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Final Report: Technoeconomic Evaluation of Underground Coal Gasification (UCG) for Power Generation and Synthetic Natural Gas

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Technoeconomic Evaluation of Underground Coal Gasification (UCG) for Power Generation and Synthetic Natural Gas

Final Report, June 15, 2011

UCG Group, E-Program,

Global Security Directorate, Lawrence Livermore National Laboratory

Executive Summary

LLNL has evaluated the economics of utilizing syngas from Underground Coal Gasification UCG for two scenarios:

- Power generation from UCG (at the 485 MW net output scale)
- Synthetic Natural Gas (SNG) from UCG (66 Trillion BTU/yr scale [34,500 Barrels Oil Per Day (BOPD) Equivalent])

For both scenarios, while the economics are not quite competitive at currently prevailing U.S. prices, they may be competitive for locations with higher prevailing energy and natural gas prices (e.g. Central Europe, Japan) or in the future if natural gas and electricity prices rise substantially in the U.S. The economics of power production become significantly more favorable after the depreciation period. Costs associated with sales taxes and corporate income taxes are not included in our cost estimates.

For the synthetic natural gas from UCG option, we see significant challenges meeting pipeline specifications for content of nitrogen and other impurities. We have found that even with low percentages of nitrogen (<1.8%) in the feed syngas, separation processes in the gas cleanup increase the percentage of nitrogen in the syngas as CO₂ is removed and CO and H₂ converted to methane. Minor intrusions of nitrogen in the underground formation, combined with nitrogen within the combusted coal itself, may cause the percentage of nitrogen in the product SNG to exceed specifications of <5%. Hence, we judge SNG from UCG as being challenging technically under current pipeline specifications.

*Table E.1
Cost Estimate Summary Table*

Cost Parameter	Power Generation	Synthetic Natural Gas
Nominal Capacity	485 MW Net	66 Trillion BTU/yr [34,500 BOPD Equivalent]
Total Fixed Capital	U.S.\$886 million	U.S. \$363 million
Production Costs including Cost of Capital	\$93/MWh	\$7.5/MMBTU
Production Costs including Cost of Capital but excluding depreciation	\$60/MWh	\$6.7/MMBTU

Introduction

Scope Note

This report concerns the technoeconomics of using Underground Coal Gasification (UCG) for power generation and for production of synthetic natural gas. Lawrence Livermore National Laboratory was retained under the Work for Others Agreement L-13208 for ExxonMobil Upstream Research Laboratoryⁱ to investigate the economics of using UCG for feedstock supply for these two scenarios. The scope included conceptual designs, mass balances, and capital & operating cost estimates.

Methodology

LLNL performed the work as follows:

- Capacity of the design scenarios was agreed with ExxonMobil. The hypothetical location used for cost estimates was the Powder River Basin, Wyoming, USA
- Likely compositions of the UCG product gas were estimated for both air-blown and O₂/steam-blown UCG operations, based on results from historic field tests
- Clean-up and use of the UCG syngas was simulated using ASPENTech process simulation software using the Predictive Soave-Redlich-Kwong (PSRK) property method. ASPENTech was also used for sizing of certain major process equipment items.
- A conceptual design of a UCG module was devised based on knowledge of previous and current UCG field tests and plans, supplemented by knowledge of geomechanical limitations
- Capital cost estimates were generated using published correlations, published cost & prices, and vendor quotes where available
- Operating costs were estimated using vendor quotes, published prices, and labor costs typical for the projected location from the U.S. Bureau of Labor Statistics. Numbers of operators were estimated subjectively using the rule-of-thumb of 1 operator per 2-3 major process equipment items. For the UCG field operations, it was similarly assumed that one UCG field operator would be needed for every three UCG modules in operation.
- Subjective cost factors (depreciation lifetime, discount rate, contingency percentage) were agreed between the client and the LLNL team

Capital Cost Estimation

Vendor quotes were solicited from GE Power (for the cost of the power generation combined-cycle package plant). The cost of the Claus desulfurization unit was estimated by extrapolating published capital cost by Linde. Other capital equipment items were estimated using correlations in Peters & Timmerhaus, 5th Edition.ⁱⁱ Capital costs were separated into Battery Limits Investment (BLI) including equipment cost and installation of process equipment handling process streams, and Outside Battery Limits Investment (OBLI) which includes utilities, tankage, and general service facilities.

Operating Cost Estimation

As drilling of new UCG modules would continue throughout the lifetime of the UCG operation, the cost of drilling was treated as an operating expense pro-rated annually rather than a capital cost. Quotes for

drilling costs were solicited from Mitchell Drilling of Australia, a firm with extensive experience in UCG, based on a conceptual UCG design developed by LLNL. Mitchell’s cost incorporated necessary well finishing to minimize the risks of failure of the integrity of the wells that could cause contamination of the UCG site. Table 1.1 indicates the additional cost parameters used by LLNL in the capital and operating cost estimates.

Table 1.1
Cost Parameters Used

Cost Parameter	Value Used	Comments
Location	Powder River Basin, Wyoming, USA	LLNL Estimate
Discount Rate	12.5%	Client-specified
Depreciation Lifetime	7 years	Client-specified
Contingency factor for capital costs and for drilling costs	30%	Client-specified
General Service Facilities	5% of Total Fixed Capital (power generation option) 20% of Total Fixed Capital (SNG scenario)	LLNL Estimate
Waste Treatment	1% of Battery Limits Investment [BLI] (Power scenario) 5% of BLI (SNG scenario)	LLNL Estimate
Labor costs	\$31/hour	Bureau of Labor Statistics Manufacturing Wage in Wyoming, May 2009
Plant overhead	80% of Operating Labor	
Maintenance Costs (Power Scenario)	\$1.2/GWh Maintenance Labor \$1.8/GWh Maintenance Supplies	LLNL Estimate
Maintenance Costs (SNG Scenario)	Maintenance Labor : 1.6% of BLI Maintenance Supplies : 2.4% of BLI	LLNL Estimate
Taxes & Insurance	1.6% of BLI	LLNL Estimate
General, Admin, Sales & Research	5% of Plant Gate Costs ¹	LLNL Estimate
Coal Royalty Costs	\$3/tonne consumed	LLNL Estimate
Land Lease Costs	\$1,800/hectare	LLNL Estimate

UCG Gas Compositions

LLNL used weighted-averages of historical UCG tests in Wyoming and Washington states to estimate the composition of syngas. Weighting was done using the volume of coal consumed in the historical tests. It was decided that this gave compositions more rooted in empirical data than using modeling to predict

¹ Plant gate costs are defined here as the cash cost plus depreciation charges. Production cost is equal to the plant gate cost plus a charge for corporate general, sales, administration and R&D costs (GASR).

compositions. Composition and operational data was obtained from published reports on UCG testing in the U.S. ⁱⁱⁱ As each individual tests often included both air-blown and O₂/steam blown phases, these phases were separated in the calculations of compositions. It was found that some purported oxygen-blown tests in fact used a combination of oxygen and air: results from these tests were excluded from calculations of syngas compositions. Sulfur and ammonia or nitrogen oxide content in the syngas was estimated based on typical sulfur content for Powder River Basin coals.

After weighted-average compositions for air-blown and O₂-steam blown were prepared, the composition was slightly altered to account for increased methane content. The projected depth of the UCG modules in this study (480 m depth) was deeper than the UCG tests used in (most of which were conducted at 100-150 m depth). It was hypothesized that methane content would increase at depth because of the shift in equilibrium of methanation reactions towards methane product. However, it was found that the methane content only varied slightly with the depth of historical UCG tests, suggesting that methane is predominantly a result of pyrolysis reactions rather than methanation reactions. Hence, methane content was adjusted slightly (increased ~10%) and H₂, CO, and CO₂ content accordingly slightly reduced. Table 1.2 indicates the compositions used for this study.

Table 1.2
UCG Dry Product Gas Composition Estimates

Component	Dry Gas Molar Composition, Air-Blown UCG	Dry Gas Molar Composition, Oxygen Blown UCG
Nitrogen & Argon	52.1%	1.8%
Oxygen	0%	0.0%
Hydrogen	13.6%	34.1%
Methane	5.8%	10.1%
Carbon Monoxide	11.2%	10.5%
Carbon Dioxide	16%	41.1%
C2+ hydrocarbons	0.5%	0.9%
Nitrogen oxides	0.2%	0%
Sulfur oxides	0.5%	0%
Ammonia	0%	1.2%
Hydrogen sulfide	0%	0.3%

UCG Module Design

LLNL decided to assess the case of Linear Continual Retractable Injection Point (linear CRIP) for the UCG module design. LLNL and ExxonMobil agreed that a minimum depth of 1,000 feet would be considered. We assessed costs and product compositions based on a hypothetical 480 m deep, >10 m thick seam located in the Powder River Basin, Wyoming (e.g., the Big G seam).^{iv}

We assessed a 480 m deep production well with a 480 m horizontal run to be reasonable, and assumed a ~12 m thick coal seam. At this depth, we assessed an extraction percentage of 25-35% would be feasible without fracturing potentially extending upward into aquifers: the unextracted coal would form

“pillars” between the CRIP modules supporting the overburden. We assessed that the length of the horizontal run would require three injection wells over the lifetime of the CRIP module. Four instrumentation wells were also included in the design. Design parameters for the UCG modules are summarized in Table 1.3 below.

