CONCEPTUAL DESIGN AND ECONOMICS OF THE ADVANCED CO2 HYBRID POWER CYCLE

FINAL REPORT

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DISCLAIMER

ABSTRACT

Research has been conducted under United States Department of Energy Contract DE-FC26-02NT41621 to analyze the feasibility of a new type of coal-fired plant for electric power generation. This new type of plant, called the Advanced CO_2 Hybrid Power Plant, offers the promise of efficiencies nearing 36 percent, while concentrating CO_2 for 100% sequestration. Other pollutants, such as SO2 and NOx, are sequestered along with the CO_2 yielding a zero emissions coal plant.

The CO_2 Hybrid is a gas turbine-steam turbine combined cycle plant that uses CO_2 as its working fluid to facilitate carbon sequestration. The key components of the plant are a cryogenic air separation unit (ASU), a pressurized circulating fluidized bed gasifier, a CO_2 powered gas turbine, a circulating fluidized bed boiler, and a super-critical pressure steam turbine. The gasifier generates a syngas that fuels the gas turbine and a char residue that, together with coal, fuels a CFB boiler to power the supercritical pressure steam turbine. Both the gasifier and the CFB boiler use a mix of ASU oxygen and recycled boiler flue gas as their oxidant. The resulting CFB boiler flue gas is essentially a mixture of oxygen, carbon dioxide and water. Cooling the CFB flue gas to 80 deg. F condenses most of the moisture and leaves a CO_2 rich stream containing 3%v oxygen. Approximately 30% of this flue gas stream is further cooled, dried, and compressed for pipeline transport to the sequestration site (the small amount of oxygen in this stream is released and recycled to the system when the CO_2 is condensed after final compression and cooling). The remaining 70% of the flue gas stream is mixed with oxygen from the ASU and is ducted to the gas turbine compressor inlet. As a result, the gas turbine compresses a mixture of carbon dioxide (ca. 64%v) and oxygen (ca. 32.5%v) rather than air. This carbon dioxide rich mixture then becomes the gas turbine working fluid and also becomes the oxidant in the gasification and combustion processes. As a result, the plant provides CO_2 for sequestration without the performance and economic penalties associated with water gas shifting and separating CO_2 from gas streams containing nitrogen.

The cost estimate of the reference plant (the Foster Wheeler combustion hybrid) was based on a detailed prior study of a nominal 300 MWe *demonstration plant* with a 6F turbine. Therefore, the reference plant capital costs were found to be 30% higher than an estimate for a 425 MW fully commercial IGCC with an H class turbine (1438 $/kW vs. 1111 $/kW). Consequently, the capital cost of the CO_2 hybrid plant was found to be 25% higher than that of the IGCC with pre-combustion CO_2 removal (1892 $/kW vs. 1510 $/kW), and the levelized cost of electricity (COE) was found to be 20% higher (7.53 c/kWh vs. 6.26 c/kWh). Although the final costs for the CO_2 hybrid are higher, the study confirms that the relative change in cost (or mitigation cost) will be lower.

The conceptual design of the plant and its performance and cost, including losses due to CO_2 sequestration, is reported. Comparison with other proposed power plant CO_2 removal techniques reported by a December 2000 EPRI report is shown. This project supports the DOE research objective of development of concepts for the capture and storage of CO_2.

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EXECUTIVE SUMMARY

Introduction

The linkage between global climatic change and the emission of greenhouse gases such as carbon dioxide (CO_2) is well documented. Coal-fired power plants are some of the largest emitters of CO_2. To assure continued U.S. power generation from its abundant domestic coal resources, new coal combustion technologies for power plants with CO_2 removal and high efficiency must be developed to meet the future emissions standards, especially CO_2 sequestration. Reduction of CO_2 emission from power plants can be obtained by a new concept that has high power generation efficiency and low CO_2 mitigation cost.

Combustion of fossil fuels in a power plant will inevitably lead to CO_2 emissions. Environmental legislation is currently driving the development of power generation schemes that allow the CO_2 produced to be sequestered. Several possible alternatives exist for the CO_2 removal and sequestration. The simplest method, referred to as post-combustion capture, is merely to scrub the CO_2 from flue gas by chemical absorption. However, this is the most expensive option. One alternative is to apply pre-combustion capture in a pressurized gasification system to separate the CO_2 from a pressurized syngas stream before combustion. This method, where a water-gas shift reaction is first used to maximize the CO_2 partial pressure, allows for a physical absorption, which incurs less energy penalty than chemical absorption. Another approach is direct CO_2 removal in a CO_2 cycle, where fuel is combusted using pure O_2 so that the flue gas produced is essentially CO_2. A comparison of CO_2 removal methods is shown in Figure ES-1.

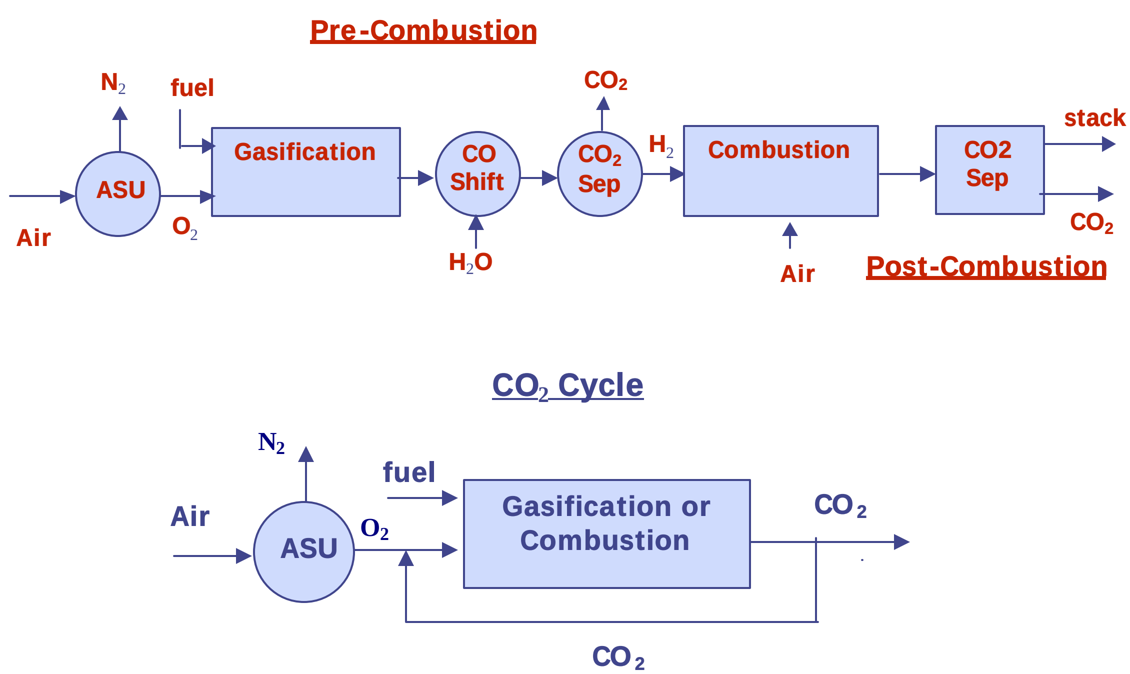

Figure ES-1. CO_2 Removal Methods

Published studies [1, 2] have shown that CO_2 removal/sequestration systems applied to the back end of a pulverized coal-fired plant can reduce its efficiency by up to 11 points with a resulting cost of $44 per tonne of CO_2 avoided (mitigation cost). For oxygen-blown gasification plants, carbon monoxide can be water gas shifted to hydrogen and CO_2 upstream of the gas turbine. The concentrated CO_2 is then separated by pre-combustion absorption, and regenerated by stripping or flashing. The resulting CO_2 stream is compressed to a pipeline pressure for sequestration.

There are both direct and indirect power (or efficiency) penalties associated with CO_2 removal. The direct penalty occurs as a result of the increase in auxiliary power requirements of CO_2 separation and compression. The indirect penalty is less obvious and refers to the gross power decrease due to CO_2 enrichment and removal processes. The systems net power and efficiency is reduced due to both direct and indirect penalties.

In the pre-combustion separation technique, the water-gas shift reaction is used to shift CO to CO_2 in order to concentrate the CO_2 in syngas. It has been well documented that this reaction reduces the syngas LHV while releasing its fuel energy as heat. Therefore, more syngas needs to be generated from gasification to compensate for the LHV loss by shifting. The low-grade heat from the shift reaction and syngas cooling before CO_2 absorption contributes to system energy loss. An efficiency loss of 6% to 7.7% and a CO_2 mitigation cost of between $21 and $23 per tonne[1] is estimated for such a power plant with a pre-combustion CO_2 removal system [1, 2]. To obtain a high CO conversion for more CO_2 to be removed, a ratio of $H_2O/CO > 2$ needs to be maintained by steam injection into the syngas. This steam can be extracted from the steam turbine, or generated from syngas cooling. Both ways reduce steam turbine power generation because of less steam flow to the bottoming cycle.

The loss of working fluid is another source of indirect penalty. In the pre-combustion separation technique, working fluid loss is caused by the removal of pressurized CO_2 from the syngas stream and by condensation of excess steam from the water gas shift. The losses of working fluids reduce power generation from the gas turbine because of less flow through the turbine. Indirect penalties can be quite significant and should be minimized to produce the maximum system efficiency.

The Advanced CO_2 Hybrid Cycle eliminates all indirect penalties by using a mixture of CO_2 recycled from the gas turbine exhaust together with oxygen as the working fluid. This facilitates straightforward concentration of CO_2 without enriching and separation processes. It eliminates the need for CO shifting, syngas cooling, absorption, and stripping and allows direct collection of CO_2 from recycled flue gas. It leads to a simpler CO_2 collection process than the conventional oxygen-blown pre-combustion CO_2 capture systems, while providing the advantages of lower energy cost and lower efficiency loss. Moreover, oxygen usage is minimized by recycling from the gas turbine exhaust most of excess oxygen along with the CO_2 back to the system.

While the CO2 Hybrid Cycle eliminates indirect energy costs, higher direct penalties are incurred due to the requirement for a larger air separation unit compared to an IGCC plant of the same net output. As the following sections will show, the net result is a 6.1% loss in cycle efficiency, and a mitigation cost (MC) of $18.3 per tonne[2] of CO_2, 11% lower than the best IGCC estimate.

The reference plant for the CO_2 hybrid cycle was chosen to be Foster Wheeler's combustion hybrid plant, known as Gasification Fluid-bed Combined Cycle (GFBCC) [3]. However, the CO_2 recycle concept can be applied just as readily to a conventional IGCC plant. This would entail enlarging the air separation unit to provide 100% stiochiometric oxygen, instead of 40 to 50%, and going to an oxygen and recycled CO_2 fired turbine leading to an enriched CO_2 stream that could be drawn off and sequestered. This is illustrated in Figure ES-2, which shows comparison of IGCC plant equipment with pre-combustion CO_2 removal (blue lines) to the CO_2 recycle concept (red lines). Compared to the CO_2 recycle plant, the IGCC plant requires the following additional processes:

[1] Cases 3A and 3E in references 1 and 2. Year 2000 dollars, 65% capacity factor.
[2] Year 2000 Dollars, 65% capacity factor

Steam extraction/generation for water gas shift
Water gas shift to enrich CO_2 content
Syngas cleanup
Syngas cooling before CO_2 absorption
Cooling water support
CO_2 absorption
CO_2 desorption
Syngas saturation

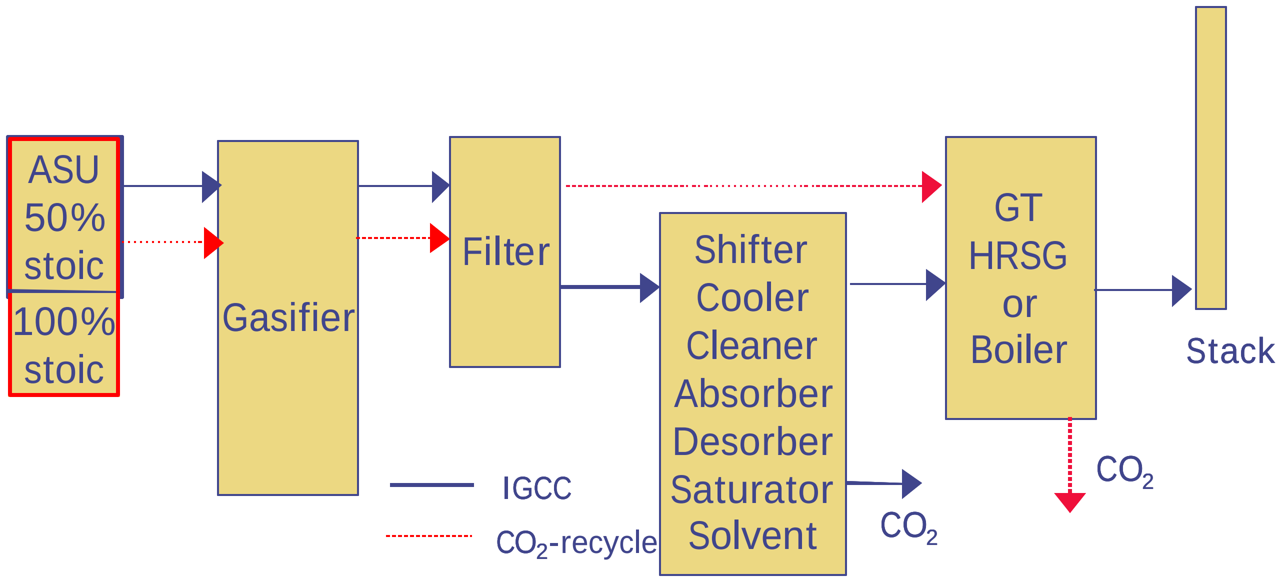

Figure ES-2. Comparison of CO_2 Removal Processes

Plant Description

The Foster Wheeler Hybrid CO_2 Cycle is a simple and efficient method of generating power while sequestering 100% of the CO_2 and other pollutants formed in the process. A Schematic of the process, based on Foster Wheeler's combustion hybrid technology called Gasification Fluid Bed Combined Cycle (GFBCC) [3], is shown on Figure ES-3.

The first step of the process is air separation, where oxygen is extracted from air for use in both the gasification and combustion processes. The oxygen is routed to the following equipment:

- In the **Partial Gasifier,** oxygen reacts with coal and steam to generate two fuel streams: Syngas and Char Residue.
- In the **Combustion Turbine (CT),** oxygen is used to burn the syngas generated by the PGM and drive the gas turbine generator.
- In the **CFB Steam Generator,** oxygen is used to burn the char generated in the PGM to make steam for the steam cycle.

To avoid the cost and energy penalty of CO_2 gas separation, the CO_2 is concentrated in the working fluid of the process by recycling the exhaust gas flow between the CFB combustor, CT, and the partial gasifier. This recirculation occurs in two main ways and provides for the exchange of energy between the topping and bottoming cycles:

- The gas turbine exhaust, which consists mostly of CO_2, O_2 and H_2O, is sent to the CFB combustor. The considerable sensible heat contained in the CT exhaust contributes to the CFB's steam production while its oxygen content is used to support the combustion of the char from the partial gasifier and the fresh coal directly sent to the CFB.

- The CO_2 rich flue gas from the CFB is first cooled to remove much of its moisture, then combined with the oxygen from the air separation unit and recycled to the gas turbine compressor inlet. Once pressurized, the recycled gas is sent to the partial gasifier as well as the CT combustor. The gasifier uses the recycle gas for its CO_2 and oxygen content to support the gasification reactions. The CT combustor uses the oxygen content of the recycled gas to support syngas combustion.

As the final step to balance the process, a portion of the recycle gas from the cooled CFB flue gas is diverted from the process for sequestration or disposal. This gas contains (by volume) about 94% CO2, 2% water, 3% oxygen, 0.7% trace gases (O_2, N_2) and .3% trace pollutants (oxide forms of sulfur and nitrogen, metals).

This exhaust stream is then compressed in three stages and further dried in intercoolers and a final dehydration unit. This process yields a dry mixture (< 50 ppmv moisture) of about 96% CO_2 and 3% O_2 at about 850 psia. At this pressure, the stream is cooled down to about 60 deg F and sent to a flash tank separator where the O_2 is removed in vapor form and reclaimed by the process. The CO_2, which is removed as condensate at the bottom of the flash tank, is pumped directly to a CO_2 pipeline and transported in liquid form to a sequestration site. In this study it has been assumed that trace pollutants will not be separated from the CO_2 (this needs further evaluation). Other then the nitrogen, and argon vent streams from the ASU and the coal ash discharge from the CFB combustor, there are no other waste streams from the process. **There is no plant stack and all waste streams including CO_2 from the process are in their most concentrated and manageable form.**

This concept, which offers an alternative method to achieve carbon sequestration from coal fired power plants, holds several unique benefits:

- A completely zero emission stackless plant that can produce power and a high pressure CO_2 exhaust stream with equal or better efficiency[3] than conventional gasification technology with pre-combustion CO_2 separation.

- CO_2 sequestration is achieved while avoiding the costly, energy-intensive CO shifting, CO_2 chemical/physical absorption, and CO_2 stripping processes used in conventional gasification technology.

- A wide range of inexpensive coals can be used as fuel since fluid bed technology is used for both the gasification and combustion process and gasification takes place at a modest 2000 deg. F

- All effluent stream from the process (SO2, CO2, NOx, N2, H2O, Metals, Ash) are concentrated for efficient reuse or disposal

- It is a simplified process offering higher reliability

[3] When the more realistic IGCC case with a water scrubber (ref. 2 case 3E) is considered, the CO2 cycle has a 0.7 percentage point (or 20%) higher efficiency, and incurs a lesser efficiency drop between no sequestration and with sequestration configurations (6.1% drop for CO2 cycle vs 7.7% for IGCC).

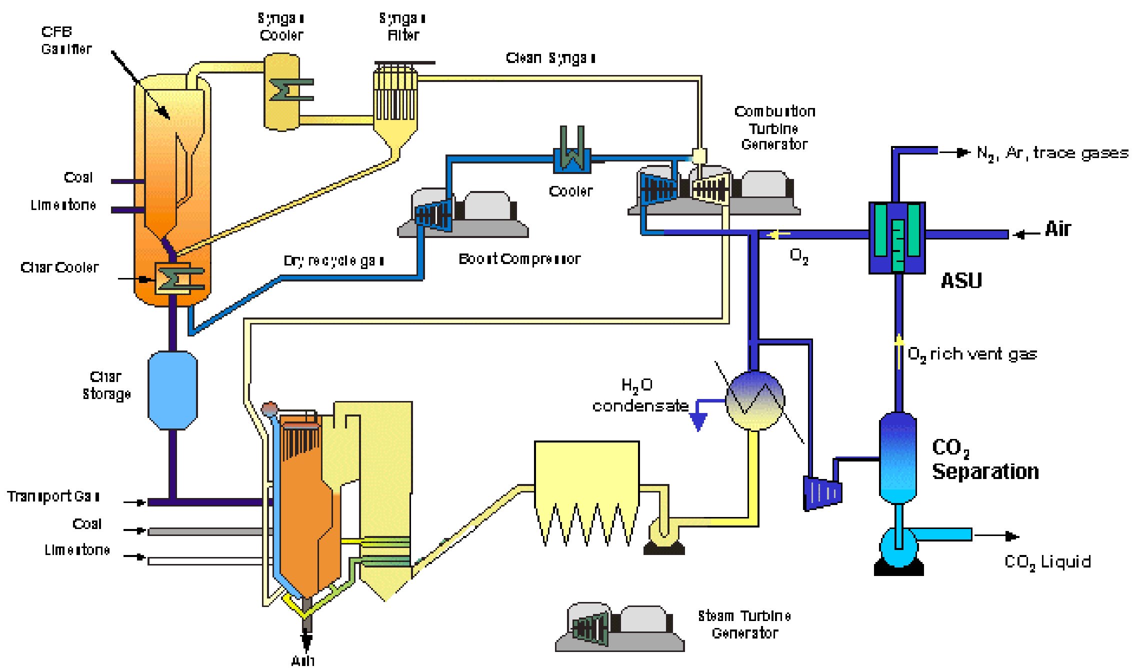

Figure ES-3. CO_2 Hybrid Cycle Process Flow Diagram

Performance and Economics

Table ES-1 shows a summary of the plant performance in comparison with IGCC plants from references [1] and [2]. The two IGCC plants shown are both e-gas gasifiers with H-class turbines employing pre-combustion CO_2 removal.

Notice that in spite of the significantly higher oxygen usage of the CO_2 cycle compared to the IGCC plants (1640 lbs/hr/MWe vs. 647 and 680), the plant efficiencies are on a par with each other. In fact, the authors of references [1], and [2] have characterized the IGCC with water scrubber as the "more realistic" case, which makes the CO_2 cycle efficiency 0.5 points better than an IGCC plant with pre-combustion CO_2 removal.

Figure ES-4 shows the loss of net efficiency when different CO_2 removal processes are added to a power plant [1, 2]. The loss of efficiency is caused by the reduction of power generation and/or the energy loss prior to turbine-generator. As expected, the PC boiler with post-combustion separation has the highest efficiency drop because of the energy intensive nature of the chemical absorption/regeneration.

The CO_2-hybrid process has among the lowest efficiency drops for CO_2 sequestration (compared to the air-blown hybrid cycle [3]) than an IGCC with pre-combustion CO_2 removal because the CO_2-hybrid cycle eliminates indirect losses such as working fluid loss, energy loss due to CO_2 enriching by water-gas shift, and CO_2 separation energy loss. The CO_2 separation penalty of the process can be further separated into two components: 1) due to CO_2 compression and 2) due to CO_2 removal processes, such as steam extraction, CO_2 enrichment and separation, and O_2/air separation as shown in Figure ES-4. CO_2

compression and air separation are direct losses, while steam extraction and CO_2 enrichment and separation are indirect losses. Note that with the exception of the CO_2 hybrid, the efficiency loss due to CO_2 separation is higher than the efficiency loss due to CO_2 compression.

Table ES-1. Comparison of the CO_2 Hybrid and IGCC Plants with CO_2 Sequestration

	CO2 Hybrid Plant	IGCC w/o Water Scrub.	IGCC with Water Scrub.
Plant Performance			
HHV Efficiency, %	35.9	37	35.4
Net Power, MWe	294.9	403.5	386.8
Gross Power, MWe			
Gas Turbine	129	345	345
Steam Turbine (3850/1050/1050/2"Hg)	270	144	127
Emissions			
SO2, lb/hr/MWe	0	0	0
NOx, lb/hr/MWe	0	0.25	0.26
CO2, lb/hr/MWe	0	162	169
Particulate	0	0	0
Oxygen Usage, lb/hr/MWe	1640	647	680
Oxygen Plant Cost, $/kWe (yr. 2000 dollars)	294	145	123
Illinois #6 Coal, lbs/hr/MWe	814	790	826

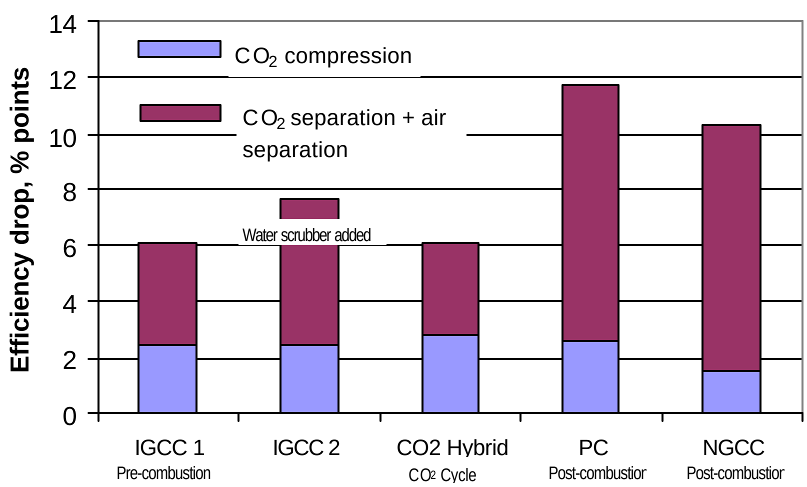

Figure ES-4. Efficiency Penalty Due to CO_2 Removal for Various Plant Types

However, in spite of its favorable energy efficiency, the capital cost of the CO_2 cycle was found to be higher than most IGCC estimates with CO_2 sequestration [1]. This is due to the higher cost of the base (reference) plant for the CO_2 hybrid cycle compared to the reference IGCC configurations. The cost estimate of the reference plant (the Foster Wheeler combustion hybrid) was based on a detailed prior study of a nominal 300 MWe *demonstration plant* with a 6F turbine. Therefore, the reference plant capital costs were found to be 30% higher than an estimate for a 425 MW fully commercial IGCC with an H class turbine (1438 $/kW vs. 1111 $/kW).

As shown on Table ES-2 in year 2000 dollars, the capital cost of the best IGCC estimate (with sequestration) was reported to be [2] $1510/kW[4] compared to $1884/kW for the CO_2 hybrid. Consequently, the levelized cost of electricity for the CO_2 hybrid is 20% higher than the best IGCC estimate ($75/MWh vs. $62.6/MWh).

Table ES-2. Plant Cost Comparisons With and Without CO_2 Capture *(year 2000 dollars, 65% capacity factor)*

	Without CO2 Removal			With CO2 Removal			Percent increase	
	Capital Cost $/kW	Net Output MW	Levelized COE c/kWh	Capital Cost $/kW	Net Output MW	Levelized COE c/kWh	Capital Cost	Levelized COE
IGCC (Audus)	1470	408	4.8	2200	382	6.9	49.66%	43.75%
IGCC (Herzog 2000)	1401		4.99	1909		6.69	36.25%	34.07%
IGCC (Simbeck)	1100	400	3.9	1474	400	5.1	34.00%	30.77%
IGCC (Parsons)	1111	425	4.77	1510	404	6.26	35.91%	31.24%
FW Combustion Hybrid	1438	300	6.09	1892	295	7.53	31.57%	23.65%

Values compiled from references [1] and [2]

This study also shows that the *increase* in both capital cost and the levelized cost of electricity in going from the reference plant to the carbon-removing configuration is lower for the CO_2 hybrid concept than for the IGCC with pre-combustion separation. Since the CO_2 recycle concept can be applied to a conventional IGCC just as readily, the merit of the concept should be judged largely based on the *change* in costs incurred in converting the reference plant to its new configuration with carbon sequestration.

Figures ES-5 and ES-6 illustrate that the relative increases, respectively, in capital cost and levelized cost of electricity incurred in going to the CO_2 capture configuration is lower for the CO_2 recycle concept.

