Quarterly Technical Progress Report

Third Report
January 16, 2000 – April 15, 2000

Title: *A Novel CO_2 Separation System*

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Contract DE-AC26-98FT40421--03

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PROJECT DESCRIPTION

Introduction

Because of concern over global climate change, new systems are needed that produce electricity from fossil fuels and emit less CO_2. The fundamental problem with current systems that recover and concentrate CO_2 from flue gases is the need to separate dilute CO_2 and pressurize it to roughly 35 atm for storage or sequestration. This is an energy intensive process that can reduce plant efficiency by 9-37% and double the cost of electricity.

There are two fundamental reasons for the current high costs of power consumption, CO_2 removal, and concentration systems: 1) most disposal, storage and sequestering systems require high pressure CO_2 (at roughly 35 atm). Thus, assuming 90% removal of the CO_2 from a typical atmospheric pressure flue gas that contains 10% CO_2, the CO_2 is essentially being compressed from 0.01 atm to 35 atm (a pressure ratio of 3,500). This is a very energy intensive process. 2) The absorption-based (amine) separation processes that are used to remove the CO_2 from the flue gas and compress it to 1 atm consume approximately 10 times more energy than the theoretical work of compression because they are heat driven cycles working over a very low temperature difference. Thus, to avoid the problems of current systems, we need a power cycle in which the CO_2 produced by the oxidation of the fuel is not diluted with a large excess of nitrogen, a power cycle which would allow us to eliminate the very inefficient thermally driven absorption/desorption step. In addition, we would want the CO_2 to be naturally available at high pressure (approximately 3 to 6 atmospheres), which would allow us to greatly reduce the compression ratio between generation and storage (from roughly 3,500 to approximately 8).

To meet this objective, we propose a power generation system in which a fossil fuel (gasified coal, petroleum fuels or natural gas) at pressure is used to reduce a metal oxide, producing metal (or a lower valance metal oxide), CO_2 and water (Figure 1). The water is condensed and its energy used to raise steam, leaving behind a stream of pure CO_2 at 3 to 6 atmospheres that can be readily stored or sequestered. The metal oxide is then “burned” or re-oxidized in air from the compressor section of a gas turbine, producing a hot high-pressure stream of air that is used to drive a gas turbine. The exhaust can in turn be used to drive a steam bottoming cycle. The oxidized particles are then recycled to the first reactor to be reduced again and repeat the cycle. The system that transfers the energy of the fuel to the air without bringing the carbon along is called a Sorbent Energy Transfer System (SETS). The cycle can be run on any fuel gas (gasified coal, oil or natural gas) and does not require the development of new hardware. The power generation cycle is essentially a standard combined cycle, except that the combustor is replaced by two fluidized beds (the SETS): one that uses fuel to reduce the particulate oxide and one that oxidizes it and heats the air entering the turbine. The system is described below using methane as the fuel (because it is simplest to explain).

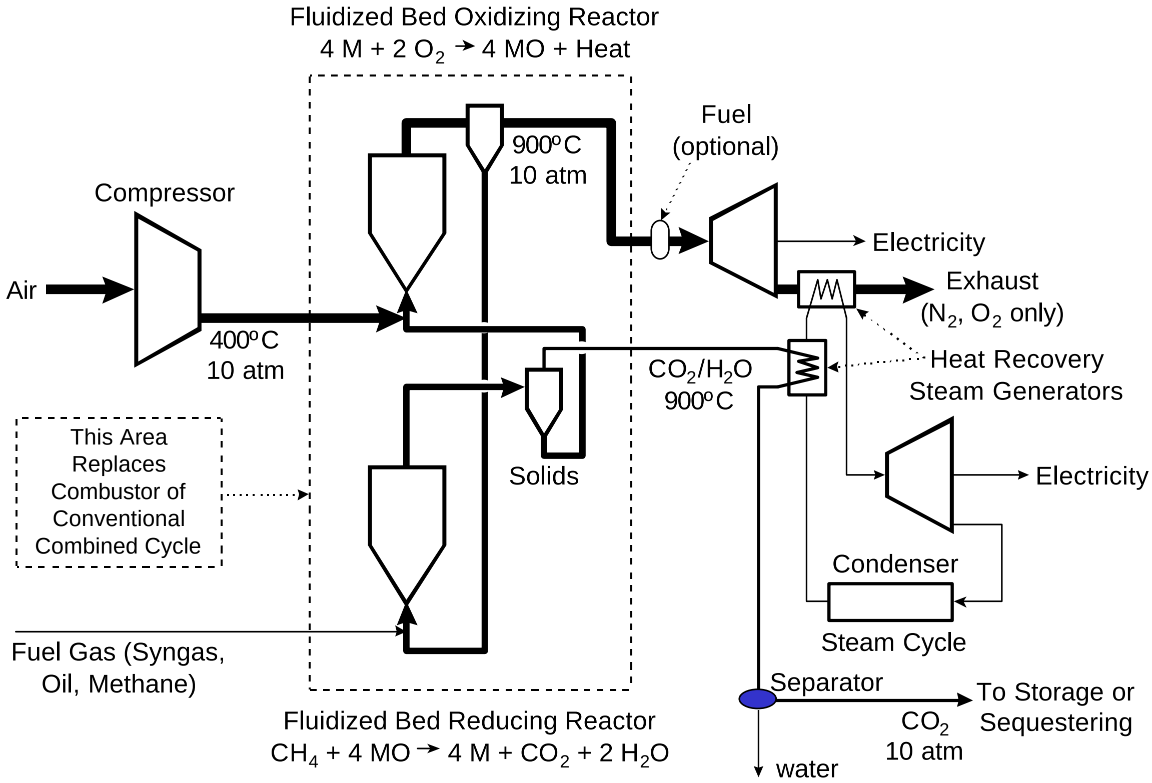

Figure 1. Sorbent energy transfer cycle schematic (M = metal).

The first step in the SETS process is to reduce a metal oxide to a metal (or a metal oxide to a lower valance metal oxide). In general, the metal (oxygen sorbent) would be supported on, or contained within, an inert support (such as alumina) that would provide a high surface area for reaction and good physical properties such as crush strength and attrition resistance. For example, if copper were the metal oxide the formula would be:

$$CH_4 + 4\ CuO \rightarrow CO_2 + 2\ H_2O + 4\ Cu$$

In this way, we convert the energy in the fuel to energy that is stored in the reduced metal, and produce a stream consisting of 33% CO_2 and water. We carry this out at pressure (10 atmospheres, for example) in a fluidized bed reactor, and remove the steam from the CO_2 by condensing it (producing valuable mid-pressure steam which can be used to generate electricity in the steam turbine of a combined cycle) and a stream of virtually 100% pure CO_2 at 10 atm. The CO_2 is sent to a storage or sequestering process with little additional compression energy.
Virtually all of the chemical energy in the original fuel gas is now incorporated in a new fuel (small particles of copper on an inert support). The reduced copper particles enter a second fluidized bed (also run at 10 atm) and are re-oxidized with air, producing large amounts of heat and heating the air to the temperatures needed to drive a gas turbine-combined cycle (900ºC or greater).

$$4\ Cu + 2\ O_2 \rightarrow 4\ CuO$$

The air entering the oxidizing bed is the hot (400°C) high-pressure (10 atm) air leaving the compressor stages of a standard gas turbine. The only CO_2 emissions from such a system would be produced if we burned some additional fuel to further boost the temperature of the high-pressure air just before it entered the expander stages of the turbine. The copper/alumina particles would then be returned to the reducing reactor and the cycle repeated.