Each UCG cavity was envisioned as a roughly teardrop-shaped cavity 24 m wide, with the final shape before a subsequent CRIP maneuver of a semicircular section of 24 m diameter and a roughly trapezoidal section 24 m wide tapering to 6 m wide over a 30 m length. This gave roughly 11 CRIP maneuvers possible in a 480 m length run. We assumed that an average 10 m of the 12 m thickness of the seam would be extracted, with the remaining 2 m remaining as char. Figures 1.1 and 1.2 give conceptual plan and side views of the UCG module.

Figure 1.1
Conceptual-level Plan View of UCG CRIP Module

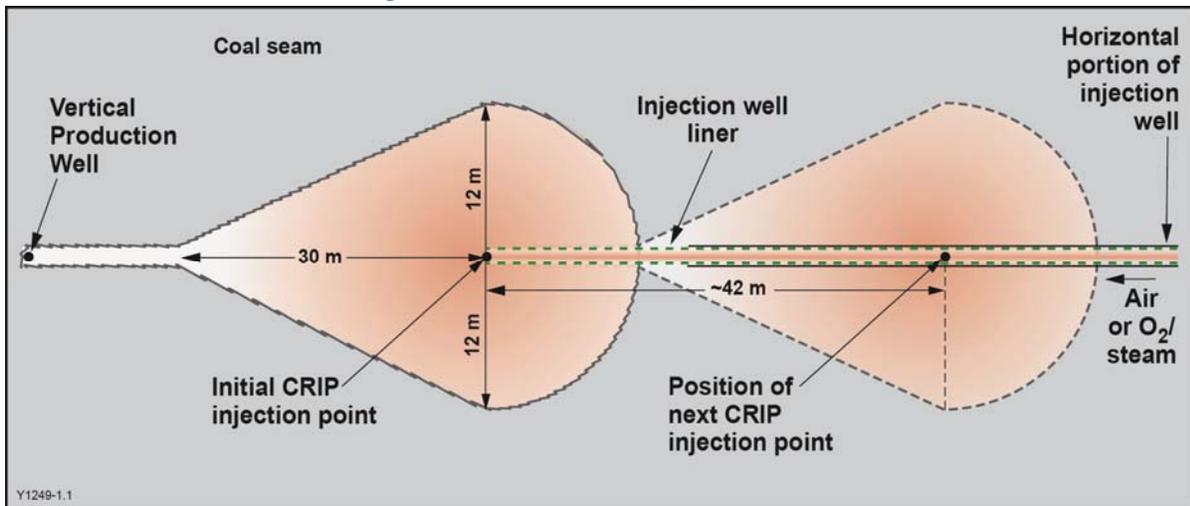
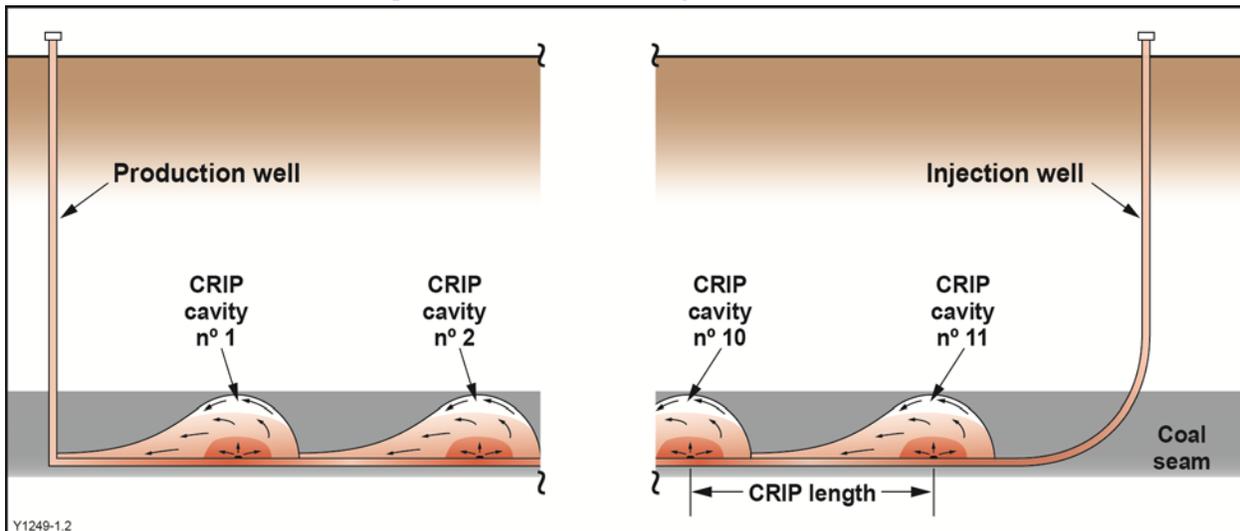


Figure 1.1
Conceptual-level Side View of UCG CRIP Module



Produced gas pressure was assumed to be 45 atm, the hydrostatic pressure of the cavity at 480 m below ground surface (bgs) with a water table of 20-30 m bgs. The pressure difference between injection and production was assumed to be ~1 atm, based on historical data from the Rocky Mountain Linear CRIP UCG tests.

Table 1.3
UCG CRIP Module Design Parameters

Parameter	Value Used	Comments
Depth to Bottom of Seam	480 m	
Seam thickness	12 m	
Module Length	480 m	
UCG Module Geometry	Linear CRIP	
Injection Cavity Length	42 m	
Maximum Cavity Width	24 m	
Depth of Char remaining	2 m	
Number of Injection Wells per CRIP module	3	
Number of instrumentation wells per CRIP module	4	
Distance between center of CRIP Modules	60-90 m	
Assumed Lifetime of one CRIP Module	~1.2 years	Assuming a ~1.12 m/day progression of cavity growth
Estimated Output per CRIP Module	43,000 tons coal	
Produced gas Pressure	45 atm	Assumed to be slightly less than the hydrostatic pressure of the UCG cavity

Case 1: Power Generation

Process Discussion

The power production flowsheet can be conceptually divided into two subsystems:

1. Raw syngas cleanup sub-system comprising
 - a. Particulate removal equipment
 - b. Waste heat recovery heat exchangers, wherein the sensible heat of the syngas is used to make steam
 - c. Acid gas removal units, comprising absorbers and scrubbers

- d. A Claus unit that converts the H₂S in the syngas to elemental sulfur
2. Power generation subsystem, comprising
 - a. gas turbines and air compressors for combustion air
 - b. Waste heat recovery heat exchangers wherein the sensible heat of the turbine exhaust is used to make steam
 - c. Steam turbines

Process Description

The process description is split into two sections, one for the gas cleanup section and one for the combined-cycle power generation system. Major equipment items are listed in Table 2.1.

Table 2.1
Power Generation from UCG
Major Process Equipment List

Item Label in Process Flow Diagram (PFD)	Size	Description	Comments
<i>Gas Cleanup Section</i>			
Reactors			
Water Gas Shift Reactor [Not Shown]	300 L	Packed Bed Multitube Reactor	304 SS
Columns			
RECTABS	14 x 5 m	Rectisol Absorber	304 SS: 20 sieve trays
RECTSTR	7.5 x 3.3 m	Rectisol Stripper	304 SS: 10 sieve trays
Pressure Vessels			
SYNFLSH1	12 x 4 m	Knockout Vessel for cooled syngas	C.S.
SYNFLSH2	10.8 x 3.6 m	Knockout Vessel for cooled syngas	C.S.
SYNFLSH3	10.8 x 3.6 m	Knockout Vessel for chilled syngas	C.S.
STRFLSH	4.5 x 1.5 m	Knockout Vessel from Stripper for Acid Gas	304 S.S.
Heat Exchangers			
WHBSYN	23 x 1000 m ²	Steam recovery from hot UCG gas	C.S. Heat Duty: 280 MW BFW
WGSCOO	3 x 1000 m ²	Syngas cooler	C.S. 65 MW Cooling Water
WGSCRYO	2 x 1000 m ²	Syngas Economizer before Rectisol Absorption	C.S. 13 MW
RECTHTX	3 x 800 m ²	pre-Rectisol Stripper Methanol Economizer	C.S. 11 MW
RECTCRY	800 m ²	Methanol Chiller	304 SS