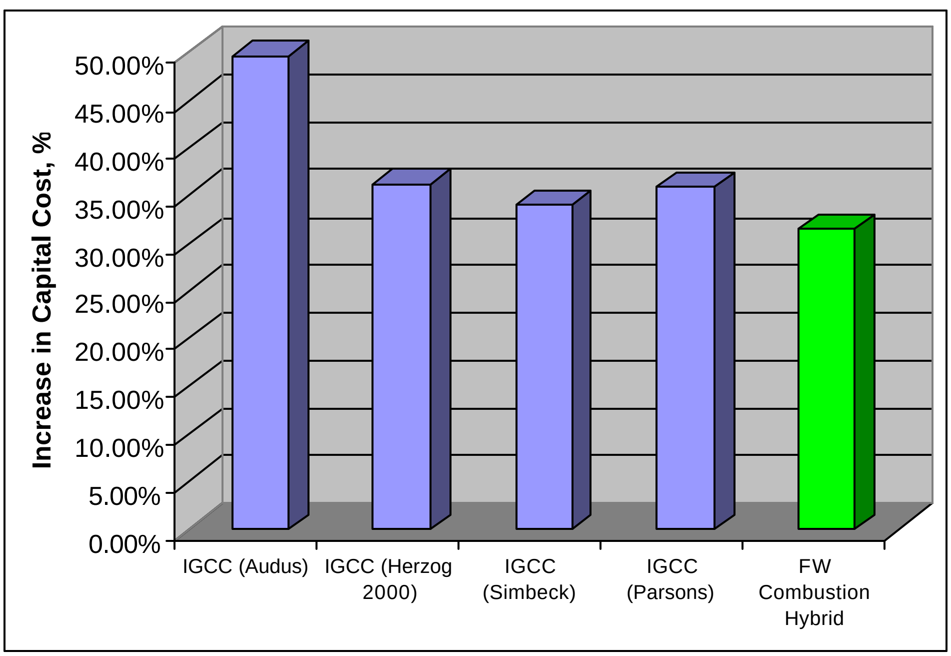

Figure ES-5. Capital Cost Increase in Going From the Base Plant to the CO_2 Capturing Configuration

[4] Case 3E, ref. 2

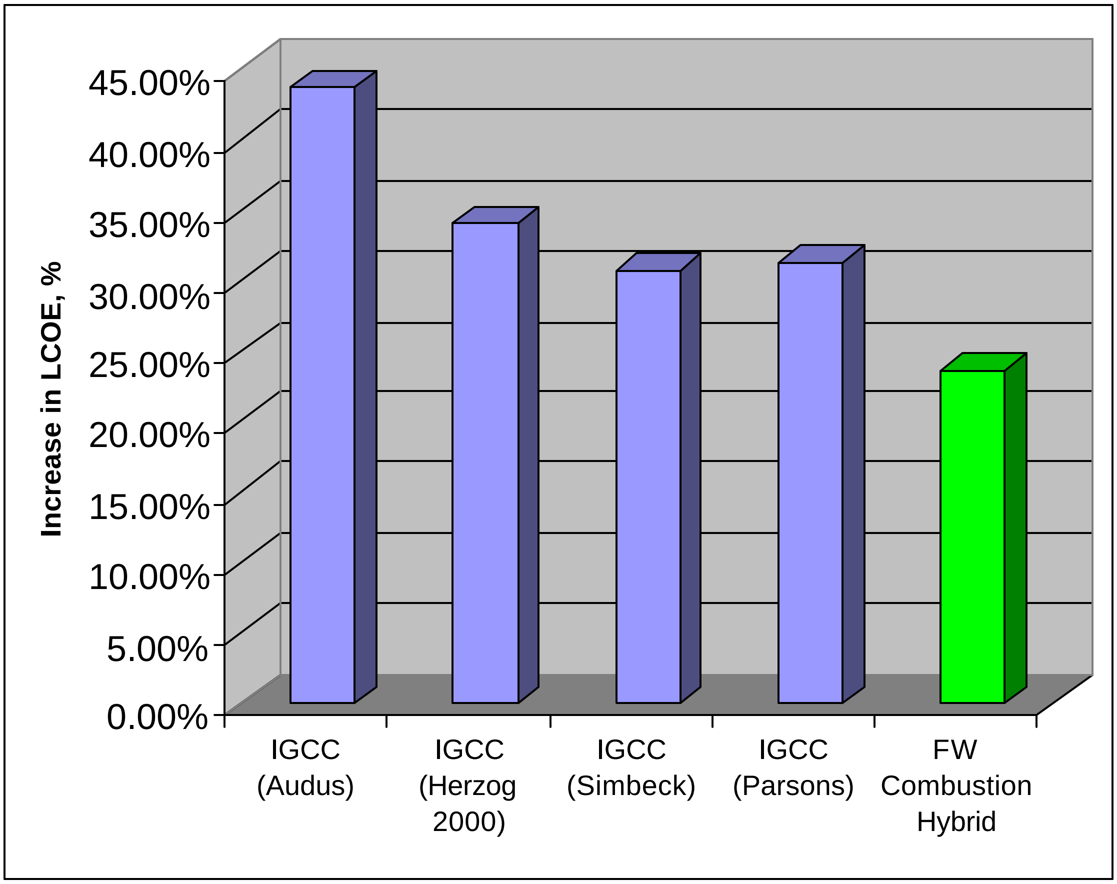

Figure ES-6. Impact of CO_2 Capture on the Levelized Cost of Electricity

Lower impact of the CO2 recycle concept on the reference plant also results in lower mitigation costs for the CO2 cycle (in $ per tonne of CO_2 avoided) compared to an IGCC with pre-combustion separation. This is illustrated in Table ES-3, which shows that the carbon mitigation cost of the CO2 cycle is 10% lower than the lowest IGCC estimate ($18.7 per tonne of CO2 compared to $20.6 per tonne for IGCC). Both cost numbers are based on year 2000 dollars and a capacity factor of 65%.

Table ES-3. CO_2 Mitigation Cost of Various Technologies *(year 2000 dollars)*

Technology	CO2 Hybrid Plant	NGCC with H Class Turbine	IGCC with H Class Turbine	IGCC[5] with H Class Turbine	Supercritical PC
$/tonne of CO_2 avoided at 65% CF	18.7	69.1	20.6	23.2	51.1
$/tonne of CO_2 avoided at 80% CF	13.5	48.2	17.5	19.5	44.1
% CO_2 removal	100	90	90	90	90

Technology Risks

An open issue in the current study is whether it is necessary to remove the sulfur oxide compounds, trace metals, and nitrogen oxide compounds from the CO_2 stream before being sent to the sequestration site. The form (liquid or gaseous) of the CO_2 is also an important factor in assessing CO_2 purity

[5] with water scrubber added

requirements, since these trace compounds can be very corrosive if mixed with liquid water. This issue has a significant impact on defining the CO_2 hybrid cycle process, plant cost, reliability, and efficiency.

There are two main areas of concern:

- **Transport Safety**. Pipeline and storage tank (stationary or mobile tanks for rail, barge, or truck transport) corrosion from trace contaminants needs investigation. Currently the U.S. Department of Transportation (DOT) regulates pipelines and transport of CO_2. Although it has not declared CO_2 as a hazardous material, the DOT has consistently adapted the language of "hazardous materials ***and*** carbon dioxide" [4]. This means that a higher level of inspection is demanded for CO_2 transport systems than for, say, crude oil transport. A compromised pipeline or tank presents the obvious hazard of releasing trace contaminants into the atmosphere along with CO_2, which is a ground hugging asphyxiate. Although Canadian experience[6] shows that up to 45% H_2S in CO_2 mixtures can be handled in pipelines there is no experience with SO_2 or other compounds at levels that might be expected from coal or coal derived syngas combustion [4]. An assessment will be conducted to determine the safety issues of transporting CO_2 with varying levels of the trace pollutants expected from the CO_2 Hybrid Power Plant.

- **Co-disposal of CO_2 with SOx, NOx, and Hg.** DOT will regulate SOx and NOx as hazardous materials. Therefore co-disposal of CO_2 with these contaminants must be investigated. Proposals to co-dispose of SO_2 and NOx are covered in 49 CFR 195.579, however, co-disposal issues need to be investigated beyond transportation safety requirements. The ecological impact of trace contaminants in the selected sequestration sinks must be understood. The main categories of CO_2 sequestration sinks are storage in terrestrial ecosystems, geologic formations and in oceans. An assessment will be conducted to determine the potential interaction between the sequestered CO_2 and the environment for each storage method.

A second and, perhaps, more significant risk is the availability of a gas turbine to meet the demands of the cycle. The aerodynamic and physical properties of the CO_2 rich working fluid necessitate a turbine with a pressure ratio of 49. This requires the use of the Advanced Inter-cooler Aero-derivative engine (ICAD). Although this engine is conceptual in nature and is not commercially available, the new LMS100 from GE comes close with a pressure ratio of 42. Considering the higher reported performance of the LMS100, it is likely that the required expansion can be achieved at a pressure ratio of 42.

Another gas turbine related issue, aside from the high pressure-ratio requirement, is that the CO_2 rich working fluid will require new airfoil and centerline designs for the compressor as well as the turbine. This is a costly product development effort and is only likely to happen if a clear market need is demonstrated. Therefore, although the LMS100 seems a close match, a commercial engine is currently not available for this technology.

In this study, the gas turbine performance was determined by making certain assumptions regarding the isentropic efficiency of the compression and expansion steps. Specifically, it was assumed that the engine, when designed for the CO_2 rich working fluid, would achieve the same efficiencies that are typical and reasonable for current air-fired units. The cost of the turbine was taken from a paper [5] that projected the development of ICAD engines capable of pressure ratios up to 49.

Conclusions

Although the potential of the CO_2 recycle concept to reduce the *incremental* costs of avoiding CO_2 emissions was demonstrated, the Foster Wheeler combustion hybrid cycle used as a reference for the

[6] The DOE, through a bilateral agreement with Natural Resources Canada is conducting a CO2 Sequestration field test at the Weyburn oil field in southeastern Saskatchewan.

study led to higher capital and levelized costs than the leading estimates for IGCC plants with pre-combustion CO2 removal.

It is believed that the cost of the reference plant is higher because the estimate is based on a demonstration plant. One can make the argument that application of the concept at a larger commercial scale could bring the cost down on a par with the IGCC cases used as comparison. However, there is currently no data available to substantiate this thought.

Other arguments in favor of the combustion hybrid concept are:

- Plant reliability would be higher due to simpler configuration compared to IGCC. This would imply that the combustion hybrid might be able to take credit for a higher capacity factor than the IGCC, changing the economic equation. Operating the CO2 hybrid cycle at 80% CF would match the cost of electricity of the best IGCC estimate derived at 65% CF.

- The combustion hybrid plant utilizes circulating fluidized beds for both gasification and combustion. Furthermore, the gasifier operates at a modest 2000 deg. F. In addition to improving availability, these attributes allow the technology to be highly fuel flexible and able to utilize a wide variety of inexpensive coals.

Gas turbine availability is also likely to be an issue in the commercialization of this concept. Gas turbine manufacturers must see a clear market demand for this technology to engage in costly development work for a high pressure-ratio engine that can efficiently use a CO2 rich working fluid. Market demand could be created if the concept can demonstrate a clear and substantial benefit compared to the IGCC with pre-combustion capture approach.

Looking Ahead

To further evaluate the merits and potential of the CO_2 recycle concept, a conventional IGCC plant should be used as a reference. The IGCC would then be converted to a CO2 cycle configuration by enlarging the air separation unit to provide 100% stiochiometric oxygen, instead of 40 to 50%, and going to an oxygen and recycled CO_2 fired turbine. This would enable an enriched CO_2 stream that could be drawn off and sequestered.

The economic feasibility of such a conversion can be directly compared with the same reference IGCC with pre-combustion CO_2 removal.

Air separation costs make up a significant portion of the cost adder to the base plant configuration. New developments in air separation, such as oxygen transport membranes (OTM) have the potential to reduce this cost. The feasibility of the concept should be evaluated with the OTM, and with customized and optimized cryogenic ASU designs integrated into the cycle (best possible cryogenic design).

Lastly, better gas turbine definition (design, performance, and economics) and gas turbine availability assessments are necessary to evaluate the future commercialization of the concept.

Section 1

INTRODUCTION

Research has been conducted under United States Department of Energy (DOE) contract DE-FC26-02NT41621 to develop the conceptual design and conduct a feasibility study of a new hybrid gasification/combustion power plant that can efficiency produce both power and a concentrated CO_2 exhaust stream for sequestration. The concept utilizes efficient combined cycle technology while avoiding the energy intensive process of separating CO_2 from syngas or flue gas. This new type of coal fired hybrid power plant promises high cycle efficiency, zero ambient air emissions, and complete capture of all of the CO_2 generated in the process.

In this plant coal is fed to a pressurized circulating fluid bed gasifier that produces a low-Btu syngas and char. After being cooled and passing through a barrier filter to remove the particulate the syngas is burned in an Advanced Inter-cooled Aero-derivative (ICAD) gas turbine. The char generated from the gasifier is depressurized and transferred to an atmospheric pressure circulating fluidized bed (ACFB) boiler, which generates supercritical steam to drive a steam turbine. The exhaust from the gas turbine is also fed to the ACFB and provides the oxidant for combustion. Both the gasification and combustion reactions use a mixture of recycled CO_2 and oxygen (generated by an air separation unit) as the oxidant, as well as the working fluid through the gas turbine and the rest of the cycle. In this manner, there is no nitrogen in the cycle and CO_2 is concentrated for direct removal from the exhaust stream. This leads to a simpler CO_2 collection process, which eliminates the need for CO shifting, syngas cooling, absorption, and stripping.

The fluid bed gasifier operates at a temperature of 2000 deg. F or less, which eliminates problems related to slag removal and allows for greater reliability, availability, and fuel flexibility compared to conventional IGCC systems.

The work has been carried out in 4 tasks:

- In Task 1 (Section 2) a conceptual design was developed and thermodynamic cycle analyses were conducted to determine the performance of the plant.

- In Task 2 (Section 3) the key components of the plant, namely the pressurized circulating bed gasifier and the atmospheric pressure circulating fluidized bed boiler, were designed in sufficient detail to allow cost estimation.

- In Task 3 (Sections 3 and 4) the balance of plant components were specified and the plant cost estimate was developed.

- Finally, in Task 4 (Section 4) an economic analysis was conducted where the plant capital cost, cost of electricity, and CO_2 mitigation cost were calculated and compared with IGCC, NGCC, and PC plants incorporating CO_2 sequestration.

Recommendations for further study include application of the CO_2 recycle concept to a conventional IGCC plant, conducting a gas turbine availability/feasibility study, and exploring the use of advanced air separation concepts.

Section 2

CONCEPTUAL DESIGN AND THERMODYNAMIC CYCLE ANALYSIS

2.1 MAJOR ASSUMPTIONS AND PROCESS CONDITIONS

During task 1 of the study the plant was conceptually designed and its performance and emissions were determined by thermodynamic cycle analyses.

A sea-level plant site was assumed with the following ambient design conditions:

Barometric pressure	14.7 psia
Dry bulb temperature	60 deg. F
Relative humidity	55%
Condenser Pressure	2” Hg

These conditions were used for performance calculations and to determine cycle efficiencies. For equipment design purposes, a temperature range of between 20 deg. F and 95 deg. F was assumed.

Illinois #6 coal (Table 2.1) was used as the fuel. It was assumed that all emissions, including SO2 and NOx would be sequestered along with the CO2. Therefore sorbent is not used in the plant, and all associated sorbent handling and feed facilities are eliminated.

Table 2-1. Design Coal Analysis – Illinois #6 Coal

Proximate Analysis	**As-received (wt%)**	**Dry Basis (wt%)**
Moisture	11.12	--
Ash	9.70	10.91
Volatile Matter	34.99	39.37
Fixed Carbon	44.19	49.72
TOTAL	100.00	100.00
HHV (Btu/lb)	11,666	13,126
Ultimate Analysis	**As-received (wt%)**	**Dry Basis (wt%)**
Moisture	11.12	--
Carbon	63.75	71.72
Hydrogen	4.50	5.06
Nitrogen	1.25	1.41
Chlorine	0.29	0.33
Sulfur	2.51	2.82
Ash	9.70	10.91
Oxygen (by difference)	6.88	7.75
TOTAL	100.00	100.00

Due to the aerodynamic and physical properties of the CO_2 rich working fluid, a turbine with a pressure ratio of 49 is required for this cycle, in order to reach the optimal combined cycle efficiency and achieve the desired gas turbine exhaust energy for the bottoming cycle. Therefore the gas turbine used in the plant design is an Advanced Inter-cooled Aero-derivative engine (ICAD) [5]. It was assumed that the compressor and turbine stages of the engine, when designed for the CO_2 rich working fluid, would achieve the same efficiencies that are typical and reasonable for current air-fired units.

The bottoming cycle is a single reheat, supercritical-steam, Rankine cycle operating at 3850/1050/1050/2” Hg.

2.2 PROCESS DESCRIPTION

A process flow diagram of the CO_2 hybrid plant is shown in Figure 2-1. The first step of the process is air separation, where oxygen is extracted for use in both the gasification and combustion processes throughout the plant. The oxygen is mixed with recycled flue gas and is routed to the partial gasification module (PGM) operating at 820 psia and 1950 deg. F, the combustion turbine (CT), and the CFB steam generator. In the partial gasification module, oxygen reacts with coal and steam to generate two fuel streams: syngas and char residue. The syngas generated by the PGM is combusted with a mixture of oxygen and recycled CO_2 to drive a gas turbine generator. In the steam generator, char generated in the PGM is burned with the oxygen in the gas turbine exhaust to generate steam for the steam cycle.

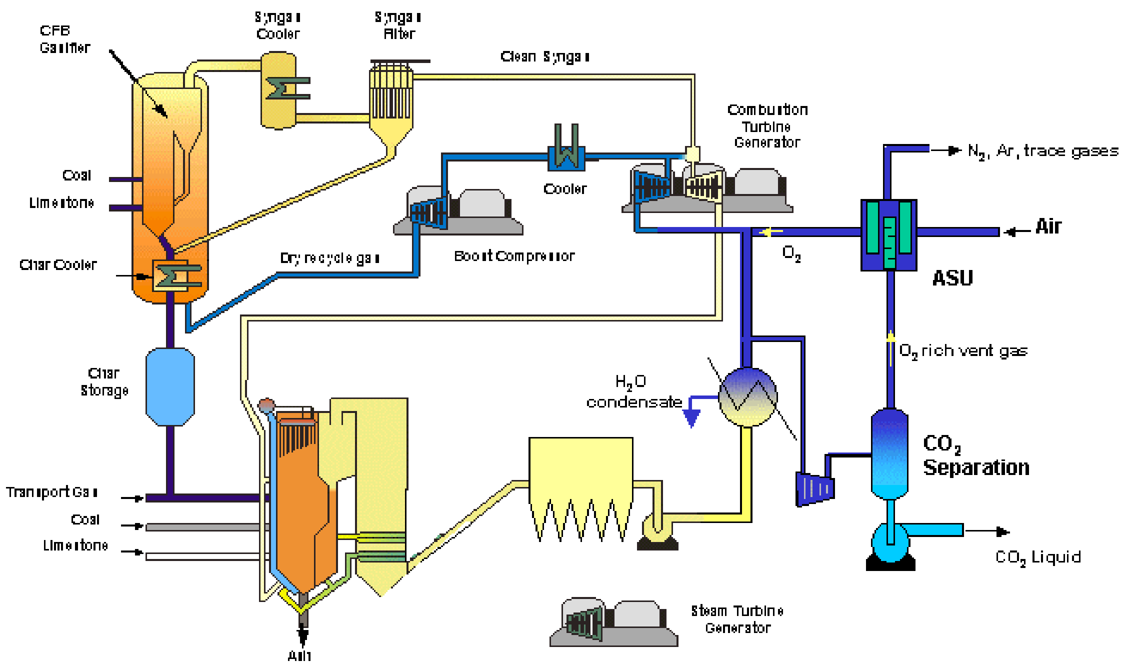

Figure 2-1. CO_2 Hybrid Cycle Process Flow Diagram

Since pure oxygen is used in the combustion process, exhaust flow recirculation is used to control temperature and maintain process velocities. This recirculation occurs in two main ways and provides for the exchange of energy between the topping and bottoming cycles. The gas turbine exhaust, which consists mostly of CO_2, O_2 and H_2O, is sent to the steam generator. This allows the steam generator to recover the considerable sensible heat contained in the GT exhaust and also to utilize its oxygen content for combustion. The CO_2 rich flue gas from the steam generator is recycled to the gas turbine compressor inlet. Once pressurized, the recycle gas goes to the PGM, and also to the GT combustor for temperature control. No additional compressor and associated power is required for O_2 pressurization since the O_2 is part of the working fluid, which is mixed with the recycled CO_2 and sent to the gas turbine compressor.

As the final step of the process, a portion of the recycled gas is diverted to a separator where CO_2 is condensed out and pumped away to the sequestration site. This is the only discharge from the plant. There is no stack and no gaseous effluents emitted to the atmosphere. The cycle is completed with zero emissions and 100% of the CO_2 sequestered. The oxygen separated from the CO_2 at the separator is recycled back into the process.

Because of the aerodynamic and physical properties of the CO2 rich working fluid the combustion turbine operates at a pressure ratio of 49. The CT inlet temperature is 2295 deg. F.

The CO_2 Hybrid Cycle is based on Foster Wheeler's combustion hybrid technology, also known as the Gasification Fluid-bed Combined Cycle (GFBCC) plant. A process flow diagram of this air-blown reference plant (without CO_2 sequestration) is shown in Figure 2-2.

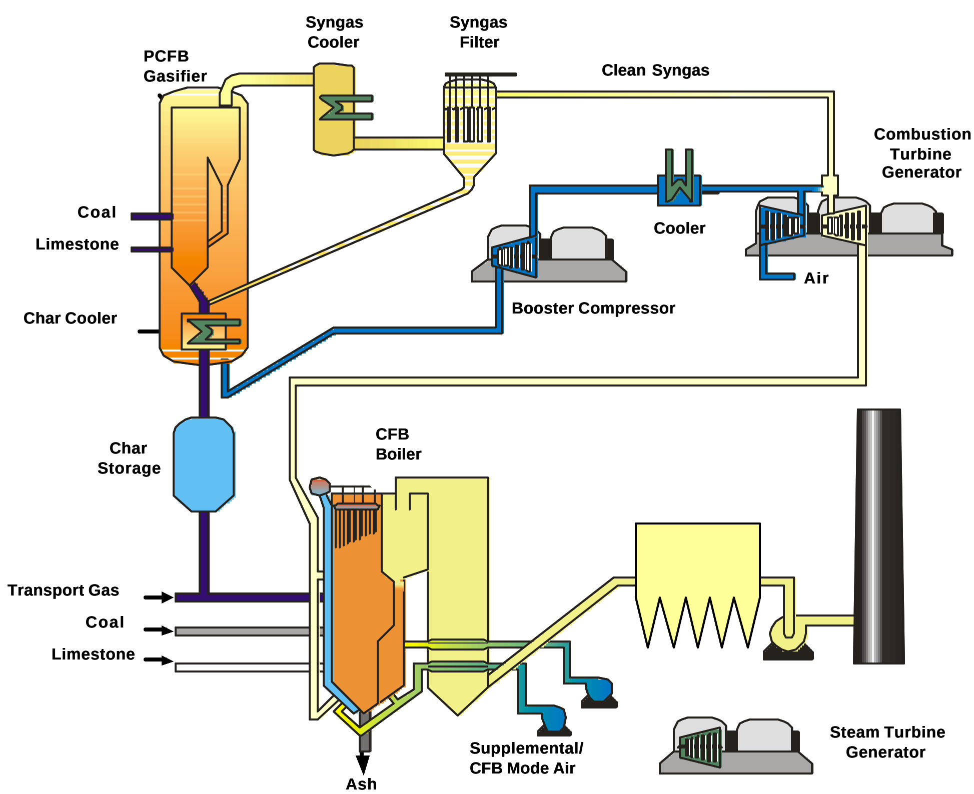

Figure 2-2. Air Fired Reference Plant Configuration Without CO_2 Sequestration

In the reference plant configuration (GFBCC) solid fuel is partially gasified in an air blown CFB gasifier operating at approximately 1800-2000 °F temperature and 350-450 psia pressure. Depending on the fuel, carbon conversion can vary between 50-90 % resulting in a syngas with a lower heating value between 125–150 btu/scf. The gasifier receives pressurized air from a separate dedicated air compressor.

The syngas is cooled to approximately 650 °F and cleaned of solids with a metallic filter before being combusted in a General Electric model 6FA combustion turbine (CT). No other gas cleaning of the syngas is performed, since the exhaust gas from the gas turbine is directed to a supercritical CFB boiler. The CFB boiler absorbs the energy of the hot (1000-1100 °F) CT exhaust gas while simultaneously cleaning it before exhausting to the environment.

The residual carbon-rich char from the gasifier is depressurized and stored in a containment vessel before being injected into the once through variable pressure supercritical CFB for final combustion. The hot gas from the combustion turbine is directed to the plenum under the CFB's fluidizing grid. This gas is used to fluidize the solids within the CFB and as the source of oxygen (contains between 10-15% Oxygen by weight) to support the CFB combustion process. The heat from the char combustion generates steam for the Rankine steam cycle. A separate fuel and limestone feed system supplies limestone and fresh fuel to the supercritical CFB power plant.

Both the combustion turbine and steam turbine generate power within the GFBCC cycle. Since the integration between the gasification process, combustion turbine and CFB steam plant is limited, a wide range of steam turbine capacities can be achieved for a given gas turbine resulting in a wide range of plant size. The CFB can be sized to fire solid fuel in addition to the char generated by the gasification process. More importantly, both the CT and the CFB can generate their full power ratings operating independently without the gasification system.

2.3 HEAT AND MATERIAL BALANCES

Thermodynamic cycle analyses of both the reference plant and the CO_2 cycle were conducted to determine the impact of 100% carbon sequestration on the base plant performance. Before the final plant configuration was chosen, several cases based on the following criteria were investigated:

- Because of anticipated gas turbine availability issues a configuration using existing industrial equipment (e.g. from Dresser-Rand or Man-Turbo) was sought.
- Two oxygen delivery options were investigated; oxygen from the ASU at near atmospheric pressure, and oxygen from the ASU at process pressure (850 psia).
- Two CO_2 removal locations were investigated; (1) high pressure removal from gas compressor discharge, and (2) low pressure removal from CFB boiler exhaust.

The following cases were run:

Case 1: Commercially available industrial turbomachinery with gas reheat is used for the topping cycle/ASU delivers oxygen at 15 psia/CO_2 is removed from the cycle at low pressure (compressor inlet).

Case 2: Deleted (a variation of case 1)

Case 3: Advanced Intercooled Aeroderivative (ICAD) engine used for the topping cycle/ASU delivers oxygen at 15 psia/CO2 is removed from the cycle at high pressure (compressor discharge)

Case 4: Advanced Intercooled Aeroderivative (ICAD) engine used for the topping cycle/ASU delivers oxygen at 850 psia/CO2 is removed from the cycle at high pressure (compressor discharge)

Case 5: Advanced Intercooled Aeroderivative (ICAD) engine used for the topping cycle/ASU delivers oxygen at 15 psia/CO2 is removed from the cycle at low pressure (compressor inlet)

Table 2-2 summarizes the process conditions and performance results for these cases in comparison with the air-fired reference (GFBCC) plant configuration. Based on net plant efficiency, case 5 emerges as the best alternative. The conceptual design presented in the previous sections pertains to this case.

ASPEN process simulation software coupled with Foster Wheeler's extensive plant performance database was used to generate the heat and material balances, and the plant performance results.

The air separation power requirements shown on Table 2-2 were obtained from Air Products based on what would be their standard offering for the given oxygen demand. Foster Wheeler believes that there may be a more efficient ASU design to deliver high-pressure cryogenic oxygen to improve Case 4 results. This issue needs further evaluation with the participation of an air separations vendor.

This section will focus on Case 5, as the design case. Heat and mass balances for the other cases can be found in Appendix A.

