to be charitable assumptions, but a more accurate approach to increasing the cost of the recycling gasifier was not developed).

Costs for slagging one-stage gasifiers were based on the BGL slagging gasifier. The current standard BGL gasifier has a 7.5-ft inside diameter gasifier and is oxygen-blown. Performance data and costs for this oxygen-blown BGL gasifier design operating at 437 psia and using Pittsburgh No. 8 coal were given in a recent EPRI report (Booras 1988) and formed the basis for the costing. It was estimated that the corresponding specific throughput for the slagging one-stage gasifier using preheated air would be 755 lb/hr/ft² at 285 psia. It appears a 12-ft diameter slagging gasifier is technically credible (and comments from BGL personnel support this), and this was taken to be a maximum internal diameter for this type of gasifier. Gasifier capacity was scaled from this point, based on product-gas mass flow varying directly with the square root of gasifier pressure. Again, no credit was taken for the potential advantages arising from the prototype gasifier design features. No contingency factor was judged to be applicable for this section.

The mass of ash output from each of these gasifiers was utilized as a scaling factor from which the cost of the land (to support and site the plant) was estimated.

6.2.4 Particulate Removal

Cyclones were used for particulate removal from the gasifier's hot, raw product gases for all system configurations. Cyclone costs for all system configurations were based on the physical-cleanup section costs associated with the Lurgi system in the GE report (Corman 1986), and are scaled with raw-gas mass flow. However, this may be optimistic for the cyclones serving the raw gas leaving the side-gas port of the two-stage gasifiers since the operating temperature is well above the normal range for Lurgi gasifiers. Cyclones and the related equipment operating from about 750 to 1200 °F (typical temperature levels for top-gas and one-stage, dry-bottom gasifier output streams) are considered to be fully developed, and no contingency factor was applicable for this section for these applications. However, at

1500 °F and above, cyclones and related hot gas equipment have been operating only at demonstration scale and have occasionally experienced problems. Therefore, a process contingency factor of 15% was utilized for these higher temperature applications. It should also be recognized that the effects of alkali loadings in the coal gas streams will probably require gas cooling to somewhat below 1600 °F prior to particulate removal, and this would limit cyclone operating temperatures to this level.

6.2.5 Tar-Cracking/Sulfur-Removal Section

The tar-cracking/desulfurization section is the least developed portion of the reference system. As mentioned earlier, it is based on a dual fluidized-bed configuration employing solids transfer (via both a lift pipe and gravity flow) at up to 1400 °F and a combined function media: zinc ferrite for desulfurization and a Y-zeolite for tar cracking. Cases 1, 1a, 1b, and 2 are exceptions in that little tar content was expected in the side-gas stream and the tar-cracking function was not included in the system model (which is probably unrealistically optimistic). As a result of this section's immaturity, no costs were available for this exact design approach. However, there are exact parallels in FCC as employed in refineries, and the Conoco Hot Desulfurization Process (Mitre Corporation 1976) contained the dual fluidized-beds and solids transfer features and operated at a slightly higher temperature. The Conoco process is reported to utilize a superficial fluidizing velocity of about 0.6 ft/s, which would be representative of the lower range of velocities being considered for the CGIA desulfurization system and permitted a calculation of the single-train capacity. Realistically, it would be expected that the tar-cracking/sulfur sorption and regeneration vessels would each be slightly smaller than the gasifier, based on FCC experience and the likelihood that their superficial velocities would exceed that of the gasifier. (This will result in one train per gasifier.) The Conoco system included a Claus plant, and its cost was deleted from the cost figures used to develop a reference basis for this plant section. The system size and cost were assumed to vary with the mass flow of the dirty gas, using a 0.7 exponential scaling factor. The fluid-bed version of the desulfurization portion of this process has been run only at laboratory- to bench-scale. While heavy hydrocarbon cracking using fluidized-beds has been proven at commercial-scale in the petrochemical industry, coal tar cracking with FCC media has been done only at laboratory-scale. Therefore, a process contingency of 70% was utilized for this section.

6.2.6 Pressurized Fluidized Bed Combustor (PFBC) Section

This is essentially a fairly small, pressurized, bubbling, fluidized-bed boiler operating at 1700 °F with a solids bed that is nearly all calcium sulfite/sulfate. The active vessel sizing was provided as described in Chapter 5 and the costs were taken from a METC contractor's report (Robertson et al. 1989), which was found to have an applicable cost breakdown. Even though the report design is specifically for a circulating fluidized-bed, it contained enough

detailed information to form a basis for the current study. Since there is almost no solid fuel in the PFBC for most of the CGIA system configurations, capacity and cost scaling were done on the basis of flue-gas mass flow leaving the PFBC. A process contingency of 15% seemed appropriate for this plant section, since a demonstration-scale PFBC is currently being built (but has not yet operated) and since the PFBC concept in this study is unlike the normal PFBC (very little coal is burned).

6.2.7 Direct Sulfur Recovery Process (DSRP)

This is the most technologically immature option considered in the CGIA study. It is a two-stage, catalytic process (i.e., two reactors and two condensers), operating at elevated temperatures (reactors at about 1025 °F and condensers at about 400 °F) and at system pressure. The feed streams are a mildly corrosive gas (hot regenerator offgas, containing about 2 to 6% SO₂) plus a small amount of gasifier product-gas, and the products are molten elemental sulfur and a nitrogen-rich inert gas that is recycled to the GT. The reference costs for this subsystem were provided by RTI (Dorchak, Gangwal, and Harkins 1990), the process developers, and costs were scaled with the regeneration-gas mass-flow rate to the 0.6 power. This process has been operated only at near bench-scale, and it has not been run as a fully integrated unit. Therefore, a contingency factor of 85% was selected for this process section.

6.2.8 Boost Air Compressor

The boost air compressor is unusual only in that its inlet stream is already hot and at considerable pressure and the driving horsepower requirements are large. However, compressors of this general type are commercially sold for applications in the petrochemical and process industries. A cost equation was developed from the compressor costs provided in the GE reference report (Corman 1986). The referenced compressor air flows ranged from 55 to 300 lb/s, which was consistent with the needs of the CGIA system configurations. No contingency factor was judged to be applicable for this section.

6.2.9 Power System

The power system consists of the GT, ST, and HRSG sections. The GT reference costs were based on specific commercial models, and consequently, were not flexible. The majority of system configurations examined were based on the relatively lower pressure STAG power cycle. For this CGIA study, the STAG cycles were considered to be based on a GE MS7000F industrial GT, and the higher pressure STIG cycles were considered to be based on the GE LM5000 aircraft-derivative GT. Costs for these turbines were taken from public literature (*Gas Turbine World Handbook 1988-89* 1988). The cited costs were for the turbine only, and were multiplied by a factor of 1.375 to obtain an installed cost. Even though this

plant section is based upon state-of-the-art components, it is complicated by a non-standard air pathway between the GT compressor discharge and turbine inlet, and the fuel gas is hotter (and, in some cases, of lower heating value) than is currently known to be utilizable; consequently, a contingency factor of 10% was assigned to this section.

The ST cost was estimated directly from cost curves provided in the KOH Systems report (1989). KOH's cost curves were for direct material and labor only, and were multiplied by a factor of 1.5 to account for other costs that make up the total installed cost (e.g., indirect cost, home office expenses, engineering). No contingency factor was judged to be applicable to this section.

The HRSG costs were developed from data in the "Design of Advanced Fossil Fuel Systems" (DAFFS) studies (Argonne National Laboratory 1983). The specific HRSG reference unit costs for the STAG systems were based on the input flue-gas mass flow, and for the STIG systems, they were based on the product of flue-gas mass-flow rate and temperature drop. An exponential cost-scaling factor of 0.6 for unit size was taken from Peters and Timmerhaus (1980).

6.2.10 Balance of Plant (BOP)

In capital cost estimates, an item called "general plant facilities" is often used to account for items not included in the cost of the process plant. The following subsystems are typically included in the BOP:

- BOP Mechanical - circulating cooling water, make-up water, wastewater treatment, distillate oil, start-up boiler, and certain yard items.
- BOP Electrical - electrical controls, general facility computer, and electrical yard items.
- BOP Civil - administrative buildings and yard civil work.

BOP was discussed in Chapter 2 under Ground Rule 19 and was fixed at 20% of installed process plant cost. As a point of comparison, the GE study (Corman 1986) developed a corresponding value of 20.18% for IGCC plants using air-blown Lurgi gasifiers and hot gas cleanup. All components within this section are of commercial status, and consequently, no contingency factor was judged to be applicable.

6.2.11 Other Capital Cost Items

As mentioned earlier, a purpose of the study was to compare the potential of various design options; this comparison becomes clouded as cost elements that are heavily based on assumptions or are project- or site-specific are included. As a result, the comparisons were

based on the Nth plant "Total Process Capital" (i.e., the amount of hardware required to build the Nth plant). However, there was a desire to develop COE values, and presenting representative "Total Capital Requirement" (TCR) values facilitates comparisons with other study results. Consequently, a representative set of assumptions was selected to allow development of a TCR and a COE value for each configuration analyzed; the assumptions were as follows:

- Engineering Fees = 10% of Total Process Capital.
- Project Contingency = 15% of Total Process Capital.
- AFDC = Based on a 2-year construction period.
- Working Capital, Royalties, etc. = 7% of Total Plant Investment.
- Land = \$4452 times the tons of ash generated per on-stream day.

These assumptions are reflected in the cost summary sheets provided in the next Section.

6.2.12 Operating and Maintenance (O&M) Costs

Total O&M costs were estimated using data from the GE study (Corman 1986). For the nominal 50-MWe STIG systems, the operating labor was estimated to be 10 people/shift. The STAG systems in this CGIA study are larger, nominally 250-MWe, (but smaller than the STAG systems considered in the GE report, which were 500-MWe with 28 people/shift), and have at least two trains of equipment compared to the single-train STIG configuration. The operating labor for the STAG systems in this study were, consequently, estimated to be 19 people/shift. The maintenance cost was assumed to be the same as that used in the GE report: 2% of the TCR. Maintenance cost is distributed as 40% for maintenance labor and 60% for maintenance material, a typical split for all generating plants. The administrative costs are estimated to be 30% of the operating labor plus the maintenance labor costs. Taxes and insurance are estimated to be 2% of the TCR. The administrative costs and tax and insurance percentages are typical of all power generation plants.

The variable costs reported in the GE report are for a very similar system configuration except that the ZnFe desulfurization unit is a moving-bed design (as opposed to the fluidized-bed, combined tar-cracking/desulfurization unit in this study). As a result, the variable O&M costs, in mills/kWh, were taken from the GE report except for the ZnFe unit. The basis for the desulfurization unit's variable cost was the zinc-ferrite/media consumption (attrition loss), which was taken as 25% of the media cost per 50 cycles of the sorbent/catalyst circulation; this is probably much too high (conservative), but is representative of early (experimental) experience with ZnFe pellets.

6.3 RESULTS OF THE SYSTEM ANALYSES

As mentioned at the outset of Chapter 6, the system analyses were much less extensive in scope than originally contemplated. As a result, a few areas of investigation were

recognized to be incomplete and some questions and issues were left unaddressed. In a few instances, this also caused the analyzed configurations to differ very slightly from the case definitions provided in the summary Table 15 and in Appendix A. The two primary examples were the re-sizing of each system to utilize an integer number of MS7000F (or LM5000) GTs and a more realistic treatment of the use of ROM coal (and the related re-estimation of the quantities of fines appearing at various locations within the various system configurations, and the resulting effects on the systems' characteristics). It should be noted that all coal-handling section and coal (fuel) costs were based on the same unit costs, which corresponded to sized coal in all cases. (While this was not realistic, it was not felt to have a significant effect on the comparisons intended for this study.) The figures and tables provided are, however, all self-consistent and describe in detail the configurations analyzed. The results are quite useful when the system configurations are considered to be a bit "rubber-like," and the performance characterizations are taken on the basis of normalized parameters, e.g., capital costs in \$/kWe.

A summary of the results is provided in Table 17. The left half of Table 17 summarizes the plant performance including emissions levels, and the right half summarizes the costs. Efficiency is presented on a lower heating value (LHV) basis and was calculated for the total plant, that is input-coal to output-electricity. The Process-Capital column represents the total installed equipment cost (consistent with the assumptions previously discussed), and the TCR column represents a typical value for the total Nth plant capital cost (i.e., this value is the Process Capital plus representative values for normal soft cost items such as fees, royalties, and AFDC). The COE was calculated in the conventional manner and reflects fuel, capital, and O&M costs. Details on the performance computations are given below; depictions of the heat and material flows for each system configuration analyzed are provided at the end of this Chapter. A corresponding discussion of the costing results is provided in Section 6.3.2.

6.3.1 System Performance Results

The results of the ASPEN simulations are depicted in Figures 59 through 70 (located at the end of this Chapter). These figures provide a "line/box" representation of the system configuration for each case analyzed. Key points on major flow streams are labeled to provide an understanding of the flowing media (type of media, physical conditions, and flow rate), and all major process boxes are similarly labeled. While the stream or box names may at first appear a bit cryptic, they provide the reader with a strongly inferred definition of each stream's characteristics and each box's functions. It should be recognized that the processes depicted are in fact simplifications of the processes actually simulated. The simplifications shown in the figures were made in order to make the dominant flows and functions more easily understood. As an example, while not explicitly shown, the heat released during the regeneration of the ZnFe media is used to preheat the high-pressure boiler feed-water upstream of the HRSG. Similarly, heat inputs to the low pressure side of the steam system from the GT compressor's discharge air or the gasifier's steam jacket were not explicitly

Table 17. Summary of System Analysis Results

Case No.	Plant Output (MWe)	LHV Effic. (%)	Emissions (lb/MBtu)		Total Plant O&M (mill/kWh)	Process Capital (\$/kWe)	TCR (\$/kWe)	COE (mill/kWh)
			SO _x	NO _x				
1	60	25.6	.25	.37	30.6	741	1046	89.3
1a	51	31.5	.25	.37	33.8	923	1295	95.2
1b	77	32.6	.25	.37	26.8	641	903	75.4
2	342	44.0	.25	.37	20.3	476	671	56.3
3	55	33.4	.26	.65	29.4	641	903	77.5
4	250	42.0	.26	.61	22.1	502	708	60.0
4a	300	42.2	.23	.43	20.9	478	673	57.7
4b	252	42.2	.26	.61	21.7	502	707	59.5
6	284	43.8	.31	.64	22.0	570	801	62.1
7	253	43.7	.30	.16	21.8	515	725	59.6
10	264	39.2	.08	.61	19.5	506	709	58.6
11	290	41.2	.13	.64	20.8	637	890	64.6

depicted. There is also a minor inconsistency in the two-stage gasifier cases (1, 1a, 1b, and 2) in that the effect of a cyclone near the gasifier on the side-gas output stream was functionally omitted. This solids separation step is needed for operational reliability; however, the only effect on the simulation would have been to add slightly more fines (i.e., fuel) to the PFBC for these cases, since a cost for this cyclone was included. Similarly, the quantities of fines separated from the gasifier output-gas streams for the recycle gasifiers, the two-stage gasifiers, or when feeding ROM coal were assumed percentages. Refining these values would have been somewhat of an academic exercise, however, since the prototype gasifiers' responses to fines in the coal feed streams would be conjecture at this point. The astute reader is consequently forewarned that energy flows and costs relating to coal fines may have to be adjusted somewhat to better represent reality. However, the figures and tables provided are all self-consistent and do provide balanced energy and material flows.