Item Label in Process Flow Diagram (PFD)	Size	Description	Comments
RECTCOOL	375 m ²	Methanol Cooler	Heat Duty :12 MW Refrigerant 304 SS Heat Duty 20 MW Cooling Water
STREBOIL	375 m ²	Rectisol Stripper Reboiler	304 SS Heat Duty: 30 MW
Compressors			
COMP1	3 x 20 MW	3 Serial Air Compressors to supply UCG Modules	Intake flow 240,000 scf/min (400,000 m ³ /hour) Outlet pressure: 46.5 atm Not shown on flow diagram
Miscellaneous Equipment			
DEASHER		Cyclone Electrostatic precipitator Ceramic filter	Misc.
Power Generation Section (Priced as Package Unit)			
COMP1	48 MW	Air Compressors for Gas Turbine	Intake flow 600,000 scf.min (1,000,000 m ³ /hour) Outlet pressure 3.6 atm
COMP2	50 MW	Air Compressors for Gas Turbine	Outlet pressure 12.6 atm
COMP3	50 MW	Air Compressors for Gas Turbine	Outlet pressure 12.6 atm
TURB1	580 MW	Gas Turbine	Misc.
STMTURB1	52 MW	Condensing Steam Turbine	Misc.
STMTURB2	42 MW	Condensing Steam Turbine	Misc.
STMTURB3	36 MW	Condensing Steam Turbine	Misc.
STMGEN1	17,250 m ²	Steam Drum	Misc.
COMPCLR1		Intercooler for Air Compression	C.S.
COMPCLR2		Intercooler for Air Compression	C.S.
TRBFLSH1	20 m ³	Condensate Separator	C.S.
TRBFLSH2	20 m ³	Condensate Separator	C.S.
CONDCOOL	15,000 m ²	Condenser: 300 MW	C.S. 300 MW Heat duty
[COMBRXR]		Combustor	
Tankage			

Item Label in Process Flow Diagram (PFD)	Size	Description	Comments
Methanol Day Tanks	2 x 800 m ³		
Methanol Surge Tanks	4 x 100 m ³		
Other			
Claus Unit [Not shown]	40 tons/day	Package Unit	Misc.
Utilities			
Refrigeration	12 MW		
Boiler Feed Water	105 tonnes/hr		
Cooling Water	42,000 tonnes/hr		

A process flowsheet showing most, but not all, of these equipment items is given in Appendix A as Figure A.1. A stream table of process flows is given in Table B.1

Raw Syngas Cleanup Section

A combination of cyclones, ceramic filters and electrostatic precipitators are used in our design to remove particulates from the raw syngas. ^y The combination of these is noted with the symbol DEASHER in the process flowsheet, with an assumed 100% removal efficiency. The units were costed based on total gas flow treated.

The particle-free syngas stream, SYNNOASH is sent to the heat exchange WHBSYN, where it exchanges heat with the boiler feed water stream BFW1. The resultant steam stream WHBSTM is sent to the flow splitter unit, STMSPLT, where it is split into two streams, MPSTEAM1, and MPSTEAM2. The cooled syngas stream, SYNMHOT, passes through a knock-out vessel, SYNFLSH1, to remove condensate as wastewater. The dry gas is cooled further in the heat exchanger unit WGSCOOOL using cooling water, CW1, and then passes through a second knock-out vessel, SYNFLSH2. The cooled syngas stream, WGHPRDC, is further cooled in the heat exchanger WGSCRYO to cryogenic temperature, using cold syngas (BALSUNC) exiting the Rectisol absorber, RECTABS. The cold syngas, WGSPRDCC, is flashed in SYNFLSH3 to remove condensate and ice (SYNCOND). The dry cold syngas BALDSYN is introduced into RECTABS, wherein cold methanol, L1CL, is used as the solvent, to remove the acid gas H₂S. The clean syngas, BALSUNC, exchanges heat with WGHPRDC stream mentioned earlier in the heat exchanger WGSCRYO, and is then sent to the gas turbine.

The rich cold methanol stream, L1CR, exits the bottom of the Rectisol absorber unit, RECTABS, and exchanges heat, in the heat exchanger RECTHTX with the lean hot methanol stream L1HL, exiting the bottom of the Rectisol stripper unit, RECTSTR. The warm rich methanol stream L1HR is sent to the Rectisol stripper RECTSTR, where the acid gas H₂S is stripped from the rich methanol stream. The now lean methanol stream, L1HL exchanges heat with the cold rich methanol stream in RECTHTX. The steam stream MPSTEAM1, generated earlier, is used to provide the heat for the reboiler (STREBOIL) of the stripping column. The warm lean methanol stream, L1WL, is sent to a cryogenic cooler (not shown in process flow diagram A-1) to produce the cold methanol for recycle. The cold methanol stream,

L1CRYOI, is used as the refrigerant to cool the solvent methanol stream, MAKEUPL1, to produce L1CL, which is introduced into RECTABS as the Rectisol solvent. The warmed L1CRYOO is sent to a refrigeration system (not shown) to regenerate L1CRYOI.

The acid gas stream from the Rectisol stripper, RECTSTRP, is sent to a Claus unit (not shown) to recover elemental sulfur.

Power Generation Subsystem

Aspen does not have a gas turbine simulation unit, so we have had to approximate it using a combination of an adiabatic stoichiometric combustion reactor, and a reverse compressor.

Combustion air, COMBAIR1, is compressed in a series of isentropic compressors, COMP1, COMP2 and COMP3, with intermediate coolers, COMPCLR1 and COMPCLR2, to produce the compressed air stream, COMPAIR3. It is mixed with the clean syngas from the unit WGSCRYO, and sent to the reactor COMBRXR, wherein all the fuel from the syngas (H₂, CO, CH₄ and higher hydrocarbons, represented by C₂H₆) are completely oxidized. The hot pressurized syngas is sent to the reverse compressor unit, TURB1, connected to a generator, not shown, to produce power.

The hot exhaust from TURB1 is sent to the heat exchanger STMGEN1, where steam is generated from the boiler feed water stream BFW4STRB. The resultant steam stream, STM4TRB, is mixed with the steam stream MPSTEAM2, from the syngas waste heat boiler, and sent to a series of steam turbines STMTURB1, STMTURB2 and STMTURB3 to produce more power. Table 2.2 summarizes the projected power generation and consumption at the plant.

Table 2.2
Power Generation from UCG
Electrical Power Generation and Consumption

Item Label in Process Flow Diagram (PFD)	Function	Power Generated (Consumed), MW	Comments
COMP1	50 MW Combustion Air Compressor	(48)	
COMP2	50 MW Combustion Air Compressor	(50)	
COMP3	50 MW Combustion Air Compressor	(50)	
UCGCOMP1	20 MW Air Compressor for UCG Modules	(20)	Not shown in Flow Diagram
UCGCOMP2	20 MW Air Compressor for UCG Modules	(20)	Not shown in Flow Diagram
UCGCOMP3	20 MW Air Compressor for UCG Modules	(20)	Not shown in Flow Diagram
TURB1	580 MW Gas Turbine	580	

Item Label in Process Flow Diagram (PFD)	Function	Power Generated (Consumed), MW	Comments
STMTURB1	52 MW Steam Turbine	52	
STMTURB2	42 MW Steam Turbine	42	
STMTURB3	36 MW Steam Turbine	36	
	Refrigeration	(5)	
	Other Estimated Electrical Power Demands (e.g. pumping)	(12)	
Net Power Output, MW		485	

Materials of Construction

We do not envision excessively corrosive conditions in the process, hence we envisioned that most of the gas cleanup process equipment would be carbon steel, with the exception of the rectisol absorber/stripper loop, where 304 Stainless steel was projected. Power generation equipment would be a mixture of different materials, but again we do not envision more severe conditions than would be normal for IGCC or NGCC power plants.

Process Discussion

The flowsheet simulated here is has not been fully optimized for heat integration. A number of improvements are possible:

- A more thorough heat integration for in-process heat exchange between hot and cold streams is likely to reduce utility consumption (cooling water, steam and refrigeration)
- A better utilization of steam is possible. For example, all the steam generated in the syngas waste heat boiler can be sent to the first two steam turbines, and a slip stream from the exhaust of the second steam turbine can be used to provide the heat to the Rectisol stripper reboiler.
- Heat from the Claus unit, not simulated here, can be used to preheat the combustion air after the last compressor to improve the efficiency of the gas turbine.
- Specific unit operations for removal of volatile metals such as mercury

We have not completely converged the recycle loop involving the recycle of the solvent. This is done for two reasons:

- The amount of the solvent flowing through the absorber/stripper system can be independently set by the designer/operator
- Calculation of the exact composition of the circulating solvent does not significantly affect the design of the process, yet creates numerical instabilities in Aspen, thus making such calculations difficult

Instead, we show that the amounts of the methanol in and out, including the makeup methanol, are in mass balance.

Likewise, we have not shown the recycle of steam, showing, instead, the steam balance.

Capital and Operating Cost

Equipment sizes were estimated using ASPEN and according to general rules of thumb for process engineering. Table 2.1 above lists the major process equipment items. Variable costs, including estimated annual utility consumptions, are listed in Table B.2. Annual drilling costs are given in Table B.3

The capital cost of the power generation section was estimated using a rule of thumb of \$1,000/kW net power (i.e. power less power loss for compression of combustion air) given by GE Power.^{vi} As the power needed for 480 net power from the UCG/Power Plant Operation would be 550 MW (because of 60 MW required to compress air for the UCG modules, 8 MW refrigeration and ~2 MW other power demands), we estimated an installed cost for the power generation island of \$550 million.