This power cycle requires no new hardware; it is essentially a standard combined cycle in which the combustor is replaced by the fluidized bed oxidation and reduction reactors. The system can dramatically reduce or eliminate CO_2 emissions. The degree of CO_2 removal is limited only by the performance of the filters used to protect the turbine. With current 900°C filters the cycle would reduce the CO_2 emissions of a coal gasifier combined cycle by 83% while suffering only a 1.5 - 4% efficiency penalty. With the advanced filters currently under development, CO_2 emissions could be totally eliminated with no additional efficiency penalty. Since the power cycle uses only standard combined cycle generating equipment and two circulating fluidized bed reactors, the major research needs are further design, sorbent development, pilot scale testing and detailed engineering and cost analyses.

Geodes for a Long Life Sorbent

To make a sorbent pellet that can hold large amounts of sorbent without being destroyed by the absorption-regeneration process, TDA has developed a new sorbent structure, which we call a geode (Figure 2). Like the geode that you buy at a gift shop, our geode has a hollow shell. The sorbent is loosely contained in, but does not fill, the hole(s) in the center. Thus, the sorbent can expand and contract indefinitely without destroying the pellet structure that surrounds it. Unlike the gem shop geode that has a single hole in the middle, our geode sorbent contains hundreds or thousands of holes in a structure that looks like a conventional catalyst support pellet on the outside. The interior structure of the geode is like a sponge, a sponge that is partially filled with the chemically active material, copper and/or iron oxides.

To make the very small (e.g., 100 µ) particles required by fluidized bed and transport reactors, TDA has modified our previous methods of making the geode. We mix the active component of the sorbent with water, a low cost inert material, and other additives. The mixture is formed by spray drying or extruding and then firing. After firing, the water and some additives evaporate and/or burn, leaving behind a complex porous structure.

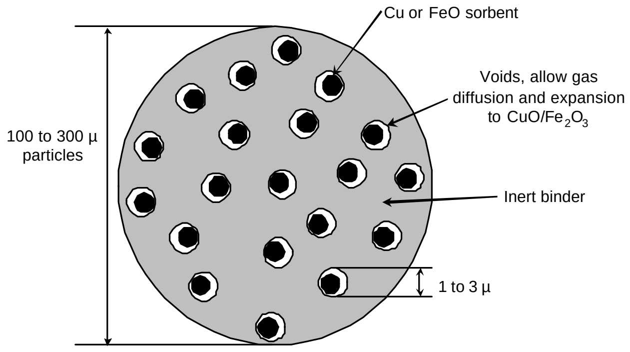

Figure 2. Structure of TDA's "geode."

Due to the nature of this mixture, it separates upon firing into two different phases which are each thermodynamically stable. Data from both published phase diagrams and our independent measurements show that the phases are physically separate. The size of each chemically active region of the geode (i.e., 1-3 µm) is controlled by the selection of the starting materials.

In the geode structure there are voids in both the mixed chemically active and inert phases. The binder is a continuous porous structure filled with extremely small pores (0.1 to 1 µm). This structure holds the sorbent in place. Because of the porosity left behind as the water evaporates, the O_2, H_2, and CO can easily diffuse in from the exterior of the pellet to the tiny pockets of sorbent. The geode structure: 1) is very strong because there is a continuous support phase, 2) effectively contains the sorbent inside small holes in the interior of the pellet, 3) allows the sorbent to expand and contract freely without disrupting the pellet structure, 4) allows the gases to diffuse quickly into the interior of the pellet, and 5) can hold large quantities of sorbent and absorb large quantities of oxygen. In addition, the process that we use to produce the geode is inherently low in cost. The geode is formed by mixed metal oxide techniques, yet it has the continuous inert structure associated with catalyst supports, which have high strength and long life.

TECHNICAL OBJECTIVES

The Phase I research will carry out preliminary proof of concept experiments to determine whether the geode structure can be used to make a low-cost, long life, redox metal oxide based sorbent for use in the SETS cycle. The specific goal of the Phase I work is to optimize the formulation of the geode in order to obtain the following properties:

- Less than 10 lbs of oxygen per 100 lbs of fresh sorbent,
- An 80 to 300 micron particle size,
- Reduction @ 700°C (1292°F) with oxidation initiated with 400°C (752°F) air, and
- Costs of less than $8/lb when produced at 50,000 lbs/year.

Technical Approach

In order to bring our new power cycle from a concept to a commercial reality we need a team with a variety of skills. Specifically, in this three phase program we need to:

1) Carry out the conceptual and thermodynamic analyses necessary to better define the cycle and optimize the reaction conditions,
2) Develop a sorbent with suitable physical and chemical properties,
3) Test the sorbent at the laboratory and pilot scale,
4) Engineer the oxidation and reduction reactors,
5) Prepare detailed engineering designs to predict the total cycle performance and cost.

To provide the necessary skills, we have assembled a team that includes TDA Research (TDA), M.W. Kellogg (Kellogg) and Louisiana State University (LSU) (Figure 3). TDA invented the new system design. TDA will manage the project, carry out the conceptual engineering and thermodynamic analysis to define the process, and develop the sorbent. Kellogg will review the initial conceptual designs, carry out detailed engineering analyses, and test the sorbent in their circulating fluidized bed pilot facility. LSU will aid in the thermodynamic analyses. Kellogg, one of the largest U.S. engineering and construction contractors (with particular experience in fluidized bed design and construction) has the ability to provide complete commercial scale facilities.

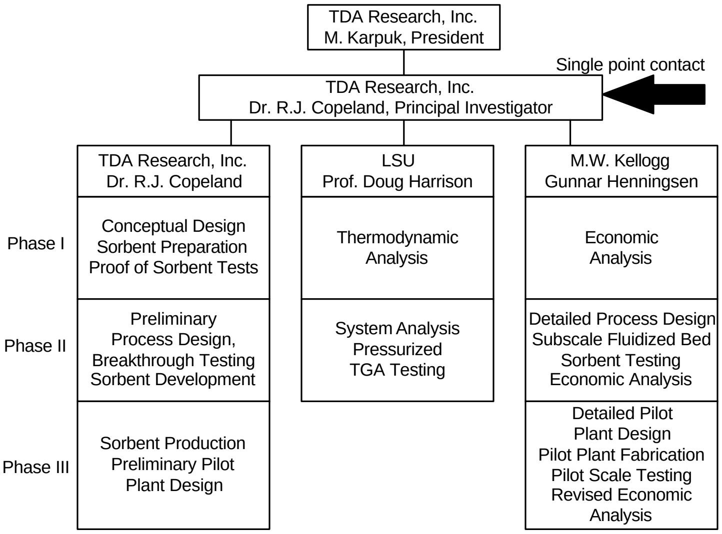

Figure 3. Project organization.

SUMMARY

We continued to evaluate sorbents for the Sorbent Energy Transfer System (SETS). Two types of tests were performed: one using an attrition tester, and one using a microbalance, where multiple reducing and oxidizing cycles were conducted and the weight of the samples continuously monitored. A fluidized bed test apparatus was set up and the instruments calibrated.

TDA is considering modifying SETS to increase the CO_2 capture to 100%; two approaches are being considered.