A myriad of insights may be drawn from this set of analyses and figures; the following is just a summary of the more significant ones:

- There is not a major impact on system performance as a result of the type of fixed-bed gasifier utilized (within the current ability to mathematically simulate the gasifiers and with the assumption of equal operability).
- There is a major impact on system performance as a function of how well the system is integrated and what technologies are employed in the various (non-gasifier) subsystems; those related to the overall fuel-gas desulfurization process, specifically, are very important.
- The air-blown, slagging-bottom, one-stage, fixed-bed gasifier has an inherently low ammonia content in its product gas, and this is quantitatively reflected in the system NO_x emissions.
- The use of the PFBC is very effective environmentally, enhances the overall system efficiency, and provides a large degree of design flexibility; but its merits are contingent upon a successful interface with the GT. As increasing amounts of fuel are fed to the PFBC, the interface with the GT is made more complex (because of the difficulty of extracting and readmitting large volumes of air). With still further increases in the fuel-feed rate, a point can be reached wherein the ability of the GT to accept more preheated combustion air is exceeded, and at this point the plant efficiency starts to degrade with increased fuel since the incremental energy is only useable by the steam side of the system.
- The tar cracking/desulfurization media temperature for all of the cases analyzed exceeded 1200 °F (e.g., cleaned gas-outlet temperatures ranged from 1202 to 1245 °F), which implies a need for somewhat greater fuel gas precooling (the gas temperature rises varied from 67 to 142 °F), a sulfur-sorbent temperature capability more representative of zinc titanate than current ZnFe formulations, or both.
- There are noticeable disincentives arising with cleaned, product-gas temperatures significantly above 1200 °F (besides the obvious hardware complications), and most of these arise as a result of a (relatively) lower gas heating value or the possibility of excessive alkali content. For example, the fluid-bed desulfurization unit is a fractional-removal device that leaves a fixed concentration (by volume) of sulfur in the cleaned fuel gas; as the fuel-gas heating value decreases, the total mass of sulfur reaching the GT increases and the sulfur emissions (per MBtu of coal input to the plant) consequently increase. The raw-gas heating value is often low as a result of the same phenomena that makes the gas hot, or the heating value may be low as a result of the overt use of a water quench (which is the most cost effective means) to cool the gas. Lowered gas heating values can also significantly complicate the GT combustion process (which cannot be readily ameliorated by increased fuel gas temperatures since the fuel gas temperature is essentially limited by the capability of the ZnFe sorbent). Similarly, at temperatures above 1600 °F, the quantity of vapor phase alkali (which is not easily removed) becomes significant to the GT.

- The DSRP is highly promising because of its excellent environmental characteristics, a cool high-pressure offgas that is useable for internal cooling of the GT, and a reasonable projected cost. However, its merit is almost totally dependent on the ability of the desulfurization subsystem to produce a regeneration offgas stream containing nearly no free oxygen (assuming successful development of the basic DSRP process). Therefore, the stoichiometric regeneration process, as assumed in all these cases, is essentially a requirement. Similarly, the system will likely require means to utilize coal fines, such as briquetting or reburning for NO_x control.
- The quantities of air utilized external to the GT and the amount of highly preheated combustion air provided to the GT from the PFBC are large enough in several of these cases to cast doubts on the ability to utilize GTs having in-line combustors (as opposed to "silo" combustors).

6.3.2 System Costing Results

The analyzed system configurations include a number of new technologies, and there is current uncertainty regarding the capital costs of plant sections utilizing them. As a result, the estimated cost of a system that utilizes several more or less experimental technologies must also be considered to be somewhat uncertain. While the absolute accuracy of the overall estimates may not be high, the costs of the individual plant sections were derived in a uniform manner for all cases studied, and contingencies and other factors were either constant or derived in a consistent manner from case to case. As a result, the costs can be useful for comparison purposes. It is expected the relative differences in the capital costs of sections will be much less uncertain than the total section costs. This applies to total plant cost as well but to a lesser degree, since the mix of new technologies varies. The reader should not assume this study has established the lowest capital cost system, but rather that an informative start has been made. Contemplation of the results presented in this Section provides insights regarding the probable influence of various technologies on overall plant cost and COE.

A summary of the costing results and the COE computations for the 12 cases analyzed is provided in Table 18. The details of the capital costing for each individual case are provided as spreadsheets in Appendix B. As mentioned earlier, Total Process Capital (Cost) represents the total installed cost of hardware and is the basis of comparison for this study. TCR provides a representative cost for the Nth plant and includes representative "soft" costs. (Both of these cost totals are also listed in the System Analysis Summary, Table 17.)

Again, a myriad of insights may be drawn from the detailed costing tables provided in Appendix B. However, the reader is cautioned to not be overly quantitative as there is a recognizable level of softness in this whole process that culminates in the cost estimates. This notwithstanding, some interesting comparisons between various system configurations are possible, and the more direct comparisons are shown in Table 19. In two instances, both Case 1 and Case 1b are utilized in the comparisons; this was done because while it is the more cost

Table 18. Summary of Detailed Costing Results

Case Number:	1	1a	1b	2	3	4	4a	4b	5	7	10	11
Total Process Cost, \$/kW	741	903	641	476	641	502	478	502	570	515	506	637
Total Capital Required, \$/kW	1046	1295	903	671	903	708	673	707	801	725	709	890
Efficiency, %	25.6	31.5	32.6	44.0	33.4	42.0	42.2	42.2	43.8	43.7	39.2	41.2
Plant Net Output, MWe	60	51	77	342	55	250	300	252	284	253	264	290
O&M Cost, \$ thousands/year:												
Fixed O&M Costs:												
Operating Labor	1890.0	1890.0	1890.0	3591.0	1890.0	3591.0	3591.0	3591.0	3591.0	3591.0	3591.0	3591.0
Maintenance Labor	502.1	528.4	556.2	1835.9	397.3	1416.0	1615.2	1425.3	1819.9	1467.4	1497.4	2064.8
Maintenance Materials	753.1	792.5	834.4	2753.8	596.0	2124.0	2422.8	2138.0	2729.8	2201.1	2246.1	3097.2
Administrative Costs	717.6	725.5	733.9	1628.1	686.2	1502.1	1561.9	1504.9	1623.3	1517.5	1536.5	1696.7
Taxes and Insurance	1258.2	1320.9	1390.6	4589.6	993.3	3540.0	4038.0	3563.3	4549.7	3668.5	3743.5	5162.0
Total Fixed Costs	5118.0	5257.3	5405.1	14396.3	4562.8	12173.1	13228.9	12222.5	14313.6	12445.5	12604.6	15611.7
Variable O&M Costs	3577.8	2916.8	4363.8	18459.0	3090.8	13933.2	18527.4	13638.9	15310.2	13621.7	11754.8	12917.2
Total O&M Cost	8695.8	8174.1	9768.9	32857.3	7653.5	26106.3	29756.3	25861.3	29623.8	26067.3	24359.3	28529.0
COE, mills/kWh:												
Capital	32.1	39.8	27.8	20.6	27.8	21.8	20.7	21.7	24.6	22.3	21.8	27.4
O&M	30.6	33.8	26.8	20.3	29.4	22.0	20.9	21.7	22.0	21.8	19.5	20.8
Fuel	26.5	21.6	20.8	15.4	20.3	16.2	16.1	16.1	15.5	15.5	17.3	16.5
Total COE	89.3	95.2	75.4	56.3	77.5	60.0	57.7	59.5	62.1	59.6	58.6	64.6
Solids Flows, tpd:												
Coal Input	768	525	768	2545	537	1949	2328	1359	2121	1880	2205	2304
Solid Waste	220	153	220	724	153	600	619	557	598	543	321	318
Limestone Usage	105	73	105	346	70	336	403	253	284	243	—	—
Sulfur for Sale	—	—	—	—	—	—	—	—	—	—	55.88	57.67

Table 19. Capital Cost Comparisons

Configuration Comparison	Total Process Capital, \$/kWe (Case)			
STIG versus STAG	STIG	(Case)	STAG	(Case)
- Two-Stage, Dry-Bottom, PFBC	741	(1)	476	(2)
- Two-Stage, Dry-Bottom, PFBC, plus ST	641	(1b)	476	(2)
- One-Stage, Dry-Bottom, PFBC	641	(3)	502	(4b)
Dry-Bottom versus Slagging	Dry-Bottom	(Case)	Slagging	(Case)
- One-Stage, PFBC, STAG	502	(4b)	515	(7)
PFBC versus DSRP	PFBC	(Case)	DSRP	(Case)
- One-Stage, Dry-Bottom, STAG	502	(4b)	506	(10)
- Recycling Two-Stage, Dry-Bottom, STAG	570	(6)	637	(11)
One-Stage versus Two-Stage	One-Stage	(Case)	Two-Stage	(Case)
- Dry-Bottom, PFBC, STIG	641	(3)	741	(1)
- Dry-Bottom, PFBC, STIG (plus ST in 1b)	641	(3)	641	(1b)
- Dry-Bottom, PFBC, STAG	502	(4b)	476	(2)
- Dry-Bottom, PFBC, STAG (max. PFBC in 4a)	478	(4a)	476	(2)
With/Without Top-Gas Recycle (TGR)	With TGR	(Case)	Without TGR	(Case)
- Two-Stage, Dry-Bottom, STAG	570	(6)	476	(2)
- One-Stage, Dry-Bottom, STAG	570	(6)	502	(4b)
Extra Fuel to PFBC	30% ROM + 5% Fines	(Case)	5% Fines	(Case)
- One-Stage, Dry-Bottom, STAG	478	(4a)	502	(4)

effective, Case 1b contains an ST and, thus, is not a pure STIG system. In the same vein, both Case 2 and Case 4b should be considered relevant comparisons to the recycling gasifier configuration of Case 6, Case 2 being the more similar in hardware and Case 4b being the more similar in function and interface with the rest of the system.

One of the more interesting comparisons possible is the one between the use of STIG and STAG power cycles. The Case 1 family and Case 2 offer a direct comparison, since the only difference in technology between the two systems is the use of STIG in the Case 1 family and STAG in Case 2. The estimated capital cost is 25 to 50% (or more) higher for the use of a STIG power system. A large portion of the disadvantage for the STIG system is caused by scale effects, and it would appear an overwhelming advantage exists for the STAG approach. However, other considerations can become important. The STIG configurations are single-train systems of relatively small size (because the only STIG turbine commercially available is in the 50-MWe range), and in the reference system configuration, the STIG systems have more steam generating capacity than is required to satisfy the GT and gasifier. This characteristic could, for example, make this an ideal configuration for a cogeneration application where export steam was desired (but that was not analyzed). The system's capital

cost is also somewhat penalized because the GT cost is relatively high (in \$/kWe) and other equipment, such as the gasifier and PFBC, must be scaled down to match the turbine; thus, the economies of scale possible in the STAG configurations cannot be achieved. It is quite likely that a larger system using multiple STIG units and maximum-sized single-train equipment elsewhere would show considerable improvement in capital costs over the Case 1 family; however, it is very doubtful that it would surpass the STAG approach, given the reference system configuration.

In comparing dry-bottom to slagging gasifiers, the very slight cost advantage shown for the dry-bottom design is felt to be within the accuracy band of the computations, and the two should be considered as equal in cost. Other independent studies have shown the slagging design to have a cost advantage in oxygen-blown systems, but this study found the cost advantage to disappear in air-blown systems. This effect is likely to be explained by the reduction of the gasifier's steam requirement (which is generally felt to be a major potential advantage for the slagging design) being much less influential when the raw gas contains large quantities of nitrogen as in an air-blown design. In any event, a more definitive statement will have to wait for a more refined analysis.

The advantage to using of a PFBC over a DSRP to treat the regeneration offgas also appears to be slight (essentially zero to 10% of capital cost). The probable main reasons are that the currently envisioned DSRP requires consumption of cleaned product gas for effective operation, and the coal fines inherent in the system become a solid waste stream. Both features lead to lower overall system efficiency and higher coal usage rates, which affect capital cost unfavorably. However, given the feasibility of near stoichiometric sorbent regeneration, minor modifications to the system configuration, e.g., the use of coal fines briquetting or the use of hydrogen-sulfide-laden reducing gas, would likely result in a cost advantage for the DSRP approach. It should also be remembered that the DSRP approach is at a very early stage of development and the cost uncertainty is high. There appears to be no clear cost discrimination between one- and two-stage gasifiers (without recycle), and further resolution will require an improved ability to simulate and estimate the costs for a two-stage gasifier design. The one-stage design leads to a less complex system and is certainly more technically mature. It is likely to hold an edge for this reason.

The system configurations employing top-gas recycle did not appear attractive. Cases 2 and 6 offer a direct comparison of systems with comparable gasifier hardware, and Cases 4b and 6 have comparable interfaces with the rest of the system (i.e., both produce a single output-gas stream). The systems without top-gas recycle appear to have a 15 to 20% capital cost advantage. A major disadvantage of the recycle design approach is that it increases gas flow through the gasifier without a corresponding increase in net gas output. This, of course, requires more gasifier hardware (plus some ancillaries) for a given production rate. In addition, it should be recognized that it is the least mature gasifier design considered, and that top-gas recycle significantly complicates the gasifier's control system and is likely to affect basic gasifier and system operability (and this was not quantified).

The apparent advantage of being able to provide a significant quantity of extra fuel to the PFBC may not be realistically achievable because of the greatly increased quantity of GT air extraction (and subsequent re-admission) that is required. However, this comparison illustrates the potential value of the PFBC as a system "trash" burner, assuming that an acceptable interface with the GT can be maintained and that the GT can utilize the energy released.

A comparison of the effects of various technologies on overall capital cost strongly suggests that the preferable technologies in the near-term would include a STAG power subsystem, a dry-bottom one-stage gasifier, and a limestone PFBC for regeneration offgas treatment. Somewhat less mature systems would use a slagging-bottom one-stage gasifier and/or DSRP technology to significantly enhance their environmental characteristics, with possible economic gains, particularly if coal fines are directly injected into the gasifier or briquetting is incorporated into the system configuration. Cases 2, 4, 4a, 4b, 7, and 10 all have comparable capital costs and contain these particular technologies. The role to be played by two-stage gasifiers is currently unclear, but it is not obvious they will have an advantage, and the level of development work required may well preclude their ever making a contribution. While not specifically evaluated, an enhanced ability to accommodate coal fines and the ability to increase gasifier throughput potentially offered by the prototype gasifier designs would intuitively further improve the attractiveness of these systems (relative to the more traditional gasifier designs).

The O&M costs range from essentially 20 to 34 mills/kWh. Larger capacity systems, such as the STAG system, generally have lower O&M costs for two reasons. First, the larger STAG systems are less labor intensive than the STIG systems on a kWh-output basis, and as a consequence, the operating labor per kWh is less for these larger systems. Second, the more capital intensive systems have a higher maintenance cost. The fixed maintenance cost, being a constant percentage of capital cost, is a larger number of dollars for the larger STAG system, but since the capital cost of the STAG system in \$/kWe is less than the capital cost of the STIG system, the maintenance portion of the O&M is lower for the larger STAG system. Since all systems are assumed to have the same capacity factor, 65%, the maintenance cost in mills/kWh is then also lower for the larger STAG systems.