Capital costs for the gas cleanup process equipment were estimated using Peters & Timmerhaus.^{vii} Costs were inflated from 2002 to 2010 costs using the *Chemical Engineering* plant cost index. Installation costs factors ranging from 0.9-1.8 (depending on equipment type) of purchased equipment costs, were used to include construction, electrical and instrumentation, and piping costs. Utility investment costs were also estimated using Peters & Timmerhaus, using an installation cost factor of 0.4. Certain elements (the power generation island and the Claus desulfurization unit) were treated as a package unit with installation costs included. Power generation island costs were obtained from GE Power, and Claus desulfurization package unit costs were obtained from publications by Lurgi.^{viii} It was found, as would be expected, that the capital cost of the gas cleanup section was much smaller than the power generation section.

Drilling costs, being incurred throughout the plant lifetime as UCG modules are expended, were treated as an annual operating expense. Cost estimates for drilling were obtained from Mitchell Group of Queensland, Australia.^{ix}

Labor costs were estimated using May 2009 hourly wage rates for manufacturing labor in Wyoming from the Bureau of Labor Statistics. We estimated 13 field operators per shift for the UCG production field and six operators for the above-ground process (approx. one operator for every 3 UCG modules, 4 operators per shift for the power generation section and 2 for the gas cleanup section.)

Discussion of Capital & Operating Costs

Total Fixed Capital for the Gas Cleanup and Power Generating Sections are given in Table 2.4. The total fixed capital cost was estimated at \$886 million. (For a net power output of 485 MW, this works out at \$1,760/kW generation capacity.) Working capital of 3 months of operating cash costs was estimated at \$27 million, for a total capital investment of \$913 million. Investment in the inventory of methanol kept on-site was not included.

Operating expenses, including return on investment and plant depreciation over a 7-year period, were estimated at \$360 million, as shown in Table B.5, with variable expenses shown in Table B.4 and Annual drilling expenses in Table B.3. Excluding depreciation, the operating costs, including return on investment, are \$233 million. These give a cost of \$93/MWh during the depreciation period and \$60/MWh post-depreciation.

Case 2: Synthetic Natural Gas

Process Review

For gas cleanup prior to methanation, the Rectisol process was used because of its ability to remove H₂S which would poison the methanation catalyst. However, as most of the CO₂ needs to be removed to meet pipeline specifications, three absorbers were used: a H₂S absorber, a CO₂ absorber, and a polishing absorber to remove CO₂ resulting from the methanation reaction.

Process Description

The SNG production flowsheet can be conceptually divided into two subsystems:

3. Raw syngas cleanup sub-system comprising
 - a. Particulate removal equipment
 - b. Waste heat recovery heat exchangers, wherein the sensible heat of the syngas is used to make steam
 - c. Acid gas removal units, comprising absorbers and scrubbers
 - d. A Claus unit (not shown) that converts the H₂S in the syngas to elemental sulfur
4. SNG generation subsystem, comprising
 - a. Methanation reactor
 - b. CO₂ absorber
 - c. CO₂ stripper and solvent regenerator (shared with the acid gas removal system, mentioned under the acid gas removal system)

The description of each subsystem in detail follows.

Table 3.1 gives the major process equipment items for the SNG scenario.

Table 3.1
Synthetic Natural Gas from UCG
Major Process Equipment List

Item Label in Process Flow Diagram (PFD)	Size	Description	Comments
<i>Gas Cleanup Section</i>			
Reactors			
WGSEQ	300 L	Water Gas Shift Reactor Packed Bed Multitube Reactor	304 SS
METHRXR		3 x Pack Bed Multitube Reactor	Misc
Columns			
H2SABS	10 x 7 m dia	Rectisol Absorber (Hydrogen Sulfide)	304 SS
CO2ABS	5.5 x 18 m dia	Rectisol Absorber (Carbon Dioxide)	304 SS. Fluor-Daniel is constructing CO ₂ Absorber/Strippers of up to 20 m diameter
H2SSTRP	5.5 x 5 m dia	Rectisol Stripper (H ₂ S)	
CO2STRP	5.5 x 20 m dia	Rectisol Stripper (CO ₂)	304 SS. Fluor-Daniel is constructing CO ₂ Absorber/Strippers of up to 20 m diameter
SNGRECT	10 x 6 m dia	Polishing Absorber	304 SS
Pressure Vessels			
SYNFLSH1	20 x 5 m	Knockout Vessel for syngas	C.S.
SYNFLSH2	20 x 5 m	Knockout Vessel for cooled syngas	C.S.
LTFLASH	8 x 2.5 m	Flash Vessel for CO ₂ absorber bottoms	304 S.S.
METHCOND	8 x 2.5 m	Knockout Vessel for cooled SNG product gas	304 S.S.
METHRCVR	6.5 x 2 m	Knockout Vessel for cooled feed to Claus Unit	304 S.S. Not shown on flow diagram
Heat Exchangers			
C1FDHTR	1,100 m ²	Heater for Methanation Input.	Misc. 73 MW High-Pressure Steam
L1CRYO1	13 x 1000 m ² 1 x 500 m ²	Cryogenic Cooler	304 S.S. tubes 143 MW Refrigerant
L2CRYO2	76 x 1000 m ²		304 S.S. tubes 122 MW Not shown on flow diagram
L3CRYO3	4 x 925 m ²	Cryogenic Heat exchanger	304 S.S. tubes 41 MW Not shown on flow diagram
RECTHTX1	2 x 720 m ²	Cryogenic Heat exchanger	304 S.S. tubes 53 MW
RECTHTX2	7 x 1,000 m ²	Heat Exchanger w/ Steam	304 S.S. tubes 72 MW
		Heat Exchanger w/ CW	

Item Label in Process Flow Diagram (PFD)	Size	Description	Comments
SYNCRY1	27 x 1000 m ²	Cryogenic Cooler	304 S.S. tubes
SYNCRY2	3 x 800 m ²	Cryogenic HX	304 S.S. tubes Not shown on flow diagram
SYNCRY3	6 x 1,000 m ²		304 S.S. tubes Not shown on flow diagram
SYNCRY4	6 x 1,000 m ²	Cryogenic HX	304 S.S. tubes Not shown on flow diagram
WGSCOOL	9 x 1000 m ²	Cryogenic HX	C.S.
WHBSYN	41 x 1000 m ²	Cooler pre-H ₂ S absorption Steam recovery from hot UCG gas	33 MW CW C.S. 590 MW BFW
WHBWGS	31 x 1000 m ²	Steam recovery from hot UCG gas	
[Not shown]	2 x 900 m ²	CO2STRP Reboiler	304 S.S. 177 MW Steam
[Not shown]	1 x 1200 m ²	H2STRP Condenser	304 S.S. 7 MW Refrigerant
[Not shown]	1 x 200 m ²	H2SSTRP Reboiler	304 S.S. 20 MW Steam
[Not shown]	6 x 1,100 m ²	Methanol Condenser from Claus feed	304 S.S. 10 MW Refrigeration
Tankage			
Methanol Day Tanks	2 x 1,000 m ³		
Methanol Surge Tanks	5 x 100 m ³		
Miscellaneous Equipment			
DEASHER		Cyclone Electrostatic precipitator Ceramic filter	Misc.
Other			
Claus Unit [Not shown]	105 tonnes/day	Package Unit	Misc.
Utilities			
Refrigeration	100 MW		
Boiler Feed Water	2,000 tonnes/hr		
Cooling Water	20,000 tonnes/hr		

Figure A.2 in Appendix A is a process flow diagram showing most of the major process equipment. Table C.1 in Appendix C is a stream table for the process with streams corresponding to the streams shown in Figure A.2.

Note that the stream table contains an erroneous value of too much CO₂ in the product stream (7%), due to a slight undersupply of methanol to the product polishing column SNGRECT and imperfections in ASPEN's property methods for methanol in Rectisol processes. However, we judge this error would not affect the accuracy of our cost estimates. We have verified that a slight increase (15%) in input methanol flow to SNGRECT would give CO₂ in the product gas of <0.5%. Because of project time constraints, we were unable to integrate that correction in the full process. However, we judge that such a minor change would not make a material difference to the capital and operating cost of the SNG process, given that the small size of SNGRECT to other columns in the process. We also have confidence, based on published studies, that with better property data, that our design could meet CO₂ specifications.^{2,x,xi}

Raw Syngas Cleanup Subsystem

A combination of many units is needed to remove particulates from the raw syngas. These include cyclones, baghouse filters, venture scrubbers and electrostatic precipitators.^{xii} The amount of information needed to rigorously design and size these various units is beyond the scope of this study, so we have lumped all these into a composite unit called DEASHER in the flowsheet, with 100% particle removal efficiency, and costed them based on their gas throughput.