Louisiana State University (LSU) prepared a preliminary Aspen model of the SETS in order to evaluate its performance (see attached report). The estimated efficiency was relatively high as compared to the Phase I results. LSU received data from General Electric (GE) on the performance of a nominal combined cycle and will revise their model to match the performance of known gas turbines and combined cycles (without SETS) before incorporating the modifications in the SETS power cycle.

Results of the new SETS combined cycle base case simulation using actual flow rates are attached in Appendix A. Our total natural gas (CH_4) feed rate is 3.15e04 kg/hr (6.95e04 lb_m/hr) compared to the GE specification of 7.02e04 lb_m/hr. The air rate is 1.52e06 kg/hr (3.35e06 lb_m/hr), the same as GE's specification. The primary reason for the difference in fuel feed rates is the difference in heating values for CH_4 and natural gas. The overall efficiency is now 53.0% (LHV) with a net power output of 232 MWe, compared to a cycle efficiency of 55.0% (LHV) and power output of 241MWe reported by GE for their natural gas fired combined cycle. The current SETS cycle provides for 49.2% carbon capture.

Sorbent tests results

Three large batches of oxide-based sorbents were prepared using the spray-dry technique, and then fired at a variety of elevated temperatures. Unlike the process for previously synthesized sorbents, an alumo-hydrogel (boehmite dispersed in an aqueous HNO_3 solution) was not used in the sorbents' preparation. An Air Jet attrition test was done on the fabricated materials to measure their strength. The results of the attrition tests, and the chemical compositions and calcination temperatures for the fabricated sorbents, are summarized in Table 1.

The tests were done for a period of 2.5 hours. Fines generated in the first hour were included in the total loss calculations. According to the ASTM method for Air Jet attrition, a five-hour test should be performed, and the first hour results discarded, in order to normalize the particle size difference between various formulations. However, all of the material in our tests are sieved to the same mesh size prior to loading, so the results of the first hour as well as the results of the abbreviated test are highly representative of the full test results. The generally accepted attrition index (A.I.) value for use in a transport reactor is 5 wt% loss/h for the ASTM test protocol; we have found that this relates to a value somewhere around 12-13 wt% loss/h on our instrument.

Table 1. Results of the air jet attrition test for synthesized sorbents.

Sorbent Composition	Firing temperature, ^oC	Attrition Loss, wt% loss/h	Bulk Density, g/cc
30% CuO / Al_2O_3	950	24.17	1.40
	1050	18.55	1.22
	1100	17.78	1.28
	1150	13.05	1.33
	1200	7.61	1.53
5% CuO/45% Fe_2O_3/ Al_2O_3	950	19.89	1.63
	1050	8.82	1.74
	1200	1.30	2.03
50% Fe_2O_3 / Al_2O_3	950	15.98	1.59
	1050	8.81	1.77
	1100	5.45	1.91

The values presented in Table 1 should be considered representative of relative strengths rather then final values as, upon examination with a low power microscope, the spheroids formed from the test spray drying were dimpled (see Figure 4). This dimpling could be eliminated with some optimization of the processing step, which should also improve the strength of the resulting particulate.

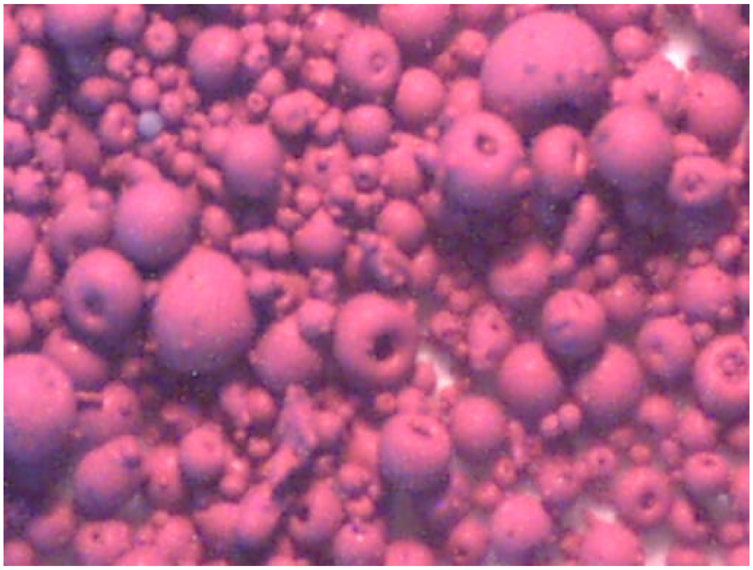
Figure 4. Example of dimpling on the test batches of spray-dried materials.

TDA estimated the life of these sorbents based on a comparison between these initial measurements and the measurements of another cycled sorbent. Under a DOE SBIR contract, TDA subcontracted to Norton Chemical Process Products (Norton) the fabrication of a zinc ferrite sorbent by techniques that were similar to large-scale production. Their sample number was Lot N9925548. Kellogg, Brown, & Root (KBR) tested that sorbent and measured the material against the Davison Index. TDA also measured the attrition index in our attrition tester. The TDA results are reported in Table 2. The zinc ferrite sorbent also had grape-like clusters and, therefore, a high attrition rate when fresh (our recently spray-dried sorbent has a similar problem). After cycling in KBR's TRTU, the attrition index improved (due to the removal of the grape-like clusters). KBR estimated that the long-term attrition rate of the N9925548 sorbent was only $0.55(10)^{-6}$ lb/lb-circulated. TDA measured an attrition index of 13 for the new Cu sorbent using the same apparatus that measured Lot N9925548. Although the new Cu sorbent had a better attrition index as fresh material, the loss mechanisms could be different; we conservatively estimated that the loss rate of this material would be $\sim 1(10)^{-6}$ lb/lb-circulated.

Table 2. Extrapolation of attrition loss rate.

Sorbent Identification	Lot N9925548	*New Cu Sorbent*
Bulk Density of fresh sorbent	82.4 lb/ft^3 (1.32 gm/cc)	*1.33 g/cc*
D_{50} after 5 cycles in MC test	115 µm	
D_{50} after 10 cycles in MC test	100 µm	
TDA attrition index for fresh[1]	17	*13*
TDA attrition index after 5 cycles in TRTU test[1]	11	
TDA attrition index after 10 cycles in TRTU test[1]	9	

[1] Since sorbent has grape-like clusters, which rapidly attrit, the 10th cycle attrition index is more representative of the sorbent long-term attrition behavior. Similar effects are seen in our new Cu sorbent

Long term attrition rate, Lot N9925548 lb/lb circulated	0.55×10^{-6}
Long term attrition rate, *new Cu sorbent lb/lb circulated*	*~1×10^{-6}*

Using the latest efficiency data reported by LSU (see Appendix A) and the attrition losses estimated in Table 2, we estimated the cost for CO_2 capture using SETS; Table 3 presents these estimates. The fuel usage and sorbent losses are significantly lower than expected, reducing the estimated cost for SETS from $25/ton (Phase I estimate) to ~$15/ton. Capital costs will also be updated by LSU but that data will not be available until sometime in the summer.

Table 3. Latest estimate of SETS cost for CO_2 capture.

Fuel costs: $3/MMBtu @2.7% fuel penalty (1.5 efficiency points) = 0.57mill/kwh

Capital costs: 194/kW$_e$ 75% CF, 10% FCR =2.95 mills/kWh Total of 3.97 mills/kWh

Sorbent cost: $1/ton CO_2 removed sorbent costs include: the expense of the manufactured material, the oxygen loading per cycle and the loss rate of sorbent per cycle.