Table 2-2. Process Conditions and Performance of Four Alternate Cycle Configurations in Comparison with the Reference Plant

	Air Fired Reference	Case 1	Case 3	Case 4	Case 5
ASU					
O_2 Flow, klb/hr		422	497	506	497
Power Consumption, MW	Air Blown	42.8	55.5	88.1	55.5
O_2 Delivery Pressure, psia		15	15	850	15
Gasifier					
Coal Feed Rate, klb/hr	107	99	180	190	164
O_2 Feed Rate, klb/hr	265 (Air)	51	89	91	82
Steam Feed Rate, klb/hr	27	9	16	15	13
Syngas Flowrate, klb/hr	373	294	582	577	434
Syngas LHV, btu/scf	133	189	156	159	205
Syngas Cooler Duty, mmbtu/hr	168	144	280	283	218
Char Flowrate, klb/hr	30	28	63	69	47
Gas Turbine					
Type	GE 6FA	2-Stg Industrial	ICAD	ICAD	ICAD
Output, MW	87.4	47	108	107	129
Exhaust Flow, klb/hr	1828	2050	2030	2091	2170
1st Stage Turbine Inlet Temp., F	2097	1000	2270	2225	2295
2nd Stage Turbine Inlet Temp., F	N/A	1600	N/A	N/A	N/A
Turbine Exhaust Temp., F	1089	925	1167	1137	1142
Steam Turbine					
Output, MW	250	258	284	281	270
Main Steam Flowrate, klb/hr	1300	1300	1300	1300	1300
Reheat Steam Flowrate, klb/hr	1340	1369	1521	1524	1451
CFB Boiler					
Coal Feed Rate, klb/hr	102	105	61	54	76
Flue Gas Flowrate, klb/hr	2563	2167	2379	2443	2274
CO_2 Compression and Dehyd.					
CO_2 Inlet Pressure, psia		15	756	838	15
CO_2 Flowrate, klb/hr	N/A	480	571	580	568
CO_2 Sequestered as		Liquid	Vapor	Liquid	Liquid
Overall System Performance					
Net Efficiency, %	42	30.9	35.7	33.2	35.9
Net Power Output, MWe	300	216	295	277	295
Auxiliary Power Consumption, MW	37.5	89.3	97.1	111.4	104.1

Figures 2-3 and 2-4 show the ASPEN heat and material flowsheets for, respectively, the air-fired reference plant and the CO2 hybrid cycle design represented by Case 5.

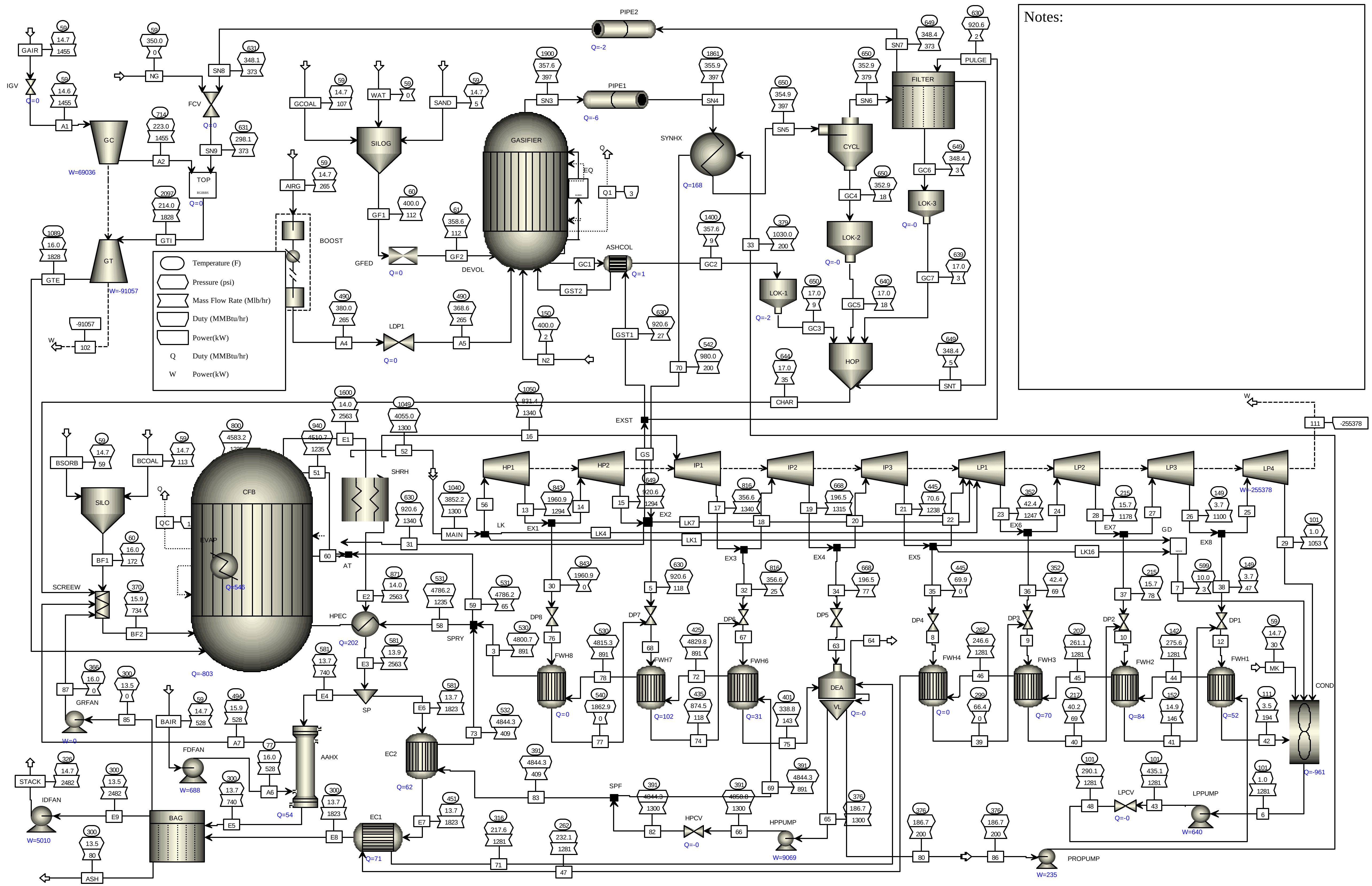

Figure 2-3. Heat And Material Balances For The Air Fired Reference Hybrid Plant (GFBCC Without CO_2 Sequestration)

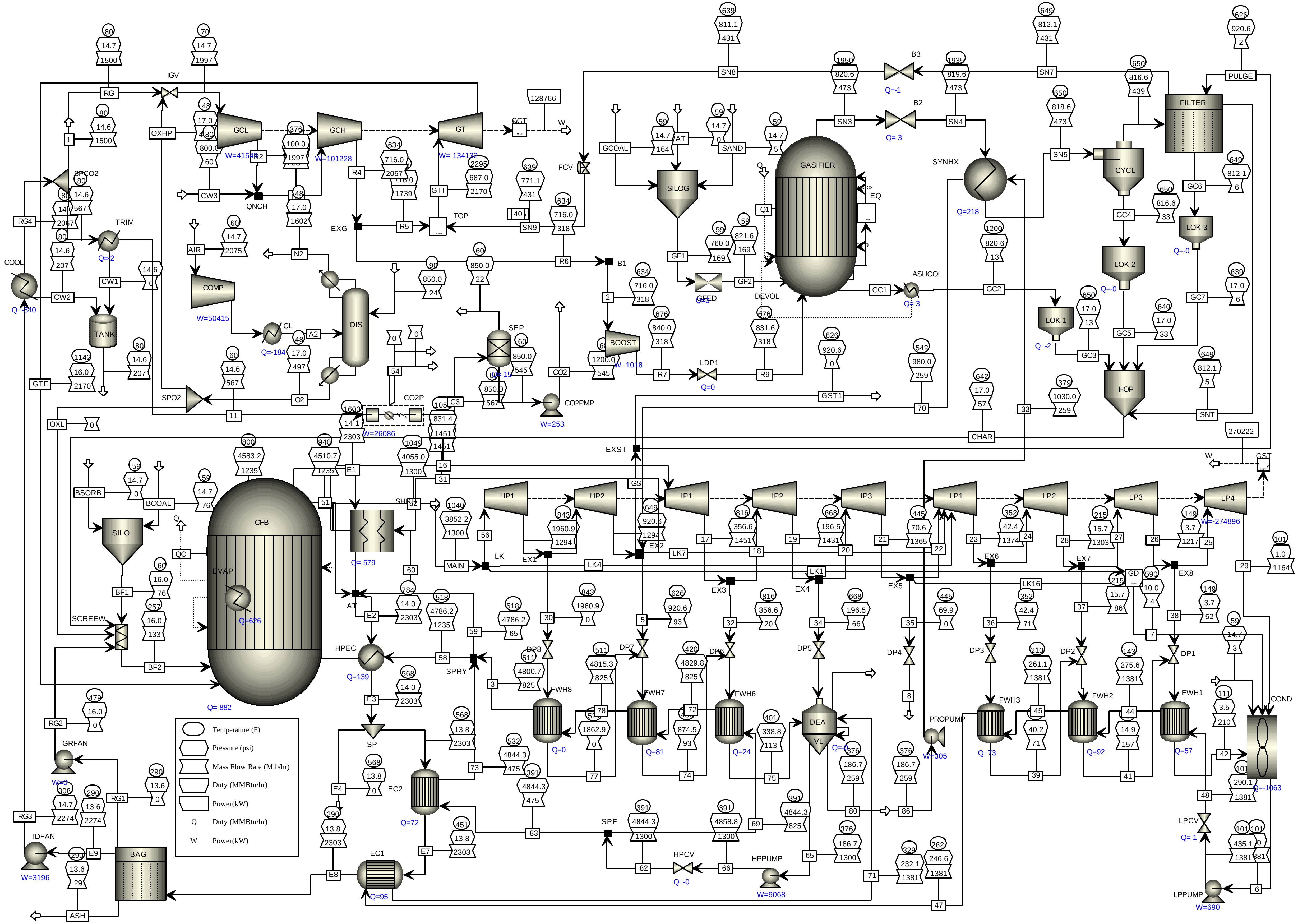

Figure 2-4. Heat And Material Balances For The CO_2 Hybrid Plant (GFBCC With CO_2 Sequestration)

2.4 POWER PENALTIES ASSOCIATED WITH CO_2 SEQUESTRATION

There are both direct and indirect power (or efficiency) penalties associated with CO_2 removal. Direct penalties occur as a result of an increase in auxiliary power requirements due to air separation and CO_2 separation and compression. Indirect penalties are less obvious and refer to the gross power decrease due to CO_2 enrichment and removal processes. The systems net power and efficiency is reduced due to both direct and indirect penalties.

Pre-combustion (i.e. the Selexol process for an IGCC) and post-combustion (i.e. amine absorption of CO2 from the flue gas of a PC or NGCC plant) have large indirect losses. In the pre-combustion separation technique, the water-gas shift reaction is used to shift CO to CO_2 in order to concentrate the CO_2 in syngas. It has been well documented that this reaction reduces the syngas LHV while releasing its fuel energy as heat. Therefore, more syngas needs to be generated from gasification to compensate for the LHV loss by shifting. The low-grade heat from the shift reaction and syngas cooling before CO_2 absorption contributes to system energy loss. An efficiency loss of 6% to 7.7% and a CO_2 mitigation cost of between $21 and $23 per tonne[7] is estimated for such a power plant with a pre-combustion CO_2 removal system [1, 2]. To obtain a high CO conversion for more CO_2 to be removed, a ratio of $H_2O/CO > 2$ needs to be maintained by steam injection into the syngas. This steam can be extracted from the steam turbine, or generated from syngas cooling. Both ways reduce steam turbine power generation because of less steam flow to the bottoming cycle.

The loss of working fluid is another source of indirect penalty. In the pre-combustion separation technique, working fluid loss is caused by the removal of pressurized CO_2 from the syngas stream and by condensation of excess steam from the water gas shift. The losses of working fluids reduce power generation from the gas turbine because of less flow through the turbine. Indirect penalties can be quite significant and should be minimized to produce the maximum system efficiency.

The Advanced CO_2 Hybrid Cycle eliminates all indirect penalties by using a mixture of CO_2 recycled from the gas turbine exhaust together with oxygen as the working fluid. This facilitates straightforward concentration of CO_2 without enriching and separation processes. It eliminates the need for CO shifting, syngas cooling, absorption, and stripping and allows direct collection of CO_2 from recycled flue gas. It leads to a simpler CO_2 collection process than the conventional oxygen-blown pre-combustion CO_2 capture systems, while providing the advantages of lower energy cost and lower efficiency loss. Moreover, oxygen usage is minimized by recycling from the gas turbine exhaust most of excess oxygen along with the CO_2 back to the system.

While the CO_2 Hybrid Cycle eliminates indirect energy costs, higher direct penalties are incurred due to the requirement for a larger air separation unit compared to an IGCC plant of the same net output. Table 2-3 shows a comparison of direct and indirect losses for the CO_2 cycle in comparison to an IGCC plant with pre-combustion CO_2 separation.

The net result is the same 6.1% loss in cycle efficiency due to CO_2 sequestration, for both the CO_2 cycle and the IGCC with CO_2 sequestration. Figure 2-5 illustrates the direct vs. indirect efficiency losses for the two technologies.

As Section 4 (Economic Analysis) will show, due to the fewer equipment changes and simpler operation of the CO_2 recycle concept the actual dollar costs of CO_2 mitigation is lower, even though the energy costs are the same.

[7] Cases 3A and 3E in references 1 and 2. Year 2000 dollars, 65% capacity factor.

Table 2-3. Summary of Power Penalties Incurred Due to CO_2 Removal

Plant Type	**IGCC**		**FW Hybrid Cycle**	
	With CO_2 Sequestration[8]	Without CO_2 Sequestration[9]	With CO_2 Sequestration	Without CO_2 Sequestration
Net Power	403.5	424.5	294.9	300
Aux Power	86.9	49.5	104	37.5
Net Cycle Efficiency, %	37	43.1	35.9	42
CO2 emissions, lb/hr/MWe	162	1582	0	1709
CO2 removal, %	90	na	100	na
Oxygen Usage, lbs/hr/MWe	647	568	1640	0
Power Penalties, kWh/lb of CO2 avoided				
Direct (due to aux power increase)	70	--	133	--
Indirect (due to gross power decrease)*	65	--	0	--

*Corrected for coal heat input differences

$$\text{Direct Losses} = \frac{\text{(Aux Power)/(Net Power)} - \text{(Aux Power/Net Power)}_{\text{Reference Plant}}}{\text{(klbs/hr CO2 Emitted/Net Power)}_{\text{Reference Plant}} - \text{(klbs/hr CO2 Emitted/Net Power)}}$$

$$\text{Indirect Losses} = \frac{\text{Corrected Gross Power of Reference Plant} - \text{Gross Power with CO2 Sequestration}}{\text{Klbs/hr CO2 Emissions from Reference Plant} - \text{klbs/hr CO2 Emissions with CO2 Sequestration}}$$

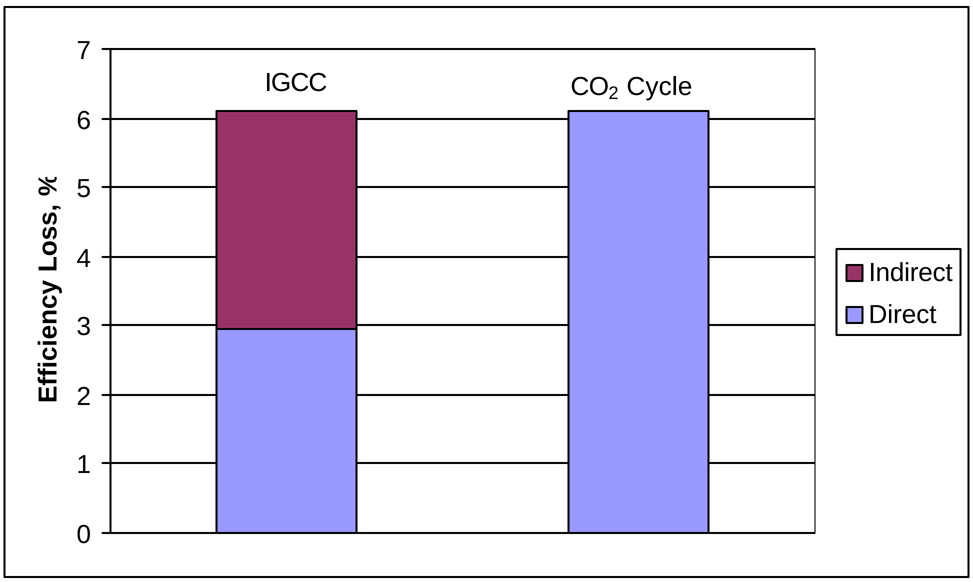

Figure 2-5. Direct and Indirect Components of Efficiency Loss

An updated calculation for the IGCC with pre-combustion CO_2 sequestration from reference [2] indicates that a more realistic plant design should incorporate a water scrubber (Case 3E, ref. [2]). If this updated case is considered the IGCC plant efficiency drop goes up to 7.7%.

2.5 PLANT OXYGEN USAGE

As mentioned in the previous sections, a significant cost adder to the reference plant is the air separation unit. The CO_2 Hybrid Plant uses 2.5 times the amount of oxygen required by an IGCC with pre-combustion CO_2 capture (Table 2-2). Consequently, 54% of the efficiency losses of the CO_2 cycle are

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[8] Reference 1, Chapter 4, case 3A

[9] Reference 1, Chapter 4, case 3B

due to air separation. It is clear that future reductions in the cost of air separation will benefit the CO_2 cycle more than it would any alternate CO_2 mitigation techniques for fossil fuels. The relative simplicity of operating and maintaining an air separation plant compared to, say, a Selexol absorption/regeneration plant will also impact the technology costs favorably.

A point worthy of note is that the power consumption values used in references [1] and [2] for IGCC plants are about 12% lower (in kWh/lb of oxygen generated) than the values used here. For this study, power consumption values communicated by Air Products were used. We have confidence in our values and believe that they reflect the energy costs more accurately than the values used in references [1] and [2].

Section 3

PLANT DESIGN

3.1 DESIGN BASIS

The primary objective of this study is to determine the feasibility of a zero emissions coal plant that utilizes the CO_2 Hybrid Cycle concept. As such, design of the components need to be detailed enough to determine the cost of building the plant. Component designs and a detailed cost estimate for an air fired Gasification Fluid-bed Combined Cycle (GFBCC) plant had previously been carried out during the front-end engineering work for a commercial plant demonstration project[10]. This demonstration plant work, which was conducted in year 2001, produced detailed designs for the CFB and gasifier units, and equipment definitions for the balance of plant. Since the CO_2 Hybrid cycle is based on the GFBCC, this earlier design and costing study was used as a reference in this study.

Taking the existing GFBCC design as the reference, the CFB and gasifier were redesigned based on the new process conditions listed on Table 2-2 (see previous section). Balance of plant equipment was reviewed to see what changes are required for conversion to the CO_2 sequestering plant configuration. Equipment was then removed, added, or altered, as required.

The plant site for the GFBCC demonstration was a Midwestern US location at an elevation of 4500 ft. Although the CO_2 hybrid plant design was based on this reference, appropriate corrections for the site elevation and ambient conditions were made to approach the following site characteristics, used in References 1 and 2, for proper comparison of the results with IGCC plants:

Level Topography
0 ft Elevation[11]
60 deg. F Dry Bulb Temperature
55% Relative Humidity
2” Hg Condenser Pressure

The feedstocks were Illinois#6 coal and Greer limestone. The coal analysis is given in Table 2-1 of the previous section. The composition of the Greer limestone is shown in Table 3-1.

Table 3-1. Greer Limestone Analysis

	Dry Basis, %
Calcium Carbonate, $CaCO_3$	80.40
Magnesium Carbonate, $MgCO_3$	3.50
Silica, SiO_2	10.32
Aluminum Oxide, Al_2O_3	3.16
Iron Oxide, Fe_2O_3	1.24
Sodium Oxide, Na_2O	0.23
Potassium Oxide, K_2O	0.72
Balance	0.43

Table 3-2 shows the original system design parameters used for the GFBCC plant in the 2001 study and Table 3-3 shows the design parameters for the CO2 Hybrid Cycle for the present study.

[10] The demonstration plant was not built because the utility customer decided that the electricity demand they were forecasting was not going to materialize
[11] 500 ft elevation was used in references [1] and [2]. This has a minor effect on performance and cost.

Table 3-2. GFBCC Design Basis

TITLE: GFBCC DESIGN BASIS			Date: 3/22/2001
PROJECT: GFBCC Demonstration		FILE:	REV.
ORIGINATOR: SMPM/PE PM	APPROVED:	Job#	Rev. Description:

PROJECT DESCRIPTION............... GFBCC Power Plant Consisting of Gasification Island for GE 6FA GT & ACFB Supercritical/Once-Through Unit, 1 x 250MWe (gross)

Steam Condition BOILER LOAD %MCR	UNITS	PEAK %	MCR 100%	CONTROL 70%	MIN 40%
Main Steam Flow	klb/hr		1378		
Main Steam Press.	psig		3916		
Main Steam Temp.	F		1045		
LP Feedwater Press.	psig		196		
HP1/2 Feedwater Press.	psig		4626/4566		
LP Feedwater Temp.	F		267		
HP1/2 Feedwater Temp.	F		428/521		
SH Spray	%		5		
Spray Water Press.	psig		4566		
Spray Water Temp.	F		521		
RH Steam Flow	klb/hr		1386		
RH Inlet Press.	psig		924		
RH Outlet Press.	psig		895		
RH inlet Temp.	F		644		
RH Outlet Temp.	F		1055		

* Assumes 5 °F Critical Piping Temperature Drop

Design Parameters Ambient air = 0.013 lb water/lb dry air

Ambient Temp.	F	85	85	85	85
Reference Temp.	F	105	105	105	105
Relative Humidity	%	50	50	50	50
Plant Elevation	ft	4000	4000	4000	4000
Boiler Exit Gas Temp.	F	300	300	300	300
Bottom Ash Temp.	F	550	550	550	550
DeNOx Reagent	aqueous ammonia				
Comments:	All information represents Basis of Design. Actual Site Conditions may be different.				

Emissions Guarantees For Load Range 65 to 100% MCR

	lb/mbtu	ppmdv	% O2 dry	% Red.	mg/MJ
SO_2	0.065			95	
NOx	0.1				
CO	0.15				
VOC	0.005				
Particulates	0.015				
Opacity	10%				
Ammonia Slip		15	3		

Steam Cycle Information

ST Condenser Pressure	in HG	2.5	
ST Condenser Temperature		101.2	
Condenser Type		Surface	
Cooling Water System		Mechanical Draft Cooling Tower	

Fuel Name/Type - Approx 55% Coal, 20% Char, 25% GT exhaust

Constituent	units	Perf Coal	Worst Fuel	Range Max.	Range Min.
Fuel Ultimate Analysis, Wt. % as Fired					
C	% Wt.	65.30	60.30	71.20	60.30
H_2	% Wt.	4.75	4.00	5.80	4.00
O_2	% Wt.	9.00	6.00	11.20	6.00
N_2	% Wt.	1.20	1.70	1.70	0.70
CL	ppm	250	428	428	63
S	% Wt.	0.75	1.26	1.26	0.37
H_2O	% Wt.	10.00	16.90	16.90	5.80
Ash	% Wt.	9.00	11.80	11.80	5.80
HHV		11,800	10,500	12,600	10,500
LHV					
Fuel Poximate Analysis, Wt. % as Fired					
Fixed Carbon	% Wt.	45.00	41.10	49.10	41.10
Volatile Matter	% Wt.	36.00	32.80	40.40	32.80
Ash	% Wt.	9.00	11.80	11.80	5.80
Moisture	% Wt.	10.00	16.90	16.90	5.80
Ash Analysis					
Silicon, SiO_2	% Wt.	58.0	51.40	65.6	51.4
Aluminum, Al_2O_3	% Wt.	22.2	11.70	33.0	11.7
Iron, Fe_2O_3	% Wt.	5.0	8.00	8.0	1.9
Calcium, CaO	% Wt.	8.5	2.00	15.0	2.0
Potassium, K_2O	% Wt.	0.7	1.10	1.1	0.2
Sodium, Na_2O	% Wt.	2.1	3.80	3.8	0.4
Sulfur, SO_3	% Wt.	3.6	6.00	6.0	1.1
		100.0	84.0	132.5	68.7

Limestone - assumed by PM (good limestone)

Type -		Perf.	Range Max.	Range Min.
$CaCO_3$	Wt. %	95.00		
$MgCO_3$	Wt. %	1.00		
Inert	Wt. %	3.00		
H_2O	Wt. %	1.00		
Maximum Utilization	Wt. %			
TGA react rate (CuFt/lb mol/s)				
Rectivity Index, RI		1.7	2.0	1.5

Gas Turbine

Model		6FA
Inlet Evaporative Cooling		Required
Start-Up Fuel		Natural Gas or Fuel Oil
NOx Abatement (Simple Cycle)		Steam Injection

Table 3-3. CO_2 Hybrid Plant Design Parameters*. *Fuel Basis: Illinois #6 Coal.*

Air Separation Unit	
Oxygen Yield, klb/hr	508
Oxygen Delivery Pressure, psia	15
Gasifier	
Coal Feed Rate, klb/hr	190
Oxygen Usage, klb/hr	91
Process Steam Usage, klb/hr	15
Design Pressure, psia	900
Design Temperature, deg. F	2200
Syngas Yield, klb/hr	577
Lower Heating Value of Syngas, btu/scf	159
Syngas Cooler Duty, MMbtu/hr	283
Char Yield, klb/hr	69
Gas Turbine	
Power Output, MW	130
Turbine Inlet Temperature, F	2300
Exhaust Gas Flowrate, klb/hr	2200
Exhaust Gas Temperature, F	1142
Boiler	
Coal Feed Rate, klb/hr	163
Flue Gas, klb/hr	2480
Main Steam Flowrate, klb/hr	1300
Main Steam Temperature, F	1050
Main Steam Pressure, psig	4040
Reheat Steam Flowrate[12], klb/hr	1525
Reheat Steam Temperature, F	1050
Reheat Steam Pressure, psig	830
Steam Turbine	
Power Output, MW	270

**Note: Equipment design parameters do not necessarily reflect base case process design values.*

3.2 GASIFIER DESIGN

The gasifier will utilize Foster Wheeler's circulating fluid bed technology for highest fuel flexibility, scalability, and operational simplicity. Operating at 800 psig, it will be a compact unit at just 4ft in internal diameter for 13 ft/sec nominal superficial gas velocity. Coal will be fed at the base of the unit along with oxygen, recycled CO_2, and process steam. A cyclone and solids return leg will recycle solids back to the base of the gasifier to maintain temperature uniformity throughout the gasifier and to enhance carbon conversion rates. Coal and char particles will travel up the riser through a height of 80 ft, which will provide a residence time of about 6 seconds.

The gasifier will generate about 575 klb/hr of syngas for the topping cycle and 69 klbs/hr of char to be burned in the char combustor. The syngas will be generated at about 1950 degrees F. A fire-tube type syngas cooler will be used to bring this temperature down to about 650 degrees F so that the syngas can be put through sintered metal filters for fine particulate removal before it is burned in the gas turbine combustor. The syngas cooler will generate IP steam for the steam cycle. The details of the gasifier design, including the barrier filter, are summarized on Table 3-4. An elevation drawing of the gasifier showing outline dimensions is shown on Figure 3-1.