The particle-free syngas stream, SYNNOASH is sent to the heat exchanger WHBSYN, where it exchanges heat with the boiler feed water stream BFW1. The resultant steam stream MPSTEAM0 is sent to a flow splitter unit, STMSPL1, where it is split into two streams, MPSTEAM2, and MPSTEAM3. The warm syngas stream, SYNMHOT, is flashed in the unit SYNFLSH1 to remove condensate, which is sent to a wastewater treatment unit. The dry gas is mixed with MPSTEAM3 and a portion of it is passed through the water gas shift reactor, WGSEQ. The product stream WGSPRDH is mixed with the portion not passed through the reactor to produce the balanced syngas stream BALSYN1. The hot BALSYN1 is used to generate more steam, MPSTEAM1, in the waste heat boiler WHBWGS. The warm syngas from WHBWGS is cooled further in the heat exchanger unit WGSCOOOL using cooling water, CW1. The cooled syngas is sent to the flash drum SYNFLSH2 to remove condensed water, and is further cooled in the heat exchanger SYNCRY2 (not shown on flow diagram) with the cold stream from the top of the CO2ABS, and then to cryogenic temperature in the cryogenic cooler SYNCRY1. The cold syngas, SYN2SABS, is introduced into H2SABS, wherein cold methanol, L1TOSABS, is used as the solvent, to remove the acid gas H₂S. The H₂S-free syngas, BALSYN1, is sent to CO2ABS where the CO₂ remaining in the syngas is removed using cold

² Published studies indicate that with the right thermodynamic package, a more accurate Rectisol model can be developed, and that the Rectisol system is capable of producing high purity syngas and SNG. Weiss describes a 5-column Rectisol scheme used to purify syngas wherein he feeds a syngas containing 34% CO₂ to the Rectisol system and gets a clean gas containing 10 ppm CO₂ and H₂S content of 0.24 % to 0.1 ppm. Preston has modeled a six-column Rectisol system using the SRK thermodynamic model that was modified by specifying the binary interaction parameters for all the important binary pairs from measured data. With the use of this model, she was able to get excellent agreement with field data, lowering the CO₂ concentration to <2.5 % in a column with less than ten theoretical stages.

methanol stream, L1TOCABS. The rich liquid stream from CO2ABS is flashed in LTFLSH to remove dissolved light gases H₂, CO and CH₄. The gas stream from LTFLSH is mixed with BALSYN2, the gas stream from CO2ABS, and the combined stream, TOC1RXR, exchanges heat with WGHPDC2 stream mentioned earlier in the heat exchanger SYNCRY2 (not shown) and is then sent to the methanator reactor METHRXR after passing through the fired heater C1FDHTR.

The H₂S-rich rich methanol stream is heated in the heater RECTHTX1 and then sent to the H₂S stripper H2SSTRP. The top product from the stripper containing H₂S is cooled to -40 and flashed to remove methanol (not shown on flowsheet) and then mixed with air/oxygen and sent to a Claus unit (not shown). Likewise, the cold bottom product of the LTFLSH, stream L1TOCST2, is mixed with the cold bottom stream from another Rectisol unit, SNGRECT (described later), and the combined cold stream L1CO2RCH is used to cool the recycle methanol stream further, in the heat exchanger L1CRYO2 (not shown in flow diagram). The now warm rich stream is sent to the CO₂ stripper CO2STRP where the CO₂ is stripped out from the liquid. The bottoms products of the two strippers, namely, L1WL and LEANL1 are mixed and recycled back to the front of the process where a cryogenic cooler L1CRYO1 (L1CRYO3, not shown in flow diagram) lowers the temperature of the recycle stream to the design temperature of the Rectisol unit. The top of the CO₂ absorber can be disposed of as tail gas, or sent to sequestration after further processing.

SNG Generation Subsystem

The heated balanced gas stream from C1FDHTR is sent to the methanation reactor METHRXR, simulated as an equilibrium reactor. The product of the reactor is cooled and flashed in the flash drum METHCOND to remove the condensate formed during cooling. The dry stream is sent to the Rectisol column SNGRECT to remove CO₂ from the raw product. The bottom stream from SNGRECT is mixed with the liquid stream from LTFLSH mentioned earlier to form the cold stream L1CO2RCH, which, after passing through the heat exchanger L1CRYO2 (not shown on flow diagram), is sent to CO2STRP where the CO₂ is stripped out, thus regenerating the solvent. The top product from SNGRECT is HPPRDSNG, high pressure SNG.

Process Discussion

The heating value of the SNG product is below the acceptable pipeline minimum of 950 BTU/scft (Foss, 2004).^{xiii} A major reason for this is that it contains ~8% N₂, even though the raw syngas fed to the process has only about 1.2% N₂.³ This indicates that the quality of the oxygen used as an oxidant in the upstream UCG process needs to be very high, and that minor intrusions of nitrogen (e.g. nitrogen in the coal being converted to nitrogen gas, or intrusions of air into the UCG chamber) into the product will cause great difficulty in meeting pipeline specification.

It should be further noted that the flowsheet simulated here is has not been optimized. A number of improvements are possible:

³ For the product gas with 89% methane, 8% nitrogen and ~3% other gases (~1% each of ethane, hydrogen and CO₂), we estimate the HHV would be ~920 BTU/scf.

- A more thorough heat integration for in-process heat exchange between hot and cold streams is likely to reduce utility consumption (cooling water, steam and refrigeration)
- We have not modeled a use for exported steam. For example, any excess steam can be used to generate power before being used as a heat source, thus reducing electrical power consumption in the cryogenic systems.
- Instead of using methanol as the stripping agent in the CO₂ stripper, nitrogen from the air separation unit can be used to reduce energy consumption, in scenario where the CO₂ stream can be released to the atmosphere.
- Heat from the Claus unit, not simulated here, can be used to preheat the combustion air after the last compressor to improve the efficiency of the gas turbine.
- We have not modeled purification and liquefaction of the CO₂ stream to carbon capture and storage quality. As this CO₂ stream contains some methane, purification of the CO₂ stream would likely have a beneficial effect on the economics.
- We have assumed that oxygen is delivered from the toll air separation unit (ASU) at the required pressure of ~45 atm.

We have avoided, on purpose, closing of the recycle loop involving the recycle of the solvent. This is done for two reasons:

- The amount of the solvent flowing through the absorber/stripper system can be independently set by the designer/operator
- Calculation of the exact composition of the circulating solvent does not significantly affect the design of the process, yet creates numerical instabilities in Aspen, thus making such calculations difficult.

Instead, we show that the amounts of the methanol in and out, including the makeup methanol, are in mass balance.

Likewise, we have not shown the recycle of steam, showing, instead, the steam balance.

Materials of Construction

As with the power generation option, we do not envision excessively corrosive conditions in the process, hence we envisioned that most of the gas cleanup process equipment would be carbon steel, with the exception of the rectisol absorber/stripper loop, where 304 Stainless steel was projected.

Capital and Operating Cost

Equipment sizes were estimated using ASPEN and according to general rules of thumb for process engineering. Table 3.1 lists the major process equipment items. Variable costs, including estimated annual utility consumptions, are listed in Table C.2.

Capital costs for the gas cleanup process equipment were estimated using Peters & Timmerhaus.^{xiv} Costs were inflated from 2002 to 2010 costs using the *Chemical Engineering* plant cost index. Installation costs factors ranging from 0.6-1.5 (depending on equipment type) of purchased equipment costs, were used to include construction, electrical and instrumentation, and piping costs. Utility investment costs were also estimated using Peters & Timmerhaus, using an installation cost factor of 0.4. Certain elements (the Claus desulfurization unit) were treated as a package unit with installation costs included. The methanation and water-gas shift reactors were costed using correlations for heat exchangers with a multiplier for complexity of construction.

Drilling costs, being incurred throughout the plant lifetime as UCG modules are expended, were treated as an annual operating expense. Cost estimates for drilling were obtained from Mitchell Group of Queensland, Australia.^{xv} Drilling cost estimates for the SNG scenario are given in Table C.3

Labor costs were estimated using May 2009 hourly wage rates for manufacturing labor in Wyoming from the Bureau of Labor Statistics. We estimated one operator for every 3 UCG modules, and 9 operators per shift for the gas cleanup and methanation plant.

Discussion of Capital & Operating Costs

Total Fixed Capital for the Gas Cleanup and Power Generating Sections are given in Table C.4. The total fixed capital cost was estimated at \$363 million. Working capital of 3 months of operating cash costs was estimated at \$90 million, for a total capital investment of \$453 million. This capital investment does not include the cost of the air separation unit (ASU) or capital investment for productive use of exported steam, or investment in the inventory of methanol kept on-site.

Annual operating expenses, including return on investment and plant depreciation over a 7-year period, were estimated at \$489 million, as shown in Table C.5, with variable expenses shown in Table C.2. Excluding depreciation, the operating costs, including return on investment, are \$437 million. These give a cost of \$7.5/MMBTU during the depreciation period and \$6.7/MWh post-depreciation.

Uncertainties

For the power generation option, the greatest uncertainty is in the capital cost of the power generation island. While GE power supplied a rough cost using a rule of thumb of \$1,000-1,200/kW capacity, we could not completely clarify with GE Power what assumptions underlay their cost estimate. GE Power also believed the syngas product was marginal at the combustion ratios we specified. Use of a richer syngas:air mixture would reduce the gas turbine output and adversely affect the economics.

For the SNG option, the greatest uncertainty is the price of oxygen from an air separation unit and the price of steam exported. Because of the requirement for a low percentage of nitrogen on the raw syngas from the UCG modules, cryogenic oxygen rather than pressure-swing-adsorption (PSA) oxygen would be needed. Therefore a relatively high estimate of oxygen costs was used. Also, we have assumed that excess steam not needed in the process can be exported offsite for revenue (e.g. to a co-located steam turbine electricity generation plant). A more rigorous analysis would be to include the ASU in the capital and operating costs, and inclusion of a steam turbine to consume the exported steam to supply power to the ASU and to the rest of the SNG process.