Overall: $15/ton of CO_2 removed or 16% increase in COE

TGA cycling results

A number of cycles were conducted for each of the fabricated sorbents in order to 1) understand the reaction kinetics, 2) evaluate the ability of a sorbent to withstand multiple oxidation/reduction cycling, and 3) evaluate the repeatability of the results. In this test, a sorbent was first exposed to the reducing gas mixture (50% H_2 in CO_2) and then re-oxidized in air. The cycling was conducted isothermally at 900°C. The experiments were run in a fully automatic mode, with all the experimental parameters, including temperature and gas composition, being computer controlled. The resulting weight change was also computer monitored and recorded every 10 seconds.

Figure 5. Shimadzu micro-balance/TGA.

These tests were conducted for ~ 80 cycles on each the 10 sorbents (i.e., combination of composition and firing temperature). The cycling results are summarized in Table 4. These data show that the sorbent retained chemical activity with multiple cycling at the highest firing temperatures in all cases. Therefore TDA will be focusing our fluidized bed tests on the sorbents fired at the highest temperatures, since those sorbents have the best attrition resistance (i.e., longest life and lowest sorbent replacement costs).

Although the most attrition resistant sorbents are those fired at the highest temperatures, they are also the densest sorbents. The TGA tests are able to measure capacity effects on cycling but not the rapid reaction rates which are needed in fluidized bed/transport reactors. Sorbents fired at less than the maximum temperature may be needed to achieve the fast reaction rates needed.

Table 4. Results of the multi-cycling tests for the synthesized sorbents.

Sorbent Composition	**Firing temperature, °C**	**Weight change,%**	**Comments**
30% CuO / α–Al_2O_3	1050	6.0	Survived
	1100	5.9	Survived
	1200	6.2	Survived
5% CuO/45% Fe_2O_3/α–Al_2O_3	950	4.0	Survived
	1050	2.8	Barely survived
	1100	4.2	Survived
	1200	5.0	Survived
50% Fe_2O_3 / α–Al_2O_3	950	3.9	Barely survived
	1050	4.2	Survived
	1100	4.0	Survived

Fluidized Bed Testing at TDA

TDA designed a fluidized bed reactor for testing the spray dried, fluidizable geodes, which became operational in April. We installed and calibrated new gas analysis equipment for continuous measurement of CO, CO_2, and CH_4 (these instruments were purchased by TDA and not charged to the contract). We calibrated the gas chromatograph for hydrogen (using a N_2 carrier gas) to allow high H_2 areas and rapid cycling. We will be able to collect H_2 data every two minute; the other gases will be recorded every 30 seconds.

100% CO_2 Capture with SETS

This month TDA initiated a study to evaluate the potential for SETS to capture 100% of the CO_2 produced in the power plant. Two basic approaches are being considered: 1) use of a membrane to separate hydrogen from reformed natural gas (or oil or gasified coal). The H_2, instead of the natural gas, would be burned in the combustor to reach the peak temperatures required in a high efficiency combined cycle; and 2) inclusion of a fuel cell in the combined cycle.

Figure 6 illustrates the basic system for the hydrogen membrane approach. The natural gas is reformed by partial oxidation with O_2 (either from a cryogenic oxygen separation plant or from an Ion Transfer Membrane, ITM). The membrane removes most of the hydrogen, and the remaining CO, CO_2, H_2, and CH_4 are delivered to SETS, which fully oxidizes these gases to CO_2 and H_2O. The hydrogen is cooled and recompressed and used to provide the remaining heat required in the combined cycle power plant.

Figure 7 presents a hybrid fuel cycle power plant of the type TDA is considering. In this system no CO_2 is captured. Although 75% to 85% of the fuel is reacted in the fuel cells, the unreacted fuel, including the unburned CO and CH_4, are oxidized to CO_2 and released into the environment.

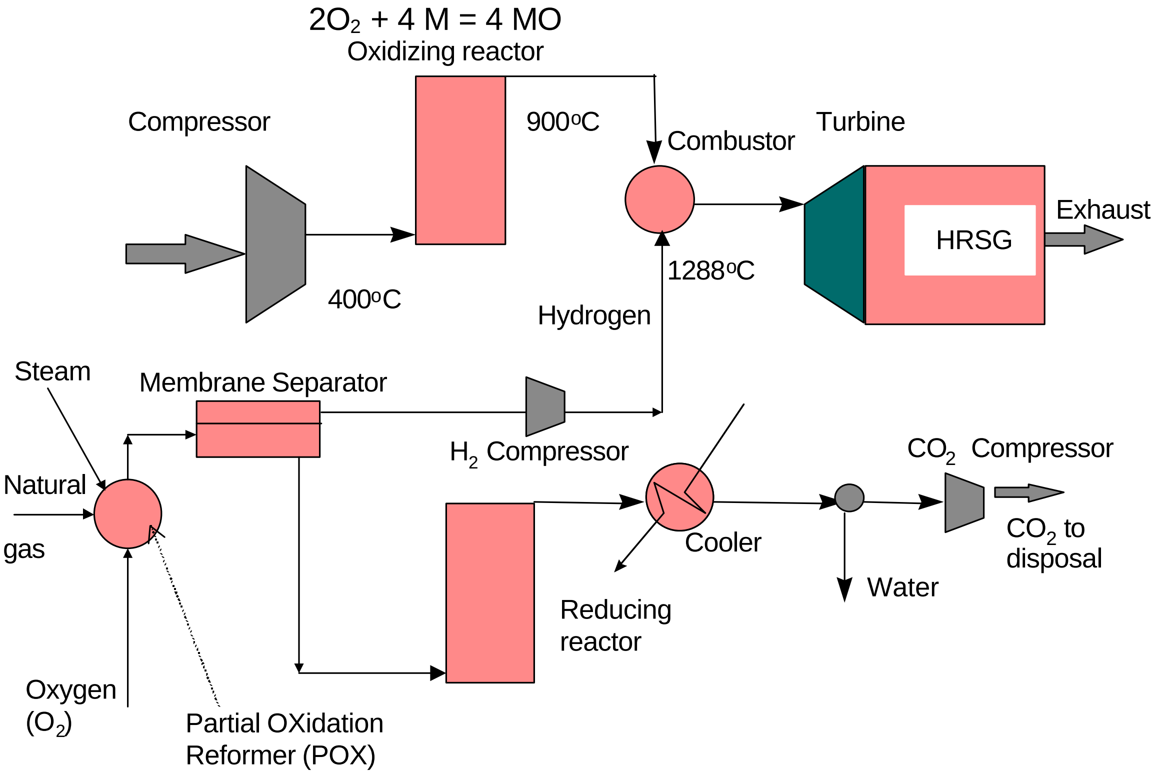

Figure 6. SETS with POX Reforming.

Figure 8 illustrates the same hydrid fuel cycle, but modified so that the fuel leaving the fuel cell is fully oxidized by SETS. The SETS still deliverers the heat value of the fuel to the power cycle, so the efficiency impact should be minimal. However, we capture 100% of the CO_2.