Other components required for the gasifier island are listed on Table 3-5.

[12] Includes IP steam generated by the gasifier syngas cooler

Table 3-4. Gasifier Design

Riser			**Cyclone**			**Downcomer leg**		
gas flow	klb/hr	577	gas flow	klb/hr	577	solid flow	klb/hr	2885
gas flow	kcf/hr	600	gas flow	kcf/hr	600	solid flow	kcf/hr	58
gas flow	ft/s	13	gas flow	ft/s	100	solid flow	ft/s	2
ID	ft	4.0	ID	ft	3.0	ID	ft	3.2
OD	ft	6.5	OD	ft	5.5	OD	ft	5.6
H	ft	80	H	ft	9	H	ft	75
design P	psia	900	design P	psia	900	design P	psia	900
design T	F	2200	design T	F	2200	design T	F	2200
thichness	inch	2.6	thichness	inch	2.8	thichness	inch	2.4
Candle filter vessel			**Pipe**			**Syngas cooler**		
gas flow	klb/hr	577	gas flow	klb/hr	577	**shell side**		
gas flow	kcf/hr	277	gas flow	kcf/hr	277	gas flow	klb/hr	577
gas flow	ft/s		gas flow	ft/s	50	gas flow	kcft/hr	600
ID	ft	6.7	ID	ft	1.4	gas flow in	ft/s	98
OD	ft	9.4	OD	ft	1.7	P	psia	820
H	ft	50	H	ft	200	duty	Mbtu	283
design P	psia	900	design P	psia	900	shell ID	ft	3.9
design T	F	800	design T	F	800	shell OD	ft	4.2
thichness	inch	4.2	thichness	inch	2.0	shell H	ft	93
Candles						shell thick	inch	2.0
gas flow	kcf/hr	277				**tube side**		
face vel	ft/min	2				gas flow out	kcft/hr	277
length	ft	27				gas vel out	ft/s	48
OD	inch	3				length	ft	54
candles	-	109				ID	inch	1.4
						tubes	-	170

Table 3-5. Gasifier Island Major Components

Component	Type	Design Basis
Booster Compressor	Vendor design	635 deg. F and 50 bar inlet. Medium: 30.5% O2, 69% CO2
Syngas System	Refractory lined pipe	2200 deg. F, 900 psia, 500,000 cu. Ft./hr of syngas
Char system	Lock hopper with pneumatic transport	60 klbs/hr char continuous transport basis, 650 deg. F max char temperature and 850 psia max inlet pressure
Hot gas filter	Sintered Metal Candles	Included in gasifier design specifications (Table 3-3)
Syngas cooler	Shell and tube	Included in gasifier design specifications (Table 3-3)

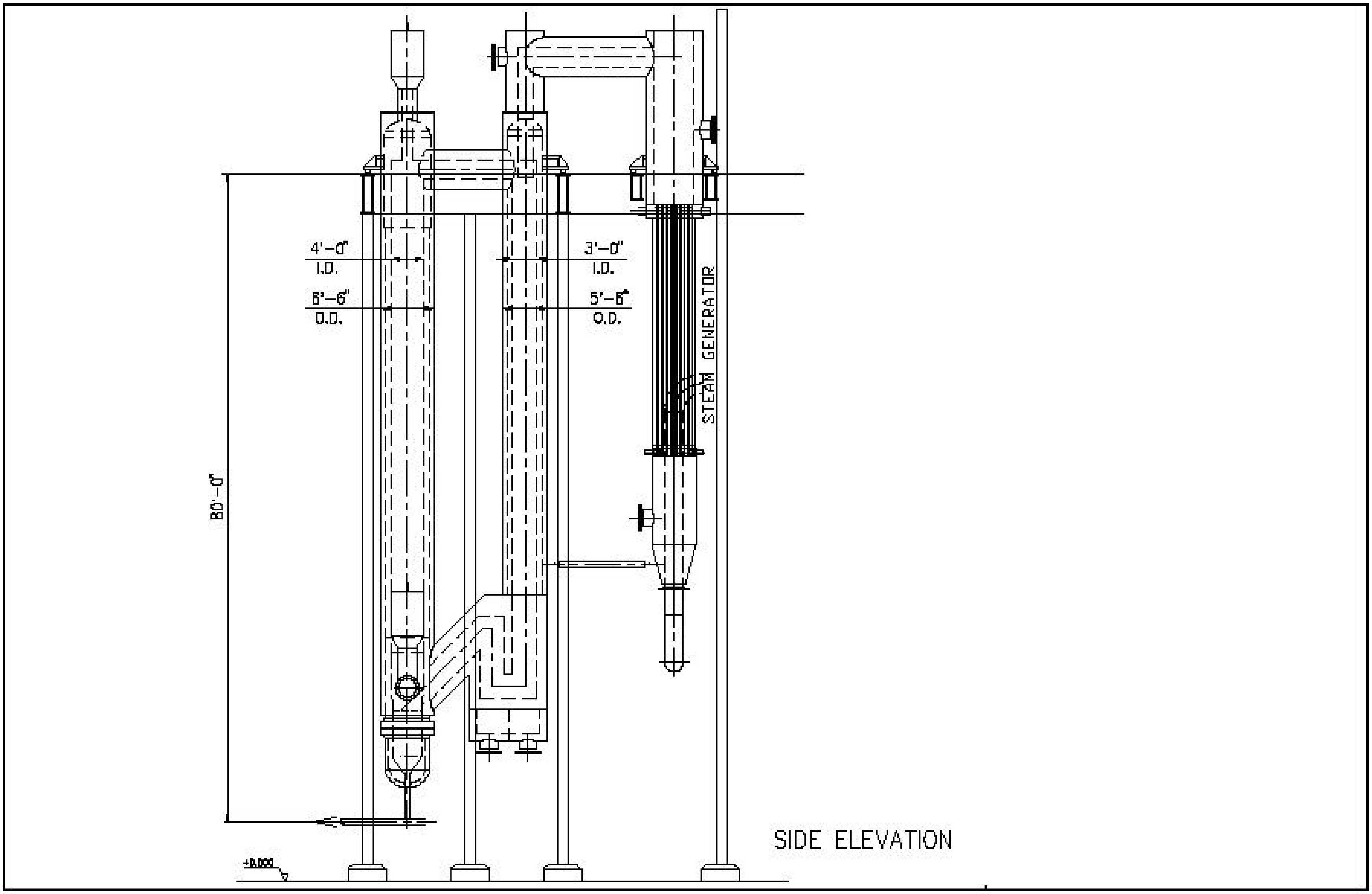

Figure 3-1. Gasifier Side Elevation

3.3 CFB DESIGN

The char combustor will utilize Foster Wheeler's circulating fluidized bed combustion technology. It will generate 1300 klbs/hr of supercritical steam at 3850 psig and 1040 degrees F. This steam will then go through the HP stages of the steam turbine. The HP section discharge will combine with about 225 klbs/hr of intermediate pressure steam generated by the gasifier syngas cooler. The combination, about 1525 klbs/hr of steam, will return to the reheat section of the CFB char combustor. Reheater outlet steam will be at 830 psig and 1050 degrees F.

The CFB furnace, with its major dimensions, is shown on Figure 3-2. The main furnace section will be 58 ft wide by 28 ft deep. The overall height of the structure will be about 236 ft from grade. The design will feature wing wall superheaters as well as in-duct HRA superheaters.

For maximum operational flexibility, the plant was designed with a "CFB stand-alone" option. This entails running the CFB when the gasifier is down for repairs/maintenance, or due to an upset event (e.g. a gas turbine trip). In the standalone mode the CFB is operated with oxygen and recycled CO2, using the gas recirculation fan instead of the GT exhaust as the oxidant. With this design feature, the facility can still operate (albeit at a lower efficiency) and generate about 68% of the combined cycle power output.

Design parameters for the normal and standalone operating modes, compared to the air-fired reference unit are shown in Table 3-6.

In determining the pressure parts (waterwall and steam surfaces) for cost analysis a comparative approach was taken. The metal weighs for each tube bank of the GFBCC reference CFB was adjusted by using suitable criteria (e.g. bank width, heat transfer area, # of units, perimeter, etc.). This resulted in revised metal masses, which could then be used to determine cost differentials. Table 3-7 shows the pressure parts adjustments and adjustment bases. As the table shows, the net result was a 29% reduction in the pressure parts used. This is due to the more compact nature of the O2/CO2 fired CFB compared to the air-fired reference unit. Higher oxygen concentrations and higher working fluid density allow the use of less surface area and hence an overall lighter unit.

Other major components required for the CFB island are shown on Table 3-8.

3.4 GAS TURBINE

An Advanced Inter-cooled Aero-derivative (ICAD) engine was used for the plant, as described in reference [5]. The components required for the ICAD turbine plant are listed in Table 3-9. The gas turbine is treated as a "black box" in this study and requires further investigation.

3.5 STEAM TURBINE AND BALANCE OF PLANT DESIGN

The steam plant remains essentially unchanged from the air-fired reference. The steam turbine and balance of plant equipment were specified relative to the air-fired (GFBCC) reference plant. For the steam turbine components, a cost adjustment factor, equal to the ratio of steam turbine gross power for the CO2 cycle to that of the reference GFBCC (270/250), was applied to the installed cost. A list of equipment for the steam turbine and balance of plant is given in Table 3-10.

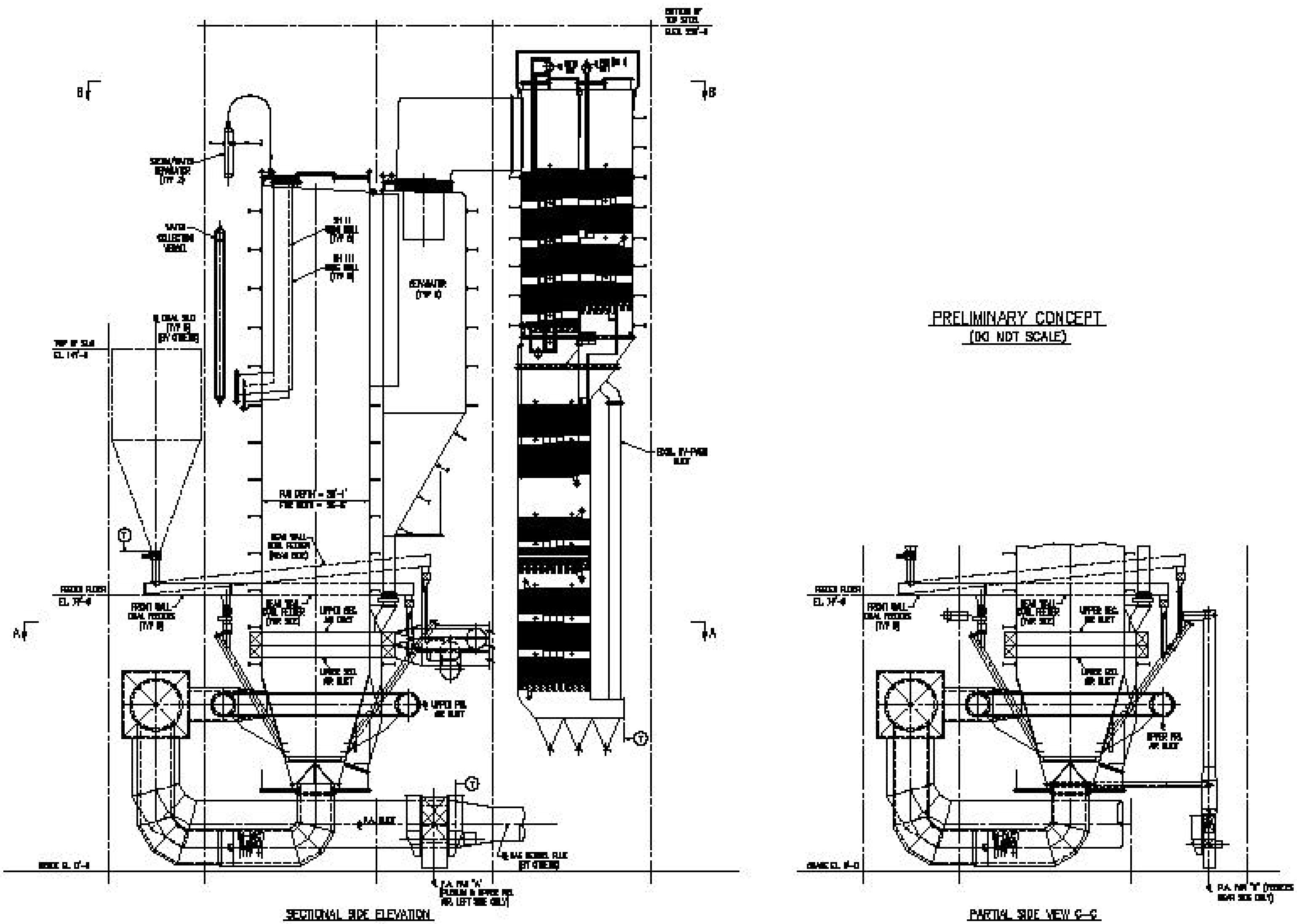

Figure 3-2. CFB Elevation Drawing

Table 3-6. CFB Design Parameters

		Base-GFBCC	CO2-Hybrid	CO2-Hybrid Stand-Alone
Steam Turbine Power	MWe	250	270	270
Main Steam Flow	M lb/h	1404.5	1300	1300
Main Steam Temperature	F	1045	1049	1049
Main Steam Pressure	psig	3904.4	4040.3	4040.3
Reheat Steam Flow	M lb/h	1379.4	1451	1451
Hot Reheat Steam Temperature	F	1055	1050	1050
Hot Reheat Steam Pressure	psig	901.8	867.5	867.5
Feedwater Temperature	F	531	518	518
Boiler Duty	E6 Btu/h	1632.5	1590.7	1590.7
Gas Turbine Exhaust Temp.	F	1073	1142	NA
FEGT	F	1578	1584	1584
Boiler Flue Gas Flow	M lb/h	2478	2274.5	2107.41
Gas Recirculation	%	0	0	338
Gas Molecular Weight		29.29	38.52	34.87
Plant Elevation	ft.	6500	0	0
Ambient Pressure	psia	11.76	14.7	14.7
Furnace Height	ft.	155	160	160
Furnace Width	ft.	93.5	54	54
Furnace Depth	ft.	29	28.1	28.1
Number of Separators		4	2	2
Separator Type		Original	D70	D70
Separator Depth	ft.	22	20.7	20.7
Reheat Pass Gas	%	57.3	64.9	65.3
Velocities:				
Furnace	m/s	4.91	4.91	5.03
HRA Reheater	ft/s	54.4	51	51.5
HRA Primary Superheater	ft/s	42.6	38.2	37.7
Lower Economizer	ft/s	47	47.8	47.8
Economizer Gas Temp. Out	F	581.6	562.3	560.6
Evaporator Outlet Enthalpy	Btu/lb	1099	1120	1119
Flow Rates:				
Char	M lb/h	17.99	46.98	0
Coal	M lb/h	122.96	76.02	162.59
GT Exhaust	M lb/h	1632.61	2170.6	0
Oxygen	M lb/h	0	0	336.56
Air	M lb/h	716.91	0	0
Syngas	M lb/h	0	5.02	0
Flue Gas	M lb/h	2477.54	2274.52	2107.4
Limestone	M lb/h	17.94	0	0
Sand	M lb/h	5	5	0
Ash	M lb/h	38.96	29.1	17.62

Table 3-7. CFB Pressure Parts

Description	Weight	Wt. Adustment Basis	GFBCC	CO2 Hybrid	Factor	WEIGHT CO2 Hybrid
ECONOMIZER HEADERS	128,932	HRA Width	55	42	0.76	98,457
ECONOMIZER COILS/TUBING	3,712,364	H.T. Area	169,656	110,896	0.65	2,426,376
ECONOMIZER TRANSFERS	50,521					50,521
ECONOMIZER MISCELLANEOUS	0					0
WATERWALL LARGE TRANSFERS	39,738					39,738
WATERWALL PANELS	1,191,457	Area (Approx.)	37,975	28,512	0.75	894,558
WATERWALL HEADERS	155,910	Perimeter + WW	245	178	0.73	113,401
WATERWALL SMALL TRANSFERS	88,044					88,044
FURNACE ROOF SUPERHEATER HEADERS	90,850	Furnace Depth	28.96	28.13	0.97	88,237
FURNACE ROOF SUPERHEATER PANELS	61,540	Area	2,709	1,618	0.56	34,484
FURNACE ROOF SUPERHEATER TRANSFERS	24,580					24,580
SUPERHEATER WINGWALL HEADERS	111,864	Total WW Width	1,319	1,120	0.85	94,987
SUPERHEATER WINGWALL PANELS	253,043	H.T. Area	11,155	11,120	1.00	252,249
SUPERHEATER WINGWALL TRANSFERS	56,538					56,538
REHEATER HEADERS	104,517	HRA Width	55	42	0.76	79,813
REHEATER COILS/TUBING	1,399,971	H.T. Area	189,769	165,452	0.87	1,220,579
REHEATER TRANSFERS	8,816					8,816
HRA HEADERS	73,260	Perimeter + PWall	224	186	0.83	60,751
HRA PANELS/TUBING	538,098	Perim. (same H)	224	186	0.83	446,216
HRA TRANSFERS/PIPE	90,015					90,015
PSH HEADERS	98,277	HRA Width	55	42	0.76	75,048
PSH COILS/TUBING	976,272	H.T. Area	70,578	54,661	0.77	756,100
PSH TRANSFERS	100,833					100,833
INTREX / FSH COILS/TUBING	0					0
INTREX / FSH HEADERS	0					0
STRIPPER COOLER COILS/TUBING	0					0
STRIPPER COOLER HEADERS	0					0
STRIPPER COOLER TRANSFERS	0					0
WATER COLLECTING VESSEL	58,151					58,151
TANGENTIAL STEAM SEPARATOR	45,172					45,172
SUPERHEATER WINGWALL SPRAY HEADERS	7,624					7,624
REHEATER SPRAY HEADERS	6,909					6,909
PSH SPRAY HEADERS	7,244					7,244
INTREX / FSH SPRAY HEADERS	0					0
SEPARATOR W/ HEADERS	1,086,351	H.T. Area	44,167	19,086	0.43	469,448
CROSSOVER DUCT HEADERS	118,792	No. of Separators	4	2	0.50	59,396
CROSSOVER DUCT PANELS	267,969	No. of Separators	4	2	0.50	133,985
SEPARATOR TRANSFERS	314,288	No. of Separators	4	2	0.50	157,144
INTREX WATERWALL DOWNCOMERS	0					0
WALL SEAL ENCLOSURE PANELS	91,721	No. of Separators	4	2	0.50	45,861
WALL SEAL ENCLOSURE HEADERS	36,978	No. of Separators	4	2	0.50	18,489
WALL SEAL ENCLOSURE TRANSFERS	13,806	No. of Separators	4	2	0.50	6,903
FURNACE GRID NOZZLES ARROWHEAD	39,630	Grid Area	1,490	759	0.51	20,185
WALL SEAL ENCL. GRID NOZZLES ARROWHEAD	5,958	Grid Area	1,490	759	0.51	3,035
FURNACE GRID NOZZLES PIPE	12,515	Grid Area	1,490	759	0.51	6,374
WALL SEAL ENCL. GRID NOZZLES PIPE	1,882	No. of Separators	4	2	0.50	941
--						
--						
TOTALS (1) UNIT	**11,470,329**					**8,147,097**

Table 3-8. CFB Island Major Components

Component	Type	Design Basis
Primary air system	High temperature manifolded ducting	1150 deg. F, 15 psia, CO2/O2 mix
Flue gas recirculation system	ID Fan and gas recirc fan (latter is for standalone mode)	2300 kpph at 14 psia and 300 deg. F
Particulate removal system	Baghouse	2300 kpph inlet at 14 psia and 300 deg. F with 30 kpph particulate loading

Table 3-9. Gas Turbine Plant Components

Component	Type	Design Basis
GT Engine	Advanced ICAD	129 MW, 2300 F TIT, 700 psia TIP
Lube Oil System	CS reservoir and pumps, cartridge filters	
Generator Cooling System	Plate and frame	
Control package	Dedicated PLC with outputs to plant distributed control system	

Table 3-10. Steam Plant and Steam Cycle Balance of Plant Components

Component	Type	Design Basis
Steam Turbine		270 MW, 4050 psig, 1050 F/1050 F/2"Hg
Lube Oil System	Closed loop, pressure filter	
Control Fluid System	Electro-hydraulic	
Gland Steam System		
Control System	Tied to plant distributed control system	
Generator Cooling System	Plate and frame	
BOP Components		
Condensate Clean-up System		1400 klbs/hr feedwater/condensate flow, 2" Hg condenser presssure
Condensate System		
Feedwater System		
Cooling Water System		
Condenser Air Removal Sys.		
Auxiliary Steam System		
Turbine Bypass System		
Main Steam System		
Reheat Steam System		
Extraction Steam System		
Compressed Air System		
Closed Cycle Cooling Water System		
Waste Water Collection System		
Drips and Drain Collection System		
Small Bore Piping		
Natural Gas Supply & Distribution		
Building Drains (Roof & Floor)		
Seal Water Feed & Storage		
Condensate Reclaim System		
Pond Complex/Recirc System		
Aux Boiler System		

3.6 CO_2 COMPRESSION AND DEHYDRATION SYSTEM

CO_2 rich flue gas is removed from the plant at low pressure from the CFB exhaust (see Figure 2-1) and sent to a compression/dehydration system. At this point the flue gas stream is at 60 deg. F, and 14.6 psia pressure, and is composed of 3.1 v% O_2, 93.5 v% CO_2, and 1.8 v% H_2O. The gas goes through three stages of compression with inter-cooling to enhance compression performance and remove the moisture. Between second and third stages of compression a triethylene glycol dehydration package is used to dry the gas to less than 50 ppmv of moisture. After the final compression stage, which pressurizes the mixture to 850 psia, the stream is cooled down to 60 deg. F and sent to a flash tank separator. At the flash tank gaseous oxygen is recovered from the mixture and vented from the top of the tank back to the air separation unit, where it is routed back to the system. The CO2 and trace pollutants (such as SO2 and NOx) condense out and are removed from the bottom of the tank. Liquid CO2 and trace pollutants are then pumped to 1200 psia and leave the system.

The system is designed for a continuous flow of 570 klbs/hr of CO_2. The nominal inlet flow of flue gas to the system, including oxygen and moisture components is 595 kpph. The design has been based on using two identical parallel trains. A schematic of the CO2 compression and dehydration system is shown in Figure 3-3 (one of two trains shown).

The design layout has been taken from an internal Foster Wheeler commercial project and is very similar to that used by a Nexant study [6] on post-combustion CO_2 capture. The system components are shown in Table 3-11.

Table 3-11. CO_2 Compression and Dehydration System Components

Component	Type	Design Basis
Compressors	Multi-stage centrifugal (3 sections)	14 psia / 850 psia
Intercoolers		
Dehydration Package	Triethylene glycol	850 psia, 100 F
Knock-out pots		
Flash tank		1200 psia, 100 F
CO2 pump	Diaphragm	850 psia/1200 psia 570 kpph CO_2

3.7 AIR SEPARATION UNIT

The air separation unit (ASU) will be vendor supplied. The system will be designed to supply 6180 tons per day (TPD) of oxygen at 2psig (16.7 psia). The oxygen purity will be 99%. Two parallel air separation units of 3090 TPD capacity will be supplied.

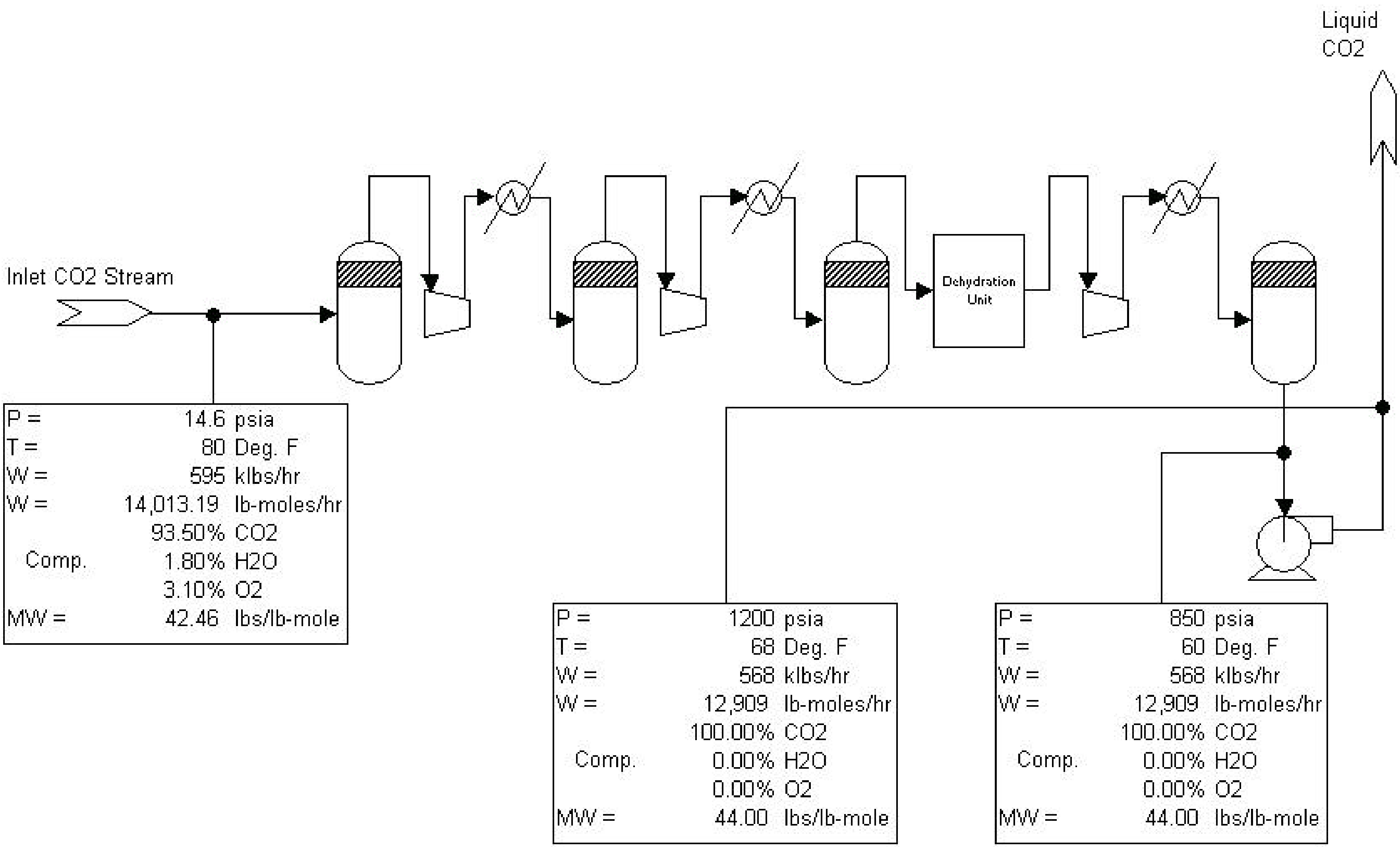

Figure 3-3. CO_2 Compression and Dehydration Package.