Two O&M cost characteristics peculiar to the desulfurization function are worth noting. The variable cost associated with the fluid-bed tar cracking and desulfurization unit is estimated to be 7.7 mills/kWh. This is significantly higher than comparable ZnFe costs, which were previously estimated for fixed- and moving-bed desulfurization units, and results from the high sorbent/catalyst attrition rates assumed in this study. The cost impact of high rates of solid media attrition (both directly and as an impact on other system components) can be significant. In addition, the O&M costs are beneficially affected by sulfur sale revenues in some cases and detrimentally by CaSO_4 disposal costs in others. The configurations utilizing DSRP systems benefit from an average O&M cost reduction of 0.7 mills/kWh because of sulfur sales, and the other systems incur an average O&M cost of 0.4 mills/kWh for CaSO_4 disposal.

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Figure 59. Case 1 System Configuration

Figure 60. Case 1a System Configuration

Figure 61. Case 1b System Configuration

Figure 62. Case 2 System Configuration

Figure 64. Case 4 System Configuration

M90002281

Figure 65. Case 4a System Configuration

Figure 66. Case 4b System Configuration

M90002284

Figure 68. Case 7 System Configuration

M90002285

Figure 69. Case 10 System Configuration

M90002286

Figure 70. Case 11 System Configuration

Chapter 7

Conclusions

This Chapter provides a synopsis of what are felt to be the principal conclusions that evolved from this study to identify the design characteristics important to a CGIA (Commercial Gasifier for IGCC Applications). Conclusions specific to the gasifier design and the IGCC system configuration are covered in specific Sections within this Chapter. The broad, general conclusions are as follows:

- A coal-based electric generation system with pollutant removal capabilities above the 97% range and thermodynamic efficiencies exceeding 40% (based on the LHV, lower heating value) is credible. This type of system can allow more than a twenty-fold growth in electric generation with no increase in pollutant burdens from acceptable levels (based on the Clean Air Act standards as proposed in late 1989). A simplified IGCC (integrated coal gasification/combined cycle) system was shown to have this capability and to be significantly less costly than traditionally considered, alternative, electric generation technologies (including *conventional* IGCC).
- Elimination of "wet" technologies for coal gas cleaning allows major cost and efficiency improvements to be realized in IGCC systems. The key to this is the removal of dust and sulfur species from the coal gas at conditions approximating those at the gasifier outlet, followed by direct coal-gas utilization in the GT (gas turbine) of a combined cycle. Technologies capable of accomplishing this hot gas cleanup, at conditions typical of fixed-bed gasification and at pressure levels consistent with contemporary GTs, are currently proven at small scale and several are clearly on the route to commercial status. The usage of an integrated hot-gas-cleanup design approach plus use of a portion of the GT compressor discharge air to supply the gasifier are the basic characteristics of the Simplified IGCC system.
- Fixed-bed gasifier designs based on one- and two-stage concepts, dry-ash and slagging bottoms, and recycling of the pyrolysis gas (top gas) to a deep-bed location have all been proven to be functional and have been incorporated into commercially offered fixed-bed gasifiers. Each of these general design features was found to have associated virtues and problems. However, no fixed-bed gasifiers have been sold (or even designed) that have features that maximize the virtues of the Simplified IGCC system concept.
- Of the existing fixed-bed designs, the Lurgi dry-bottom, BGL slagging-bottom, and METC (or GE) pilot-scale gasifiers are by far the most consistent with the Simplified IGCC system concept (and the first two are also the nearest to being commercially implemented). However, both the Lurgi and BGL designs have major potential for improvements that should result in significantly enhancing the overall performance of

a Lurgi- or BGL-based IGCC system, and neither has been operated with a hot-gas-cleanup subsystem or integrated with a GT. The METC design has most of the characteristics necessary for a Simplified IGCC system, but there is no active vendor (the same is true of the GE design).

- The manner in which the IGCC system is integrated or configured was found to be a more dominant influence on the overall system's cost and performance characteristics than the specific type of fixed-bed gasifier utilized within the system. While this may well be a truism, this finding is based on two tacit assumptions: (1) equal levels of operability and reliability for all gasifier designs (which is probably not real, but the difference was unquantifiable), and (2) the validity of the mathematical models developed for the gasifiers during this study (the models have defined limitations).
- A near-term Simplified IGCC system that is consistent with the Ground Rules of this study is very credible. The technologies favored for this system would be
 - A dry-bottom, one-stage, fixed-bed gasifier pressurized to approximately 150% of the GT compressor discharge pressure and incorporating a vertically traversing stirrer.
 - A combined function (tar cracking and sulfur sorbing) fluidized-bed unit for hot-gas chemical cleanup with media regeneration accomplished in either a lift-pipe or fluidized-bed regenerator.
 - A limestone PFBC consuming the minimum amount of fuel, but serving as a "trash-burner" and sulfur fixation device for the system.
 - A STAG (steam turbine and gas turbine combined cycle)-type power system, with the GT utilizing low-NO_x combustors, providing air to the gasifier and the PFBC (pressurized fluidized-bed combustor), and accepting flue gas from the PFBC.

This configuration resembles Cases 4 and 4b but utilizes a higher boost-compression ratio and the minimal PFBC energy release; both deviations would be expected to simplify the overall system control and the GT interface but to retain most of the other system virtues.

- An advanced Simplified IGCC system would offer significantly improved environmental characteristics and should offer improved economics (relative to the near-term system). The technologies favored for this system would be
 - A slagging-bottom, one-stage, fixed-bed gasifier pressurized to 130% of the GT compressor discharge pressure and utilizing direct reinjection of coal fines and fixed-plane bed stirring.

- A combined function (tar cracking and sulfur sorbing) fluidized-bed unit for hot-gas chemical cleanup with media regeneration accomplished in either a moving-bed or fluidized-bed regenerator to produce a nearly oxygen-free regeneration gas.
- A DSRP (Direct Sulfur Recovery Process) for sulfur fixation.
- A STAG-type power system with the GT utilizing low-NO_x combustors and providing air to the gasifier.

This configuration was not specifically analyzed, but rather borrowed characteristics from several configurations, which resulted in similarly attractive overall performances (i.e., the Case 4 family, Case 7, and Case 10); consequently, this conclusion is based on expectations rather than numerical performance predictions. This system approach is also likely to benefit noticeably from a higher temperature capability in both the coal-gas-cleanup section and the ability of the GT to accept hot fuel gas, relative to what was assumed in this study. However, it should be recognized a several-hundred-degree increase in fuel-gas temperatures could cause fuel-gas alkali levels to become a concern.

7.1 CONCLUSIONS RELATING TO FIXED-BED GASIFIER DESIGN

Conclusions relating to the gasifier's design and important to the operation of the gasifier in an economically and environmentally attractive fashion (and, in reality, the entire IGCC system) are provided in this Section.

- The current ability to mathematically model the phenomena occurring within a fixed-bed gasifier is less than that required to unequivocally compare the operational characteristics and performances of one- and two-stage gasifiers, with and without slagging bottoms and pyrolysis gas recycle. However, a depth of understanding was developed sufficient to allow useful insights to be drawn.
- The usage of more sophisticated gasifier instrumentation and control systems than is currently practiced is highly desirable if the full potential of a simplified IGCC system is to be realized. This becomes successively more important for slagging, two-stage, and two-stage recycling designs. (It is virtually mandatory for a recycling gasifier.) Direct or analogous in-bed measurements are expected to be required for these control systems, and these types of measurements are difficult to accomplish reliably. The single most important (and difficult) control function is the selection and maintenance of the appropriate ash removal rate. Gamma ray densitometry is likely the preferred technique for sensing coal-bed height.

- The current ability to mathematically model the phenomena occurring within a fixed-bed gasifier is also less than that required to design a sophisticated system to control gasifier operations. One potentially viable alternative is the use of a control-system design approach based on "artificial intelligence": specifically, "neural networks."
- Achieving particulate removal from the product coal gas without using a liquid-based gas scrubbing step essentially reduces one of the two potential process limitations on the gasifier's coal throughput to a pure economic trade-off (i.e., the cost of dealing with collected fines versus the value of increased gasifier throughput). This is particularly true for gasifiers operating with hotter combustion zones, that is, the slagging-bottom designs.
- Any enhancement of the gasifier's ability to limit the production of contaminants in its product-gas stream(s) is highly desirable. The most viable approaches to this are
 - To limit coal fines carry-over by reducing overbed gas velocities, providing overbed voidage, increasing overbed height, or a combination;
 - To concentrate contaminants into different product-gas streams that are utilized differently (e.g., the two-stage gasifier in Case 2); and
 - To crack tars and beneficially consume fines (e.g., by recycling portions of the gasifier's product stream[s] to deep-bed locations).

Earlier studies concluded that in-bed sulfur capture using the direct injection of a sorbent was not practical for fixed-bed gasifiers. (This occurs because of a combination of inherently ineffective sorbent utilization and significant potential operational problems.) Both overt reductions of gas velocities and increases in overbed height increase gasifier volume per unit of coal throughput (i.e., cost) but in a quite predictable manner. Recycling of the top gas (of a two-stage gasifier) was found to be not attractive because of cost and operational considerations. However, both the provision of overbed voidage and the deep-bed reinjection of fines were found to be feasible. Operational experiences at METC and GE (relative to overbed voidage) and BGL (relative to fines injection) implied that benefits would be achieved. The magnitudes of the benefits were not quantified, however. Reduced desulfurization and tar cracking requirements result from two-stage gasifier configurations wherein only the side gas serves as GT fuel and the tars and sulfur species (and probably fines) are concentrated in the top gas and bypass the majority of the cleanup subsystem to be directly combusted in a PFBC (Cases 1 through 2). This approach could significantly lower system costs and environmental burdens. (However, it should be noted that there were some uncertainties in the two-stage gasifier analyses).

- Reduction of alkali levels in the product-coal gas can be achieved inherently as a result of particulate removal with one-stage, dry-bottom, fixed-bed gasifiers. At

gasifier product-gas temperatures of 1200 °F and below, the vast majority of alkali species will be condensed on the coal fines to the extent that efficient particulate removal will reduce alkalis to below levels of concern for the GT. However, this is likely to be a marginal situation at 1600 °F, and at 1800 °F, alkali levels in the product gas will be of concern, independent of the effectiveness of the particulate removal step. Consequently, it can be concluded that hot product-gas streams should be cooled to somewhat below 1600 °F prior to particulate removal to ensure a lack of alkali problems in the GT. This phenomena is also likely to limit the incentive for increased temperature capability in the hot-gas desulfurization subsystem to levels somewhat below 1600 °F.

- Fixed-bed gasifier ash is not regarded as a potential environmental problem, but this is contingent upon the existence of an aggressive combustion process just prior to the ash passing out of the reaction volume of the gasifier. The environmental benignity is enhanced to the degree the combustion zone temperatures are increased (i.e., slag is more benign than ash), and it is compromised to the degree that design or process modifications lower combustion zone temperatures. Similarly, achievable gasifier coal-throughput rates are expected to vary directly with combustion zone temperatures. (Unfortunately, general operational complexity also varies in the same manner.)
- The coal-feeding subsystem and general configuration for the upper portion of the gasifier (shown conceptually in Figure 13 on page 65) potentially ameliorate all of the known problems that arise in the upper regions of fixed-bed gasifiers, and the subsystem has shown a capability to accommodate all U.S. coals.
- The chevron-injector concept (shown in Figure 16 on page 71) offers a credible approach to in-bed feeding (or extraction) of *mostly* gaseous streams.
- An in-bed stirrer is an asset in controlling coal- and char-bed porosity (and is a necessity for strongly caking coals), and it offers an effective route to accomplishing some needed in-bed measurements. Unfortunately, proper stirring protocol is currently an empirically determined parameter. An approach to obviating this potential problem is to cause the stirrer to vertically traverse as well as rotate; however, adding a vertical traversing capability significantly complicates the design. The key potential problem areas are the pressure seal at the gasifier shell and the shaft-support bearing. With a traversing stirrer, the use of an internal (to the gasifier) lateral support spider (i.e., a second, loose-fitting bearing) is likely to be a necessity for large gasifiers. It is probable that a stirrer with arms at two or more planes can be designed to function efficiently, given development testing, without vertical travel. Stirring blade shape is not critical, and an adequate design is shown in Figure 30 (page 133).
- A variable speed grate with some level of ash crushing capability and the ability to operate in the reverse direction (no more than 30 degrees of rotation is likely to be required) is recommended. The grate's ability to distribute reactant gases across the

gasifier diameter in the designed manner is quite important, and to the degree this cannot be accomplished, the designed depth of the ash layer should be increased. Similarly, the grate must withdraw ash in a reasonably areally uniform manner from the gasifier's cross-section in order to create a downward plug-flow of solids. In a commercial-scale gasifier, the active bed depth will have a height-to-diameter ratio slightly over 1.0, and it is likely that both the stirrer's motions and the grate's motions will interact and could effect the desired plug-flow nature of the solids. This possibility should be considered in the design of both the grate and stirrer. One proven approach to grate design at the 42-inch-diameter scale (and used with diameters up to 10 feet in atmospheric pressure gasifiers) is provided in Figure 18 (page 81).

- The current preferred gasifier wall design utilizes a pressurized steam jacket to cool a steel membrane wall enclosing the coal/char bed. (The gasifier's pressure envelope then becomes the exterior wall of the steam jacket.) In the future, improvements in the gasifier control systems may allow the use of refractory-walled, fixed-bed gasifier designs, which should further improve IGCC system efficiency. (The simplified IGCC systems generally have appreciable steam raising capability and the gasifier's contribution is not really necessary. To the extent the reduced gasifier heat loss does not have to be offset by increased steam input for gasifier operability, a lowered heat loss represents an increased system efficiency.) However, refractory wall designs are too operationally unforgiving with current operating techniques and instrumentation.
- Slagging-bottom, air-blown, fixed-bed gasifier designs are credible for IGCC applications. They offer approximately a three-fold increase in coal throughput for a given size gasifier, an apparent capability to directly utilize whatever level of coal fines might realistically exist within the system, inherently lower flue-gas NO_x levels, and an ash that is exceptionally benign (and, in many instances, can be sold for value). The potential problems mostly relate to the comparative lack of design maturity and understanding, e.g., the "art" in controlling slag levels in the gasifier, tuyere design and operation, and a high (predicted) product-gas temperature. The principal technology reservoirs for this approach are BGL and the designers of iron ore reducing/steel making equipment. Raw product-gas temperatures in the range of 1800 °F were predicted during this study and are inconsistent with both soon-to-be-available gas cleanup technology and adequately low alkali levels following hot particulate removal. (They were also inconsistent with the Ground Rules in this study.) Consequently, a direct quench for gas cooling was incorporated in Case 7 that in many respects is counter to the Simplified IGCC concept philosophy of making each system component's capabilities consistent with upstream or downstream gas characteristics and requirements. If the air-blown, slagging gasifier product-gas can be made to fall into the 1500 to 1600 °F range for otherwise desirable operating conditions, this is likely to be a very desirable design approach.
- Two-stage, pressurized, air-blown, fixed-bed gasifiers are not as potentially attractive for the Simplified IGCC system as the one-stage designs. However, they have

operational aspects that could make them attractive: specifically, more tolerance of caking coals and very disparate product-gas characteristics in the two product-gas streams (e.g., condensibles loading, temperature, heating value, and fines loading). In the systems analyses, there were no resulting obvious distinctions between the economic and environmental performances related to the two types of gasifiers; however, this assumed that the mathematical models represented them correctly and that the gasifiers operated with equal reliability (both of which were major assumptions). Again, this approach results in a hot product-gas stream (side-gas temperatures of 1600 °F were predicted), which raises concerns regarding alkali levels similar to those discussed above for the slagging-bottom gasifiers. The generic merit of this gasifier design is exemplified by the Case 1 through 2 configurations wherein most of the contaminants are sequestered for use in a PFBC combustor, and where the GT fuel-gas stream is, consequently, more easily dealt with. The realistically achievable performance of a pressurized version of this gasifier may make it better or worse than was predicted in the analyses. Unfortunately, a significant amount of R&D (research and development) is needed to bring this gasifier-design approach to fruition, and the realities of this need cast doubt on whether this approach will actually become credible for a Simplified IGCC application.