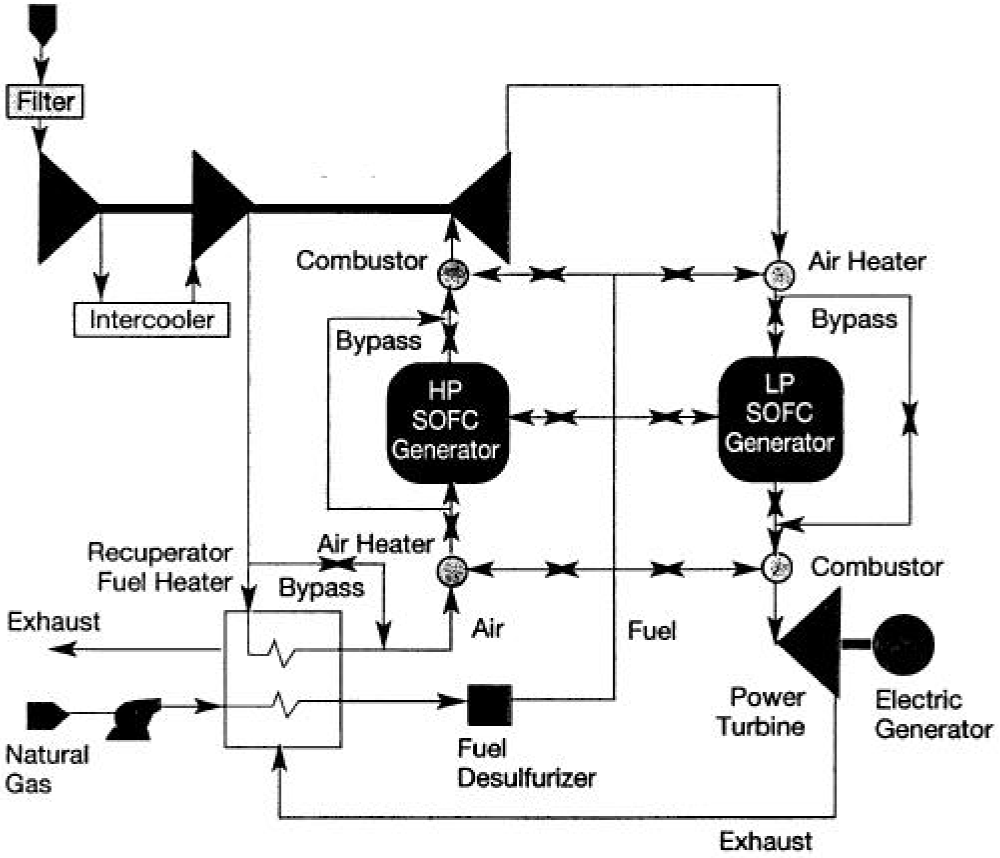

Figure 7. High-efficiency hybrid cycle (De Biasi *Gas Turbine World* July-August 1999).

As of this writing, TDA is discussing these approaches to increased CO_2 capture with DOE and we plan to select one approach based on the goals of the DOE greenhouse program.

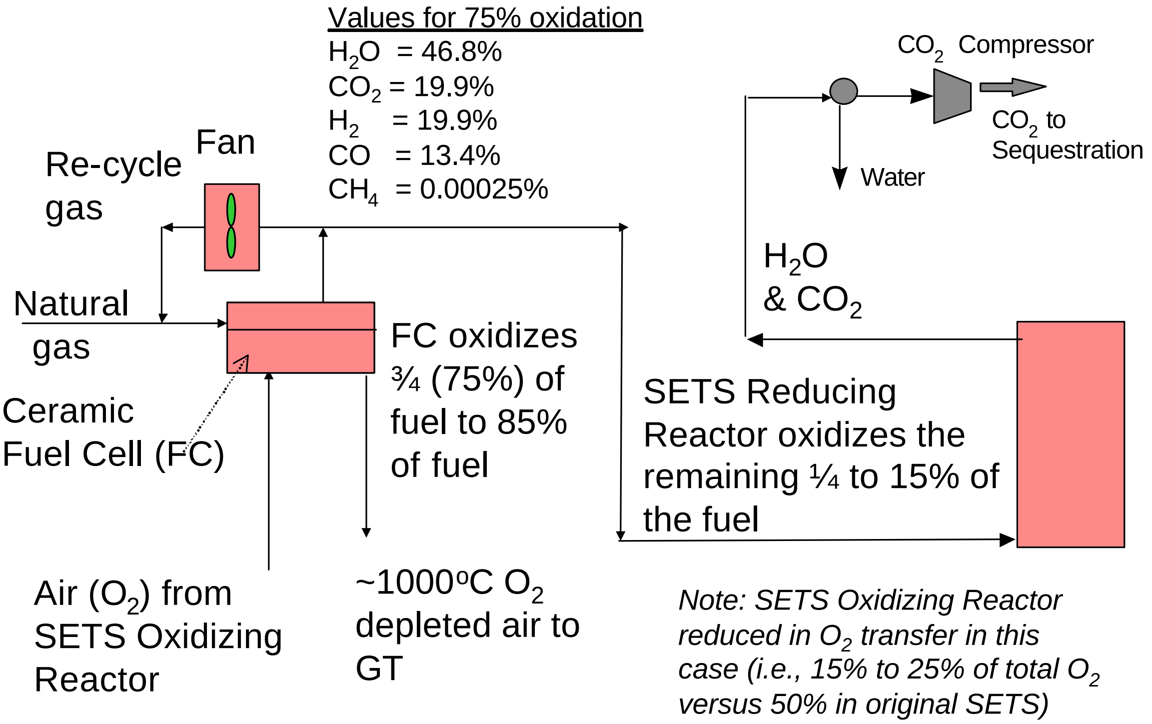

Figure 8. Fuel Cell with SETS.

LSU Subcontract

Professors Corripio and Harrison of LSU traveled to TDA for an April 13, 2000 project review meeting. Appendix A presents their latest progress report summarizing current analysis results and planned future analyses.

Base Case Simulation

Additional minor changes were made in our base case simulation. In addition, the simulation was scaled up to actual flow rates reported by GE for the MS7001FA combined cycle. Our previous simulation of the SETS combined cycle used arbitrary flow rates in order to evaluate overall cycle efficiency. The scale up was in preparation for design calculations associated with sizing specific units of the SETS combined cycle. The base case changes include raising the steam reheat temperature to 538^{o}C (1000^{o}F), increasing the low-pressure steam to 90 psi, and specifying the pump efficiency to be 0.90.

Results of the new SETS combined cycle base case simulation using actual flow rates are attached. Our total natural gas (CH_4) feed rate is 3.15e04 kg/hr (6.95e04 lb_m/hr) compared to the GE specification of 7.02e04 lb_m/hr. The air rate is 1.52e06 kg/hr (3.35e06 lb_m/hr), the same as GE's specification. The primary reason for the difference in fuel feed rates is the difference in heating values for CH_4 and natural gas. The overall efficiency is now 53.0% (LHV) with a net power output of 232 MWe, compared to a cycle efficiency of 55.0% (LHV) and power output of 241MWe reported by GE for their natural gas fired combined cycle. The current SETS cycle provides for 49.2% carbon capture. Power requirements associated with individual SETS combined cycle units are summarized below (all units are MWe).

Gas Turbine	320.8
Air Compressor	-166.0
CO_2 Compressor	-0.8
HP Steam Turbine	19.2
IP1 Steam Turbine	4.6
IP2 Steam Turbine	11.2
LP Steam Turbine	43.6
Pumps (combined)	-0.7

Future Activities

We will test our best formulations for chemical activity and durability after multiple cycling in the fluidized bed reactor and conduct multiple cycle tests (up to 50) on the most promising sorbents. We will also select one method of increasing CO_2 capture with SETS based on input from DOE.