One of two identical trains is shown. The data indicates total flowrate so each train would receive half of the indicated flow.

3.8 PLANT START-UP SYSTEMS

Although the CO_2 hybrid is a “stackless” plant concept, a small vent (or start-up stack) is required to purge air out of the system during early stages of start-up, until flue gas recirculation can be initiated. This purge will be done by liquid CO_2 stored on site in pressurized tanks. Since the plant concentrates and liquefies CO_2, the start-up tanks will be refurbished during normal operation.

The start-up stack is designed to have 15% of the full flow capacity of the plant. Based on 5 volume changes and with 20% spare capacity, 8000 ft^3 of liquid CO_2 will be available for start-up. This will amount to six 10,000-gallon storage tanks designed for 1000 psia at 100 deg. F.

3.9 PLANT AUXILIARY POWER REQUIREMENTS

Table 3-12 lists the auxiliary power use for the CO2 hybrid plant in comparison with the air-fired reference plant. Auxiliary power use is seen to nearly triple from 37.5 MW to 104.1 MW between the reference plant and the CO2 hybrid. This is largely due to air separation (56 MW) and CO2 compression (26 MW) duties.

Table 3-12. Plant Auxiliary Power Usage

	Reference Plant	**CO_2 Hybrid**
ID and FD fans	5.7	3.2
Feedwater Pumps	9.9	10.1
Booster Compressor	15.1	1.0
Other Auxiliaries	6.7	6.8
Air Separation	0	55.5
CO2 Compression	0	27.5
Total	37.5	104.1

Section 4

PLANT COST AND ECONOMIC ANALYSIS

This section presents the plant costs and economic evaluation of the CO_2 Hybrid Plant. The results are expressed in terms of a levelized cost of electricity (COE), as well as the cost in dollars per tonne of CO_2 avoided (also called the mitigation cost). Component level contributions of the capital and operating costs, including fuel, are presented based on a 20-year book life and a constant dollar basis.

The results are compared with those from references [1] and [2] for IGCC and PC plants with pre and post-combustion CO_2 mitigation. The objective was to assess the feasibility of the CO_2 Hybrid Cycle concept as a means of power generation from coal without stack emissions. For comparison purposes, costs were calculated in year 2000 dollars, to match the plant investment year in references [1] and [2]. For the sake of completeness, results were also calculated in year 2004 dollars, to illustrate what would be the cost of building such a plant today. The main results of the economic evaluation are summarized in table 4-1 below. As the table shows, we estimate an appreciable increase between 2000 and 2004 due to rising steel prices and unfavorable changes in currency exchange.

Table 4-1. Summary of Capital Costs and Economics of the CO_2 Hybrid Plant*

Item	Year 2000 Dollars	Unit Cost	Year 2004 Dollars	Unit Cost
Total Plant Cost (TPC)	557,940,247	1892 $/kW	657,119,597	2289 $/kW
Operating and Maintenance	19,420,685	65.9 $/kW-yr	23,132,631	78.4 $/kW-yr
Consumables	1,454,364	0.56 mills/kWh	1,606,200	0.62 mills/kWh
Fuel	19,789,212	11.8 mills/kWh	19,997,000	11.9 mills/kWh
Levelized Busbar COE	75.3 mills/kWh		79.0 mills/kWh	
CO_2 Mitigation Cost	18.7 $/tonne of CO2 avoided		21.9 $/tonne of CO2 avoided	

* *Based on a net plant output of 295 MWe and a 65% capacity factor. COE levelized over 20 years.*

The COE and CO_2 mitigation cost are the most salient results of the analysis and allow direct comparison with other technologies for CO_2 sequestration. The percentage contribution of components that make up the COE is shown in Figure 4-1. The carrying charge component, by far the largest contributor, is calculated by a revenue requirement analysis that sets the net present value of the project to zero based on given costs of equity and debt capital. The total capital requirement (TCR) for the investment is obtained by adding the total plant cost (TPC) with allowance for funds during construction (AFDC), royalty allowance, pre-production costs, inventory capital, and land costs.

This report focuses on results calculated in year 2000 dollars. Details of the year 2004 results can be found in Appendix B.

4.1 MAIN ASSUMPTIONS

The economic analysis was carried out based on the EPRI Technical Assessment Guide (TAG) methodology. Plant capital costs were compiled under the Code of Accounts developed by EPRI and used in references [1] and [2].

The estimate basis and major assumptions are listed below:

- Total plant costs are estimated in January 2000 dollars.
- Plant book life is 20 years.
- The net power output for the hybrid plant without CO2 sequestration is 300 MWe, that for the hybrid plant with sequestration is 295 MWe (see Section 2).
- Capacity factor is 65%. The plant will operate at 100% load at 65% of the time.
- Cost of electricity (COE) was determined on a levelized constant dollar basis.
- Average annual ambient air conditions for material balances, thermal efficiencies and other performance related parameters are at a dry bulb temperature of 60 deg. F and an air pressure of 14.7 psia. An ambient temperature range of 20 deg. F to 95 deg. F was used for equipment sizing.
- The coal assumed in the analysis is Illinois #6 coal (see Table 2-1 for analysis). The sorbent (for the reference plant) is Greer limestone (Table 3-1).
- Terms used are consistent with the EPRI TAG.

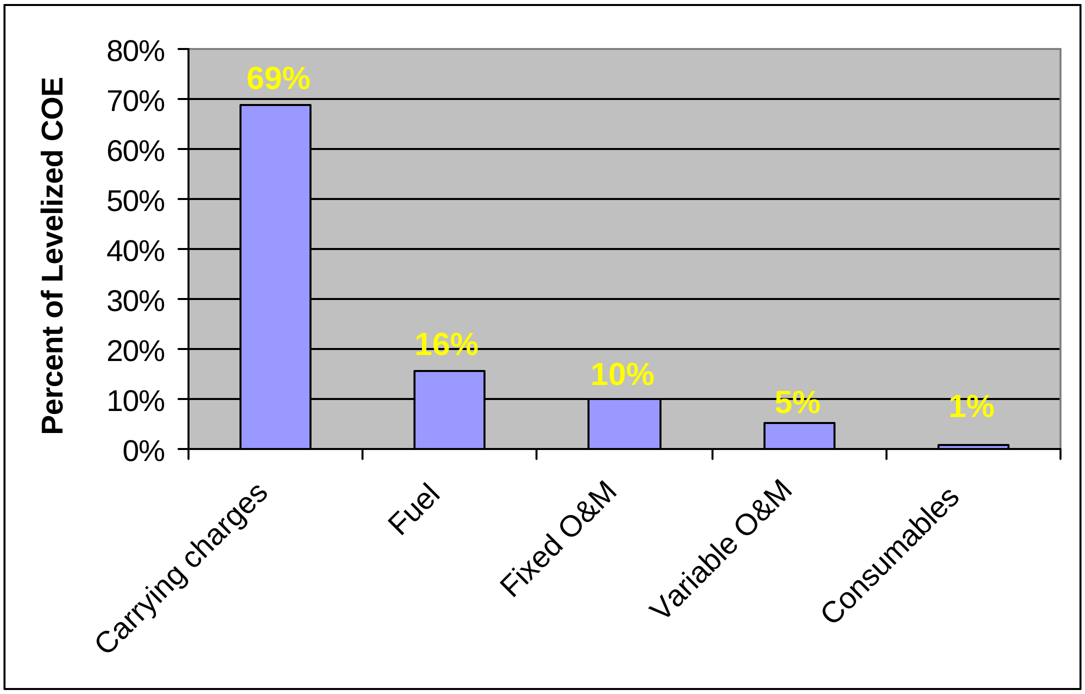

Figure 4-1. Components of Levelized Plant COE

4.2 TOTAL PLANT COST

The total plant cost (TPC), also referred to as the plant capital cost is comprised of the following elements:

- Bare erected plant cost.

- Overheads and fee for engineering and home office.

- Project and process contingencies.

A code of accounts was developed based on the EPRI account structure [1,2] for the estimate. The code allows direct comparison of individual systems costs among various clean coal technologies, with and without CO_2 mitigation, and provides a basis for future evaluation. Table 4-2 lists the code of accounts used for the CO_2 hybrid TPC evaluation.

Table 4-2. Code of Accounts for the CO_2 Hybrid

Account Number	Account Description
1	Coal and Sorbent Handling
2	Coal and Sorbent Preparation and Feed
3	Feedwater and Miscellaneous Balance-of-Plant Systems
4	Partial gasifier (Including Hot Gas Cleanup), Atmospheric CFB
5A	Gas Cleanup and Piping[13]
5B	CO_2 Compression and Dehydration
6	Combustion Turbine and Accessories
7	Ductwork and Stack
8	Steam Turbine Generator
9	Cooling Water System
10	Ash and Spent Sorbent Handling System
11	Accessory Electric Plant
12	Instrumentation and Control
13	Improvements to Site
14	Building and Structures

4.2.1 Bare Erected Cost

The bare erected cost is the sum of the costs of the plant equipment, field materials and supplies, engineering, installation, and commissioning labor costs. Equipment and major systems costs are obtained from estimates provided by technology vendors. In this study, detailed cost estimates from an earlier demonstration plant project[14] conducted in year 2001 were used as a reference. During that project, Foster Wheeler worked with a utility partner to develop plant cost information. This plant cost, adjusted for the investment year and contingencies was then used as the reference plant cost (GFBCC without CO_2 sequestration) for this study. As discussed in Section 3, the design of major components for the CO_2 hybrid plant (i.e. GFBCC ***with*** CO2 sequestration) was done in a manner to determine cost differentials relative to the reference plant.

The following sources have been used for costing the plant components:

Gasifier and related equipment: Foster Wheeler
Supercritical CFB Boiler and related equipment: Foster Wheeler
Sintered Metal Candle Filters: Pall Advanced Separation Systems
Combustion Turbine Package: Collaborative Advanced Gas Turbine (CAGT) Program [5]
Steam Turbine/Generator: 2001 GFBCC demonstration project utility partner
Balance-of-Plant (BOP) Major Systems: 2001 GFBCC demonstration project utility partner
Air Separation Unit (ASU): Air Products
CO_2 Compression and Dehydration Package: Foster Wheeler & Nexant Study [6]

For the steam turbine components, a cost adjustment factor, equal to the ratio of steam turbine gross power for the CO2 cycle to that of the reference GFBCC (270/250), was applied to the installed cost. The BOP items, virtually unchanged from the reference plant were taken as is from the 2001 estimate.

For the gas turbine, a capital cost estimate of $240/kW was used. This is the top end of the $200/kW – $240/kW range reported in reference [5]. Additional contingencies were applied, as shown in section 4.2.3 below.

[13] This typically includes gas cleanup systems for IGCC plants and is not applicable to this plant.
[14] Nominal 300 MWe GFBCC plant demonstration, front end engineering work. Completed in July 2001.

The July 2001 study values for equipment, engineering, installation, and commissioning costs were de-escalated to December 1999 by using an annual escalation rate of 3%.

4.2.2 Overheads and Fee for Engineering and Home Office

These costs were estimated at 6% of the bare erected cost. The 6% adder was applied uniformly to all accounts.

4.2.3 Contingencies

To account for the risk of cost overruns, project and process contingencies were applied to the plant cost estimate. Project contingencies refer to expected additional costs because of insufficient design detail available at the time of cost estimation. The less the level of detail, the greater should be the level of project contingencies.

Process contingencies, on the other hand account for commissioning and operational issues relating to components that are not fully commercial. In or case these include, among others, the partial gasifier, sintered metal barrier filter, and the ICAD combustion turbine.

Table 4-3 shows the project and process contingencies used for the study. Project contingencies were based on EPRI TAG guidelines, using the level of information that was available for the estimate. The process contingencies were derived based on the level of maturity of the technology in question, using FW's own experience.

Table 4-3. Project and Process Contingencies

		Contingencies	
Acct. No.	**Item/Description**	Process	Project
1	Coal and Sorbent Handling	0	15%
2	Coal and Sorbent Prep and Feed	5.0%	15%
3	Feedwater and Misc. BOP systems	0	15%
4	Gasifier and Accessories		
4.1 - 4.2	Gasifier and Auxiliaries	15%	15%
4.3	Air Separation Unit	0%	5%
4.4	CFB Boiler	0%	10%
4.4 - 4.9	Other Gasification Equipment	15%	15%
5A	Gas Cleanup and Piping		
5B	CO2 Compression	0	15%
6	Combustion Turbine/Accessories		
6.1	Combustion Turbine Generator	25%	15%
6.2 - 6.9	CT Accessories	0	5%
7	HRSG, Ducting, and Stack		
7.1	Heat Recovery Steam Generator	0	10%
7.2-7.9	CFB Accessories, Ductwork & Piping	0	15%
8	Steam Turbine Generator		
8.1	STG & Accessories	0	15%
8.2 - 8.9	Turbine Plant Auxiliaries and Steam Piping	0	15%
9	Cooling Water System	0	15%
10	Ash/Spent Sorbent Handling System	5%	15%
11	Accessory Electric Plant	5%	15%
12	Instrumentation and Control	5%	15%
13	Improvements to Site	0	15%
14	Buildings and Structures	0	15%

4.2.4 Capital Cost Results for the FW Hybrid Plant With and Without CO_2 Sequestration

Itemized capital costs for the hybrid plant with and without CO_2 removal are presented respectively in Tables 4-4 and 4-5. show the TPC summary for respectively the CO_2 hybrid plant (GFBCC with CO_2 sequestration) and GFBCC without CO_2 sequestration. Figure 4-2 illustrates the contributions of various systems to the TPC, with and without sequestration. It is clear that the air separation unit makes up a considerable portion of the investment, comparable to the cost of the entire gasification system.

Table 4-4. CO_2 Hybrid Plant Total Plant Cost Summary

CO2 Hybrid Plant Cost Summary Date: 10/14/2004
Cost Base 1999 (dec)
Plant Net Power Output: **294.9 MW**

Acct. No.	Item/Description	Equipment	Engineering	Installation	Comissioning	Bare Erected Cost	Eng'g H.O. Ovhd. & Fee	Contingency Process	Contingency Project	Total Cost	$/kW
1	Coal and Sorbent Handling	$ 1,731,583.21		$ 1,325,967.56		***$ 3,057,550.76***	$ 183,453.05	$ -	$ 458,632.61	***$ 3,699,636.42***	***13***
2	Coal and Sorbent Prep and Feed	$ 5,112,619.19		$ 3,915,010.93		***$ 9,027,630.12***	$ 541,657.81	$ 451,381.51	$ 1,354,144.52	***$ 11,374,813.95***	***39***
3	Feedwater and Misc. BOP systems	$ 14,681,861.60	$ 12,595,435.46	$ 11,387,002.38		***$ 38,664,299.45***	$ 2,319,857.97	$ -	$ 5,799,644.92	***$ 46,783,802.33***	***159***
4	Gasifier and Accessories										
4.1 - 4.2	Gasifier and Auxiliaries	$ 32,056,446.21	$ 9,039,508.11		$ 3,108,278.69	$ 44,204,233.01	$ 2,652,253.98	$ 6,630,634.95	$ 6,630,634.95	$ 60,117,756.89	204
4.3	Air Separation Unit	$ 78,199,703.40	w/equipment	w/equipment	w/equipment	$ 78,199,703.40	$ 4,691,982.20	$ -	$ 3,909,985.17	$ 86,801,670.77	294
4.4	CFB Boiler	$ 44,308,375.19	$ 8,428,923.00	$ 39,070,220.48	$ 2,555,742.97	$ 94,363,261.65	$ 5,661,795.70	$ -	$ 9,436,326.16	$ 109,461,383.51	371
4.4 - 4.9	Other Gasification Equipment	$ 8,001,059.48	$ -	$ 8,989,288.44	$ -	$ 16,990,347.91	$ 1,019,420.87	$ 2,548,552.19	$ 2,548,552.19	$ 23,106,873.16	78
	SUBTOTAL 4					***$ 233,757,545.96***				***$ 279,487,684.33***	***948***
5A	Gas Cleanup and Piping	$ -	$ -	$ -	$ -						
5B	CO2 Compression	$ 16,876,555.02	$ 5,348,174.46	$ 5,348,174.46	w/equip	***$ 27,572,903.95***	$ 1,654,374.24	$ -	$ 4,135,935.59	***$ 33,363,213.78***	***113***
6	Combustion Turbine/Accessories										
6.1	Combustion Turbine Generator	$ 30,960,000.00	w/equip	w/equip	w/equip	$ 30,960,000.00	$ 1,857,600.00	$ 7,740,000.00	$ 4,644,000.00	$ 45,201,600.00	153
6.2 - 6.9	CT Accessories	$ 3,725,172.22		$ 6,467,490.16		$ 10,192,662.38	$ 611,559.74	$ -	$ 509,633.12	$ 11,313,855.24	38
	SUBTOTAL 6					***$ 41,152,662.38***				***$ 56,515,455.24***	***192***
7	HRSG, Ducting, and Stack										
7.1	Heat Recovery Steam Generator					$ -	$ -	$ -	$ -	$ -	-
7.2-7.9	CFB Accessories, Ductwork & Piping	$ 9,688,913.07	$ -	$ 7,313,544.67	$ -	$ 17,002,457.75	$ 1,020,147.46	$ -	$ 2,550,368.66	$ 20,572,973.87	70
	SUBTOTAL 7					***$ 17,002,457.75***				***$ 20,572,973.87***	***70***
8	Steam Turbine Generator										
8.1	STG & Accessories	$ 24,822,721.30		$ 1,858,906.50	$ 1,372,712.49	$ 28,054,340.29	$ 1,683,260.42	$ -	$ 4,208,151.04	$ 33,945,751.75	115
8.2 - 8.9	Turbine Plant Auxiliaries and Steam Piping	$ 320,092.84		$ 2,889,585.44		$ 3,209,678.29	$ 192,580.70	$ -	$ 481,451.74	$ 3,883,710.73	13
	SUBTOTAL 8					***$ 31,264,018.58***				***$ 37,829,462.48***	***128***
9	Cooling Water System	$ 2,857,438.64		$ 2,527,405.76		***$ 5,384,844.40***	$ 323,090.66	$ -	$ 807,726.66	***$ 6,515,661.73***	***22***
10	Ash/Spent Sorbent Handling System	$ 1,482,602.01		$ 1,316,604.16		***$ 2,799,206.16***	$ 167,952.37	$ 139,960.31	$ 419,880.92	***$ 3,526,999.76***	***12***
11	Accessory Electric Plant	$ 15,599,801.05		$ 4,646,874.00		***$ 20,246,675.06***	$ 1,214,800.50	$ 1,012,333.75	$ 3,037,001.26	***$ 25,510,810.57***	***87***
12	Instrumentation and Control	$ 1,732,942.61		$ 510,553.63		***$ 2,243,496.23***	$ 134,609.77	$ 112,174.81	$ 336,524.44	***$ 2,826,805.25***	***10***
13	Improvements to Site	$ 2,812,588.94		$ 12,749,465.96		***$ 15,562,054.91***	$ 933,723.29	$ -	$ 2,334,308.24	***$ 18,830,086.44***	***64***
14	Buildings and Structures	$ 4,733,494.11		$ 4,442,407.18		***$ 9,175,901.29***	$ 550,554.08	$ -	$ 1,376,385.19	***$ 11,102,840.56***	***38***
		$ 299,703,970.08	***$ 35,412,041.04***	***$ 114,758,501.71***	***$ 7,036,734.15***	***$ 456,911,246.99***	***$ 27,414,674.82***	***$ 18,635,037.52***	***$ 54,979,287.39***	***$ 557,940,246.71***	***1,892***

Table 4-5. Total Plant Cost Summary for Hybrid Plant Without CO_2 Sequestration

GBBCC Combustion Hybrid Plant Cost Estimate Date: 10/14/2004
Cost Base 1999 (dec)
Plant Net Power Output: **300 MW**

						Bare Erected	Eng'g H.O.	Contingency		Total	
Acct. No.	Item/Description	Equipment	Engineering	Installation	Comissioning	Cost	Ovhd. & Fee	Process	Project	Cost	$/kW
1 & 2	Coal and Sorbent Handling	$ 2,310,802.41		$ 1,656,625.50		***$ 3,967,427.91***	$ 238,045.67	$ -	$ 595,114.19	***$ 4,800,587.78***	***16***
2	Coal and Sorbent Prep and Feed	$ 6,822,803.95		$ 4,891,301.39		***$ 11,714,105.34***	$ 702,846.32	$ 585,705.27	$ 1,757,115.80	***$ 14,759,772.73***	***49***
3	Feedwater and Misc. BOP systems	$ 14,681,861.60	$ 12,595,435.46	$ 11,387,002.38		***$ 38,664,299.45***	$ 2,319,857.97	$ -	$ 5,799,644.92	***$ 46,783,802.33***	***156***
4	Gasifier and Accessories										
4.1 - 4.2	Gasifier and Auxiliaries	$ 27,129,911.89	$ 9,039,511.24		$ 3,108,278.61	$ 39,277,701.74	$ 2,356,662.10	$ 5,891,655.26	$ 5,891,655.26	$ 53,417,674.37	178
4.3	Air Separation Unit	$ -	w/equipment	w/equipment	w/equipment	$ -	$ -	$ -	$ -	$ -	-
4.4	CFB Boiler	$ 54,307,948.03	$ 8,554,421.20	$ 39,087,711.03	$ 2,592,870.08	$ 104,542,950.35	$ 6,272,577.02	$ -	$ 10,454,295.03	$ 121,269,822.41	404
4.4 - 4.9	Other Gasification Equipment	$ 10,056,829.29	$ -	$ 8,989,288.44	$ -	$ 19,046,117.72	$ 1,142,767.06	$ 2,856,917.66	$ 2,856,917.66	$ 25,902,720.10	86
	SUBTOTAL 4					***$ 162,866,769.81***				***$ 200,590,216.88***	***669***
5A	Gas Cleanup and Piping	$ -	$ -	$ -	$ -						
5B	CO2 Compression	$ -	$ -	$ -	w/equip	***$ -***	$ -	$ -	$ -	***$ -***	***-***
6	Combustion Turbine/Accessories										
6.1	Combustion Turbine Generator	$ 22,959,128.81	w/equip	w/equip	w/equip	$ 22,959,128.81	$ 1,377,547.73	$ 2,295,912.88	$ 3,443,869.32	$ 30,076,458.74	100
6.2 - 6.9	CT Accessories	$ 3,728,951.87		$ 6,470,891.84		$ 10,199,843.71	$ 611,990.62	$ -	$ 509,992.19	$ 11,321,826.52	38
	SUBTOTAL 6					***$ 33,158,972.52***				***$ 41,398,285.26***	***138***
7	HRSG, Ducting, and Stack										
7.1	Heat Recovery Steam Generator					$ -	$ -	$ -	$ -	$ -	-
7.2-7.9	CFB Accessories, Ductwork & Piping	$ 9,688,913.07	$ -	$ 7,313,544.67	$ -	$ 17,002,457.75	$ 1,020,147.46	$ -	$ 2,550,368.66	$ 20,572,973.87	69
	SUBTOTAL 7					***$ 17,002,457.75***				***$ 20,572,973.87***	***69***
8	Steam Turbine Generator										
8.1	STG & Accessories	$ 22,984,001.20		$ 1,721,209.73	$ 1,434,945.55	$ 26,140,156.48	$ 1,568,409.39	$ -	$ 3,921,023.47	$ 31,629,589.34	105
8.2 - 8.9	Turbine Plant Auxiliaries and Steam Piping	$ 296,382.26		$ 2,675,542.08		$ 2,971,924.34	$ 178,315.46	$ -	$ 445,788.65	$ 3,596,028.45	12
	SUBTOTAL 8					***$ 29,112,080.82***				***$ 35,225,617.79***	***117***
9	Cooling Water System	$ 2,857,438.64		$ 2,527,405.76		***$ 5,384,844.40***	$ 323,090.66	$ -	$ 807,726.66	***$ 6,515,661.73***	***22***
10	Ash/Spent Sorbent Handling System	$ 1,542,661.01		$ 1,316,604.16		***$ 2,859,265.17***	$ 171,555.91	$ 142,963.26	$ 428,889.78	***$ 3,602,674.11***	***12***
11	Accessory Electric Plant	$ 15,599,801.05		$ 4,646,874.00		***$ 20,246,675.06***	$ 1,214,800.50	$ 1,012,333.75	$ 3,037,001.26	***$ 25,510,810.57***	***85***
12	Instrumentation and Control	$ 1,732,942.61		$ 510,553.63		***$ 2,243,496.23***	$ 134,609.77	$ 112,174.81	$ 336,524.44	***$ 2,826,805.25***	***9***
13	Improvements to Site	$ 2,812,588.94		$ 11,856,929.83		***$ 14,669,518.77***	$ 880,171.13	$ -	$ 2,200,427.82	***$ 17,750,117.72***	***59***
14	Buildings and Structures	$ 4,733,494.11		$ 4,442,407.18		***$ 9,175,901.29***	$ 550,554.08	$ -	$ 1,376,385.19	***$ 11,102,840.56***	***37***
		$ 204,246,460.76	***$ 30,189,367.91***	***$ 109,493,891.61***	***$ 7,136,094.24***	***$ 351,065,814.52***	***$ 21,063,948.87***	***$ 12,897,662.89***	***$ 46,412,740.29***	***$ 431,440,166.57***	***1,438***

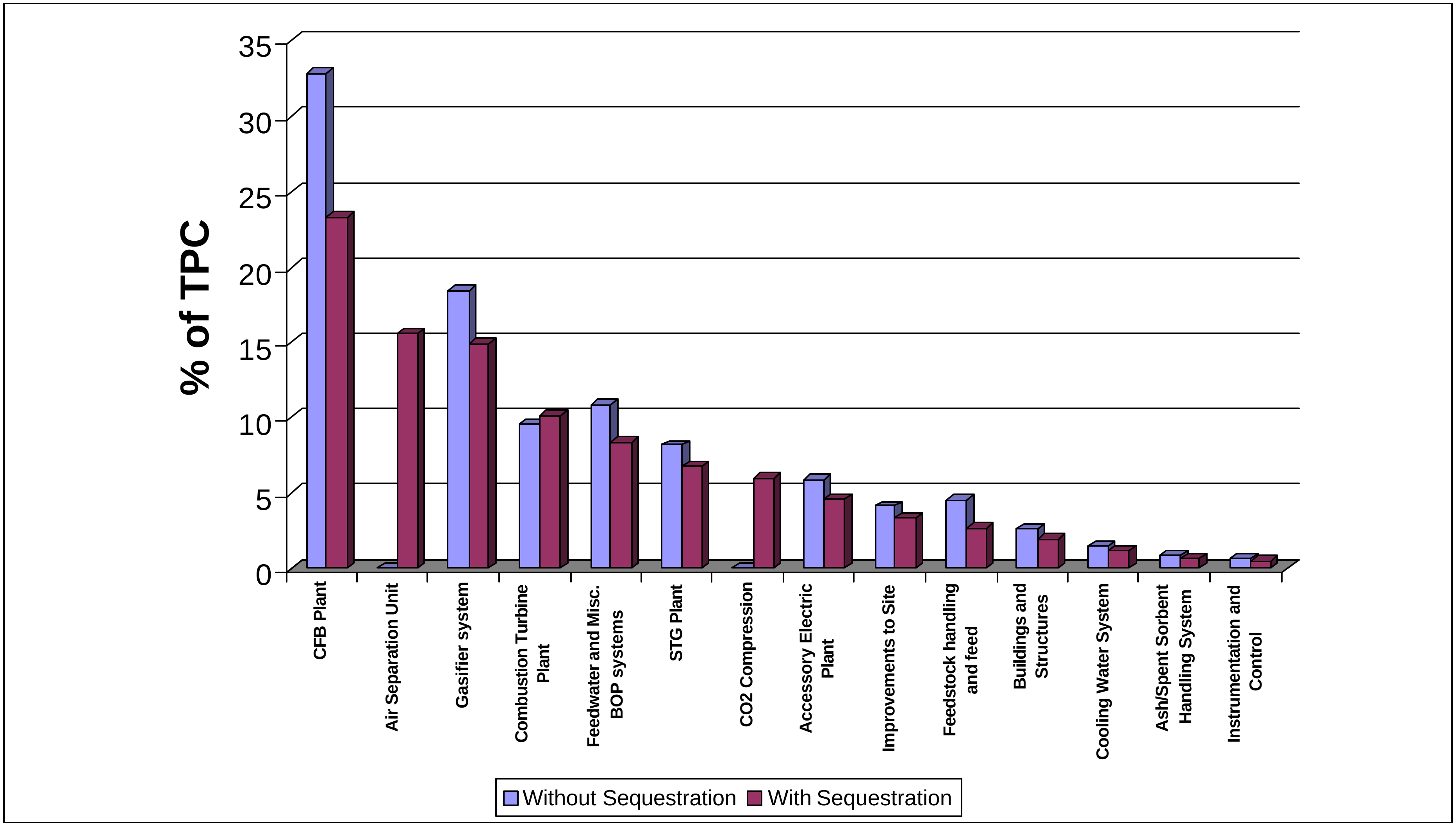

Figure 4-2. Components of Total Plant Cost

4.3 TOTAL PLANT INVESTMENT (TPI)

To determine the total investment required at date of start-up, the TPC is escalated by the average interest rate over the construction period. Unlike the TPC, which assumes instantaneous construction, TPI ensures that escalation of construction costs and allowance for funds used during construction is properly taken into account. The construction period was estimated to be 4 years. Assuming uniform cash flow over the construction period, the TPI was calculated as follows:

$TPI = TPC[1+i_{avg}]$

where

i_{avg} = Average interest rate over construction period
= (Interest rate)(Construction Period in Years)/2 = 10%

The annual interest rate was taken as 5%.