- Two-stage, pressurized, air-blown, fixed-bed gasifiers that recycle the top gas to an in-bed location are not attractive for this IGCC application. The designs are immature, the operational complexities are significant, the gasifiers are significantly larger for a given coal throughput, and the predicted performances of systems based on recycling designs were poor relative to the other gasifier design approaches considered.
- A prototype gasifier-design concept is provided in Figure 29 (page 130). This concept incorporates all the features felt to be valuable in a commercial gasifier for IGCC applications. The design could be operated as either a one- or two-stage gasifier, and while it could be representative of first units, the design is best suited to providing the information necessary to develop a mature gasifier configuration.

7.2 CONCLUSIONS RELATING TO SIMPLIFIED IGCC SYSTEM DESIGN

Conclusions relating to the entire IGCC system configuration and the design and operation of its major components (exclusive of the gasifier) are provided in this Section.

- Briquetting is a reasonably inexpensive means of accommodating coal fines within the overall system, and it is likely to represent less than half of 1% of the total system's capital cost. It will be the agglomeration method of choice, assuming an agglomeration step is incorporated within the system.

- Using a particulate media to combine the heavy hydrocarbon chemical-cracking and hot-gas-desulfurization functions appears credible. The performance of these functions in a single vessel operating in the fluidized-bed mode is preferred because of the potential for improved thermal control, reduced vessel size (i.e., cost), and the ensuing simplifications that can be realized in the subsequent regeneration process. This step would be based on fluidized catalytic cracking experience in the petrochemical industry and zinc-ferrite R&D conducted by METC. Zinc ferrite is an adequate sorbent for desulfurizing all the product-gas streams considered in the study; however, some more advanced sorbents with increased temperature capability (e.g., zinc titanate) offer potential for further improvements. It is expected that virtually all of the heavy hydrocarbon condensibles can be cracked to lighter molecules within the residence times consistent with credible desulfurization-vessel design. Similarly, achieving sulfur reductions to 100 ppm or below in the fuel gas should be no problem, and this translates to system SO₂ emission levels of only about 0.25 lb/MBtu - about a 98% SO₂ emissions reduction in most cases. (If required, sulfur emission levels can be further reduced by using a modestly increased tar-cracking/desulfurization vessel.) It would be expected that the tar-cracking/desulfurization vessel will be similar in size to the gasifier, and that there will normally be one for each gasifier. (In some of the larger-scale system Cases - e.g., Case 2 as shown in Table 23 in Appendix B - this relationship came out to be slightly more than one gasifier per desulfurization vessel. This is likely to make the process train arrangements more complex than desired and would probably not be done.) As configured, it is not expected this step will have any effect on NO_x emissions; however, this concept provides a good basis for accommodating any successful ammonia-cracking-catalyst developments that may arise in the future.
- The techniques involved in regeneration of the tar cracking and desulfurization media and disposition of the regeneration offgas are the most critical aspects of the whole Simplified IGCC system design. Essentially all the sulfur in the feedstock coal and much of the carbon in the cracked condensibles are present on the media as a fuel to be "burned" in the regeneration process. However, the maximum allowable media temperature rise is in the vicinity of 100 °F and any resulting flue-gas emissions constitute SO₂ emissions for the entire system. While other regeneration approaches are credible, two were particularly attractive: one based on a fluidized-bed regenerator vessel, and the other on a riser-tube regenerator. The fluidized-bed regenerator vessel is half of the two-vessel subsystem concept shown in Figure 41 (page 163). It can easily limit the media temperature rise, should be capable of producing a regeneration offgas with little or no excess oxygen, and has no unusual height requirements; however, it will add another vessel to the system, and this vessel can approach the tar cracking/desulfurization vessel in size (and, thus, could also be comparable to the gasifier in size). The lift-pipe regenerator shown in Figure 44 (page 169) essentially utilizes an excess air stream to provide the media lifting function and to accomplish media regeneration at the same time. This approach results in less hardware, but the regeneration offgas is inherently oxygen-rich and there is a limit to the amount of regeneration that can be accomplished in a given height. (It should be noted that

realistic heights correspond to adequate levels of regeneration.) The air requirements to accomplish carbon combustion typically exceed the requirements for sulfur combustion and a 3-to-1 relationship would be representative. (In reality, this ratio is strongly coal-dependent and gasifier-dependent; that is, a two-stage gasifier could result in much less carbon on the media.) The physical requirements on the solid media are expected to be similar for both approaches. The air requirements for either approach can be formidable, and careful design is required in this area. (The system analyses are representative of only the fluidized-bed approach for this specific process area.) The fluidized-bed regenerator is clearly the preferred choice for systems incorporating a DSRP unit for sulfur fixation; however, the combination of the riser-tube regenerator and a limestone PFBC for sulfur fixation is likely the lowest cost approach overall (but was not specifically analyzed).

- The limestone-based PFBC was found to offer a very flexible means to accommodate modest quantities of "trash fuels" and to serve as a means of fixing the total system's sulfur in the form of a high quality gypsum. The PFBC is also expected to reduce NO_x emission levels from what would otherwise result if these fuels were combusted outside of the PFBC. As the amount of fuel consumed in the PFBC is increased beyond that required to maintain 1700 °F, the integration with the GT becomes more complex (because of the need to extract and re-admit an increased GT air flow). As the amount of fuel consumed in the PFBC is increased beyond that which can be accommodated as 1700 °F-preheated combustion air to the GT, the system's performance is degraded because the incremental energy flow is restricted to only the steam side of the system.
- The DSRP represents a potentially very attractive approach to sulfur fixation from the regeneration offgas. While not yet mature, the DSRP approach appears to be potentially low in cost, result in still lower system-level SO₂ emissions, produce elemental sulfur of saleable quality, and to simplify the interface with the GT. The negative aspects are that it removes all capabilities to directly utilize "trash fuels" within the system, and the system performance is degraded to the degree free oxygen is present in the regeneration offgas.
- The viability of the reference IGCC system configuration is fundamentally dependent upon an effective means to accommodate the regeneration offgas. Consequently, a third, totally state-of-the-art process was identified as a contingency approach. This approach is a combination of the BSRP (Beavon Sulfur Recovery Process) and the Lo-Cat process; it is somewhat more costly than either the limestone PFBC or DSRP approaches, but is comparable in performance. The BSRP/Lo-Cat combination was found to be much lower in cost than other conventional gas treatment alternatives.
- While no overt steps to remove ammonia downstream of the gasifier and prior to the GT were evaluated in this study, it appeared that relatively state-of-the-art low-NO_x combustion techniques could achieve currently acceptable emission levels. In addition,

post-combustion NO_x conversion technologies, SCR or "reburning" (perhaps even coal-fueled), are compatible with the simplified IGCC system concept, and if required, could achieve lower emission levels but at some economic penalty. Catalytic cracking of ammonia in the fuel gas is likely to be the eventual preferred approach; however, this is dependent upon future developments and was felt to be a system enhancement.

- Total Process Capital costs in the range of \$475 to 525/kWe for an Nth plant appear to be achievable. With representative assumptions for "soft" cost items, this translates to a Total Capital Requirement in the vicinity of \$700/kWe and a COE of about 59 mills/kWh. A summary of performance results for the various systems analyzed is provided in Table 17 (page 219).
- The impact that the Air Toxic Standards will have is currently not clear. The standards do not distinguish between the condition when the toxic element is present as an element or in a compound form, the data base is unclear or ill-defined, and the interpretations that will be applied to the standards are unknown. What is known is that fixed-bed coal gasification is likely to place a lesser quantity of trace elements into the product-gas stream than virtually any other form of coal utilization. Further, a disposable Nahcolite sorbent bed is capable of removing chlorine (the element of most potential concern) down to 1 ppm levels for almost negligible costs. In addition, it is expected a significant fraction of the defined toxic elements will be sorbed on, and removed with, the dust particles. It would appear the potential National benefit of the Simplified IGCC technology, from both economic and environmental standpoints, far exceeds the potential consequences of trace level, toxic element emissions.

Abbreviations and Acronyms

AEP	American Electric Power
AFDC	Allowance for funds during construction
ASPEN	Advanced System for Process ENgineering
BCURA	British Coal Utilization Research Association
BGL	British Gas/Lurgi
BOP	Balance of plant
BSRP	Beavon Sulfur Recovery Process
CGIA	Commercial Gasifier for IGCC Applications
COE	Cost of electricity
CURL	Coal Utilization Research Laboratory
CWS	Coal water slurry
DAFFS	"Design of Advanced Fossil Fuel Systems"
DOE	(U.S.) Department of Energy
DSRP	Direct Sulfur Recovery Process
EPA	(U.S.) Environmental Protection Agency
FBN	Fuel-bound nitrogen
FCC	Fluidized catalytic cracking
FGD	Flue gas desulfurization
FSI	Free-swelling index
GE	General Electric (Company)
GT	Gas turbine
HHV	Higher heating value
HRSG	Heat recovery steam generator
IGCC	Integrated coal gasification/combined cycle
ISTIG	Intercooled, steam-injected gas turbine
KGN	Kohlegas Nordrhein
LHV	Lower heating value
LSU	Louisiana State University
METC	Morgantown Energy Technology Center
NCB	National Coal Board (Great Britain)
NSPS	New Source Performance Standards
O&M	Operation and maintenance (costs)
OSHA	Occupational Safety and Health Administration
P	Pressure
PC-FGD	Pulverized coal (with) flue gas desulfurization
PFBC	Pressurized fluidized-bed combustion (or combustor)
PNL	Pacific Northwest Laboratories
POC	Products of combustion
R&D	Research and Development
ROM	Run-of-mine

RTI	Research Triangle Institute
SCR	Selective catalytic reduction
ST	Steam turbine
STAG	Steam turbine and gas turbine combined cycle
STIG	Steam-injected gas turbine
T	Temperature
TCR	Total capital requirement
TDH	Transport disengagement height
TGR	Top-gas recycle
tpd	Tons per day
tph	Tons per hour
V	Velocity

**Process Flow-Diagram Abbreviations Used in Case System Configurations
(Figures 59-70)**

GASIFSTM	Steam supplied to gasifier
HOTVAP	Intermediate stage of gas flow through DSRP
HPSTM	High-pressure steam
LAIR	Air supplied to PFBC
LIFTSTM	Steam used for sorbent transport
LPOC	PFBC tail gas sent to GT combustor
LPSTM	Low-pressure steam
REC2VAP	DSRP offgas supplied to GT
REDGAS	Reducing gas supplied to DSRP
REGENAIR	Air supplied for regeneration
SHFTSTM	Steam used to supply shift reaction
SHSTEAM	Superheated steam to ST
SPSORB	Spent sorbent
TAIL2PFB	Tail gas to PFBC

Appendix A

System Configuration Assumptions

CASE 1 - TWO-STAGE GASIFIER WITH STIG CYCLE

The system configuration for Case 1 contains a two-stage, dry-bottom, prototype gasifier operating at 40-atm pressure; a PFBC fueled by raw top-gas from the gasifier plus 5% of the coal entering the gasifier (resulting from the assumed coal-dust carry-over in the top-gas stream); and a STIG power cycle. The configuration analyzed did not include a hot cyclone in the side-gas stream, but that would be preferred in a real system. Two sub-options were considered for this configuration and are presented subsequently. The following were the assumptions and key operating conditions pertinent to the ASPEN analysis of Case 1:

- The reference coal composition is used as input to the gasifier.
- The math model's computations (see Chapter 4, Section 4.6) define the two-stage gasifier's gas compositions and yields of 32% top-gas and 68% side-gas at temperatures of 800 and 1600 °F, respectively.
- The operating pressure of the gasifier is 612 psia.
- The steam-to-coal ratio equals .84, and the air-to-coal ratio equals 2.12.
- Five percent of the coal fed to the gasifier is elutriated as fines, which are subsequently combusted in the PFBC.
- ZnFe sulfur capture is not completely effective, and 100 ppm of sulfur exits in the fuel-gas stream. ZnFe sorbent utilization is 40%. No tar cracking was considered to be required.
- A 100 °F temperature approach is assumed in the desulfurizer for the water-gas shift reaction. Gas temperature at the desulfurizer inlet is held down to 1140 °F by water quenching.
- Regeneration of the ZnFe sorbent is accomplished stoichiometrically in the regenerator, which operates at 1400 °F.
- No sulfate formation occurs within the ZnFe during regeneration.
- The quantity of steam needed in the ZnFe-sorbent media lift-pipe is 0.085 lb of steam/lb of solids being transported.

- Twenty-percent excess air is utilized in the PFBC.
- Solid feed rates corresponding to a Ca/S ratio of 1.5 result in 96% sulfur capture within the PFBC operating at 1700 °F.
- A waste-steam stream is produced by the PFBC in order to preclude excessive GT combustor temperatures.
- Flue gas from the PFBC is sent to the GT and is expanded through the turbine.
- The GT is an LM5000 with a firing temperature of 2200 °F.
- Injection steam flow is set at 15% of the compressor air flow. The split between turbine and combustor steam-injection flows is 79 and 21%, respectively.
- The GT combustor injection-steam temperature is held at 514 °F to provide a heat sink in order to avoid overheating the combustor.
- The stack-gas temperature, including the contribution from the waste-steam stream, is 296 °F.

Case 1a

Case 1a was a duplicate of Case 1 except that all of the GT compressor discharge air (exclusive of that needed for internal cooling) is assumed to be utilized in some fashion external to the GT. The incentive for this sub-option was that it could simplify the ducting interface to the GT, could reduce the magnitude of the waste-steam stream (needed to keep the temperature of the preheated combustion air to the GT within reason), and would enhance the application of the PFBC. In effect, all of the incremental air is utilized in the PFBC, and this results in 385% excess air within the PFBC. The following were the assumptions and key operating conditions pertinent to the ASPEN analysis of Case 1a:

- The reference coal composition is used as input to the gasifier.
- The math model's computations (see Chapter 4, Section 4.6) define the two-stage gasifier's gas compositions and yields of 32% top-gas and 68% side-gas at temperatures of 800 and 1600 °F, respectively.
- The operating pressure of the gasifier is 612 psia.
- The steam-to-coal ratio equals .84, and the air-to-coal ratio equals 2.12.