The properties method used in the ASPEN model, even between those recommended for Rectisol processes, can make a difference in the modeled performance of the absorption/stripping columns, particularly for the final polishing column in the SNG process, leading to over an order-of-magnitude difference in the percentage of CO₂ remaining in the product gas.

Other uncertainties are the exact volume of coal extractable from each UCG run. Larger volumes of coal extractable in a run would improve economics, especially for the SNG option where drilling costs alone are almost half of the operating expenses.

Field tests indicate that conditions (pressure, temperature, composition) of the product gas from an individual UCG module can fluctuate radically due to conditions in the subsurface (e.g. spalling of overburden into the UCG cavity, changing composition of coal burned, intrusion of groundwater into the cavity, startup/shutdown between CRIP maneuvers, etc.). As dozens of UCG modules would be operating in parallel, we have assumed that such excursions from average conditions would be largely be 'smoothed out'. However, some surge capacity in the gas cleanup may be needed to

We have costed the gas cleanup as a single-train Rectisol absorption system. However, for the SNG option, although columns of 20 m diameter have been constructed for CO₂/amine absorption, these may not be practical for the absorption pressures used (~45 atm), and multiple parallel absorbers may be used instead. Also, because of the large flows of methanol, separate absorption trains may be advisable for health & safety reasons. This would cause a modest increase in the capital and operating costs.

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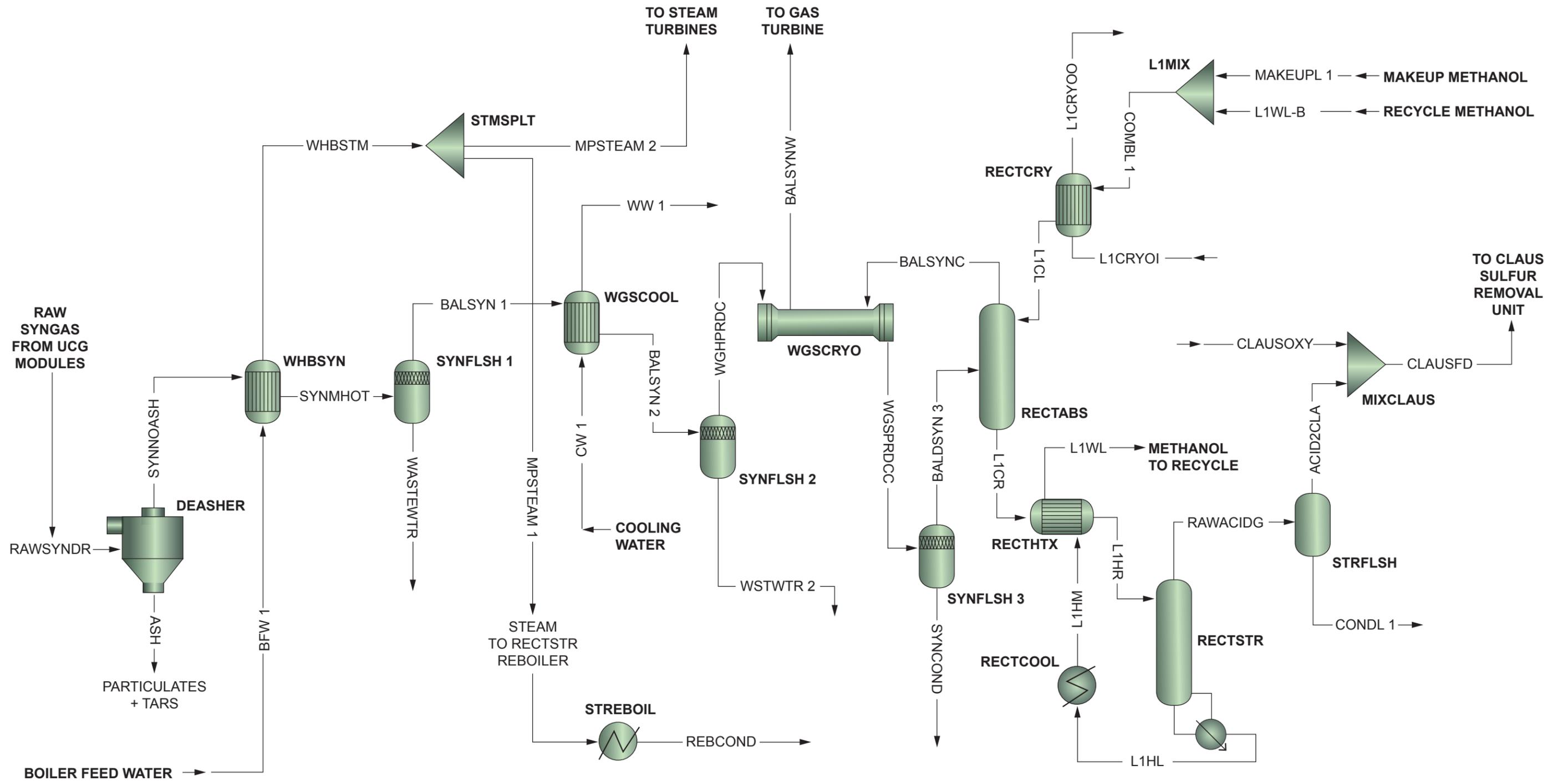
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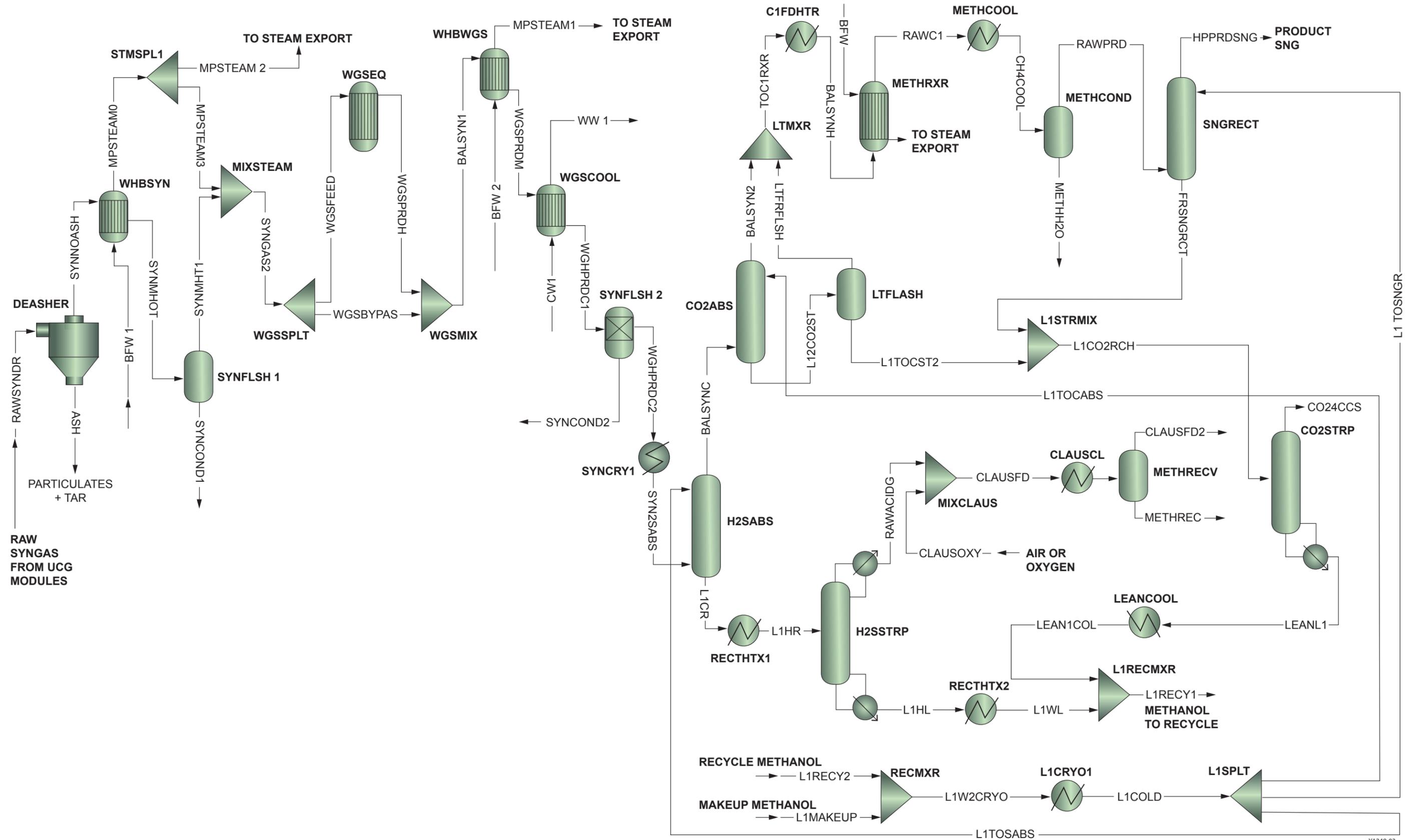
Appendix A

Figure A.1 Process Flow Diagram Power Generation from UCG. (cont.)