4.4 TOTAL CAPITAL REQUIREMENT (TCR)

The TCR includes all capital required to complete the project. TCR is the sum of TPI, pre-paid royalties, pre-production (start-up) costs, inventory capital, and land cost:

- Royalties costs are assumed inapplicable to the CO_2 hybrid plant
- Pre-production costs cover operator training, equipment checkout, major changes in plant equipment, extra maintenance, and inefficient use of fuel and other materials during start-up. They are estimated as follows:
  - 1 month of fixed operating costs, operating and maintenance labor, administrative and support labor, and maintenance materials.
  - 1 month of variable operating costs at full capacity (excluding fuel) – includes chemicals, water, and other consumables and waste disposal charges.
  - 25% of full capacity fuel cost for 1 month – covers inefficient operation that occurs during the start-up period.
  - 2% of TPI – covers expected changes and modifications to equipment that will be needed to bring the plant up to full capacity.
- Inventory capital is the capital required for initial inventories of fuel and other consumables, which are capitalized and included in the inventory capital account. The inventory capital is estimated as follows: Fuel and other consumables inventory (except water) is based on full-capacity operation for 15 days. In addition, an allowance of ½ percent of the TPC equipment cost is included for spare parts.
- Initial catalyst and chemical charge covers the initial cost of any catalysts or chemicals that are contained in the process equipment, but not in storage. In this study, this small charge was included with the equipment capital costs and is a part of the TPC.
- Land cost is based on 100 acres at $10,000/acre.

The TPI and the TCR cost components are shown on Tables 4-8a and 4-8b.

4.5 OPERATING COSTS AND EXPENSES

Operating costs were expressed in terms of the following categories:

- Operating Labor

- Maintenance Cost
    - Maintenance labor
    - Maintenance materials

- Administrative and Support Labor
- Consumables
- Fuel Cost

These values were calculated in consistence with EPRI TAG methodology. All costs were based on a first year basis with January 2000 dollars. The first year costs do not include start-up expenses, which are included in the TCR.

The cost categories listed above are calculated, on a dollars per year basis, as follows:

- Operating labor is calculated by multiplying the number of operating personnel with the average annual (burdened) compensation per person.
- Maintenance costs are estimated to be 2% of the TPC and are divided into maintenance labor and maintenance materials
    - Maintenance labor is estimated to be 40% of the total maintenance cost
    - Maintenance materials are estimated to be 60% of the total maintenance cost
- Administrative and support labor is estimated to be equal to 25% of the sum of operating and maintenance labor.
- Consumables are feedstock and disposal costs calculated from the annual usage at 100% load and 65% capacity factor. The costs is expressed in year 2000 dollars and levelized over 20 years on constant dollar basis.
- Fuel cost is calculated based on the assumed net cost for delivered coal, which is \$1.24/MMBtu. This is the same value used in references [1] and [2] for other clean coal plants. Fuel cost is determined on a first year basis and levelized over 20 years on a constant dollar basis. The calculation of first year fuel costs is done as follows:

$$\text{Fuel (tons/day)} = \frac{\text{(Plant Heat Rate)x (net capacity in kW) x 24 h/d}}{\text{HHV x 2000 lb/t}}$$

$$\text{Fuel Unit Cost (\$/ton)} = \text{HHV x 2000 lb/t x (Fuel Cost in \$/MMBtu) x } 10^6$$

$$\text{Fuel Cost (1}^{st}\text{ year)} = \text{Fuel (tons/day) x Fuel Unit Cost (\$/ton) x 365 d/yr x 0.65 (CF)}$$

The operating and maintenance costs, excluding fuel and consumables, are then combined and divided into two components: Fixed O&M, which is independent of power generation, and variable O&M, which is proportional to power generation. These are calculated as follows:

$$\text{Fixed O\&M (\$/yr)} = \text{Operating Labor + Maintenance Labor + Administrative and Support Labor}$$

$$\text{Fixed O\&M (\$/kW-yr)} = \frac{\text{Fixed O\&M(\$/year)}}{\text{Net Power (kW)}}$$

$$\text{Variable O\&M (\$/yr)} = \text{Maintenance Materials}$$

$$\text{Variable O\&M (¢/kWh)} = \frac{\text{Variable O\&M (\$/yr)}}{\text{Net Power (kW) x CF x 8760}}$$

Where CF is the plant capacity factor and 8760 is the total number of hours in one year.

The operating and maintenance costs for the hybrid plant with and without CO_2 sequestration are shown on Table 4-6 below. The "total production cost" shown at the bottom of the table expresses the charge of operating and maintaining the baseline plant (including fuel and consumable costs) in terms of cents per kilowatt-hour.

Table 4-6. Hybrid Plant Operating and Maintenance Costs and Expenses

With CO2 Sequestration					
Plant Net Power = 294.9 MW					
Capacity Factor = 65%					
Operating and Maintenance Costs					
			Unit Cost	**$/year**	**$/KW-yr**
Operating labor	72	people	$79,400.00	$5,716,800.00	$19.39
Maintenance Cost (2% of TPC)					
Maintenance Labor (40%)				$4,463,521.97	$15.14
Maintenance Materials (60%)				$6,695,282.96	$22.70
Administrative Support and Labor	25 % of O&M Labor			$2,545,080.49	$8.63
Consumable Operating Costs (Except Fuel)			**Unit Cost**	**$/year**	**¢/KWh**
Water	2800	kgals/day	$0.70	$465,010.00	0.018
Water Chemicals				$342,679.35	0.013
Limestone					
Start-up Fuel	5 starts/year			$30,000.00	0.001
Startup Electricity	6300	MWh/yr	$30	$189,000.00	0.007
Start-up Sand (Gasifier)	125	ton/yr	$5	$625.00	0.000
Ash disposal	15	ton/hr	$5	$427,050.00	0.017
					0.056
Fuel Cost (1999 Dollars)					
Heat Rate = 9452 Btu/kWh			**Unit Cost**	**$/year**	**¢/KWh**
	2802.78	MMBtus/hr	$1.24	$19,789,211.70	1.179
Total Production Cost					2.39 ¢/kWh

Table 4-6. (Continued from previous page)

Without CO2 Sequestration

Plant Net Power = 300 MW

Capacity Factor = 65%

Operating and Maintenance Costs		Unit Cost	$/year	$/KW-yr
Operating labor	60 people	$79,400.00	$4,764,000.00	$15.88
Maintenance Cost (2% of TPC)				
Maintenance Labor (40%)			$3,451,521.33	$11.70
Maintenance Materials (60%)			$5,177,282.00	$17.56
Administrative Support and Labor	25 % of O&M Labor		$2,053,880.33	$6.96

Consumable Operating Costs (Except Fuel)		Unit Cost	$/year	¢/KWh
Water	2545.4545 kgals/day	$0.70	$422,736.36	0.016
Water Chemicals			$479,271.81	0.018
Limestone	25 tons/hr	$12.00	$1,708,200.00	0.065
Start-up Fuel	5 starts/year		$30,000.00	0.001
Startup Electricity	6300 MWh/yr	$30	$189,000.00	0.007
Start-up Sand (Gasifier)	125 ton/yr	$5	$625.00	0.000
Ash disposal	80 ton/hr	$5	$2,277,600.00	0.087

Fuel Cost (1999 Dollars)

Heat Rate = 9452 Btu/kWh		Unit Cost	$/year	¢/KWh
	2437.14 MMBtus/hr	$1.24	$17,207,593.37	1.007

Total Production Cost 2.12 ¢/kWh

4.6 COST OF ELECTRICITY (COE)

The cost of electricity is the most salient result of economic analyses done on power generating plants. It assembles widely differing costs associated with dissimilar components in a single parameter that can be compared with other alternatives. COE is expressed in cents per kilowatt-hours or mills per kilowatt hours and is the levelized coal pile-to-busbar cost of energy that satisfies the plant revenue requirements.

The COE value is made up of contributions from the capital cost, called the carrying charge, the operating and maintenance costs, consumables, and fuel costs. The following relationship is used to calculate COE from these cost components:

$$COE = LCC + LFOM \times 100/(8760 \times CF) + LVOM + LCM + LFC$$

LCC = Levelized carrying charge, ¢/kWh
LFOM = Levelized fixed O&M, $/kW-yr
LVOM = Levelized variable O&M, ¢/kWh
LCM = Levelized consumables, ¢/kWh
LFC = Levelized fuel costs, ¢/kWh
CF = plant capacity factor (0.65)

The basis for calculating the capital investment and revenue requirements for the reference plant and the CO2 hybrid plant is given in Tables 4-7a and 4-7b, respectively. The basis and financial criteria are identical to those used in references [1] and [2]. The main output of this economic analysis, which is the capital investment and revenue requirements summary is given in Table 4-8a for the CO_2 hybrid plant and in Table 4-8b for the reference hybrid plant without CO_2 sequestration.

As Tables 4-8a and 4-8b show, the levelized COE for 65% capacity factor was calculated at 7.53 ¢/kWh with CO_2 sequestration, and at 6.09 ¢/kWh without CO_2 sequestration.

4.7 CO_2 MITIGATION COST (MC)

Another useful parameter to compare CO_2 capture technologies, besides the COE, is the mitigation cost expressed in \$/tonne of CO_2 avoided. This value shows the cost impact, in dollars per tonne of CO_2 that would otherwise be emitted, of going from the reference plant to a configuration that allows CO_2 capture.

The MC is calculated as follows:

$$MC = \frac{COE_{\text{with removal}} - COE_{\text{reference}}}{E_{\text{reference}} - E_{\text{with removal}}} \times 0.01\ \$/¢$$

COE = Cost of electricity in ¢/kWh
E = CO_2 emission in tonnes/kWh

As seen on Table 4-8a, the mitigation cost for the CO2 hybrid technology was calculated at 18.7 \$/tonne of CO_2 removed, for a capacity factor of 65%.

Table 4-7a. Estimate Basis/Financial Criteria for Revenue Requirement Calculations: CO_2 Hybrid

GENERAL DATA/CHARACTERISTICS

Plant Type:	Combustion Hybrid With CO2 Sequestration (Advanced CO2 Cycle)			
Plant Size:	294.9	MW, net		
Location:	Sea Level, Middletown USA			
Fuel: Primary/Secondary	Illinois #6		--	
Energy from Primary/Secondary Fuels:	9,504	Btu/kWh	--	Btu/kWh
Levelized Capacity Factor / Preproduction (equivalent months):	65%		1	months
Capital Cost Year Dollars (Reference Year Dollars):	2000 (January)			
Delivered Cost of Primary/Secondary Fuel:	1.24	$/MBtu	--	$/MBtu
Design/ Construction Period:	4	years		
Plant Start-up Date (1st year Dollars):	2004 (January)			
Land Area/Unit Cost:	100	acres	$10,000	/ Acre

FINANCIAL CRITERIA

Project Book Life:	20	years		
Book Salvage Value:	-	%		
Project Tax Life:	20	years		
Tax Depreciation Method:	Accel. Based on ACRS Class			
Property Tax Rate:	1.0	%		
Insurance Tax Rate:	1.0	%		
Federal Income Tax Rate:	34.0	%		
State Income Tax Rate:	4.2	%		
Investment Tax Credit/% Eligible	-	%	-	%
Economic Basis:	Over Book	Constant Dollars		

Capital Structure	% of Total		Cost(%)
Common Equity	45		12.0
Preferred Stock	10		8.5
Debt	45		9.0
Weighted Cost of Capital: (after tax)		8.81%	

Table 4-7b. Estimate Basis/Financial Criteria for Revenue Requirement Calculations: Reference Plant

GENERAL DATA/CHARACTERISTICS

Plant Type:	Combustion Hybrid without CO2 Sequestration (GFBCC Cycle)			
Plant Size:	300	MW, net		
Location:	Sea Level, Middletown USA			
Fuel: Primary/Secondary	Illinois #6		--	
Energy from Primary/Secondary Fuels:	8,124	Btu/kWh	--	Btu/kWh
Levelized Capacity Factor / Preproduction (equivalent months):	65%		1	months
Capital Cost Year Dollars (Reference Year Dollars):	2000 (January)			
Delivered Cost of Primary/Secondary Fuel:	1.24	$/MBtu	--	$/MBtu
Design/ Construction Period:	4	years		
Plant Start-up Date (1st year Dollars):	2004 (January)			
Land Area/Unit Cost:	100	acres	$10,000	/ Acre

FINANCIAL CRITERIA

Project Book Life:	20	years		
Book Salvage Value:	-	%		
Project Tax Life:	20	years		
Tax Depreciation Method:	Accel. Based on ACRS Class			
Property Tax Rate:	1.0	%		
Insurance Tax Rate:	1.0	%		
Federal Income Tax Rate:	34.0	%		
State Income Tax Rate:	4.2	%		
Investment Tax Credit/% Eligible	-	%	-	%
Economic Basis:	Over Book	Constant Dollars		

Capital Structure	% of Total		Cost(%)
Common Equity	45		12.0
Preferred Stock	10		8.5
Debt	45		9.0
Weighted Cost of Capital: (after tax)		8.81%	

Table 4-8a. Capital Investment and Revenue Requirement Summary: CO_2 Hybrid

Case:	CO2 Hybrid Plant		
Plant size:	294.9 MW net	**Net Cycle Efficiency:**	35.90%
Primary/Secondary Fuel (type):	Illinois #6	**Net Plant Heat Rate (NPHR):**	9,504 (Btu/kWh)
Design/Construction:	4 (years)	**Cost:**	1.24 ($/MMBtu)
TPC (Plant Cost) Year:	1999 (dec.)	**Book Life:**	20 (years)
Capacity Factor:	65 (%)	**TPI Year:**	2000 (Jan)
		CO2 Removed:	1,617,096 (tons/year)

CAPITAL INVESTMENT		**$x1000**		**$/kW**
Process Capital and Facilities		456,911		1,549.4
Engineering (Incl. C.M., H.O., and Fee)		27,415		93.0
Process Contingency		18,635		63.2
Project Contingency		54,979		186.4
TOTAL PLANT COST (TPC)		$ 557,940		1,892.0
TOTAL CASH EXPENDED	$ 557,940			
AFDC	$ 49,489			
TOTAL PLANT INVESTMENT (TPI)		$ 607,430		
Royalty Allowance				
Preproduction Cost		$14,476		
Inventory Capital		$2,750		
Initial Chemicals and Catalysts (w/equip)				
Land Cost		$1,000		
TOTAL CAPITAL REQUIREMENT (TCR)		$ 625,655		
OPERATING & MAINTENANCE COSTS (1999 DOLLARS)		**$x1000**		**$/kW-yr**
Operating Labor		5,717		19.4
Maintenance Labor		4,464		15.1
Maintenance Materials		6,695		22.7
Administrative Support and Labor		2,545		8.6
TOTAL OPERATION & MAINTENANCE		$ 19,421		65.9
FIXED O & M			43.15 $/kW-yr	
VARIABLE O & M			0.40 c/kWh	
CONSUMABLE OPERATING COSTS less Fuel (1999 DOLLARS)		**$x1000**		**c/kWh**
Water		465		0.02
Chemicals		343		0.01
Other Consumables		220		0.01
Waste Disposal		427		0.02
TOTAL CONSUMABLE OPERATING COST		$ 1,454		0.06
FUEL COST (1999 Dollars)		$ 19,789		1.18
		Levelized (Over Book Life $)		
PRODUCTION COST SUMMARY		**$/tonne CO2 avoided**		**c/kWh**
Fixed O&M		2.02		0.76
Variable O&M		1.23		0.40
Consumables		(1.78)		0.06
By-product Credit		-		0
Fuel		2.21		1.18
TOTAL PRODUCTION COST		3.68		2.39
LEVELIZED CARRYING CHARGES (Capital)		14.98		$5.14
LEVELIZED (Over Book Life) BUSBAR COST OF POWER		18.65		$7.53

Table 4-8b. Capital Investment and Revenue Requirement Summary: Reference Plant

Case:	CO2 Hybrid Plant		
Plant size:	300 MW net	**Net Cycle Efficiency:**	42.0%
Primary/Secondary Fuel (type):	Illinois #6	**Net Plant Heat Rate (NPHR):**	8,124 (Btu/kWh)
Design/Construction:	4 (years)	**Cost:**	1.24 ($/MMBtu)
TPC (Plant Cost) Year:	1999 (dec.)	**Book Life:**	20 (years)
Capacity Factor:	65 (%)	**TPI Year:**	2000 (Jan)
		CO2 Removed:	- (tons/year)

CAPITAL INVESTMENT		**$x1000**		**$/kW**
Process Capital and Facilities		351,066		1,170.2
Engineering (Incl. C.M., H.O., and Fee)		21,064		70.2
Process Contingency		12,898		43.0
Project Contingency		46,413		154.7
TOTAL PLANT COST (TPC)		$ 431,440		1,438.1
TOTAL CASH EXPENDED	$ 431,440			
AFDC	$ 38,269			
TOTAL PLANT INVESTMENT (TPI)		$ 469,709		
Royalty Allowance				
Preproduction Cost		$11,458		
Inventory Capital		$2,109		
Initial Chemicals and Catalysts (w/equip)				
Land Cost		$1,000		
TOTAL CAPITAL REQUIREMENT (TCR)		$ 484,276		

OPERATING & MAINTENANCE COSTS (1999 DOLLARS)	**$x1000**		**$/kW-yr**
Operating Labor	4,764		15.9
Maintenance Labor	3,452		11.5
Maintenance Materials	5,177		17.3
Administrative Support and Labor	2,054		6.8
TOTAL OPERATION & MAINTENANCE	$ 15,447		51.5
FIXED O & M		34.23 $/kW-yr	
VARIABLE O & M		0.30 c/kWh	

CONSUMABLE OPERATING COSTS less Fuel (1999 DOLLARS)	**$x1000**	**c/kWh**
Water	423	0.02
Chemicals	2187	0.08
Other Consumables	220	0.01
Waste Disposal	2278	0.09
TOTAL CONSUMABLE OPERATING COST	$ 5,107	0.19
FUEL COST (1999 Dollars)	$ 17,208	1.01

	Levelized (Over Book Life $)	
PRODUCTION COST SUMMARY	**$/tonne CO2**	**c/kWh**
Fixed O&M		0.60
Variable O&M		0.30
Consumables		0.19
By-product Credit		0
Fuel		1.01
TOTAL PRODUCTION COST		2.11
LEVELIZED CARRYING CHARGES (Capital)		$3.98
LEVELIZED (Over Book Life) BUSBAR COST OF POWER		6.09

Section 5

RESULTS AND DISCUSSION

The primary objective of this study was to investigate the viability of a promising method to generate power from coal and sequester 100% of the carbon dioxide and other gaseous pollutants. The CO_2 recycle method, applied in this study to the Foster Wheeler hybrid cycle (GFBCC), was targeted to achieve:

- Lowest efficiency penalty for carbon sequestration than alternative methods
- Lowest CO2 mitigation cost compared to alternatives
- Lowest cost of electricity (COE) among coal fired alternatives

As the data and graphs presented in this section will show, the first two of these objectives have been demonstrated. The CO_2 recycle method results in the lowest efficiency drop (6.1%) among leading alternatives, matching[15] or below[16] that of the IGCC design with pre-combustion CO_2 separation. The CO_2 recycle method also results in the lowest cost to convert the reference plant to capture 100% of the CO_2.

This section will also reiterate that the third objective, lowest COE among alternatives, has not been satisfied with the CO_2 hybrid cycle technology. Although the cost *difference* between the reference plant and the CO_2 sequestering configuration is the lowest among alternative technologies the base cost of the reference plant (the air fired GFBCC) is 30% higher than the lowest IGCC estimate (1438 $/kW vs. 1111 $/kW). This leads to higher cost of electricity for both the reference plant and the CO2 hybrid plant.

5.1 PERFORMANCE RESULTS

As discussed in detail in section 2.3, the CO_2 hybrid cycle eliminates all indirect losses, such as those created by water-gas shift, CO_2 absorption, and solvent regeneration, trading them off with greater direct losses due to the larger air separation unit. The CO_2 hybrid cycle uses up to 2.5 times the oxygen required by an IGCC with pre-combustion CO_2 separation (see section 2.5). The net result of this trade-off is an efficiency penalty of 6.1%.

Figure 5-1 compares the efficiency penalties of CO2 sequestration by various methods. The data is from reference [1] for all cases except IGCC case 3E, which is from reference [2]. The following plants are used for the comparison:

- Supercritical PC with post-combustion CO2 removal with amine (Reference [1], case 7A)
- Natural Gas Combined Cycle (NGCC) plant with post-combustion CO2 removal using amine (Ref. [1], case 1A)
- IGCC with pre-combustion CO2 removal using a double-stage selexol unit (Ref. [1], case 3A)
- IGCC with pre-combustion CO2 removal with water scrubber using a double-stage selexol unit (Ref. [2], case 3E)

As the chart shows, the CO2 hybrid plant matches the efficiency drop of the best IGCC design with pre-combustion CO2 removal. The second IGCC design is different in that it incorporates a water scrubber

15 Reference [1], case 3A
16 Reference [2], case 3E, described as a more “realistic” version of case 3A from reference [1].

for the water-gas shift reaction. The authors [2] state that the inclusion of the water scrubber makes the IGCC design more realistic for a commercial application.

The post-combustion technologies, namely the PC and NGCC, where the CO2 is scrubbed out of the flue gases using a solvent, suffer the highest penalties. This is not surprising, since the flue gas partial pressure of CO2 (a driving force for the absorption) is far lower than that in the IGCC syngas before combustion in the gas turbine.

A point worthy of note is that the power consumption values used in references [1] and [2] for IGCC plants are about 12% lower (in kWh/lb of oxygen generated) than the values used here. For this study, power consumption values communicated by Air Products were used. We have confidence in our values and believe that they reflect the energy costs more accurately than the values used in references [1] and [2].

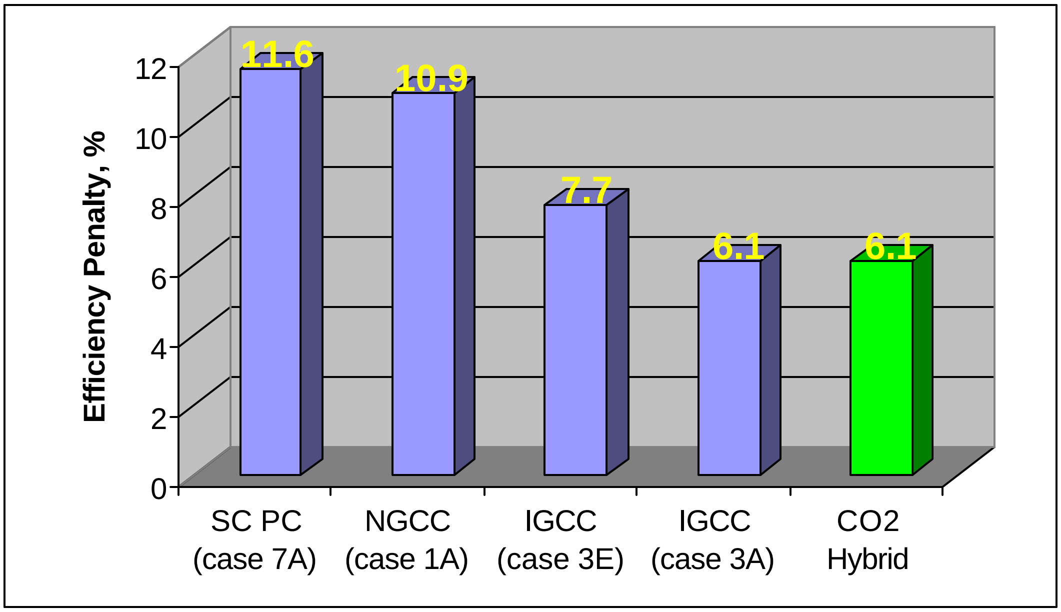

Figure 5-1. Efficiency Penalty Associated With CO2 Sequestration For Various Capture Methods. *Supercritical PC, NGCC, and IGCC case 3A taken from reference [1]. IGCC case 3E taken from reference [2].*

The energy efficiency penalties impact the bottom line of plant operations and reduce the net cycle efficiencies of the plants utilizing these CO2 mitigation methods. Figure 5-2 shows the net plant efficiencies of the technologies compared in Figure 5-1. Although the NGCC with post combustion CO2 removal ha the highest plant efficiency at 39.2%, this is a meager value compared to its without sequestration value of 50.1%. The CO2 hybrid retains a reasonable 35.9% efficiency, which is on a par with the best modern conventional coal plants of today.

Since the main objective in developing these technologies is emissions mitigation, another performance measure to use for ranking them is the emissions per unit of net power produced. The CO2 hybrid cycle is the only truly "zero emission" plant, where not only the CO2 but all of the other emissions (e.g. NOx and SO_2) are 100% sequestered. An emissions comparison is shown on Table 5-1.