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Appendix A
LSU Progress Report for April 2000

ASPEN MODELING OF

A NOVEL CO_2 SEPARATION SYSTEM

Progress Report 08

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April 2000

Project Review

Professors Corripio and Harrison traveled to TDA for an April 13, 2000 project review meeting. Three questions about our current natural gas-fired combined cycle process based on the GE MS7001FA turbine arose during the discussion. We are now trying once again to get specific information from GE concerning the actual steam pressures specified for the combined cycle, the intermediate pressure steam reheat temperature, and whether or not a regenerative heat exchanger using gas turbine exhaust to preheat incoming air is used. Also, the three base case steam pressures reported in that meeting were in error. The pressures are 1500, 300, and 90 psi instead of 2500, 450, and 90 psi reported.

Base Case Simulation

Additional minor changes were made in our base case simulation. In addition, the simulation was scaled up to actual flow rates reported by GE for the MS7001FA combined cycle. Our previous simulation of the SETS combined cycle used arbitrary flow rates in order to evaluate overall cycle efficiency. The scale up was in preparation for design calculations associated with sizing specific units of the SETS combined cycle. The base case changes include upping the steam reheat temperature to 538^{o}C (1000^{o}F), increasing the low-pressure steam to 90 psi, and specifying pump efficiency to be 0.90.

Results of the new SETS combined cycle base case simulation using actual flow rates are attached. Our total natural gas (CH_4) feed rate is 3.15e04 kg/hr (6.95e04 lb_m/hr) compared to the GE specification of 7.02e04 lb_m/hr. The air rate is 1.52e06 kg/hr (3.35e06 lb_m/hr), the same as GE's specification. The primary reason for the difference in fuel feed rates is the difference in heating values for CH_4 and natural gas. The overall efficiency is now 53.0% (LHV) with a net power output of 232 MWe, compared to a cycle efficiency of 55.0% (LHV) and power output of 241MWe reported by GE for their natural gas fired combined cycle. The current SETS cycle provides for 49.2% carbon capture. Power requirements associated with individual SETS combined cycle units are summarized below (all units are MWe).

Gas Turbine	320.8
Air Compressor	-166.0
CO_2 Compressor	-0.8
HP Steam Turbine	19.2
IP1 Steam Turbine	4.6
IP2 Steam Turbine	11.2
LP Steam Turbine	43.6
Pumps (combined)	-0.7

Equipment Design

Equipment design using the full-scale SETS process has been initiated. A schematic diagram of the reactors is shown in the attachment and preliminary dimensions are summarized in Table 1. The three SETS reactors were sized based on gas velocity and contact times of 12.5 m/s and 2 s, respectively, with both values taken from earlier TDA reports. Thus, the gas-sorbent contact height (H2) is 25m in all three reactors. A 5 m disengaging height (H2) and 2 m height below the gas distributor (H1) were then added to produce overall reactor heights of 32 m. Flow diameter (D1) was determined by dividing the maximum volumetric flow rate (at the reactor exit) by the gas velocity. A six-inch thick layer of firebrick (t1) was added to protect the reactor walls from the high reactor temperature. Finally the wall thickness (t) was calculated from the equation

$$t = \frac{PR}{SE_J - 0.6P} + C_c$$

where: P = reactor pressure, psi

R = inside radius of reactor, m

S = maximum allowable working stress, psi 13,700 (Carbon Steel, -20 to 650 °F)

E_J = joint efficiency 0.85

C_c = corrosion allowance, in 0.125

Reactor mass was then calculated from the known dimensions.

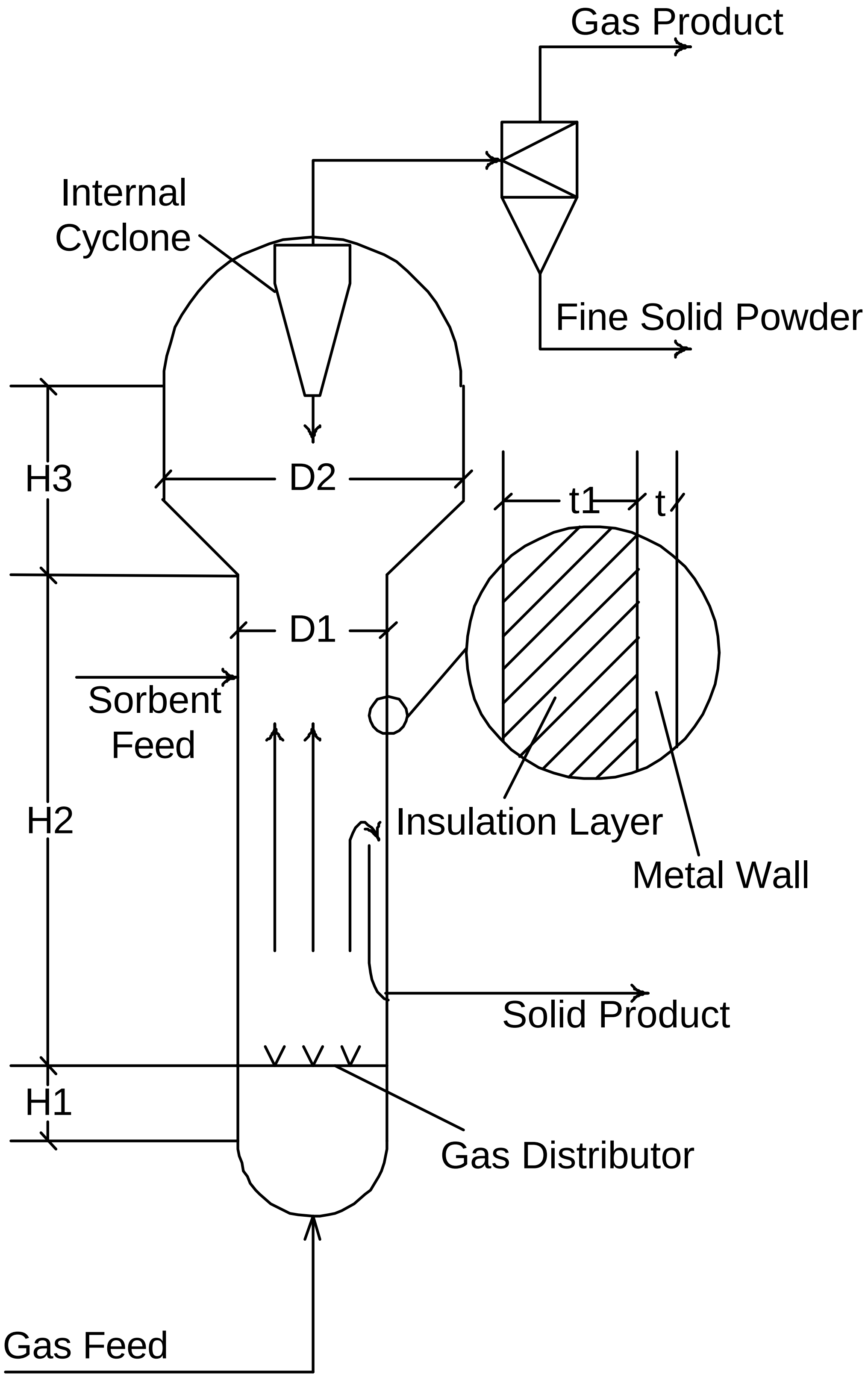
Tentative overall heat transfer coefficients based on the phases of the hot and cold fluids have been chosen from the literature and used to determine total heat transfer surface area in the three exchangers associated with the SETS cycle. Heat duties and heat transfer areas for the three SETS cycle exchangers are presented in Table 2.