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Figure A.2 Process Flow Diagram Synthetic Natural Gas from UCG.



Appendix B

Table B-1
Power Generation from UCG
Stream Table

Stream Name	BALDSYN3	BALSYN1	BALSYNC	BALSYNW	BFW1	BFW4STRB
Temperature K	263	418	249	306	298	298
Pressure atm	44.7	44.9	44.7	44.7	45	45
Component Mole Flow						
<i>kmol/hr</i>						
H2	3465	3466	3450	3450	--	--
CO	2854	2854	2835	2835	--	--
H2O	4	3239	Negl.	Negl.	20000	15300
CO2	4062	4072	2881	2881	--	--
CH4	1478	1478	1442	1442	--	--
N2	13252	13252	13183	13183	--	--
C2H6	137	137	118	118	--	--
H2S	50	51	Negl.	Negl.	--	--
NH3	26	134	Negl.	Negl.	--	--
O2	--	--	--	--	--	--
SO2	--	--	--	--	--	--
CH3OH	--	--	7	7	--	--
Component Mass Flow						
<i>kg/hr</i>						
H2	6986	6986	6955	6955	--	--
CO	79943	79955	79411	79411	--	--
H2O	68	58352	Negl.	Negl.	360306	275634
CO2	178788	179192	126785	126785	--	--
CH4	23712	23715	23134	23134	--	--
N2	371225	371242	369304	369304	--	--
C2H6	4125	4130	3555	3555	--	--
H2S	1718	1727	7	7	--	--
NH3	443	2279	Negl.	Negl.	--	--
O2	--	--	--	--	--	--
SO2	--	--	--	--	--	--
CH3OH	--	--	233	233	--	--
Total Flow kmol/hr	25329	28683	23917	23917	20000	15300
Total Flow kg/hr	667008	727579	609384	609384	360306	275634
Total Flow l/min	196898	361623	175585	223816	7972	6099
Vapor Frac	1.00	1.00	1.00	1.00	0.00	0.00
Liquid Frac	0.00	0.00	0.00	0.00	1.00	1.00

Table B-1
Power Generation from UCG
Stream Table

Stream Name	CLAUSFD	CLAUSOXY	COLDFLUE	COMBAIR1	COMBL1	COMBPRD
Temperature K	270	298	457	298	273	1707
Pressure atm	2	2	1.1	1	45	45
Component Mole Flow						
<i>kmol/hr</i>						
H2	15	--	--	--	--	--
CO	19	--	--	--	--	--
H2O	Negl.	--	6704	--	--	6704
CO2	1181	--	7402	--	--	7402
CH4	36	--	--	--	--	--
N2	69	--	46020	32836	--	46020
C2H6	19	--	--	--	--	--
H2S	27	--	--	--	--	--
NH3	Negl.	--	--	--	--	--
O2	36	36	2277	8729	--	2277
SO2	--	--	Negl.	--	--	Negl.
CH3OH	11	--	--	--	11000	--
Component Mass Flow						
<i>kg/hr</i>						
H2	31	--	--	--	--	--
CO	532	--	--	--	--	--
H2O	Negl.	--	120769	--	--	120769
CO2	51964	--	325743	--	--	325743
CH4	578	--	--	--	--	--
N2	1922	--	1289170	919863	--	1289170
C2H6	570	--	--	--	--	--
H2S	907	--	--	--	--	--
NH3	3	--	--	--	--	--
O2	1167	1167	72861	279307	--	72861
SO2	--	--	14	--	--	14
CH3OH	358	--	--	--	352464	--
Total Flow kmol/hr	1413	36	62402	41565	11000	62402
Total Flow kg/hr	58031	1167	1808550	1199170	352464	1808550
Total Flow l/min	257841	7430	35488300	16949500	9602	3252330
Vapor Frac	1.00	1.00	1.00	1.00	0.00	1.00
Liquid Frac	0.00	0.00	0.00	0.00	1.00	0.00

Table B-1
Power Generation from UCG
Stream Table

Stream Name	COMPAIR1	COMPAIR2	COMPAIR3	COND2	COOLAIR1	COOLAIR2
Temperature K	439	453	453	442	308	308
Pressure atm	3.56	12.6736	45	7.515	3.56	12.6736
Component Mole Flow						
<i>kmol/hr</i>						
H2	--	--	--	--	--	--
CO	--	--	--	--	--	--
H2O	--	--	--	3647	--	--
CO2	--	--	--	--	--	--
CH4	--	--	--	--	--	--
N2	32836	32836	32836	--	32836	32836
C2H6	--	--	--	--	--	--
H2S	--	--	--	--	--	--
NH3	--	--	--	--	--	--
O2	8729	8729	8729	--	8729	8729
SO2	--	--	--	--	--	--
CH3OH	--	--	--	--	--	--
Component Mass Flow						
<i>kg/hr</i>						
H2	--	--	--	--	--	--
CO	--	--	--	--	--	--
H2O	--	--	--	65703	--	--
CO2	--	--	--	--	--	--
CH4	--	--	--	--	--	--
N2	919863	919863	919863	--	919863	919863
C2H6	--	--	--	--	--	--
H2S	--	--	--	--	--	--
NH3	--	--	--	--	--	--
O2	279307	279307	279307	--	279307	279307
SO2	--	--	--	--	--	--
CH3OH	--	--	--	--	--	--
Total Flow kmol/hr	41565	41565	41565	3647	41565	41565
Total Flow kg/hr	1199170	1199170	1199170	65703	1199170	1199170
Total Flow l/min	7016160	2045820	585690	1670	4923190	1385230
Vapor Frac	1.00	1.00	1.00	0.00	1.00	1.00
Liquid Frac	0.00	0.00	0.00	1.00	0.00	0.00

Table B-1
Power Generation from UCG
Stream Table

Stream Name	CW1	HOTFLUE	INSTM2	INTSTM	L1CL	L1CR	L1CRYOI
Temperature K	298	827	380	442	228	259	193
Pressure atm	1	1.1	1.255005	7.515	45	44.7	10
Component Mole Flow							
<i>kmol/hr</i>							
H2	--	--	--	--	--	15	--
CO	--	--	--	--	--	19	--
H2O	3825000	6704	29653	33300	--	4	--
CO2	--	7402	--	--	--	1182	--
CH4	--	--	--	--	--	36	--
N2	--	46020	--	--	--	69	--
C2H6	--	--	--	--	--	19	--
H2S	--	--	--	--	--	50	--
NH3	--	--	--	--	--	26	--
O2	--	2277	--	--	--	--	--
SO2	--	Negl.	--	--	--	--	--
CH3OH	--	--	--	--	11000	10993	100000
Component Mass Flow							
<i>kg/hr</i>							
H2	--	--	--	--	--	31	--
CO	--	--	--	--	--	532	--
H2O	68908500	120769	534205	599909	--	68	--
CO2	--	325743	--	--	--	52003	--
CH4	--	--	--	--	--	578	--
N2	--	1289170	--	--	--	1922	--
C2H6	--	--	--	--	--	570	--
H2S	--	--	--	--	--	1710	--
NH3	--	--	--	--	--	443	--
O2	--	72861	--	--	--	--	--
SO2	--	14	--	--	--	--	--
CH3OH	--	--	--	--	352464	352231	3204220
Total Flow kmol/hr	3825000	62402	29653	33300	11000	12412	100000
Total Flow kg/hr	68908500	1808550	534205	599909	352464	410088	3204220
Total Flow l/min	1525840	64156500	11125300	2287930	9254	10403	82243
Vapor Frac	0.00	1.00	0.92	0.89	0.00	0.00	0.00
Liquid Frac	1.00	0.00	0.08	0.11	1.00	1.00	1.00

**Table B-1
Power Generation from UCG
Stream Table**

Stream Name	L1CRYOO	L1HL	L1HR	L1WL	LPSTM	MAKEUPL1	MPSTEAM1
Temperature K	198	383	295	274	442	318	531
Pressure atm	10	5	44.7	4.946598	7.515	45	45
Component Mole Flow							
<i>kmol/hr</i>							
H2	--	--	15	--	--	--	--
CO	--	--	--	--	--	--	--
H2O	--	4	4	4	29653	--	2000
CO2	--	Negl.	1182	Negl.	--	--	--
CH4	--	Negl.	36	Negl.	--	--	--
N2	--	Negl.	69	Negl.	--	--	--
C2H6	--	Negl.	19	Negl.	--	--	--
H2S	--	24	50	24	--	--	--
NH3	--	26	26	26	--	--	--
O2	--	--	--	--	--	--	--
SO2	--	--	--	--	--	--	--
CH3OH	100000	10967	10993	10967	--	38	--
Component Mass Flow							
<i>kg/hr</i>							
H2	--	--	31	--	--	--	--
CO	--	--	532	--	--	--	--
H2O	--	68	68	68	534205	--	36031
CO2	--	5	52003	5	--	--	--
CH4	--	Negl.	578	Negl.	--	--	--
N2	--	Negl.	1922	Negl.	--	--	--
C2H6	--	Negl.	570	Negl.	--	--	--
H2S	--	801	1710	801	--	--	--
NH3	--	439	443	439	--	--	--
O2	--	--	--	--	--	--	--
SO2	--	--	--	--	--	--	--
CH3OH	3204220	351401	352231	351401	--	1218	--
Total Flow kmol/hr	100000	11020	12412	11020	29653	38	2000
Total Flow kg/hr	3204220	352714	410088	352714	534205	1218	36031
Total Flow l/min	82492	11220	11032	9628	2286260	35	27276
Vapor Frac	0.00	0.00	0.00	0.00	1.00	0.00	1.00
Liquid Frac	1.00	1.00	1.00	1.00	0.00	1.00	0.00