- Five percent of the coal fed to the gasifier is elutriated as fines, which are subsequently combusted in the PFBC.
- ZnFe sulfur capture is not completely effective, and 100 ppm of sulfur exits in the fuel gas stream. ZnFe sorbent utilization is 40%. No tar cracking was considered to be required.
- A 100 °F temperature approach is assumed in the desulfurizer for the water-gas shift reaction. Gas temperature at the desulfurizer inlet is held down to 1140 °F by water quenching.
- Regeneration of the ZnFe sorbent is accomplished stoichiometrically in the regenerator, which operates at 1400 °F.
- No sulfate formation occurs within the ZnFe during regeneration.
- The quantity of steam needed in the ZnFe-sorbent media lift-pipe is 0.085 lb of steam/lb of solids being transported.
- In order to minimize steam generation by the PFBC, all GT compressor discharge air (less that required internally) is sent to the PFBC as a heat sink (which resulted in 385% excess air within the PFBC).
- Solid feed rates corresponding to a Ca/S ratio of 1.5 result in 96% sulfur capture within the PFBC operating at 1700 °F.
- A waste-steam stream is produced by the PFBC in order to preclude excessive GT combustor temperatures.
- Flue gas from the PFBC is sent to the GT and is expanded through the turbine.
- The GT is an LM5000 with a firing temperature of 2200 °F.
- Injection steam flow is set at 15% of the compressor air flow. The split between turbine and combustor steam-injection flows is 79 and 21%, respectively.
- The GT combustor injection-steam temperature is held at 514 °F to provide a heat sink in order to avoid overheating the combustor.
- The stack-gas temperature, including the contribution from waste-steam stream, is 293 °F.

Case 1b

Case 1b was a duplicate of Case 1 except that a steam turbine power cycle was added downstream of the HRSG -- effectively making this a STIG-based, combined-cycle system. The incentive for this sub-option was that it provided a means to utilize the large heat release occurring in the PFBC and avoid the waste-steam stream required in Cases 1 and 1a, and to reduce the size of the PFBC and associated ducting of Case 1a. The following were the assumptions and key operating conditions pertinent to the ASPEN analysis of Case 1b:

- The reference coal composition is used as input to the gasifier.
- The math model's computations (see Chapter 4, Section 4.6) define the two-stage gasifier's gas compositions and yields of 32% top-gas and 68% side-gas at temperatures of 800 and 1600 °F, respectively.
- The operating pressure of the gasifier is 612 psia.
- The steam-to-coal ratio equals .84, and the air-to-coal ratio equals 2.12.
- Five percent of the coal fed to the gasifier is elutriated as fines, which are subsequently combusted in the PFBC.
- ZnFe sulfur capture is not completely effective, and 100 ppm of sulfur exits in the fuel gas stream. ZnFe sorbent utilization is 40%. No tar cracking was considered to be required.
- A 100 °F temperature approach is assumed in the desulfurizer for the water-gas shift reaction. Gas temperature at the desulfurizer inlet is held down to 1140 °F by water quenching.
- Regeneration of the ZnFe sorbent is accomplished stoichiometrically in the regenerator, which operates at 1400 °F.
- No sulfate formation occurs within the ZnFe during regeneration.
- The quantity of steam needed in the ZnFe-sorbent media lift-pipe is 0.085 lb of steam/lb of solids being transported.
- Twenty-percent excess air is utilized in the PFBC.
- Solid feed rates corresponding to a Ca/S ratio of 1.5 result in 96% sulfur capture within the PFBC operating at 1700 °F.

- The waste-steam stream from the PFBC is superheated and is expanded through a steam turbine.
- Flue gas from the PFBC is sent to the GT and is expanded through the turbine.
- The GT is an LM5000 with a firing temperature of 2200 °F.
- Injection steam flow is set at 15% of the compressor air flow. The split between turbine and combustor steam injection flows is 79 and 21%, respectively.
- The GT combustor injection-steam temperature is held at 514 °F to provide a heat sink in order to avoid overheating the combustor.
- The stack-gas temperature is 284 °F.

CASE 2 - TWO-STAGE GASIFIER WITH STAG CYCLE

The system configuration for Case 2 contains a two-stage, dry-bottom, prototype gasifier operating at 20-atm pressure; a PFBC fueled by raw top-gas from the gasifier plus 5% of the coal entering the gasifier (the assumed dust carry-over in the top-gas stream); and a STAG power cycle. The configuration analyzed did not include a hot cyclone in the side-gas stream, but that would be preferred in a real system. The following were the assumptions and key operating conditions pertinent to the ASPEN analysis of Case 2:

- The reference coal composition is used as input to the gasifier.
- The math model's computations (see Chapter 4, Section 4.6) define the two-stage gasifier's gas compositions and yields of 32% top-gas and 68% side-gas at temperatures of 800 and 1600 °F, respectively.
- The operating pressure of the gasifier is 285 psia.
- The steam-to-coal ratio equals .84, and the air-to-coal ratio equals 2.12.
- Five percent of the coal fed to the gasifier is elutriated as fines, which are subsequently combusted in the PFBC.
- ZnFe sulfur capture is not completely effective, and 100 ppm of sulfur exits in the fuel-gas stream. ZnFe sorbent utilization is 40%. No tar cracking was considered to be required.

- A 100 °F temperature approach is assumed in the desulfurizer for the water-gas shift reaction. Gas temperature at the desulfurizer inlet is held down to 1140 °F by water quenching.
- Regeneration of the ZnFe sorbent is accomplished stoichiometrically in the regenerator, which operates at 1400 °F.
- No sulfate formation occurs within the ZnFe during regeneration.
- The quantity of steam needed in the ZnFe-sorbent media lift-pipe is 0.085 lb of steam/lb of solids being transported.
- Twenty-percent excess air is utilized in the PFBC.
- Solid feed rates corresponding to a Ca/S ratio of 1.5 result in 96% sulfur capture within the PFBC operating at 1700 °F.
- Flue gas from the PFBC is sent to the GT and expanded through the turbine.
- The GT is an MS7000F with a firing temperature of 2300 °F.
- The ST throttle conditions are 1450 psia and 1000 °F, and the ST includes a single reheat to 1000 °F.
- The stack-gas temperature is 284 °F.

CASE 3 - ONE-STAGE GASIFIER WITH STIG CYCLE

The system configuration for Case 3 is a plant buying only sized coal and containing a one-stage, dry-bottom, prototype gasifier operating at 40-atm pressure; a PFBC fueled by 5% of the coal entering the gasifier (the resulting dust carry-over); and a STIG power cycle. The following were the assumptions and key operating conditions pertinent to the ASPEN analysis of Case 3:

- The reference coal composition is used as input to the gasifier.
- The one-stage gasifier output-gas yield, temperature of 1100 °F, and product-gas composition are derived from measurements made on the METC gasifier while gasifying Pittsburgh No. 8 coal.
- The operating pressure of the gasifier is 612 psia.
- The steam-to-coal ratio equals .84, and the air-to-coal ratio equals 2.12.

- Five percent of the coal fed to the gasifier is elutriated as fines, which are subsequently combusted in the PFBC.
- ZnFe sulfur capture is not completely effective, and 100 ppm of sulfur exits in the fuel-gas stream. The ZnFe sorbent utilization is 40%. The sorbent is incorporated on a zeolite pellet, which provides the tar cracking function.
- A 100 °F temperature approach is assumed in the desulfurizer for the water-gas shift reaction.
- Regeneration of the ZnFe sorbent is accomplished stoichiometrically in the regenerator, which operates at 1400 °F.
- No sulfate formation occurs within the ZnFe during regeneration.
- The quantity of steam needed in the ZnFe-sorbent media lift-pipe is 0.085 lb of steam/lb of solids being transported.
- Total cracking of tar occurs in the tar cracking/desulfurization vessel. The resulting carbon that is laid down on the sorbent media is completely burned off in the regenerator.
- In order to minimize the PFBC size, only fines separated from the product-gas stream are combusted in the PFBC. No other fines are available.
- Twenty-percent excess air is utilized in the PFBC.
- Solid feed rates corresponding to a Ca/S ratio of 1.5 result in 96% sulfur capture within the PFBC operating at 1700 °F.
- Flue gas from the PFBC is sent to the GT and is expanded through the turbine.
- The GT is an LM5000 with a firing temperature of 2200 °F.
- Injection steam flow is set at 15% of the compressor air flow. The split between turbine and combustor steam injection flows is 79 and 21%, respectively.
- The GT combustor injection steam temperature is held at 514 °F to provide a heat sink in order to avoid overheating the combustor.
- The stack-gas temperature, including the contribution from the waste steam stream, is 292 °F.

CASE 4 - ONE-STAGE GASIFIER WITH STAG CYCLE

The system configuration for Case 4 is a plant buying only sized coal and containing a one-stage, dry-bottom, prototype gasifier operating at 20-atm pressure; a PFBC fueled by 5% of the coal entering the gasifier (the resulting dust carry-over), which is operating with solids feed rates corresponding to a Ca/S ratio of 2.0 and at 50% excess air (which was likely unnecessary but it assured the flue-gas oxygen content and temperature levels were more consistent with the GT's needs); and a STAG power cycle. Two sub-options were considered for this configuration and are presented subsequently. The following were the assumptions and key operating conditions pertinent to the ASPEN analysis of Case 4:

- The reference coal composition is used as input to the gasifier.
- The one-stage gasifier output-gas yield, temperature of 1100 °F, and product-gas composition are derived from measurements made on the METC gasifier while gasifying Pittsburgh No. 8 coal.
- The operating pressure of the gasifier is 285 psia.
- The steam-to-coal ratio equals .84, and the air-to-coal ratio equals 2.12.
- Five percent of the coal fed to the gasifier is elutriated as fines, which are subsequently combusted in the PFBC.
- ZnFe sulfur capture is not completely effective, and 100 ppm of sulfur exits in the fuel-gas stream. The ZnFe sorbent utilization is 40%. The sorbent is incorporated on a zeolite pellet, which provides the tar cracking function.
- A 100 °F temperature approach is assumed in the desulfurizer for the water-gas shift reaction.
- Regeneration of the ZnFe sorbent is accomplished stoichiometrically in the regenerator, which operates at 1400 °F.
- No sulfate formation occurs within the ZnFe during regeneration.
- The quantity of steam needed in the ZnFe-sorbent media lift-pipe is 0.085 lb of steam/lb of solids being transported.
- Total cracking of tar occurs in the tar cracking/desulfurization vessel. The resulting carbon that is laid down on the sorbent media is completely burned off in the regenerator.

- In order to minimize the PFBC size, only fines separated from the product-gas stream are combusted in the PFBC. No other fines are available.
- Fifty-percent excess air is utilized in the PFBC.
- Solid feed rates corresponding to a Ca/S ratio of 2.0 result in 96% sulfur capture within the PFBC operating at 1700 °F.
- Flue gas from the PFBC is sent to the GT and is expanded through the turbine.
- The GT is an MS7000F with a firing temperature of 2300 °F.
- The ST throttle conditions are 1450 psia and 1000 °F, and the ST includes a single reheat to 1000 °F.
- The stack-gas temperature is 284 °F.

Case 4a

Case 4a was a duplicate of Case 4 except that it assumed normal ROM coal was purchased by the plant, and the plant had a coal screening capability such that only sized coal was fed to the gasifier. A result of this was that the PFBC was fueled by 30% of the coal delivered to the plant (an assumed fines removal fraction), which represented 42.9% of the gasifier's feed rate, plus 5% of the sized coal that entered the gasifier (the resulting dust carry-over), and again operated with a Ca/S ratio of 2.0 and at 50% excess air. The incentive for this sub-option was that it could lower fuel costs (on an otherwise fairly attractive system), and it provided a good comparison case to a briquetting case (which was not actually analyzed). The following were the assumptions and key operating conditions pertinent to the ASPEN analysis of Case 4a:

- The reference coal composition is used as input to the gasifier.
- The one-stage gasifier output-gas yield, temperature of 1100 °F, and product-gas composition are derived from measurements made on the METC gasifier while gasifying Pittsburgh No. 8 coal.
- The operating pressure of the gasifier is 285 psia.
- The steam-to-coal ratio equals .84, and the air-to-coal ratio equals 2.12.
- Five percent of the coal fed to the gasifier is elutriated as fines, which are subsequently combusted in the PFBC.

- ZnFe sulfur capture is not completely effective, and 100 ppm of sulfur exits in the fuel-gas stream. The ZnFe sorbent utilization is 40%. The sorbent is incorporated on a zeolite pellet, which provides the tar cracking function.
- A 100 °F temperature approach is assumed in the desulfurizer for the water-gas shift reaction.
- Regeneration of the ZnFe sorbent is accomplished stoichiometrically in the regenerator, which operates at 1400 °F.
- No sulfate formation occurs within the ZnFe during regeneration.
- The quantity of steam needed in the ZnFe-sorbent media lift-pipe is 0.085 lb of steam/lb of solids being transported.
- Total cracking of tar occurs in the tar cracking/desulfurization vessel. The resulting carbon that is laid down on the sorbent media is completely burned off in the regenerator.
- Thirty percent of the coal to the plant is fed to the PFBC.
- Fifty-percent excess air is utilized in the PFBC.
- Solid feed rates corresponding to a Ca/S ratio of 2.0 result in 96% sulfur capture within the PFBC operating at 1700 °F.
- Flue gas from the PFBC is sent to the GT and is expanded through the turbine.
- The GT is an MS7000F with a firing temperature of 2300 °F.
- The ST throttle conditions are 1450 psia and 1000 °F, and the ST includes a single reheat to 1000 °F.
- The stack-gas temperature is 284 °F.

Case 4b

Case 4b was a duplicate of Case 4 except that the PFBC was operated with a Ca/S ratio of 1.5 and at 20% excess air. The incentive for this sub-option was to lower the cost of the PFBC and associated ducting and to be consistent with most of the other cases (and because in most instances it was felt the GT could accommodate the preheated combustion air resulting from these PFBC conditions). The following were the assumptions and key operating conditions pertinent to the ASPEN analysis of Case 4b:

- The reference coal composition is used as input to the gasifier.
- The one-stage gasifier output-gas yield, temperature of 1100 °F, and product-gas composition are derived from measurements made on the METC gasifier while gasifying Pittsburgh No. 8 coal.
- The operating pressure of the gasifier is 285 psia.
- The steam-to-coal ratio equals .84, and the air-to-coal ratio equals 2.12.
- Five percent of the coal fed to the gasifier is elutriated as fines, which are subsequently combusted in the PFBC.
- ZnFe sulfur capture is not completely effective, and 100 ppm of sulfur exits in the fuel-gas stream. The ZnFe sorbent utilization is 40%. The sorbent is incorporated on a zeolite pellet, which provides the tar cracking function.
- A 100 °F temperature approach is assumed in the desulfurizer for the water-gas shift reaction.
- Regeneration of the ZnFe sorbent is accomplished stoichiometrically in the regenerator, which operates at 1400 °F.
- No sulfate formation occurs within the ZnFe during regeneration.
- The quantity of steam needed in the ZnFe-sorbent media lift-pipe is 0.085 lb of steam/lb of solids being transported.
- Total cracking of tar occurs in the tar cracking/desulfurization vessel. The resulting carbon that is laid down on the sorbent media is completely burned off in the regenerator.
- In order to minimize the PFBC size, only fines separated from the product-gas stream are combusted in the PFBC. No other fines are available.
- Twenty-percent excess air is utilized in the PFBC.
- Solid feed rates corresponding to a Ca/S ratio of 1.5 result in 96% sulfur capture within the PFBC operating at 1700 °F.
- Flue gas from the PFBC is sent to the GT and is expanded through the turbine.
- The GT is an MS7000F with a firing temperature of 2300 °F.

- The ST throttle conditions are 1450 psia and 1000 °F, and the ST includes a single reheat to 1000 °F.
- The stack-gas temperature is 284 °F.