Table 1. SETS Reactor Dimensions

	REACT1	REACT2	REACT3
Gas-Sorbent Contact Time, s	2	2	2
Maximum Gas Velocity, m/s	12.5	12.5	12.5
Gas-Sorbent Contact Height (H2), m	25	25	25
Sorbent Disengaging Height (H3), m	5	5	5
Gas Distribution Height (H1), m	2	2	2
Total Reactor Height, m	32	32	32
Flow Diameter (D1), m	0.95	1.02	3.14
Disengaging Zone Diameter (D2), m	1.91	2.04	6.28
Firebrick Thickness (6-inch), m	0.152	0.152	0.152
Pressure Vessel Inside Diameter, m	1.26	1.32	3.44
Vessel Wall Thickness, m	0.013	0.013	0.034
Reactor Mass, kg	23,600	26,900	207,800

Table 2 Heat Exchanger Specifications

Exchanger	Heat Duty, 10^6 kcal/hr	Overall U, Btu/ft^2 hr ^{o}F	LMTD, ^{o}C	Area, m^2
Reformer	3.75	150	215	23.7
HRSG1	28.2	31	217	858
Recuperator	8.52	150	22.2	584


Gas Product
Internal
Cyclone
Fine Solid Powder
H3
D2
t1
t
D1
Sorbent
Feed
H2
Insulation Layer
Metal Wall
Solid Product
H1
Gas Distributor
Gas Feed

	G01	G02	G03	G04	G05	G06	G07	G08	G09	G10	G11	G12	G13	G14	G15	G16	G17	G18	G19	G20
Temperature C	15	843	900	843	616	843	540	130	40	39.9	125.1	551	774.3	15	872.3	1288	589.4	559.8	496.7	468.2
Pressure bar	13.679	15.057	15.401	15.057	16.436	15.057	14.712	14.367	14.022	13.678	35.464	16.091	15.746	15.057	14.712	14.712	1.038	1.036	1.033	1.031
Vapor Frac	1	1	1	1	1	1	1	0.411	0.335	1	1	1	1	1	1	1	1	1	1	1
Mole Flow kmol/hr	965.342	2896.03	5792.06	5792.06	3861.37	2896.03	2896.03	2896.03	2896.03	969.391	969.391	4132.25	5790.25	998.432	53316.9	53316.9	54769	54769	54769	54769.1
Mass Flow kg/hr	15486.7	77266.3	154533	154533	92753.1	77266.3	77266.3	77266.3	77266.3	42542	42542	92753.1	142177	16017.6	1482830	1482830	1524580	1524580	1524580	1524590
Volume Flow cum/hr	1646.64	17872.6	36743.8	35745.3	17376.2	17872.6	13259.5	2719.07	1724.7	1734.34	849.754	17615.7	32083.2	1543.14	346513	471740	3784210	3662840	3393770	3275650
Enthalpy MMkcal/hr	-17.32	-178.89	-354.04	-357.79	-196.21	-178.89	-188.42	-216.68	-223.35	-90.937	-90.231	-192.46	-316.82	-17.919	133.943	133.95	-176.02	-188.89	-215.97	-228.12
Mole Flow kmol/hr																				
CU																				
CUO																				
FEO																				
FE2O3																				
H2O		1930.68	3861.36	3861.36	1930.68	1930.68	1930.68	1930.68	1930.68	4.634	4.634	1736.69	3322.38		3409.54	5406.39	5420.91	5420.91	5420.91	5420.94
O2		0.003	0.006	0.006	0.003	0.003	0.003	0.003	0.003	0.003	0.003				8853.47	6856.62	7161.57	7161.57	7161.57	7161.57
CH4	965.342	trace	trace	trace	965.342	trace	trace	trace	trace			829.899	0.902	998.432	998.432					
CO2		965.34	1930.68	1930.68	965.34	965.34	965.34	965.34	965.34	964.748	964.748	1023.9	1698.3			998.424	998.424	998.424	998.424	998.424
N2															40055.4	40055.4	41188.1	41188.1	41188.1	41188.1
AL2O3																				
H2		0.004	0.007	0.007	0.004	0.004	0.004	0.004	0.004	0.004	0.004	464.884	537.18		< 0.001	0.016	0.016	0.016	0.016	0.016
CO		0.002	0.004	0.004	0.002	0.002	0.002	0.002	0.002	0.002	0.002	76.888	231.484			0.009	0.008	0.008	0.008	0.008
C																				

	G21	G22	G23	G24	G25	G26	G27	G28	G29	G30	G31	G33	G34	G35	G36	G37	G38	G39	H01	H02
Temperature C	340.7	279.9	234.1	227.7	200.5	171.1	136.5	15	392.4	495.9	900	392.4	471.7	666	659.1	392.4	89.1	1286	497.7	833.1
Pressure bar	1.028	1.026	1.023	1.021	1.018	1.016	1.014	1.013	15.401	15.401	14.712	15.401	15.401	1.038	1.038	15.401	1.014	14.712	15.746	15.401
Vapor Frac	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1	1
Mole Flow kmol/hr	54769	54769	54769	54769.1	54769.1	54769.1	54769	52805.3	52805.3	54249.2	52318.5	1452.15	51353.1	53316.9	54769	51353.1	54769	53316.9	2896.04	2896.04
Mass Flow kg/hr	1524580	1524580	1524580	1524590	1524590	1524590	1524580	1518170	1518170	1528600	1466820	41749.8	1476420	1482830	1524580	1476420	1524580	1482830	52172.9	52172.9
Volume Flow cum/hr	2718960	2455640	2257580	2234330	2118270	1991470	1840620	1248160	190946	226482	348210	5251.02	207738	4010630	4089800	185695	1627240	471129	11640	17273.2
Enthalpy MMkcal/hr	-281.29	-306.08	-324.56	-327.14	-337.99	-349.69	-363.38	-34.281	108.475	-10.248	151.862	2.983	136.05	-148.45	-145.46	105.492	-382.02	132.995	-155.82	-146.3
Mole Flow kmol/hr																				
CU																				
CUO																				
FEO																				
FE2O3																				
H2O	5420.9	5420.9	5420.9	5420.95	5420.95	5420.95	5420.9	528.053	528.053	3409.57	3409.54	14.521	513.531	5406.39	5420.91	513.531	5420.9	5406.39	2896.04	2896.04
O2	7161.57	7161.57	7161.57	7161.57	7161.57	7161.57	7161.57	11089.1	11089.1	10784.2	8853.47	304.95	10784.2	6856.62	7161.57	10784.2	7161.57	6856.62		
CH4																				
CO2	998.424	998.424	998.424	998.424	998.424	998.424	998.424							998.424	998.424		998.424	998.424		
N2	41188.1	41188.1	41188.1	41188.1	41188.1	41188.1	41188.1	41188.1	41188.1	40055.4	40055.4	1132.67	40055.4	40055.4	41188.1	40055.4	41188.1	40055.4		
AL2O3																				
H2	0.016	0.016	0.016	0.016	0.016	0.016	0.016				< 0.001			0.016	0.016		0.016	0.016		
CO	0.008	0.008	0.008	0.008	0.008	0.008	0.008							0.008	0.008		0.008	0.008		
C																				