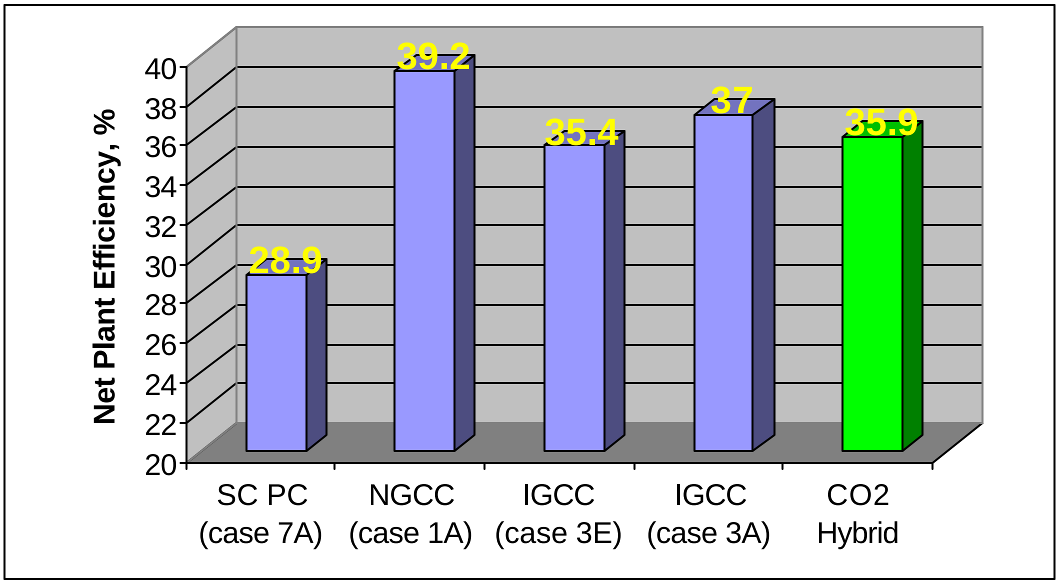

Figure 5-2. Net Cycle Efficiencies of Power Plants With CO_2 Sequestration.
Supercritical PC, NGCC, and IGCC case 3A taken from reference [1]. IGCC case 3E taken from reference [2].

Table 5-1. Emissions From Various Power Plants with CO_2 Sequestration.
Supercritical PC, NGCC, and IGCC case 3A taken from reference [1]. IGCC case 3E taken from reference [2].

Technology	Emissions (lbs/MWh)			
	CO_2	NOx	SO_2	Particulate
SC PC (case 7A)	237	1.85	1.01	0.12
NGCC (case 1A)	98.86	0.28	Negligible	Negligible
IGCC (case 3A & 3E)	162	0.25	Negligible	Negligible
CO_2 Hybrid	Negligible	Negligible	Negligible	Negligible

As these results clearly show, on a sheer performance basis, the CO_2 hybrid cycle is competitive with the IGCC with pre-combustion CO_2 capture, and can beat it on an emissions basis.

5.2 ECONOMICS RESULTS

There are two main parameters that are salient to this economic analysis: (1) the cost of electricity (COE) of the technology and (2) the cost of CO_2 mitigation. The first expresses, in a single parameter, what the cost of the *technology* is in terms of what a plant owner must charge for the electricity in order to satisfy the returns to debt and equity that are demanded by investors. The second describes what the cost of the sequestration *method* would be to implement on any given technology platform. The first depends on the capital and operating costs of the base technology while the second looks only at the cost *difference* between the base technology with and without sequestration.

In evaluating the CO_2 recycle *method* the second parameter is the more relevant because this method can be applied to various technologies including IGCC and Supercritical PC. However, in evaluating the feasibility of the CO_2 hybrid plant technology (i.e. the CO_2 recycle method applied specifically to the GFBCC), the COE would be most relevant.

The economic analysis determined that the CO2 Hybrid Cycle has one of the highest capital costs and the highest COE among alternative technologies reported [1,2]. Table 5-2 compares the capital costs

(TPC per kW of net power) of alternative technologies with CO2 sequestration. The IGCC estimate varies quite widely based on the source of the data. All of the data on Table 5-2, except the CO2 Hybrid, were taken from references [1] and [2].

Table 5-2. Capital Costs of Carbon Sequestering Technologies

Technology	Study	Without Sequestration			With Sequestration			
		Capital Cost	Net Output	Levelized COE	Method	Capital Cost	Net Output	Levelized COE
		$/kW	MWe	¢/kWh		$/kW	MWe	¢/kWh
IGCC	Audus/IEA[17]	1470	408	4.8	Pre-comb.	2200	382	6.9
IGCC	Herzog/MIT[18]	1401		4.99	Pre-comb.	1909		6.69
IGCC	Simbeck/SFA Pacific[19]	1100	400	3.9	Pre-comb.	2075	400	7.0
IGCC	EPRI/Parsons[20]	1111	425	4.77	Pre-comb.	1510	404	6.26
SC PC	EPRI/Parsons[21]	1143	460	5.15	Post-comb.	1981	330	8.56
NGCC	EPRI/Parsons[22]	505	510	3.42	Post-comb.	1010	400	5.79
CO_2 Hybrid	Foster Wheeler (this study)	1438	300	6.09	CO_2 recycle	1892	295	7.53

While this table shows the CO_2 Hybrid plant to be the second highest COE, it is noteworthy that the *difference* between the cost of the technology with sequestration and that without sequestration is lowest with the CO_2 recycle method. This is illustrated in Figure 5-3, which shows the % increase in the capital cost and the COE of power plants as a result of incorporating CO_2 sequestration.

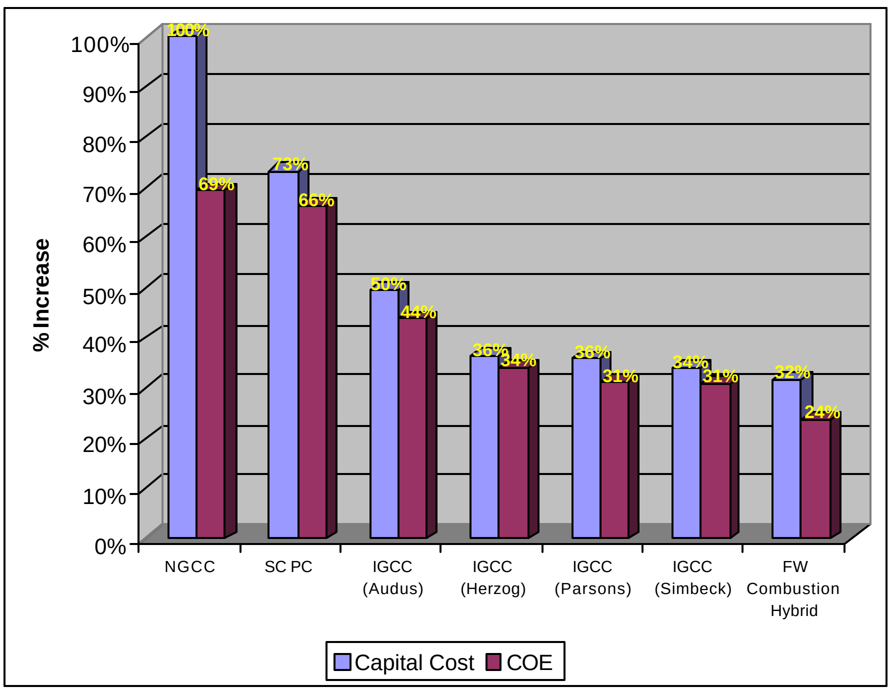

Figure 5-3. Increase In Plant Costs as a Result of Building CO_2 Sequestration Capability

[17] Coal $1.5/GJ
[18] Coal $1.18/GJ
[19] Coal $0.95/GJ
[20] Case 3B (without sequestration) and case 3E (with sequestration). Coal $1.22/GJ
[21] Case 7C (without sequestration) and case 7A (with sequestration). Coal $1.22/GJ
[22] Case 1C (without sequestration) and case 1A (with sequestration). Coal $1.22/GJ

This is because of the symplicity of the CO_2 recycle method and the relatively few components and systems required to convert the base plant. Fewer systems leads to lower capital and operating cost additions. Not surprisingly, this results in the lowest CO_2 mitigation cost, approaching $10/tonne of CO_2 avoided at 85% capacity factor, for the CO_2 hybrid plant. This is shown in Figure 5-4 below.

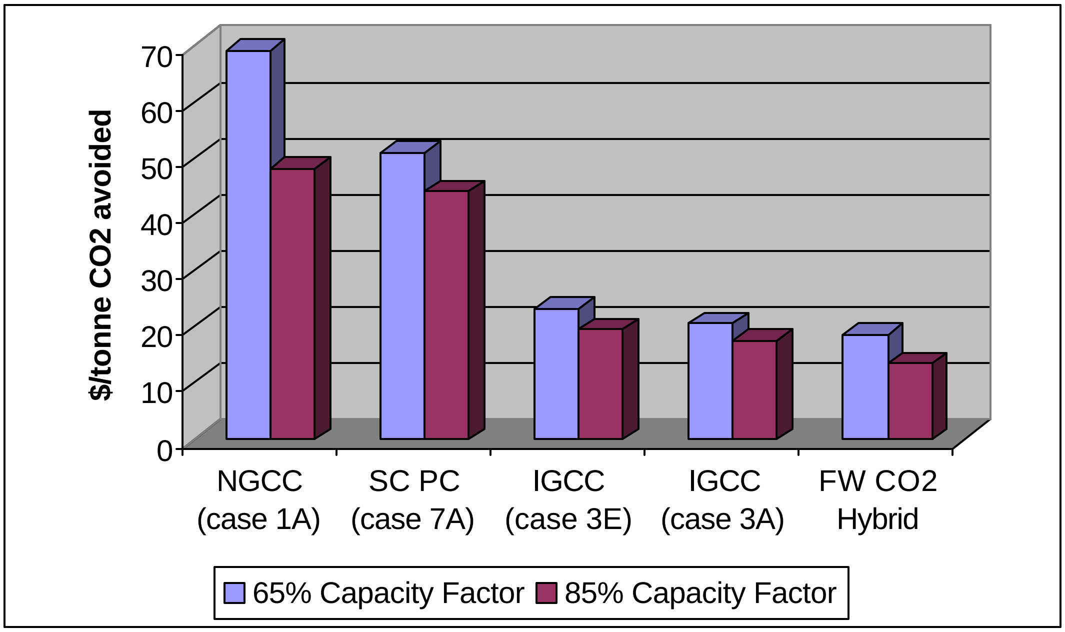

Figure 5-4. Cost of CO_2 Mitigation For Various Plant Technologies

5.3 PLANT AVAILABILITY ARGUMENTS

The Foster Wheeler combustion hybrid technology, also called the gasification fluid bed combined cycle (GFBCC), is believed to offer certain operational advantages that were not taken into account in this economic evaluation. Two main advantages are fuel flexibility (ability to use both low and high rank coals) and greater anticipated reliability of the GFBCC compared to an IGCC plant. Both of these advantages exist because circulating fluid bed technology is used for both the gasification and char combustion processes. The fluid bed gasifier operates at a modest 2000 deg. F, virtually eliminating ash slagging and refractory problems associated with high temperature operation in IGCC gasifiers.

It is, therefore, reasonable to assume that the GFBCC and the CO2 hybrid plant will achieve greater plant availabilities as developed technologies and may even be considered “base-loaded” plants. A sensitivity analysis, done using the capacity factor as a variable, showed that the CO_2 hybrid operating at 82% capacity factor breaks even with the best IGCC[23] result reported in reference [2]. This breakeven is illustrated in Figure 5-5.

[23] Ref. [2] case 3E operating at 65% capacity factor

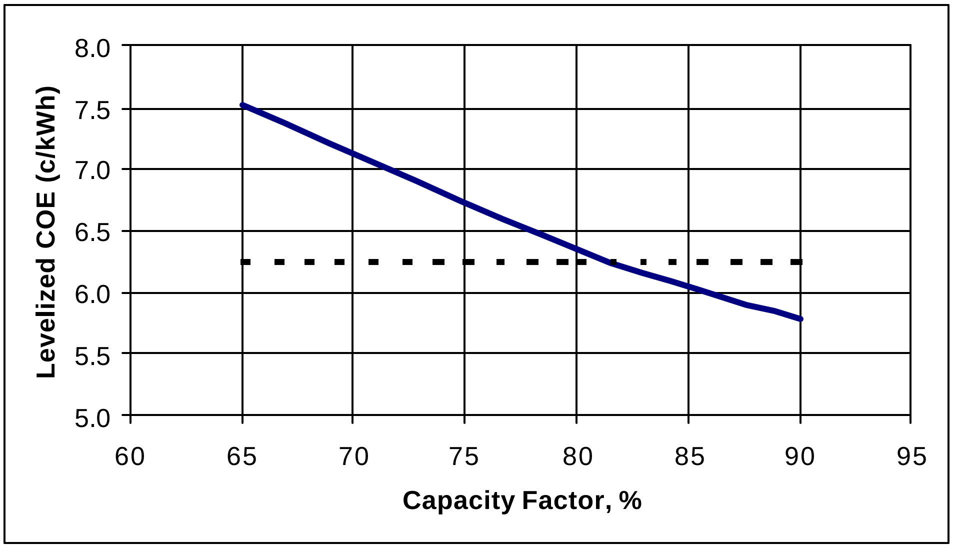

Figure 5-5. CO_2 Hybrid Plant COE as a Function of Capacity Factor.
The dashed line shows the COE of an IGCC operating at 65% CF (ref. [2] case 3E).

Section 6

CONCLUSIONS

Although the potential of the CO_2 recycle concept to reduce the *incremental* costs of avoiding CO_2 emissions was demonstrated, the Foster Wheeler combustion hybrid cycle used as a reference for the study led to higher capital and levelized costs than the leading estimates for IGCC plants with pre-combustion CO_2 removal.

It is believed that the cost of the reference plant is higher because the estimate is based on a demonstration plant. One can make the argument that application of the concept at a larger commercial scale could bring the cost down on a par with the IGCC cases used as comparison. However, there is currently no data available to substantiate this thought.

Other arguments in favor of the combustion hybrid concept are:

- Plant reliability would be higher due to simpler configuration compared to IGCC. This would imply that the combustion hybrid might be able to take credit for a higher capacity factor than the IGCC, changing the economic equation. Operating the CO2 hybrid cycle at 80% CF would match the cost of electricity of the best IGCC estimate derived at 65% CF.

- The combustion hybrid plant utilizes circulating fluidized beds for both gasification and combustion. Furthermore, the gasifier operates at a modest 2000 deg. F. In addition to improving availability, these attributes allow the technology to be highly fuel flexible and able to utilize a wide variety of inexpensive coals.

Gas turbine availability is also likely to be an issue in the commercialization of this concept. Gas turbine manufacturers must see a clear market demand for this technology to engage in costly development work for a high pressure-ratio engine that can efficiently use a CO2 rich working fluid. Market demand could be created if the concept can demonstrate a clear and substantial benefit compared to the IGCC with pre-combustion capture approach.

Looking Ahead

To further evaluate the merits and potential of the CO_2 recycle concept, a conventional IGCC plant should be used as a reference. The IGCC would then be converted to a CO2 cycle configuration by enlarging the air separation unit to provide 100% stiochiometric oxygen, instead of 40 to 50%, and going to an oxygen and recycled CO_2 fired turbine. This would enable an enriched CO_2 stream that could be drawn off and sequestered.

The economic feasibility of such a conversion can be directly compared with the same reference IGCC with pre-combustion CO_2 removal.

Air separation costs make up a significant portion of the cost adder to the base plant configuration. New developments in air separation, such as oxygen transport membranes (OTM) have the potential to reduce this cost. The feasibility of the concept should be evaluated with the OTM, and with customized and optimized cryogenic ASU designs integrated into the cycle (best possible cryogenic design).

Lastly, better gas turbine definition (design, performance, and economics) and gas turbine availability assessments are necessary to evaluate the future commercialization of the concept.

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APPENDIX A

HEAT AND MATERIAL BALANCES
CASES 1, 3, AND 4

Case 1: Commercially available industrial turbomachinery with gas reheat is used for the topping cycle/ASU delivers oxygen at 15 psia/CO_2 is removed from the cycle at low pressure (compressor inlet).

Case 2: Deleted (a variation of case 1)

Case 3: Advanced Intercooled Aeroderivative (ICAD) engine used for the topping cycle/ASU delivers oxygen at 15 psia/CO2 is removed from the cycle at high pressure (compressor discharge)

Case 4: Advanced Intercooled Aeroderivative (ICAD) engine used for the topping cycle/ASU delivers oxygen at 850 psia/CO2 is removed from the cycle at high pressure (compressor discharge)

Note: As mentioned in Section 2 of the report, these cases were examined but not used in the study. Case 5, detailed in the report, is the reference case for plant conceptual design. Heat and material balances for cases 1, 3, and 4 are presented here as additional information.

CASE 1

ASU	
O_2 Flow, klb/hr	422
Power Consumption, MW	42.8
O_2 Delivery Pressure, psia	15
Gasifier	
Coal Feed Rate, klb/hr	99
O_2 Feed Rate, klb/hr	51
Steam Feed Rate, klb/hr	9
Syngas Flowrate, klb/hr	294
Syngas LHV, btu/scf	189
Syngas Cooler Duty, mmbtu/hr	144
Char Flowrate, klb/hr	28
Gas Turbine	
Type	2-Stg Industrial
Output, MW	47
Exhaust Flow, klb/hr	2050
1st Stage Turbine Inlet Temp., F	1000
2nd Stage Turbine Inlet Temp., F	1600
Turbine Exhaust Temp., F	925
Steam Turbine	
Output, MW	258
Main Steam Flowrate, klb/hr	1300
Reheat Steam Flowrate, klb/hr	1369
CFB Boiler	
Coal Feed Rate, klb/hr	105
Flue Gas Flowrate, klb/hr	2167
CO_2 Compression and Dehyd.	
CO_2 Inlet Pressure, psia	15
CO_2 Flowrate, klb/hr	480
CO_2 Sequestered as	Liquid
Overall System Performance	
Net Efficiency, %	30.9
Net Power Output, MWe	216
Auxiliary Power Consumption, MW	89.3

The main feature in this case is the use of an industrial CO_2 compressor and a two stage industrial turbine. The HP turbine section takes the gas in at 700 psia and 1000 deg.F (its maximum operating temperature) and releases it at 250 psia and 800 deg. F. The gas is then re-heated by firing with syngas to 1600 deg. F, and passes through the IP section of the turbine.

This case represents and alternative that uses existing compressor and gas turbine technology for the topping cycle. However, the limitation of the HP section inlet temperature to 1000 deg. F limits the cycle efficiency to around 31%.

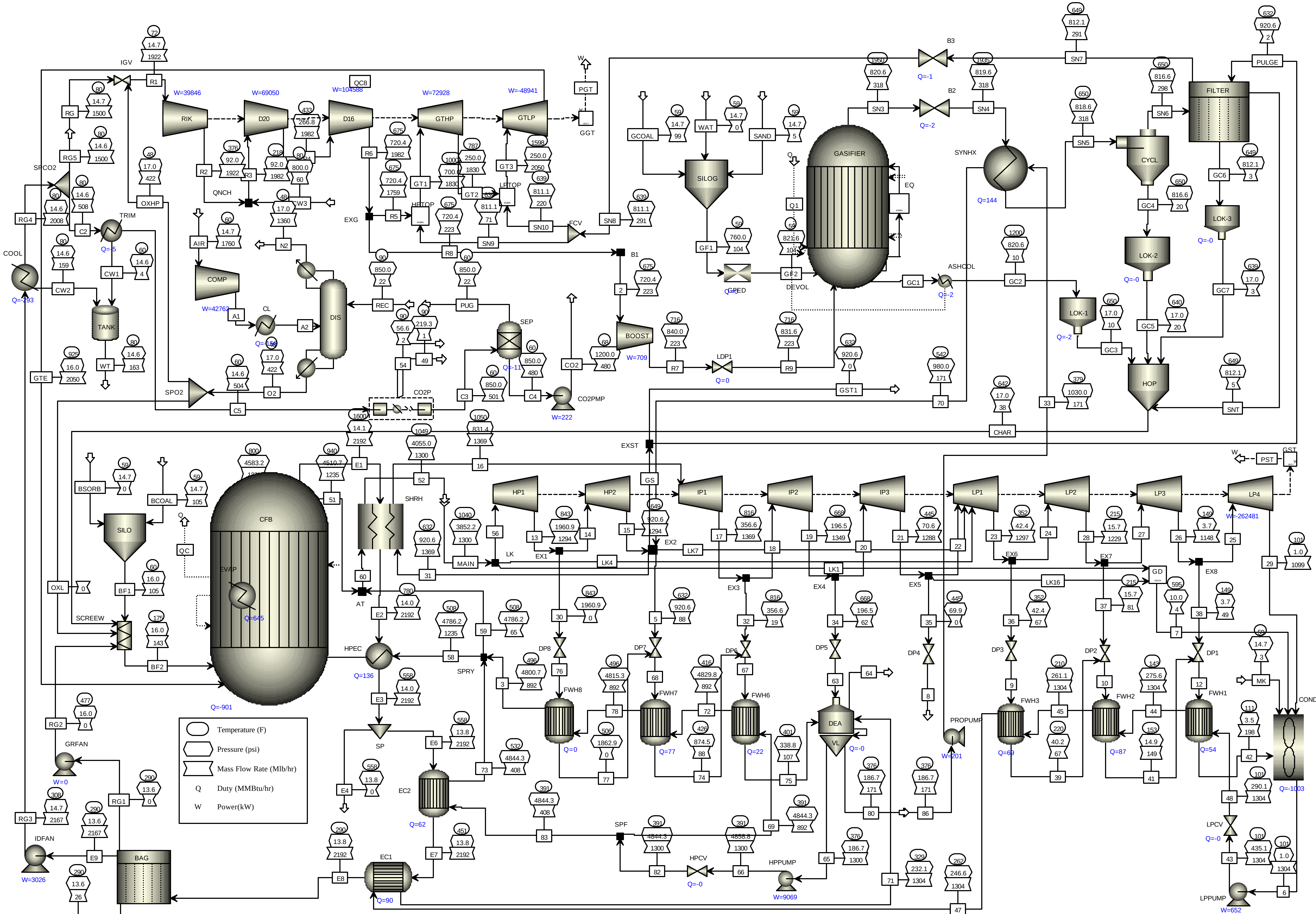

Figure A-1. CO_2 Hybrid Cycle Case 1

CASE 3

ASU	
O_2 Flow, klb/hr	497
Power Consumption, MW	55.5
O_2 Delivery Pressure, psia	15
Gasifier	
Coal Feed Rate, klb/hr	180
O_2 Feed Rate, klb/hr	89
Steam Feed Rate, klb/hr	16
Syngas Flowrate, klb/hr	582
Syngas LHV, btu/scf	156
Syngas Cooler Duty, mmbtu/hr	280
Char Flowrate, klb/hr	63
Gas Turbine	
Type	ICAD
Output, MW	108
Exhaust Flow, klb/hr	2030
Turbine Inlet Temp., F	2270
Turbine Exhaust Temp., F	1167
Steam Turbine	
Output, MW	284
Main Steam Flowrate, klb/hr	1300
Reheat Steam Flowrate, klb/hr	1521
CFB Boiler	
Coal Feed Rate, klb/hr	61
Flue Gas Flowrate, klb/hr	2379
CO_2 Compression and Dehyd.	
CO_2 Inlet Pressure, psia	756
CO_2 Flowrate, klb/hr	571
CO_2 Sequestered as	Vapor
Overall System Performance	
Net Efficiency, %	35.7
Net Power Output, MWe	295
Auxiliary Power Consumption, MW	97.1

This case evaluated the possibility of removing the CO_2 for sequestration at a pressurized state from the booster compressor discharge. This configuration eliminated gas compression costs but required a separate oxygen compressor so that the main compressor discharge could have concentrated CO_2 gas for sequestration.

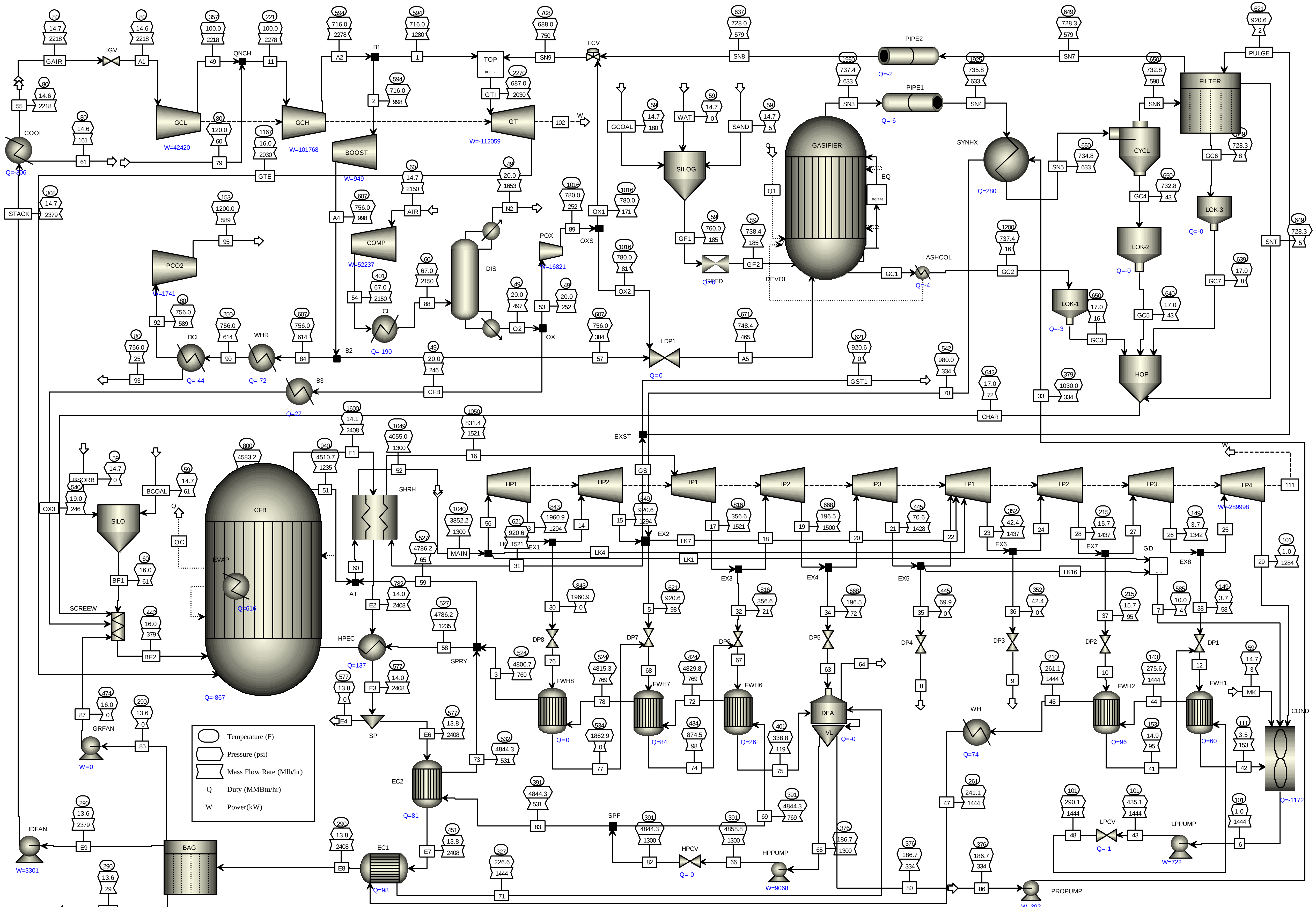

Figure A-2. CO_2 Hybrid Cycle Case 3

CASE 4

ASU	
O_2 Flow, klb/hr	506
Power Consumption, MW	88.1
O_2 Delivery Pressure, psia	850
Gasifier	
Coal Feed Rate, klb/hr	190
O_2 Feed Rate, klb/hr	91
Steam Feed Rate, klb/hr	15
Syngas Flowrate, klb/hr	577
Syngas LHV, btu/scf	159
Syngas Cooler Duty, mmbtu/hr	283
Char Flowrate, klb/hr	69
Gas Turbine	
Type	ICAD
Output, MW	107
Exhaust Flow, klb/hr	2091
Turbine Inlet Temp., F	2225
Turbine Exhaust Temp., F	1137
Steam Turbine	
Output, MW	281
Main Steam Flowrate, klb/hr	1300
Reheat Steam Flowrate, klb/hr	1524
CFB Boiler	
Coal Feed Rate, klb/hr	54
Flue Gas Flowrate, klb/hr	2443
CO_2 Compression and Dehyd.	
CO_2 Inlet Pressure, psia	838
CO_2 Flowrate, klb/hr	580
CO_2 Sequestered as	Liq.
Overall System Performance	
Net Efficiency, %	33.2
Net Power Output, MWe	277
Auxiliary Power Consumption, MW	111.4

This case evaluated the possibility of removing the CO_2 for sequestration at a pressurized state from the booster compressor discharge. This configuration eliminates gas compression costs but ordinarily requires a separate oxygen compressor so that the main compressor discharge can have concentrated CO_2 gas for sequestration.