CASE 6 - RECYCLING, TWO-STAGE GASIFIER WITH STAG CYCLE

The system configuration for Case 6 contains a two-stage, dry-bottom, prototype gasifier operating at 20-atm pressure wherein the raw top-gas is recycled to a deep bed location within the gasifier; a PFBC fueled by 5% of the coal entering the gasifier (resulting from the assumed coal-dust carry-over in the single, mid-level, product-gas stream); and a STAG power cycle. The following were the assumptions and key operating conditions pertinent to the ASPEN analysis of Case 6:

- The reference coal composition is used as input to the gasifier.
- All of the 800 °F top-gas is recycled to a location just above the combustion zone in the gasifier, and raw product-gas exits from a single mid-level port in the gasifier sidewall. The math model's computations (see Chapter 4, Section 4.7) define the gasifier's output-gas composition, yield, and temperature of 1600 °F.
- The eductor steam is at 823 °F and 570 psia and is used to recycle the top-gas. The eductor steam-to-top-gas weight ratio is 0.4.
- Indirect cooling is used to cool the gasifier output gas to 1140 °F prior to the hot-gas desulfurization unit. A direct water quench was not utilized since the resulting fuel-gas heating value would be too low.
- The operating pressure of the gasifier is 285 psia.
- The steam-to-coal ratio equals 1.21, and the air-to-coal ratio equals 3.06.
- Five percent of the coal fed to the gasifier is elutriated as fines, which are subsequently combusted in the PFBC.
- ZnFe sulfur capture is not completely effective, and 100 ppm of sulfur exits in the fuel-gas stream. The ZnFe sorbent utilization is 40%. The sorbent is incorporated on a zeolite pellet, which provides the tar cracking function.
- A 100 °F temperature approach is assumed in the desulfurizer for the water-gas shift reaction.

- Regeneration of the ZnFe sorbent is accomplished stoichiometrically in the regenerator, which operates at 1400 °F.
- No sulfate formation occurs within the ZnFe during regeneration.
- The quantity of steam needed in the ZnFe-sorbent media lift-pipe is 0.085 lb of steam/lb of solids being transported.
- Total cracking of any existing tar occurs in the tar cracking and desulfurization vessel. The resulting carbon that is laid down on the sorbent media is completely burned off in the regenerator.
- In order to minimize the PFBC size, only fines separated from the product-gas stream are combusted in the PFBC. No other fines are available.
- Twenty-percent excess air is utilized in the PFBC.
- Solid feed rates corresponding to a Ca/S ratio of 1.5 result in 96% sulfur capture within the PFBC operating at 1700 °F.
- Flue gas from the PFBC is sent to the GT and is expanded through the turbine.
- The GT is an MS7000F with a firing temperature of 2300 °F.
- The ST throttle conditions are 1450 psia and 1000 °F, and the ST includes a single reheat to 1000 °F.
- The stack-gas temperature is 284 °F.

CASE 7 - SLAGGING, ONE-STAGE GASIFIER WITH STAG CYCLE

The system configuration for Case 7 contains a one-stage, prototype gasifier operating in an ash slagging mode at 20-atm pressure; all elutriated coal fines are recycled back through the tuyeres to the slagging zone of the gasifier; a PFBC, which is fired by a small coal-fines stream (approximately 1% of the gasifier feed rate) and a correspondingly small amount of GT compressor discharge air, provides just enough heat to adequately support the limestone calcination/sulfation reactions; and a STAG power cycle. The following were the assumptions and key operating conditions pertinent to the ASPEN analysis of Case 7:

- The reference coal composition is used as input to the gasifier.

- The one-stage gasifier operates in an ash-slugging mode using preheated blast air. The math model's computations (Chapter 4, Section 4.5) define the gasifier's output gas composition, yield, and temperature of 1800 °F.
- A direct water quench is used to cool the gasifier product gas to 1140 °F prior to the hot-gas desulfurization unit.
- The operating pressure of the gasifier is 285 psia.
- The steam-to-coal ratio equals 0.30, and the air-to-coal ratio equals 3.30.
- All elutriated fines are recycled back through the tuyeres to the slugging zone of the gasifier.
- ZnFe sulfur capture is not completely effective, and 100 ppm of sulfur exits in the fuel-gas stream. The ZnFe sorbent utilization is 40%. The sorbent is incorporated on a zeolite pellet, which provides the tar cracking function.
- A 100 °F temperature approach is assumed in the desulfurizer for the water-gas shift reaction.
- Regeneration of the ZnFe sorbent is accomplished stoichiometrically in the regenerator, which operates at 1400 °F.
- No sulfate formation occurs within the ZnFe during regeneration.
- The quantity of steam needed in the ZnFe-sorbent media lift-pipe is 0.085 lb of steam/lb of solids being transported.
- Total cracking of any existing tar occurs in the tar cracking and desulfurization vessel. The resulting carbon that is laid down on the sorbent media is completely burned off in the regenerator.
- The PFBC does not generate any steam. A small amount of fines is obtained from the coal handling plant section and is sent to the PFBC to provide the energy necessary for calcination and sulfation.
- Twenty-percent excess air is utilized in the PFBC.
- Solid feed rates corresponding to a Ca/S ratio of 1.5 result in 96% sulfur capture within the PFBC operating at 1700 °F.
- Flue gas from the PFBC is sent to the GT and is expanded through the turbine.

- The GT is an MS7000F with a firing temperature of 2300 °F.
- The ST throttle conditions are 1450 psia and 1000 °F, and the ST includes a single reheat to 1000 °F.
- The stack-gas temperature is 284 °F.

CASE 10 - ONE-STAGE GASIFIER WITH DSRP SUBSYSTEM AND STAG CYCLE

The system configuration for Case 10 is a plant buying only sized coal and containing a one-stage, dry-bottom, prototype gasifier operating at 20-atm pressure; the fueled limestone PFBC unit of the previous Cases is replaced with a DSRP subsystem; 5% of the coal entering the gasifier (the resulting dust carry-over) is not utilized within the plant and becomes a solid waste stream; and a STAG power cycle. The following were the assumptions and key operating conditions pertinent to the ASPEN analysis of Case 10:

- The reference coal composition is used as input to the gasifier.
- The one-stage gasifier output gas yield, temperature of 1100 °F, and product-gas composition are derived from measurements made on the METC gasifier while gasifying Pittsburgh No. 8 coal.
- The operating pressure of the gasifier is 285 psia.
- The steam-to-coal ratio equals .84, and the air-to-coal ratio equals 2.12.
- Five percent of the coal fed to the gasifier is elutriated as fines. These fines are not further utilized in the system.
- ZnFe sulfur capture is not completely effective, and 100 ppm of sulfur exits in the fuel-gas stream. The ZnFe sorbent utilization is 40%. The sorbent is incorporated on a zeolite pellet, which provides the tar cracking function.
- A 100 °F temperature approach is assumed in the desulfurizer for the water-gas shift reaction.
- Regeneration of the ZnFe sorbent is accomplished stoichiometrically in the regenerator, which operates at 1400 °F.
- No sulfate formation occurs within the ZnFe during regeneration.

- The quantity of steam needed in the ZnFe-sorbent media lift-pipe is 0.085 lb of steam/lb of solids being transported.
- Total cracking of tar occurs in the tar cracking/desulfurization vessel. The resulting carbon that is laid down on the sorbent media is completely burned off in the regenerator.
- The first stage of the DSRP operates at 1025 °F and the second stage at 480 °F. The inlet temperature to these stages is controlled by indirect heat exchange such that both stages are adiabatic.
- The amount of reducing gas required by the DSRP subsystem is determined from regenerator offgas composition such that enough CO and H₂ are available for conversion of 92% of the SO₂ to elemental sulfur in the first stage.
- Also in the first DSRP stage, SO₂ is reacted with H₂, CO, and H₂O to generate just enough H₂S for the second stage.
- In the second stage of the DSRP subsystem, 98% of the SO₂ left from the first stage is reacted with H₂S to generate elemental sulfur.
- Steam generated from the recovery of the molten sulfur from the DSRP subsystem is sent to the steam cycle.
- Offgas from the DSRP subsystem is sent to the GT and is expanded through the gas turbine.
- The GT is an MS7000F with a firing temperature of 2300 °F.
- The ST throttle conditions are 1450 psia and 1000 °F, and the ST includes a single reheat to 1000 °F.
- The stack-gas temperature is 284 °F.

CASE 11 - RECYCLING, TWO-STAGE GASIFIER WITH DSRP SUBSYSTEM AND STAG CYCLE

The system configuration for Case 11 contains a two-stage, dry-bottom, prototype gasifier operating at 20-atm pressure wherein the raw top-gas is recycled to a deep bed location within the gasifier; the fueled limestone PFBC unit of previous Cases is replaced with a DSRP subsystem; 5% of the coal entering the gasifier (the assumed dust carry-over in the single mid-level product-gas stream) is not utilized within the plant and becomes a solid

waste stream; and a STAG power cycle. The following were the assumptions and key operating conditions pertinent to the ASPEN analysis of Case 11:

- The reference coal composition is used as input to the gasifier.
- All of the 800 °F top-gas is recycled to a location just above the combustion zone in the gasifier, and raw product-gas exits from a single mid-level port in the gasifier sidewall. The math model's computations (see Chapter 4, Section 4.7) define the gasifier's output gas composition, yield, and temperature of 1600 °F.
- The eductor steam is at 823 °F and 570 psia and is used to recycle the top-gas. The eductor steam-to-top-gas weight ratio is 0.4.
- Indirect cooling is used to cool the gasifier output gas to 1140 °F prior to the hot-gas desulfurization unit. A direct water quench was not utilized since the resulting fuel-gas heating value would be too low.
- The operating pressure of the gasifier is 285 psia.
- The steam-to-coal ratio equals 1.21, and the air-to-coal ratio equals 3.06.
- Five percent of the coal feed to the gasifier is elutriated as fines. These fines are not further utilized in the system.
- ZnFe sulfur capture is not completely effective, and 100 ppm of sulfur exits in the fuel-gas stream. The ZnFe sorbent utilization is 40%. The sorbent is incorporated on a zeolite pellet, which provides the tar cracking function.
- A 100 °F temperature approach is assumed in the desulfurizer for the water-gas shift reaction.
- Regeneration of the ZnFe sorbent is accomplished stoichiometrically in the regenerator, which operates at 1400 °F.
- No sulfate formation occurs within the ZnFe during regeneration.
- The quantity of steam needed in the ZnFe-sorbent media lift-pipe is 0.085 lb of steam/lb of solids being transported.
- Total cracking of any existing tar occurs in the tar cracking and desulfurization vessel. The resulting carbon that is laid down on the sorbent media is completely burned off in the regenerator.

- The first stage of the DSRP operates at 1025 °F and the second stage at 480 °F. The inlet temperature to these stages is controlled by indirect heat exchange such that both stages are adiabatic.
- The amount of reducing gas required by the DSRP subsystem is determined from regenerator offgas composition such that enough CO and H₂ are available for conversion of 92% of the SO₂ to elemental sulfur in the first stage.
- Also in the first DSRP stage, SO₂ is reacted with H₂, CO, and H₂O to generate just enough H₂S for the second stage.
- In the second stage of the DSRP subsystem, 98% of the SO₂ left from the first stage is reacted with H₂S to generate elemental sulfur.
- Steam generated from the recovery of the molten sulfur from the DSRP subsystem is sent to the steam cycle.
- Offgas from the DSRP subsystem is sent to the GT and is expanded through the gas turbine.
- The GT is an MS7000F with a firing temperature of 2300 °F.
- The ST throttle conditions are 1450 psia and 1000 °F, and the ST includes a single reheat to 1000 °F.
- The stack-gas temperature is 284 °F.

Appendix B

Capital Cost Details

Tables 20 through 31 provide details for the costing analyses of the 12 systems. In each table, a reference design size, cost, and applicable cost/size exponential scaling factor were utilized with the (ASPEN-derived) total flow of the reference media through the particular plant section to arrive at the number of parallel trains (where applicable) and the basic cost for each section. A process contingency factor (which was applicable to the first plant and limited the learning-curve reductions) was defined, and the first plant and Nth plant section costs were developed. A 16% reduction in construction costs was taken from the Nth plant section cost subtotal, based on the use of improved construction techniques. A BOP cost was then added to arrive at the Total Process Capital. The Total Process Capital (Cost) represents the installed cost of all the purchased hardware and is the proper basis for system comparison in this study. Typical additions for items such as Project Contingency, Engineering Fees, Royalties, and AFDC were then calculated and summed to calculate the Total Capital Requirement (TCR) for the configuration. The TCR represents a *representative* cost for the Nth plant, which may or may not be applicable to a specific situation; the primary role for the TCR was as a basis with which to compute the COE.

Table 20. Case 1 System Capital Cost Details

(2-stage dry-bottom gasifier, ZnFe, maximum air to PFBC, STIG, 60 MWe power output)

Reference Plant Costs (per train)									
Section Description	Reference Flow	Stream Description	Ref. Cost (\$ millions)	Scaling Exponent	Max. Size Train	Total Flow Trains	Basic Section Cost (\$ millions)	Process Contingency (%)	1st Plant Estimated Section Cost (\$ millions)
Coal Handling	10000 tpd	Sized Coal	44	0.8	33000	768	5.65	0	5.65
Limestone Handling	1178 tpd	Limestone	3.24	0.7	3000	105	0.60	0	0.60
Gasification and Ash	467 k lb/hr	Total Gas	10.44	0.7	467	241.5	6.58	25	8.22
Fines Removal, Top	208.8 k lb/hr	Top Gas	0.468	0.7	467	78	0.23	0	0.23
Fines Removal, Side	208.8 k lb/hr	Side Gas	0.47	0.7	381.6	163.6	0.39	15	0.45
Sulfur Removal/Regen.	78.2 k lb/hr	Quenched Gas	1.33	0.7	467	181.5	2.40	70	4.07
PFBC	2.69 M lb/hr	Flue Gas	23.9	0.7	2.69	0.37	5.96	15	6.86
Gas Turbine	1 IM5000-STIG	Flue Gas	14.2	1	1	1	14.20	10	15.62
HPSG	439.5 k lb deg/s	Flue Gas	8.3	0.6	2000	203.2	5.22	0	5.22
Booster Compressor	100 lb/s	Air	1.80	0.95	300	117	2.06	0	2.06
Subtotal							43.29		48.99
Nth Plant Learning Curve Cost Reduction = 16% Due to Improved/Integrated Construction Methods:									
Subtotal After Learning:									
Balance of Plant Cost = 20% Subtotal Without Process Contingency:									
TOTAL PROCESS CAPITAL COST:									
Engineering Fees = 10% of Total Process Capital:									
Project Contingency = 15% of Total Process Capital:									
TOTAL PLANT INVESTMENT:									
AFDC Costs are Based on 3 Years Construction Time for First Plant, 2 Years for Nth Plant, Excluding Permit Time:									
Working Capital, Royalties, Miscellaneous = 7% of Total Plant Investment:									
The Land Cost is based on an Ash Output Rate: Land Cost = \$4452/tpd of ash:									
TOTAL CAPITAL REQUIREMENT:									

Table 21. Case 1a System Capital Cost Details

(2-stage dry-bottom gasifier, ZnFe, maximum air to PFBC, STIG, 51 MWe power output)