	H03	H04	H05	H06	H07	H09	H10	H11	H12	H13	H14	H15	H16	H17	H18	H19	H20	H21	H22	H23
Temperature C	20	39.9	161.2	412.2	232.5	312.2	312.2	345.4	537.8	537.8	34.1	213.6	284.3	497.7	161.4	214.5	162.4	497.7	161.4	161.2
Pressure bar	6.55	13.678	6.205	103.766	6.205	104.11	104.11	6.205	20.339	103.421	0.04	20.684	20.684	15.746	21.029	104.455	104.8	15.746	21.029	6.205
Vapor Frac	0	0	0.72	1	1	1	0	1	1	1	0.865	0	1	1	0	0	0	1	0	0
Mole Flow kmol/hr	2896.03	1926.64	2896.03	9282.22	12448.4	9282.22	9282.14	9552.39	11455.8	9282.22	12448.4	2173.61	11455.8	11455.8	2173.61	9282.14	9282.14	8559.8	9282.14	1240.73
Mass Flow kg/hr	52172.7	34724.3	52172.7	167222	224262	167222	167220	172089	206380	167222	224262	39158.1	206380	206380	39158.1	167220	167220	154207	167220	22352.1
Volume Flow cum/hr	52.26	35.464	11723.6	4377.59	82502.7	3073.33	267.296	78270.4	37479.9	5650.58	6872880	49.913	24258.4	46044.2	46.089	213.471	197.113	34404.2	196.817	26.301
Enthalpy MMkcal/hr	-200.23	-132.42	-171.97	-512.67	-698.64	-524.81	-577.98	-526.68	-612.3	-499.79	-736.89	-141.21	-639.38	-616.38	-143.8	-602.78	-613.63	-460.56	-614.06	-82.091
Mole Flow kmol/hr																				
CU																				
CUO																				
FEO																				
FE2O3																				
H2O	2896.03	1926.05	2896.03	9282.22	12448.4	9282.22	9282.14	9552.39	11455.8	9282.22	12448.4	2173.61	11455.8	11455.8	2173.61	9282.14	9282.14	8559.8	9282.14	1240.73
O2		trace																		
CH4																				
CO2		0.591																		
N2																				
AL2O3																				
H2		trace																		
CO		trace																		
C																				

	H24	H25	H26	H27	H28	H29	H30	H31	H32	H33	H34	H35	H36	H37	H38	H39	H40	H41	H42	H43
Temperature C	366.3	161.4	161.2	161.2	213.6	161.2	300.9	29.1	29.1	161.2	161.2	213.6	213.6	312.2	312.2	111	161.2	161.2	110.9	101
Pressure bar	6.205	21.029	6.205	6.205	20.684	6.205	20.684	0.04	1.379	6.205	6.205	20.684	20.684	104.11	104.11	6.55	6.205	6.205	1.379	1.379
Vapor Frac	1	0	0	0	1	1	1	0	0	0	1	0	1	0	1	0	1	1	0	0
Mole Flow kmol/hr	8559.8	11455.7	11455.7	12696.5	2173.62	1240.74	9282.22	12448.4	12448.4	1240.73	1240.73	2173.62	2173.62	9282.2	9282.2	12696.5	992.591	248.148	12696.5	12448.4
Mass Flow kg/hr	154207	206378	206378	228730	39158.4	22352.2	167222	224262	224262	22352.2	22352.2	39158.4	39158.4	167221	167221	228730	17881.8	4470.45	228730	224262
Volume Flow cum/hr	72597.1	242.906	242.841	269.142	3886.05	6969.94	20354.3	226.634	226.634	26.301	6970	49.914	3886.05	267.297	3073.34	252.415	5575.95	1393.99	252.403	244.606
Enthalpy MMkcal/hr	-470.36	-757.86	-757.96	-840.05	-122.74	-70.396	-516.64	-858.35	-858.34	-82.092	-70.396	-141.21	-122.74	-577.98	-524.81	-853.74	-56.317	-14.079	-853.77	-839.7
Mole Flow kmol/hr																				
CU																				
CUO																				
FEO																				
FE2O3																				
H2O	8559.8	11455.7	11455.7	12696.5	2173.62	1240.74	9282.22	12448.4	12448.4	1240.73	1240.73	2173.62	2173.62	9282.2	9282.2	12696.5	992.591	248.148	12696.5	12448.4
O2																				
CH4																				
CO2																				
N2																				
AL2O3																				
H2																				
CO																				
C																				

M1	M2	M3
900	774.3	900
15.401	15.746	14.712
1	1	1
5792.057	5790.246	52318.456
154532.62	142176.71	1466820
36743.781	32083.182	348210.39
-354.038	-316.815	151.862
3861.36	3322.383	3409.539
0.006		8853.474
trace	0.902	
1930.679	1698.298	
		40055.443
0.007	537.18	< 0.001
0.004	231.484	
1031850	970074.57	5025530
-2198.58	-2037.015	-7375.199
900	774.3	900
15.401	15.746	14.712
0	0	0
8193.904	10078.58	30890.94
877321.51	827897.86	3558710
332.157	205.421	1455.372
-1844.542	-1720.2	-7527.061
301.105	1505.523	
1204.418		6022.09
942.338	4711.69	
1884.676		9423.38
3861.367	3861.367	15445.47

	S1	S2	S3	S4	S5	S6
Enthalpy MMkcal/hr	-1881.765	-1844.542	-1720.2	-5645.296	-7365.496	-7527.061
Temperature C	900	900	774.3	900	874.4	900
Pressure bar	14.712	15.401	15.746	14.712	15.057	14.712
Vapor Frac	0	0	0	0	0	0
Mole Flow kmol/hr	7722.735	8193.904	10078.58	23168.205	33246.785	30890.94
Mass Flow kg/hr	889677.42	877321.51	827897.86	2669030	3496930	3558710
Volume Flow cum/hr	363.843	332.157	205.421	1091.529	1256.131	1455.373
Enthalpy MMkcal/hr	-1881.765	-1844.542	-1720.2	-5645.296	-7365.496	-7527.061
Mole Flow kmol/hr						
CU		301.105	1505.523		1505.523	
CUO	1505.523	1204.418		4516.568	4516.568	6022.09
FEO		942.338	4711.69		4711.69	
FE2O3	2355.845	1884.676		7067.535	7067.535	9423.38
H2O						
O2						
CH4						
CO2						
N2						
AL2O3	3861.367	3861.367	3861.367	11584.102	15445.47	15445.47
H2						
CO						
C						

	Q1	Q2
QCALC MMkcal/hr	3.7493052	30.55843

	W01	W02	W03	W04	W05	W06
POWER kW	-320763.4	166022.12	820.54729	-43586.77	-4649.368	-11166.07
	W07	W08	W09	W10	W11	
POWER kW	-19202.36	9.3658567	111.10311	508.8772	40.283753	

Power Generated kW	399367.98
Power Consumed kw	167512.3
Simple Cycle Power kW	154741.29
Net Power kW	231855.69
HHV Efficiency	0.4773723
LHV Efficiency	0.5297465