To avoid the additional oxygen compressor this case investigated the possibility of getting pressurized oxygen from the ASU (pumped as cryogenic liquid and vaporized). Foster Wheeler's original calculation of ASU power consumption requirements showed favorable results for this case. The ASU power consumption values were revised using the values reported by Air Products in their quote for this configuration. The corrected values were much higher than originally anticipated and made this approach less than optimal.

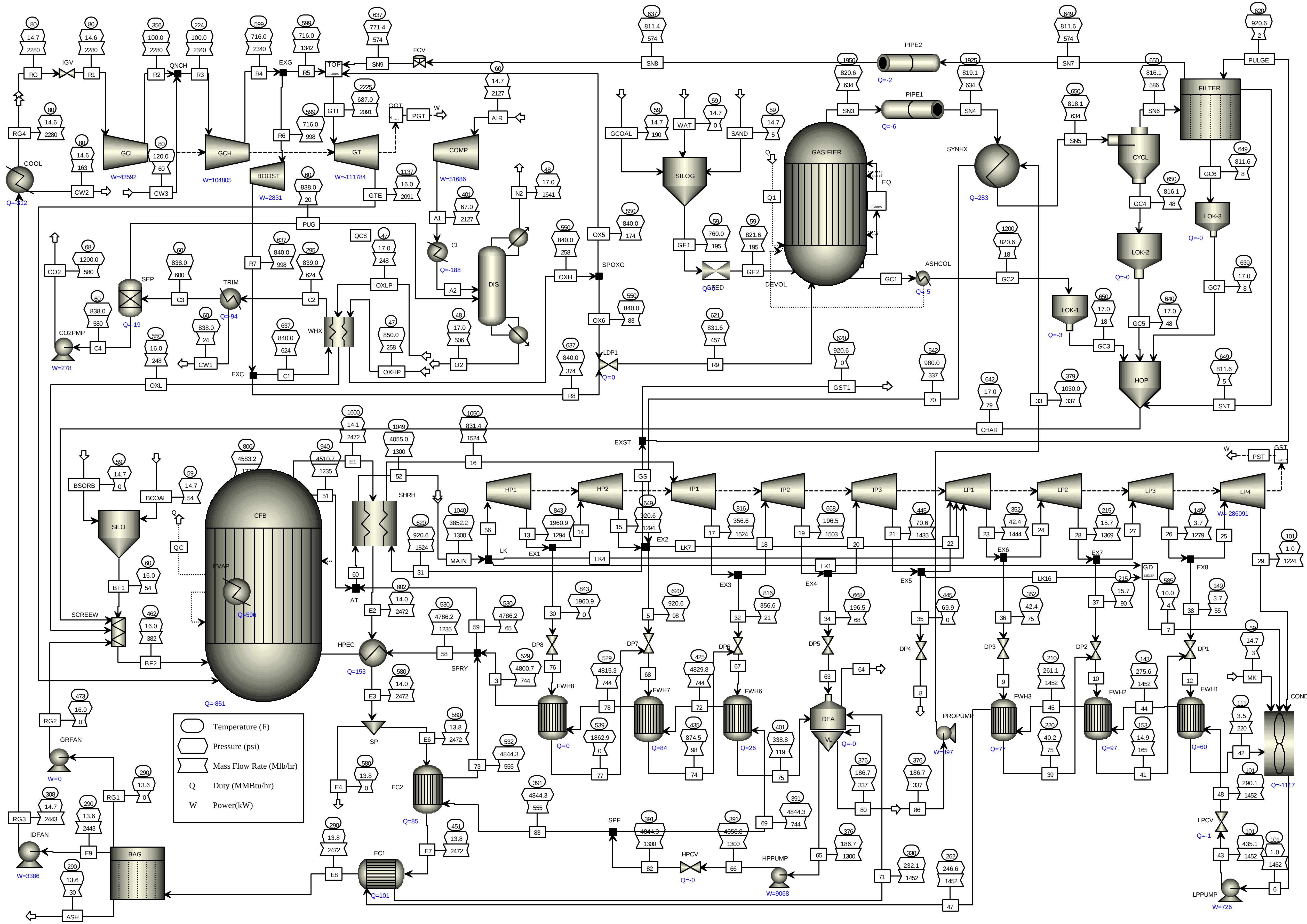

Figure A-3. CO_2 Hybrid Cycle Case 4

APPENDIX B

SUMMARY ECONOMIC DATA FOR ANALYSIS BASED ON YEAR 2004 DOLLARS

This Appendix contains the following data:

Hybrid Plant Without CO_2 Sequestration (reference plant)
- Total plant cost tabulation
- Estimate basis/financial criteria for revenue requirement calculations
- Capital investment and revenue requirement summary

CO_2 Hybrid Plant
- Total plant cost tabulation
- Estimate basis/financial criteria for revenue requirement calculations
- Capital investment and revenue requirement summary

A July 2001 cost estimate of the GFBCC (Hybrid plant without CO2 sequestration) was used for the estimate. Based on guidance from Foster Wheeler commercial estimating department, the 2001 costs were escalated as follows:

- Equipment costs were escalated at 5% a year to account for the rise in steel prices in recent years
- All labor costs, including engineering, construction and operation were escalated at 3.5% per year.

HYBRID PLANT WITHOUT CO_2 SEQUESTRATION (REFERENCE PLANT)

Table B-1. Total Plant Cost Summary for Hybrid Plant Without CO_2 Sequestration (Reference Plant)

GBBCC Combustion Hybrid Plant Cost Estimate
Cost Base 2004
Plant Net Power Output: **300 MW**

						Bare Erected	Plant Comiss. &	Contingency		Total	
Acct. No.	**Item/Description**	**Equipment**	**Engineering**	**Installation**	**Comissioning**	**Cost**	**EPC Mgt.**	**Process**	**Project**	**Cost**	**$/kW**
1 & 2	Coal and Sorbent Handling	$ 2,796,317.92		$ 1,892,308.13		***$ 4,688,626.05***	$ 281,317.56	$ -	$ 703,293.91	***$ 5,673,237.52***	***19***
2	Coal and Sorbent Prep and Feed	$ 8,256,322.08		$ 5,587,170.63		***$ 13,843,492.72***	$ 830,609.56	$ 692,174.64	$ 2,076,523.91	***$ 17,442,800.82***	***58***
3	Feedwater and Misc. BOP systems	$ 17,766,621.90	$ 14,387,346.34	$ 13,006,993.48		***$ 45,160,961.72***	$ 2,709,657.70	$ -	$ 6,774,144.26	***$ 54,644,763.69***	***182***
4	Gasifier and Accessories										
4.1 - 4.2	Gasifier and Auxiliaries	$ 32,830,093.35	$ 10,325,532.56		$ 3,550,483.10	$ 46,706,109.02	$ 2,802,366.54	$ 7,005,916.35	$ 7,005,916.35	$ 63,520,308.26	212
4.3	Air Separation Unit	$ -	w/equipment	w/equipment	w/equipment	$ -	$ -	$ -	$ -	$ -	-
4.4	CFB Boiler	$ 65,718,422.19	$ 9,771,430.37	$ 44,648,590.17	$ 2,961,749.12	$ 123,100,191.84	$ 7,386,011.51	$ -	$ 12,310,019.18	$ 142,796,222.54	476
4.4 - 4.9	Other Gasification Equipment	$ 12,169,838.43	$ -	$ 10,268,164.72	$ -	$ 22,438,003.15	$ 1,346,280.19	$ 3,365,700.47	$ 3,365,700.47	$ 30,515,684.28	102
	SUBTOTAL 4					***$ 192,244,304.00***				***$ 236,832,215.07***	***789***
5A	Gas Cleanup and Piping	$ -	$ -	$ -	$ -						
5B	CO2 Compression	$ -	$ -	$ -	w/equip	***$ -***	$ -	$ -	$ -	***$ -***	***-***
6	Combustion Turbine/Accessories										
6.1	Combustion Turbine Generator	$ 27,783,000.00	w/equip	w/equip	w/equip	$ 27,783,000.00	$ 1,666,980.00	$ 2,778,300.00	$ 4,167,450.00	$ 36,395,730.00	121
6.2 - 6.9	CT Accessories	$ 4,512,430.35		$ 7,391,484.18		$ 11,903,914.54	$ 714,234.87	$ -	$ 595,195.73	$ 13,213,345.13	44
	SUBTOTAL 6					***$ 39,686,914.54***				***$ 49,609,075.13***	***165***
7	HRSG, Ducting, and Stack										
7.1	Heat Recovery Steam Generator					$ -	$ -	$ -	$ -	$ -	-
7.2-7.9	CFB Accessories, Ductwork & Piping	$ 11,724,620.48	$ -	$ 8,354,018.44	$ -	$ 20,078,638.92	$ 1,204,718.34	$ -	$ 3,011,795.84	$ 24,295,153.10	81
	SUBTOTAL 7					***$ 20,078,638.92***				***$ 24,295,153.10***	***81***
8	Steam Turbine Generator										
8.1	STG & Accessories	$ 27,813,098.25		$ 1,966,080.53	$ 1,639,090.50	$ 31,418,269.28	$ 1,885,096.16	$ -	$ 4,712,740.39	$ 38,016,105.83	127
8.2 - 8.9	Turbine Plant Auxiliaries and Steam Piping	$ 358,654.22		$ 3,056,182.58		$ 3,414,836.80	$ 204,890.21	$ -	$ 512,225.52	$ 4,131,952.53	14
	SUBTOTAL 8					***$ 34,833,106.08***				***$ 42,148,058.36***	***140***
9	Cooling Water System	$ 3,457,806.20		$ 2,886,971.40		***$ 6,344,777.60***	$ 380,686.66	$ -	$ 951,716.64	***$ 7,677,180.89***	***26***
10	Ash/Spent Sorbent Handling System	$ 1,575,825.13		$ 1,503,913.07		***$ 3,079,738.19***	$ 184,784.29	$ 153,986.91	$ 461,960.73	***$ 3,880,470.12***	***13***
11	Accessory Electric Plant	$ 18,877,426.76		$ 5,307,969.37		***$ 24,185,396.13***	$ 1,451,123.77	$ 1,209,269.81	$ 3,627,809.42	***$ 30,473,599.13***	***102***
12	Instrumentation and Control	$ 2,097,045.79		$ 583,188.40		***$ 2,680,234.19***	$ 160,814.05	$ 134,011.71	$ 402,035.13	***$ 3,377,095.08***	***11***
13	Improvements to Site	$ 3,403,533.26		$ 13,543,775.95		***$ 16,947,309.22***	$ 1,016,838.55	$ -	$ 2,542,096.38	***$ 20,506,244.15***	***68***
14	Buildings and Structures	$ 5,728,033.84		$ 5,074,413.73		***$ 10,802,447.57***	$ 648,146.85	$ -	$ 1,620,367.14	***$ 13,070,961.56***	***44***
		$ 246,869,090.16	***$ 34,484,309.26***	***$ 125,071,224.80***	***$ 8,151,322.72***	***$ 414,575,946.94***	***$ 24,874,556.82***	***$ 15,339,359.89***	***$ 54,840,990.99***	***$ 509,630,854.63***	***1,699***

Table B-2. Estimate Basis/Financial Criteria for Revenue Requirement Calculations: Reference Plant

GENERAL DATA/CHARACTERISTICS

Plant Type:	Combustion Hybrid Without CO2 Sequestration (GFBCC Cycle)			
Plant Size:	300	MW, net		
Location:	Sea Level, Middletown USA			
Fuel: Primary/Secondary	Illinois #6		--	
Energy from Primary/Secondary Fuels:	8,124	Btu/kWh	--	Btu/kWh
Levelized Capacity Factor / Preproduction (equivalent months):	65%		1	months
Capital Cost Year Dollars (Reference Year Dollars):	2004 (July)			
Delivered Cost of Primary/Secondary Fuel:	1.14	\$/MBtu	--	\$/MBtu
Design/ Construction Period:	4	years		
Plant Start-up Date (1st year Dollars):	2008 (July)			
Land Area/Unit Cost:	100	acres	\$10,000	/ Acre

FINANCIAL CRITERIA

Project Book Life:	20	years		
Book Salvage Value:	-	%		
Project Tax Life:	20	years		
Tax Depreciation Method:	Accel. Based on ACRS Class			
Property Tax Rate:	1.0	%		
Insurance Tax Rate:	1.0	%		
Federal Income Tax Rate:	34.0	%		
State Income Tax Rate:	4.2	%		
Investment Tax Credit/% Eligible	-	%	-	%
Economic Basis:	Over Book	Constant Dollars		

Capital Structure	% of Total	Cost(%)
Common Equity	40	12
Preferred Stock	0	8.5
Debt	60	7
Weighted Cost of Capital: (after tax)	7.456%	

Table B-3. Capital Investment and Revenue Requirement Summary: Reference Plant

CAPITAL INVESTMENT AND REVENUE REQUIREMENT SUMMARY

Case:	CO2 Hybrid Plant		
Plant size:	300 MW net	**Net Cycle Efficiency:**	42.0%
Primary/Secondary Fuel (type):	Illinois #6	**Net Plant Heat Rate (NPHR):**	8,124 (Btu/kWh)
Design/Construction:	4 (years)	**Cost:**	1.24 ($/MMBtu)
TPC (Plant Cost) Year:	1999 (dec.)	**Book Life:**	20 (years)
Capacity Factor:	65 (%)	**TPI Year:**	2000 (Jan)
		CO2 Removed:	- (tons/year)

CAPITAL INVESTMENT		**$x1000**		**$/kW**
Process Capital and Facilities		414,576		1,381.9
Engineering (Incl. C.M., H.O., and Fee)		24,875		82.9
Process Contingency		15,339		51.1
Project Contingency		54,841		182.8
TOTAL PLANT COST (TPC)		$ 509,631		1,698.8
TOTAL CASH EXPENDED	$ 509,631			
AFDC	$ 50,963			
TOTAL PLANT INVESTMENT (TPI)		$ 560,594		
Royalty Allowance				
Preproduction Cost		$13,501		
Inventory Capital		$2,322		
Initial Chemicals and Catalysts (w/equip)				
Land Cost		$1,000		
TOTAL CAPITAL REQUIREMENT (TCR)		$ 577,417		
OPERATING & MAINTENANCE COSTS (1999 DOLLARS)		**$x1000**		**$/kW-yr**
Operating Labor		5,520		18.4
Maintenance Labor		4,077		13.6
Maintenance Materials		6,116		20.4
Administrative Support and Labor		2,399		8.0
TOTAL OPERATION & MAINTENANCE		$ 18,112		60.4
FIXED O & M			39.99 $/kW-yr	
VARIABLE O & M			0.36 c/kWh	
CONSUMABLE OPERATING COSTS less Fuel (1999 DOLLARS)		**$x1000**		**c/kWh**
Water		459		0.02
Chemicals		2757		0.10
Other Consumables		241		0.01
Waste Disposal		2278		0.09
TOTAL CONSUMABLE OPERATING COST		$ 5,735		0.22
FUEL COST (1999 Dollars)		$ 17,208		1.01
		Levelized (Over Book Life $)		
PRODUCTION COST SUMMARY		**$/tonne CO2**		**c/kWh**
Fixed O&M				0.70
Variable O&M				0.36
Consumables				0.22
By-product Credit				0
Fuel				1.01
TOTAL PRODUCTION COST				2.29
LEVELIZED CARRYING CHARGES (Capital)				3.92
LEVELIZED (Over Book Life) BUSBAR COST OF POWER				6.20

CO_2 HYBRID PLANT YEAR 2004 ESTIMATE

Table B-4. CO_2 Hybrid Plant Total Plant Cost Summary

CO_2 Hybrid Plant Cost Summary
Cost Base 2004
Plant Net Power Output: **294.9 MW**

						Bare Erected	Plant Comiss. &	Contingency		Total	
Acct. No.	**Item/Description**	**Equipment**	**Engineering**	**Installation**	**Comissioning**	**Cost**	**EPC Mgt.**	**Process**	**Project**	**Cost**	**$/kW**
1	Coal and Sorbent Handling	$ 2,095,400.77		$ 1,536,773.22		***$ 3,632,173.99***	$ 217,930.44	$ -	$ 544,826.10	***$ 4,394,930.53***	***15***
2	Coal and Sorbent Prep and Feed	$ 6,186,815.72		$ 4,537,429.24		***$ 10,724,244.95***	$ 643,454.70	$ 536,212.25	$ 1,608,636.74	***$ 13,512,548.64***	
3	Feedwater and Misc. BOP systems	$ 17,766,621.90	$ 14,597,889.55	$ 13,197,336.73		***$ 45,561,848.19***	$ 2,733,710.89	$ -	$ 6,834,277.23	***$ 55,129,836.31***	***187***
4	Gasifier and Accessories										
4.1 - 4.2	Gasifier and Auxiliaries	$ 40,621,310.00	$ 10,476,632.00		$ 3,602,441.00	$ 54,700,383.00	$ 3,282,022.98	$ 8,205,057.45	$ 8,205,057.45	$ 74,392,520.88	252
4.3	Air Separation Unit	$ 94,630,000.00	w/equipment	w/equipment	w/equipment	$ 94,630,000.00	$ 5,677,800.00	$ -	$ 4,731,500.00	$ 105,039,300.00	356
4.4	CFB Boiler	$ 58,537,617.37	$ 9,914,425.00	$ 45,281,702.64	$ 3,005,091.66	$ 116,738,836.68	$ 7,004,330.20	$ -	$ 11,673,883.67	$ 135,417,050.55	459
4.4 - 4.9	Other Gasification Equipment	$ 12,123,619.62	$ -	$ 10,418,428.18	$ -	$ 22,542,047.80	$ 1,352,522.87	$ 3,381,307.17	$ 3,381,307.17	$ 30,657,185.01	104
	SUBTOTAL 4					***$ 288,611,267.48***				***$ 345,506,056.44***	***1,172***
5A	Gas Cleanup and Piping	$ -	$ -	$ -	$ -						
5B	CO2 Compression	$ 20,422,435.54	$ 6,200,000.00	$ 6,200,000.00	w/equip	***$ 32,822,435.54***	$ 1,969,346.13	$ -	$ 4,923,365.33	***$ 39,715,147.00***	***135***
6	Combustion Turbine/Accessories										
6.1	Combustion Turbine Generator	$ 30,960,000.00	w/equip	w/equip	w/equip	$ 30,960,000.00	$ 1,857,600.00	$ 3,096,000.00	$ 15,480,000.00	$ 51,393,600.00	174
6.2 - 6.9	CT Accessories	$ 4,507,856.58		$ 7,495,708.05		$ 12,003,564.63	$ 720,213.88	$ -	$ 600,178.23	$ 13,323,956.74	45
	SUBTOTAL 6					***$ 42,963,564.63***				***$ 64,717,556.74***	***219***
7	HRSG, Ducting, and Stack										
7.1	Heat Recovery Steam Generator					$ -	$ -	$ -	$ -	$ -	-
7.2-7.9	CFB Accessories, Ductwork & Piping	$ 11,724,620.48	$ -	$ 8,476,270.45	$ -	$ 20,200,890.93	$ 1,212,053.46	$ -	$ 3,030,133.64	$ 24,443,078.02	83
	SUBTOTAL 7					***$ 20,200,890.93***				***$ 24,443,078.02***	***83***
8	Steam Turbine Generator										
8.1	STG & Accessories	$ 31,417,756.20		$ 1,994,851.99	$ 1,663,076.81	$ 35,075,685.00	$ 2,104,541.10	$ -	$ 5,261,352.75	$ 42,441,578.85	144
8.2 - 8.9	Turbine Plant Auxiliaries and Steam Piping	$ 394,519.64		$ 3,100,906.50		$ 3,495,426.14	$ 209,725.57	$ -	$ 524,313.92	$ 4,229,465.63	14
	SUBTOTAL 8					***$ 38,571,111.14***				***$ 46,671,044.48***	***158***
9	Cooling Water System	$ 3,457,806.20		$ 2,929,219.10		***$ 6,387,025.29***	$ 383,221.52	$ -	$ 958,053.79	***$ 7,728,300.61***	***26***
10	Ash/Spent Sorbent Handling System	$ 1,794,106.90		$ 1,525,921.20		***$ 3,320,028.11***	$ 199,201.69	$ 166,001.41	$ 498,004.22	***$ 4,183,235.42***	***14***
11	Accessory Electric Plant	$ 18,877,426.76		$ 5,385,645.75		***$ 24,263,072.51***	$ 1,455,784.35	$ 1,213,153.63	$ 3,639,460.88	***$ 30,571,471.36***	***104***
12	Instrumentation and Control	$ 2,097,045.79		$ 591,722.73		***$ 2,688,768.52***	$ 161,326.11	$ 134,438.43	$ 403,315.28	***$ 3,387,848.34***	***11***
13	Improvements to Site	$ 3,403,533.26		$ 14,776,408.21		***$ 18,179,941.47***	$ 1,090,796.49	$ -	$ 2,726,991.22	***$ 21,997,729.18***	***75***
14	Buildings and Structures	$ 5,728,033.84		$ 5,148,672.27		***$ 10,876,706.12***	$ 652,602.37	$ -	$ 1,631,505.92	***$ 13,160,814.40***	***45***
		$ 366,746,526.58				***$ 548,803,078.87***	***$ 32,928,184.73***	***$ 16,732,170.32***	***$ 76,656,163.53***	***$ 675,119,597.47***	***2,289***

Table B-5. Estimate Basis/Financial Criteria for Revenue Requirement Calculations: CO_2 Hybrid Plant

GENERAL DATA/CHARACTERISTICS

Plant Type:	Combustion Hybrid With CO2 Sequestration (Advanced CO2 Cycle)			
Plant Size:	292.9	MW, net		
Location:	Sea Level, Middletown USA			
Fuel: Primary/Secondary	Illinois #6		--	
Energy from Primary/Secondary Fuels:	9,452	Btu/kWh	--	Btu/kWh
Levelized Capacity Factor / Preproduction (equivalent months):	65%		1	months
Capital Cost Year Dollars (Reference Year Dollars):	2004 (July)			
Delivered Cost of Primary/Secondary Fuel:	1.14	$/MBtu	--	$/MBtu
Design/ Construction Period:	4	years		
Plant Start-up Date (1st year Dollars):	2008 (July)			
Land Area/Unit Cost:	100	acres	$10,000	/ Acre

FINANCIAL CRITERIA

Project Book Life:	20	years		
Book Salvage Value:	-	%		
Project Tax Life:	20	years		
Tax Depreciation Method:	Accel. Based on ACRS Class			
Property Tax Rate:	1.0	%		
Insurance Tax Rate:	1.0	%		
Federal Income Tax Rate:	34.0	%		
State Income Tax Rate:	4.2	%		
Investment Tax Credit/% Eligible	-	%	-	%
Economic Basis:	Over Book	Constant Dollars		

Capital Structure	% of Total	Cost(%)
Common Equity	40	12
Preferred Stock	0	8.5
Debt	60	7
Weighted Cost of Capital: (after tax)	7.456%	

Table B-6. Capital Investment and Revenue Requirement Summary: CO_2 Hybrid Plant

CAPITAL INVESTMENT AND REVENUE REQUIREMENT SUMMARY

Case:	CO2 Hybrid Plant		
Plant size:	294.9 MW net	**Net Cycle Efficiency:**	36.10%
Primary/Secondary Fuel (type):	Illinois #6	**Net Plant Heat Rate (NPHR):**	9,452 (Btu/kWh)
Design/Construction:	4 (years)	**Cost:**	1.26 ($/MMBtu)
TPC (Plant Cost) Year:	2004 (July)	**Book Life:**	20 (years)
Capacity Factor:	65 (%)	**TPI Year:**	2004 (July)
		CO2 Removed:	1,517,096 (tons/year)

CAPITAL INVESTMENT			**$x1000**		**$/kW**
Process Capital and Facilities			548,803		1,861.0
Engineering (Incl. C.M., H.O., and Fee)			32,928		111.7
Process Contingency			16,732		56.7
Project Contingency			76,656		259.9
TOTAL PLANT COST (TPC)		$	675,120		2,289.3
TOTAL CASH EXPENDED	$ 675,120				
AFDC	$ 67,512				
TOTAL PLANT INVESTMENT (TPI)		$	742,632		
Royalty Allowance					
Preproduction Cost			$17,503		
Inventory Capital			$3,098		
Initial Chemicals and Catalysts (w/equip)					
Land Cost			$1,000		
TOTAL CAPITAL REQUIREMENT (TCR)		$	764,232		

OPERATING & MAINTENANCE COSTS (1999 DOLLARS)		**$x1000**		**$/kW-yr**
Operating Labor		6,624		22.5
Maintenance Labor		5,401		18.3
Maintenance Materials		8,101		27.5
Administrative Support and Labor		3,006		10.2
TOTAL OPERATION & MAINTENANCE	$	23,133		78.4
FIXED O & M			50.97 $/kW-yr	
VARIABLE O & M			0.48 c/kWh	

CONSUMABLE OPERATING COSTS less Fuel (1999 DOLLARS)		**$x1000**	**c/kWh**
Water		508	0.02
Chemicals		391	0.02
Other Consumables		240	0.01
Waste Disposal		467	0.02
TOTAL CONSUMABLE OPERATING COST	$	1,606	0.06
FUEL COST (1999 Dollars)	$	19,997	1.19

	Levelized (Over Book Life $)	
PRODUCTION COST SUMMARY	**$/tonne CO2**	**c/kWh**
Fixed O&M	2.49	0.90
Variable O&M	1.60	0.48
Consumables	(2.01)	0.06
By-product Credit	-	0
Fuel	2.37	1.19
TOTAL PRODUCTION COST	4.44	2.63
LEVELIZED CARRYING CHARGES (Capital)	17.48	5.27
LEVELIZED (Over Book Life) BUSBAR COST OF POWER	21.93	7.90