Section Description	Reference Plant Costs (per train)					Number of Trains	Basic Section Cost (\$ millions)	Process Contingency (%)	1st Plant Estimated Section Cost (\$ millions)	Cost Reduction Due to Learning (\$ millions)	Nth Plant Estimated Section Cost (\$ millions)	Nth Plant Estimated Section Cost (\$/kW)
	Reference Flow	Stream Description	Ref. Cost (\$ millions)	Scaling Exponent	Max. Size Train							
Coal Handling	10000 tpd	Sized Coal	44	0.8	33000	1	4.23	0	4.23	0	4.23	83
Limestone Handling	1178 tpd	Limestone	3.24	0.7	3000	1	0.46	0	0.46	0	0.46	9
Gasification and Ash	467 k lb/hr	Total Gas	10.44	0.7	467	1	5.10	25	6.38	25	4.78	94
Fines Removal, Top	208.8 k lb/hr	Top Gas	0.468	0.7	467	1	0.18	0	0.18	0	0.18	4
Fines Removal, Side	208.8 k lb/hr	Side Gas	0.47	0.7	381.6	1	0.31	15	0.35	15	0.30	6
Sulfur Removal/Regen.	78.3 k lb/hr	Quenched Gas	1.33	0.7	467	1	1.86	70	3.16	40	1.90	37
PFBC	2.69 M lb/hr	Flue Gas	23.9	0.7	2.69	1	10.60	15	12.19	15	10.36	203
Gas Turbine	1 LM5000-STIG	Flue Gas	14.2	1	1	1	14.20	10	15.62	10	14.06	276
HRSG	439.5 k lb deg/s	Flue Gas	8.3	0.6	2000	1	5.01	0	5.01	0	5.01	98
Booster Compressor	100 lb/s	Air	1.80	0.85	300	1	3.86	0	3.86	0	3.86	76
Subtotal							45.80		51.44		45.14	885
Nth Plant Learning Curve Cost Reduction = 16% Due to Improved/Integrated Construction Methods:												
Subtotal After Learning:												
Balance of Plant Cost = 20% Subtotal Without Process Contingency:												
TOTAL PROCESS CAPITAL COST:												
Engineering Fees = 10% of Total Process Capital:												
Project Contingency = 15% of Total Process Capital:												
TOTAL PLANT INVESTMENT:												
AFDC Costs are Based on 3 Years Construction Time for First Plant, 2 Years for Nth Plant, Excluding Permit Time:												
Working Capital, Royalties, Miscellaneous = 7% of Total Plant Investment:												
The Land Cost is based on an Ash Output Rate: Land Cost = \$4452/tpd of ash:												
TOTAL CAPITAL REQUIREMENT:												

Table 22. Case 1b System Capital Cost Details

(2-stage dry-bottom gasifier, ZnFe, PFBC, STIG, steam turbine, 77 MWe power output)

Section Description	Reference Plant Costs (per train)						Basic Section Cost (\$ millions)	Process Contin-gency (%)	1st Plant Estimated Section Cost (\$ millions)	Cost Reduction Due to Learning (\$)	Nth Plant Estimated Section Cost (\$ millions)	Nth Plant Estimated Section Cost (\$/kW)
	Reference Flow	Stream Description	Scaling Exponent	Ref. Cost (\$ millions)	Max. Size Train	Total Flow						
Coal Handling	10000 tpd	Sized Coal	0.8	44	33000	768	5.65	0	5.65	0	5.65	73
Limestone Handling	1178 tpd	Limestone	0.7	3.24	3000	105	0.60	0	0.60	0	0.60	8
Gasification and Ash	467 k lb/hr	Total Gas	0.7	10.44	467	241.5	6.58	25	8.22	25	6.17	80
Fines Removal, Top	208.8 k lb/hr	Top Gas	0.7	0.468	467	77.9	0.23	0	0.23	0	0.23	3
Fines Removal, Side	208.8 k lb/hr	Side Gas	0.7	0.47	381.6	163.6	0.40	15	0.46	15	0.39	5
Sulfur Removal/Regen.	78.3 k lb/hr	Quenched Gas	0.7	1.33	467	181.5	2.40	70	4.07	40	2.44	32
PFBC	2.69 M lb/hr	Flue Gas	0.7	23.9	2.69	0.365	5.90	15	6.79	15	5.77	75
Gas Turbine	1 LMS000-STIG		1	14.2	1	1	14.20	10	15.62	10	14.06	183
HRSG	439.5 k lb deg/s	Flue Gas	0.6	8.3	2000	203.2	5.22	0	5.22	0	5.22	68
Steam Turbine	30 MWe	Power	0.75	7.30	500	17	4.77	0	4.77	0	4.77	62
Booster Compressor	100 lb/s	Air	0.85	1.80	300	117	2.06	0	2.06	0	2.06	27
Subtotal							48.00		53.69		47.36	615
Nth Plant Learning Curve Cost Reduction = 16% Due to Improved/Integrated Construction Methods:												
Subtotal After Learning:												
Balance of Plant Cost = 20% Subtotal Without Process Contingency:												
TOTAL PROCESS CAPITAL COST:												
Engineering Fees = 10% of Total Process Capital:												
Project Contingency = 15% of Total Process Capital:												
TOTAL PLANT INVESTMENT:												
AFPC Costs are Based on 3 Years Construction Time for First Plant, 2 Years for Nth Plant, Excluding Permit Time:												
Working Capital, Royalties, Miscellaneous = 7% of Total Plant Investment:												
The Land Cost is Based on an Ash Output Rate: Land Cost = \$4452/tpd of ash:												
TOTAL CAPITAL REQUIREMENT:												

Table 23. Case 2 System Capital Cost Details
(2-stage dry-bottom gasifier, ZnFe, PFBC, STAG, 342 MWe power output)

Section Description	Reference Plant Costs (per train)										Nth Plant Estimated Section Cost (\$ millions)	Nth Plant Estimated Section Cost (\$/kW)
	Reference Flow	Stream Description	Ref. Cost (\$ millions)	Scaling Exponent	Max. Size Train	Total Flow	Number of Trains	Basic Section Cost (\$ millions)	Process Contingency (%)	1st Plant Estimated Section Cost (\$ millions)		
Coal Handling	10000 tpd	Sized Coal	44	0.8	33000	2545	1	14.72	0	14.72	0	43
Limestone Handling	1178 tpd	Limestone	3.24	0.7	3000	346	1	1.37	0	1.37	0	4
Gasification and Ash	246.1 k lb/hr	Total Gas	10.44	0.7	246.1	800	4	36.12	25	45.15	25	99
Fines Removal, Top	246.1 k lb/hr	Top Gas	0.847	0.7	246.1	258	2	1.08	0	1.08	0	3
Fines Removal, Side	246.1 k lb/hr	Side Gas	0.85	0.7	246.1	541.9	3	2.05	15	2.35	15	6
Sulfur Removal/Regen.	78.3 k lb/hr	Quenched Gas	1.33	0.7	246.1	600.9	3	7.70	70	13.09	40	23
PFBC	2.69 M lb/hr	Flue Gas	23.9	0.7	2.69	1.21	1	13.66	15	15.71	15	39
Gas Turbine	1 MS7000F-STAG		33.8	1	1	1	1	33.80	10	37.18	10	98
HRSG	500 k lb/s	Flue Gas	8.3	0.6	2000	1022	1	12.75	0	12.75	0	37
Steam Turbine	30 MWe	Power	7.30	0.75	500	191	1	29.26	0	29.26	0	86
Booster Compressor	100 lb/s	Air	1.80	0.85	300	387	2	6.31	0	6.31	0	18
Subtotal								158.81		178.97		456
Nth Plant Learning Curve Cost Reduction = 16% Due to Improved/Integrated Construction Methods:												
Subtotal After Learning:												
Balance of Plant Cost = 20% Subtotal Without Process Contingency:												
TOTAL PROCESS CAPITAL COST:												
Engineering Fees = 10% of Total Process Capital:												
Project Contingency = 15% of Total Process Capital:												
TOTAL PLANT INVESTMENT:												
AFDC Costs are Based on 3 Years Construction Time for First Plant, 2 Years for Nth Plant, Excluding Permit Time:												
Working Capital, Royalties, Miscellaneous = 7% of Total Plant Investment:												
The Land Cost is based on an Ash Output Rate; Land Cost = \$4452/tpd of ash:												
TOTAL CAPITAL REQUIREMENT:												

Table 24. Case 3 System Capital Cost Details

(1-stage dry-bottom gasifier, ZnFe, PFBC, STIG, 55 MWe power output)

Reference Plant Costs (per train)									
Section Description	Reference Flow	Stream Description	Ref. Cost (\$ millions)	Scaling Exponent	Max. Size Train	Total Flow	Number of Trains	Basic Section Cost (\$ millions)	Process Contin-gency (%)
Coal Handling	10000 tpd	Sized Coal	44	0.8	33000	537	1	4.24	0
Limestone Handling	1178 tpd	Limestone	3.24	0.7	3000	70	1	0.45	0
Gasification and Ash	467 k lb/hr	Raw Gas	9.496	0.7	467	167.6	1	4.63	0
Physical Cleanup	208.8 k lb/hr	Raw Gas	0.468	0.7	467	167.6	1	0.40	0
Sulfur Removal/Regen.	78.3 k lb/hr	Raw Gas	1.33	0.7	467	173.2	1	2.32	70
PFBC	2.69 M lb/hr	Flue Gas	23.9	0.7	2.69	0.071	1	1.38	15
Gas Turbine	1 LM5000-STIG	Flue Gas	14.2	1	1	1	1	14.20	10
HRSG	439.5 k lb deg/s	Flue Gas	8.3	0.6	2000	188	1	4.99	0
Booster Compressor	100 lb/s	Air	1.80	0.85	300	45	1	0.91	0
Subtotal								34.02	
Nth Plant Learning Curve Cost Reduction = 16% Due to Improved/Integrated Construction Methods:									
Subtotal After Learning:								37.35	
Balance of Plant Cost = 20% Subtotal Without Process Contingency:								6.80	
TOTAL PROCESS CAPITAL COST:								44.15	
Engineering Fees = 10% of Total Process Capital:								4.41	
Project Contingency = 15% of Total Process Capital:								6.62	
TOTAL PLANT INVESTMENT:								55.19	
AFDC Costs are Based on 3 Years Construction Time for First Plant, 2 Years for Nth Plant, Excluding Permit Time:								4.64	
Working Capital, Royalties, Miscellaneous = 7% of Total Plant Investment:								3.86	
The Land Cost is based on an Ash Output Rate: Land Cost = \$4452/tpd of ash:								0.68	
TOTAL CAPITAL REQUIREMENT:								64.37	
								49.65	903
								33.88	616
								5.42	99
								28.46	517
								6.80	124
								35.27	641
								3.53	64
								5.29	96
								44.08	801
								1.80	33
								3.09	56
								0.68	12

(1-stage dry-bottom gasifier, ZnFe, PFBC, STAG, 250 MWe power output)

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Table 26. Case 4a System Capital Cost Details

(1-stage dry-bottom gasifier, ZnFe, 30% fines to PFBC, STAG, 300 MWe power output)

Section Description	Reference Plant Costs (per train)						Basic Section Cost (\$ millions)	Process Contingency (%)	1st Plant Estimated Section Cost (\$ millions)	Cost Reduction Due to Learning (\$)	Nth Plant Estimated Section Cost (\$ millions)	Nth Plant Estimated Section Cost (\$/kW)
	Reference Flow	Stream Description	Ref. Cost (\$ millions)	Scaling Exponent	Max. Size Train	Total Flow						
Coal Handling	10000 tpd	Sized Coal	44	0.8	33000	2328	13.71	0	13.71	0	13.71	46
Limestone Handling	1178 tpd	Limestone	3.24	0.7	3000	403	1.53	0	1.53	0	1.53	5
Gasification and Ash	246.1 k lb/hr	Raw Gas	9.496	0.7	246.1	509	21.96	0	21.96	0	21.96	73
Physical Cleanup	246.1 k lb/hr	Raw Gas	0.947	0.7	246.1	509	1.96	0	1.96	0	1.96	7
Sulfur Removal/Regen.	78.3 k lb/hr	Raw Gas	1.23	0.7	246.1	505	6.82	70	11.59	40	6.95	23
PFBC	2.69 M lb/hr	Flue Gas	23.9	0.7	2.69	1.17	13.34	15	15.35	15	13.04	43
Gas Turbine	1 MS7000F-STAG		33.8	1	1	1	33.80	10	37.18	10	33.46	112
HRSG	500 k lb/s	flue Gas	8.3	0.6	2000	987	12.48	0	12.48	0	12.48	42
Steam Turbine	30 MWe	Power	7.30	0.75	500	157	26.46	0	26.46	0	26.46	88
Booster Compressor	100 lb/s	Air	1.80	0.85	300	384	6.27	0	6.27	0	6.27	21
Subtotal							138.32		148.48		137.82	459
Nth Plant Learning Curve Cost Reduction = 16% Due to Improved/Integrated Construction Methods:												
							Subtotal After Learning:					
							Balance of Plant Cost = 20% Subtotal Without Process Contingency:					
							TOTAL PROCESS CAPITAL COST:					
							Engineering Fees = 10% of Total Process Capital:					
							Project Contingency = 15% of Total Process Capital:					
							TOTAL PLANT INVESTMENT:					
							ATDC Costs are Based on 3 Years Construction Time for First Plant, 2 Years for Nth Plant, Excluding Permit Time:					
							Working Capital, Royalties, Miscellaneous = 7% of Total Plant Investment:					
							The Land Cost is based on an Ash Output Rate: Land Cost = \$4452/tpd of ash:					
							TOTAL CAPITAL REQUIREMENT:					
							256.84					
							201.93					
							673					

Table 27. Case 4b System Capital Cost Details

(1-stage dry-bottom gasifier, ZnFe, PFBC, STAG, 252 MWe power output)

Section Description	Reference Plant Costs (per train)					Number of Trains	Basic Section Cost (\$ millions)	Process Contingency (%)	1st Plant Estimated Section Cost (\$ millions)	Cost Reduction Due to Learning (\$)	Nth Plant Estimated Section Cost (\$ millions)	Nth Plant Estimated Section Cost (\$/kW)
	Reference Flow	Stream Description	Ref. Cost (\$ millions)	Scaling Exponent	Max. Size Train							
Coal Handling	10000 tpd	Sized Coal	44	0.8	33000	1	11.94	0	11.94	0	11.94	47
Limestone Handling	1178 tpd	Limestone	3.24	0.7	3000	1	1.10	0	1.10	0	1.10	4
Gasification and Ash	246.1 k lb/hr	Raw Gas	9.496	0.7	246.1	3	24.98	0	24.98	0	24.98	99
Physical Cleanup	246.1 k lb/hr	Raw Gas	0.847	0.7	246.1	3	2.23	0	2.23	0	2.23	9
Sulfur Removal/Regen.	78.3 k lb/hr	Raw Gas	1.33	0.7	246.1	3	7.76	70	13.13	40	7.91	31
PFBC	2.69 M lb/hr	Flue Gas	23.9	0.7	2.69	0.328	1	5.48	6.30	15	5.36	21
Gas Turbine	1 MS7000F-STAG		33.8	1	1	1	33.80	10	37.13	10	33.46	133
HRSG	500 k lb/s	Flue Gas	8.3	0.6	2000	985	1	12.47	0	0	12.47	49
Steam Turbine	30 MWe	Power	7.30	0.75	500	108	1	19.08	0	0	19.08	76
Booster Compressor	100 lb/s	Air	1.80	0.85	300	181	1	2.98	0	0	2.98	12
Subtotal							121.82		131.45		121.51	482
Nth Plant Learning Curve Cost Reduction = 16% Due to Improved/Integrated Construction Methods:												
Subtotal After Learning:												
Balance of Plant Cost = 20% Subtotal Without Process Contingency:												
TOTAL PROCESS CAPITAL COST:												
Engineering Fees = 10% of Total Process Capital:												
Project Contingency = 15% of Total Process Capital:												
TOTAL PLANT INVESTMENT:												
AFDC Costs are Based on 3 Years Construction Time for First Plant, 2 Years for Nth Plant, Excluding Permit Time:												
Working Capital, Royalties, Miscellaneous = 7% of Total Plant Investment:												
The Land Cost is based on an Ash Output Rate; Land Cost = \$4452/tpd of ash:												
TOTAL CAPITAL REQUIREMENT:												

Table 28. Case 6 System Capital Cost Details

(2-stage dry-bottom gasifier with top gas recycle, ZnFe, PFBC, STAG, 284 MWe power output)

Reference Plant Costs (per train)									
Section Description	Reference Flow	Stream Description	Ref. Cost (\$ millions)	Scaling Exponent	Max. Size Train	Total Flow	Number of Trains	Basic Section Cost (\$ millions)	Process Contingency (%)
Coal Handling	10000 tpd	Sized Coal	44	0.8	33000	2121	1	12.73	0
Limestone Handling	11" tpd	Limestone	3.24	0.7	3000	284	1	1.20	0
Gasification and Ash	246.1 k lb/hr	Internal Gas	10.44	0.7	246.1	1197.8	5	51.23	25
Fines Removal, Side	246.1 k lb/hr	Raw Side Gas	0.85	0.7	246.1	985	5	3.62	15
Sulfur Removal/Regen.	78.3 k lb/hr	Cooled Gas	1.33	0.7	246.1	977	4	11.80	70
PFBC	2.69 M lb/hr	Flue Gas	23.9	0.7	2.69	0.23	1	4.27	15
Gas Turbine	1 MS7000F-STAG	Flue Gas	33.8	1	1	1	1	33.80	10
HRSG	500 k lb/s	Flue Gas	8.3	0.6	2000	1032	1	12.82	0
Steam Turbine	30 MWe	Power	7.30	0.75	500	116	1	20.13	0
Booster Compressor	100 lb/s	Air	1.80	0.85	300	209	1	3.37	0
Raw Gas Cooler	439.5 k lb deg/s	Raw Gas	8.30	0.7	2000	124.7	1	3.44	0
Subtotal								158.40	
Nth Plant Learning Curve Cost Reduction = 16% Due to Improved/Integrated Construction Methods:									
Subtotal After Learning:									
Balance of Plant Cost = 20% Subtotal Without Process Contingency:									
TOTAL PROCESS CAPITAL COST:									
Engineering Fees = 10% of Total Process Capital:									
Project Contingency = 15% of Total Process Capital:									
TOTAL PLANT INVESTMENT:									
AFDC Costs are Based on 3 Years Construction Time for First Plant, 2 Years for Nth Plant, Excluding Permit Time:									
Working Capital, Royalties, Miscellaneous = 7% of Total Plant Investment:									
The Land Cost is based on an Ash Output Rate: Land Cost = \$445/tpd of ash:									
TOTAL CAPITAL REQUIREMENT:									

1st Plant Estimated Section Cost (\$ millions)	Process Contingency (%)	Basic Section Cost (\$ millions)	Number of Trains	Total Flow	Max. Size Train	Ref. Cost (\$ millions)	Scaling Exponent	Stream Description	Reference Flow	Section Description
12.73	0	12.73	1	2121	33000	44	0.8	Sized Coal	10000 tpd	Coal Handling
1.20	0	1.20	1	284	3000	3.24	0.7	Limestone	11" tpd	Limestone Handling
64.03	25	51.23	5	1197.8	246.1	10.44	0.7	Internal Gas	246.1 k lb/hr	Gasification and Ash
4.17	15	3.62	5	985	246.1	0.85	0.7	Raw Side Gas	246.1 k lb/hr	Fines Removal, Side
20.05	70	11.80	4	977	246.1	1.33	0.7	Cooled Gas	78.3 k lb/hr	Sulfur Removal/Regen.
4.91	15	4.27	1	0.23	2.69	23.9	0.7	Flue Gas	2.69 M lb/hr	PFBC
37.18	10	33.80	1	1	1	33.8	1	Flue Gas	1 MS7000F-STAG	Gas Turbine
12.82	0	12.82	1	1032	2000	8.3	0.6	Power	500 k lb/s	HRSG
20.13	0	20.13	1	116	500	7.30	0.75	Air	30 MWe	Steam Turbine
3.37	0	3.37	1	209	300	1.80	0.85	Raw Gas	100 lb/s	Booster Compressor
3.44	0	3.44	1	124.7	2000	8.30	0.7	Raw Gas	439.5 k lb deg/s	Raw Gas Cooler
134.03										Subtotal
154.92										
24.79	87									
130.13	458									
31.68										
31.68										
215.70										
161.81	570									
16.18	57									
24.27	85									
202.26	712									
8.27	29									
14.16	50									
2.66	9									
313.82										

AFDC Costs are Based on 3 Years Construction Time for First Plant, 2 Years for Nth Plant, Excluding Permit Time:

Working Capital, Royalties, Miscellaneous = 7% of Total Plant Investment:

The Land Cost is based on an Ash Output Rate: Land Cost = \$445/tpd of ash:

TOTAL PROCESS CAPITAL COST:

Engineering Fees = 10% of Total Process Capital:

Project Contingency = 15% of Total Process Capital:

TOTAL PLANT INVESTMENT:

TOTAL CAPITAL REQUIREMENT:

Table 29. Case 7 System Capital Cost Details

(1-stage slagging gasifier, ZnFe, PFBC, STAG, 253 MWe power output)

Section Description	Reference Plant Costs (per train)					Basic Section Cost (\$ millions)	Process Contin-gency (%)	1st Plant Estimated Section Cost (\$ millions)	Cost Reduction Due to Learning (\$)	Nth Plant Estimated Section Cost (\$ millions)	Nth Plant Estimated Section Cost (\$/kW)
	Reference Flow	Stream Description	Ref. Cost (\$ millions)	Scaling Exponent	Max. Size Train						
Coal Handling	10000 tpd	Sized Coal	44	0.8	33000	11.56	0	11.56	0	11.56	46
Limestone Handling	1178 tpd	Limestone	3.24	0.7	3000	1.07	0	1.07	0	1.07	4
Gasification and Ash	205.3 k lb/hr	Raw Gas	11.13	0.7	525.6	32.24	0	32.24	0	32.24	127
Physical Cleanup	246.1 k lb/hr	Raw Gas	0.947	0.7	525.6	2.16	15	2.49	15	2.11	8
Sulfur Removal/Regen.	78.3 k lb/hr	Quenched Gas	1.33	0.7	525.6	8.91	70	15.15	40	9.09	36
PFBC	2.69 M lb/hr	Flue Gas	23.9	0.7	2.69	2.71	15	3.11	15	2.65	10
Gas Turbine	1 MS7000F-STAG	Flue Gas	33.8	1	1	33.80	10	37.13	10	33.46	132
HRSG	500 k lb/s	Flue Gas	8.3	0.6	2000	12.66	0	12.66	0	12.66	50
Steam Turbine	30 MWe	Power	7.30	0.75	500	17.47	0	17.47	0	17.47	69
Booster Compressor	100 lb/s	Air	1.80	0.85	300	2.90	0	2.90	0	2.90	11
Subtotal						125.47		135.82		125.20	495
Nth Plant Learning Curve Cost Reduction = 16% Due to Improved/Integrated Construction Methods:											
Subtotal After Learning:											
Balance of Plant Cost = 20% Subtotal Without Process Contingency:											
TOTAL PROCESS CAPITAL COST:											
Engineering Fees = 10% of Total Process Capital:											
Project Contingency = 15% of Total Process Capital:											
TOTAL PLANT INVESTMENT:											
AFDC Costs are Based on 3 Years Construction Time for First Plant, 2 Years for Nth Plant, Including Permit Time:											
Working Capital, Royalties, Miscellaneous = 7% of Total Plant Investment:											
The Land Cost is based on an Ash Output Rate: Land Cost = \$452/tpd of ash:											
TOTAL CAPITAL REQUIREMENT:											

Table 30. Case 10 System Capital Cost Details

(1-stage dry-bottom gasifier, ZnFe, DSRP, STAG, 264 MWe power output)

Reference Plant Costs (per train)									
Section Description	Reference Flow	Stream Description	Ref. Cost (\$ millions)	Scaling Exponent	Max. Size Train	Total Flow	Number of Trains	Basic Section Cost (\$ millions)	Process Contingency (%)
Coal Handling	10000 tpd	Sized Coal	44	0.8	33000	2205	1	13.13	0
Gasification and Ash	246.1 k lb/hr	Raw Gas	9.496	0.	246.1	689	3	27.14	0
Physical Cleanup	246.1 k lb/hr	Raw Gas	0.847	0.7	246.1	689	3	2.42	0
Sulfur Removal/Regen.	78.3 k lb/hr	Raw Gas	1.33	0.7	246.1	683	3	8.42	70
Gas Turbine	1 MS7000F-STAG		33.8	1	1	1	1	33.30	10
BRSG	500 k lb/s	Flue Gas	8.3	0.6	2000	989	1	12.50	0
Steam Turbine	33 MWe	Power	7.30	0.75	500	118	1	20.39	0
Booster Compressor	100 lb/s	Air	1.80	0.85	300	171	1	2.84	0
DSRP	5958 lb mol/hr	Regen. Gas	6.23	0.6	10000	7942	1	7.40	85
Subtotal								128.04	
Nth Plant Learning Curve Cost Reduction = 16% Due to Improved/Integrated Construction Methods:									
Subtotal After Learning:									
Balance of Plant Cost = 20% Subtotal Without Process Contingency:									
TOTAL PROCESS CAPITAL COST:									
Engineering Fees = 10% of Total Process Capital:									
Project Contingency = 15% of Total Process Capital:									
TOTAL PLANT INVESTMENT:									
AFDC Costs are Based on 3 Years Construction Time for First Plant, 2 Years for Nth Plant, Excluding Permit Time:									
Working Capital, Royalties, Miscellaneous = 7% of Total Plant Investment:									
The Land Cost is based on an Ash Output Rate; Land Cost = \$4452/tpd of ash:									
TOTAL CAPITAL REQUIREMENT:									

Section Description	Reference Flow	Stream Description	Ref. Cost (\$ millions)	Scaling Exponent	Max. Size Train	Total Flow	Number of Trains	Basic Section Cost (\$ millions)	Process Contingency (%)	1st Plant Estimated Section Cost (\$ millions)	Cost Reduction Due to Learning (\$)	Nth Plant Estimated Section Cost (\$ millions)	Nth Plant Estimated Cost (\$/kW)
Coal Handling	10000 tpd	Sized Coal	44	0.8	33000	2205	1	13.13	0	13.13	0	13.13	50
Gasification and Ash	246.1 k lb/hr	Raw Gas	9.496	0.	246.1	689	3	27.14	0	27.14	0	27.14	103
Physical Cleanup	246.1 k lb/hr	Raw Gas	0.847	0.7	246.1	689	3	2.42	0	2.42	0	2.42	9
Sulfur Removal/Regen.	78.3 k lb/hr	Raw Gas	1.33	0.7	246.1	683	3	8.42	70	14.32	40	8.59	33
Gas Turbine	1 MS7000F-STAG		33.8	1	1	1	1	33.30	10	37.18	10	3.46	127
BRSG	500 k lb/s	Flue Gas	8.3	0.6	2000	989	1	12.50	0	12.50	0	12.50	47
Steam Turbine	33 MWe	Power	7.30	0.75	500	118	1	20.39	0	20.39	0	20.39	77
Booster Compressor	100 lb/s	Air	1.80	0.85	300	171	1	2.84	0	2.84	0	2.84	11
DSRP	5958 lb mol/hr	Regen. Gas	6.23	0.6	10000	7942	1	7.40	85	13.69	40	8.22	31
Subtotal								128.04		143.61		128.69	487
Nth Plant Learning Curve Cost Reduction = 16% Due to Improved/Integrated Construction Methods:													
Subtotal After Learning:													
Balance of Plant Cost = 20% Subtotal Without Process Contingency:													
TOTAL PROCESS CAPITAL COST:													
Engineering Fees = 10% of Total Process Capital:													
Project Contingency = 15% of Total Process Capital:													
TOTAL PLANT INVESTMENT:													
AFDC Costs are Based on 3 Years Construction Time for First Plant, 2 Years for Nth Plant, Excluding Permit Time:													
Working Capital, Royalties, Miscellaneous = 7% of Total Plant Investment:													
The Land Cost is based on an Ash Output Rate; Land Cost = \$4452/tpd of ash:													
TOTAL CAPITAL REQUIREMENT:													

Table 31. Case 11 System Capital Cost Details

(2-stage dry-bottom gasifier with top-gas recycle, ZnFe, DSRP, STAG, 290 MWe power output)

Section Description	Reference Plant Costs (per train)									
	Reference Flow	Stream Description	Ref. Cost (\$ millions)	Scaling Exponent	Max. Size Train	Total Flow	Number of Trains	Basic Section Cost (\$ millions)	Process Contingency (%)	1st Plant Estimated Section Cost (\$ millions)
Coal Handling	10000 tpd	Sized Coal	44	0.8	33000	2304	1	13.60	0	13.60
Gasification and Ash	246.1 k lb/hr	Internal Gas	10.44	0.7	246.1	1301.5	6	57.34	25	71.68
Fines Removal, Side	246.1 k lb/hr	Raw Side Gas	0.95	0.7	246.1	1070	5	3.84	15	4.42
Sulfur Removal/Regen.	78.3 k lb/hr	Cooled Gas	1.33	0.7	246.1	1062	5	13.37	70	22.73
Gas Turbine	1 NS7000F-STAG		33.8	1	1	1	1	33.80	10	37.18
HRSG	500 k lb/s	Hot Gas	8.3	0.6	2000	1035	1	12.84	0	12.84
Steam Turbine	30 MWe	Lower	7.30	0.75	500	120	1	20.65	0	20.65
Booster Compressor	100 lb/s	Air	1.80	0.85	300	197	1	3.20	0	3.20
DSRP	5958 lb mol/hr	Regen. Gas	20.50	0.6	10000	4332	1	16.93	85	31.32
Raw Gas Cooler	439.5 k lb deg/s	Raw Side Gas	8.30	0.7	2000	135.5	1	3.64	0	3.64
Subtotal								179.22		221.26
Nth Plant Learning Curve Cost Reduction = 16% Due to Improved/Integrated Construction Methods:										
Subtotal After Learning:										
Balance of Plant Cost = 20% Subtotal Without Process Contingency:										
TOTAL PROCESS CAPITAL COST:										
Engineering Fees = 10% of Total Process Capital:										
Project Contingency = 15% of Total Process Capital:										
TOTAL PLANT INVESTMENT:										
ATDC Costs are Based on 3 Years Construction Time for First Plant, 2 Years for Nth Plant, Excluding Permit Time:										
Working Capital, Royalties, Miscellaneous = 7% of Total Plant Investment:										
The Land Cost is based on an Ash Output Rate: Land Cost = \$4452/tpd of ash:										
TOTAL CAPITAL REQUIREMENT:										

END

**DATE
FILMED**

08/08/91