Table B-1
Power Generation from UCG
Stream Table

Stream Name	MPSTEAM2	PWRSTM	RAWACIDG	RAWSYNDR	STM4TRB	SYNCOND
Temperature K	531	534	287	873	536	263
Pressure atm	45	45	5	45	45	44.7
Component Mole Flow						
<i>kmol/hr</i>						
H2	--	--	15	3467	--	--
CO	--	--	19	2855	--	--
H2O	18000	33300	--	12747	15300	68
CO2	--	--	1182	4079	--	1
CH4	--	--	36	1479	--	Negl.
N2	--	--	69	13256	--	Negl.
C2H6	--	--	19	138	--	Negl.
H2S	--	--	27	51	--	Negl.
NH3	--	--	Negl.	178	--	10
O2	--	--	--	--	--	--
SO2	--	--	--	--	--	--
CH3OH	--	--	26	--	--	--
Component Mass Flow						
<i>kg/hr</i>						
H2	--	--	31	6989	--	Negl.
CO	--	--	532	79976	--	1
H2O	324275	599909	Negl.	229634	275634	1224
CO2	--	--	51998	179512	--	42
CH4	--	--	578	23721	--	Negl.
N2	--	--	1922	371360	--	1
C2H6	--	--	570	4139	--	Negl.
H2S	--	--	909	1738	--	1
NH3	--	--	4	3039	--	168
O2	--	--	--	--	--	--
SO2	--	--	--	--	--	--
CH3OH	--	--	830	--	--	--
Total Flow kmol/hr	18000	33300	1392	38250	15300	79
Total Flow kg/hr	324275	599909	57374	900108	275634	1439
Total Flow l/min	245483	457450	106307	1012530	211963	32
Vapor Frac	1.00	1.00	1.00	1.00	1.00	0.00
Liquid Frac	0.00	0.00	0.00	0.00	0.00	1.00

**Table B-1
Power Generation from UCG
Stream Table**

Stream Name	SYNMHOT	SYNNOASH	WASTEWTR	WGHPDC	WGSPDC	WW1
Temperature K	418	873	418	313	263	299
Pressure atm	44.95	45	44.9	44.8	44.75	1
Component Mole Flow						
<i>kmol/hr</i>						
H2	3467	3467	1	3465	3465	--
CO	2855	2855	1	2854	2854	--
H2O	12747	12747	9508	72	72	3825000
CO2	4079	4079	7	4063	4063	--
CH4	1479	1479	Negl.	1478	1478	--
N2	13256	13256	4	13252	13252	--
C2H6	138	138	Negl.	137	137	--
H2S	51	51	Negl.	50	50	--
NH3	178	178	45	36	36	--
O2	--	--	--	--	--	--
SO2	--	--	--	--	--	--
CH3OH	--	--	--	--	--	--
Component Mass Flow						
<i>kg/hr</i>						
H2	6989	6989	3	6986	6986	--
CO	79976	79976	22	79945	79945	--
H2O	229634	229634	171282	1292	1292	68908500
CO2	179512	179512	320	178830	178830	--
CH4	23721	23721	6	23712	23712	--
N2	371360	371360	118	371226	371226	--
C2H6	4139	4139	9	4125	4125	--
H2S	1738	1738	10	1719	1719	--
NH3	3039	3039	760	611	611	--
O2	--	--	--	--	--	--
SO2	--	--	--	--	--	--
CH3OH	--	--	--	--	--	--
Total Flow kmol/hr	38250	38250	9567	25408	25408	3825000
Total Flow kg/hr	900108	900108	172529	668447	668447	68908500
Total Flow l/min	365490	1012530	4258	241129	196718	1526690
Vapor Frac	0.75	1.00	0.00	1.00	1.00	0.00
Liquid Frac	0.25	0.00	1.00	0.00	0.00	1.00

Table B.2
Power Generation from UCG
Annual Estimated Variable Costs

Costs in 2011 US\$

<i>Materials Consumed</i>	<i>Number/year</i>	<i>Unit Cost</i>	<i>Costs</i>
Coal Royalty Costs	1,750,000 tonnes	\$3 per tonne	\$ 5,250,000
Estimated Tar By-product	7,000 tonnes	\$ (40.00) per tonne	\$ (280,000)
Methanol Losses	9590 tonnes	\$ 300 per tonne	<u>\$ 2,877,000</u>
		Raw Material Costs	\$ 8,127,000
	<i>Subtotal, Drilling Contractor Turnkey Costs</i>		\$ 11,004,000
<i>Utilities</i>			
Cooling Water	3.30E+08 cu.m	\$ 0.02 per cu.m	\$ 6,600,000
Boiler Feed Water	8.29E+05 cu.m	\$ 0.20 per cu.m	<u>\$ 166,000</u>
		Utility Costs	\$ 6,766,000
		Total Variable costs	\$ 14,893,000

Table B.3
Power Generation from UCG
Annual Estimated Drilling & Field Costs

Costs in 2011 thousand US\$

	<i>Number/run</i>	<i>Number/year</i>	<i>Unit Cost</i>	<i>Costs</i>
Production Wells	1	33 wells	\$ 374 per well	\$ 12,333
Injection Wells	3	99 wells	\$ 173 per well	\$ 17,160
Instrumentation Wells	4	132 wells	\$ 25 per well	\$ 3,300
Instrument Costs	4	64 wells	\$ 10 per well	\$ 640
Drill Waste Disposal	325	10725 tonne	\$ 0.05 per tonne	<u>\$ 536</u>
			<i>Subtotal, Drilling Contractor Turnkey Costs</i>	\$ 33,969
Drilling Program Contingency	30%			\$ 10,191
Direct Employees for Oversight of Drilling Contract		1.5 employees	\$ 61 each	<u>\$ 90</u>
			Total drilling costs	\$ 44,249
 Site Preparation Costs				
	<i>Number/run</i>	<i>Number/year</i>	<i>Unit cost</i>	
Land Lease Costs for Extraction	0.1	3.3 hectares	1.75 per hectare	\$ 6
Site Clearing and Preparation	0.1	3.3 hectares	4.5 per hectare	\$ 15
Utility Road Construction	0.4	13.2 km	8 per km	\$ 106
Field Piping & Installation	0.6	19.8 km	125 per km	<u>\$ 2,475</u>
			Site Preparation Costs	\$ 2,601
 UCG Field Operation and Maintenance				
		Number/year		
Decommissioning of spent wells		33	10 each	\$ 330
Field Piping Maintenance				\$ 500
Monitoring Well Sampling		80	1.5 per sample	\$ 120
Environmental Reporting		2	20 each	<u>\$ 40</u>
			Field Operation Costs	\$ 990
Total Annual UCG Field Operation Costs				\$ 47,840

Table B.4
Power Generation from UCG
Fixed Capital Costs

Plant Net Capacity				485 MW
<i>Costs in 2011 thousand US\$</i>				
Gas Cleanup and Power Plant				
Battery Limits Investment (BLI)		Equipment Cost	Installation Cost	Total Cost
Power Plant Package Unit				\$ 550,000
Compressors		\$ 10,000	\$ 9,000	\$ 19,000
Reactors		\$ 50	\$ 90	\$ 140
Columns		\$ 2,520	\$ 3,780	\$ 6,300
Pressure Vessels		\$ 1,310	\$ 1,570	\$ 2,880
Heat Exchangers		\$ 4,070	\$ 5,900	\$ 9,970
Claus Package Unit				\$ 8,570
Particulate Removal		\$ 730	\$ 510	\$ 1,240
			<i>Subtotal</i>	\$ 598,100
BLI Contingency	30%	of Installed Equipment Costs		\$ 179,430
Battery Limits Investment				\$ 777,530
Battery Limits Investment, Gas Cleanup Only				\$ 37,830
Tankage				
Methanol Storage Tanks				\$ 13,200
Methanol Surge Tanks				\$ 2,600
				\$ 15,800
Utilities				
Refrigeration	12 MW	Purchased Cost	Installation Cost	Investment
Boiler Feed Water	105 tonnes/hr	\$ 12,700	\$ 5,080	\$ 17,780
Cooling Water	42,000 tonnes/hr	\$ 400	\$ 160	\$ 600
		\$ 8,000	\$ 3,200	\$ 11,200
			<i>Utilities Investment Subtotal</i>	\$ 29,600
Offsite & Utility Investment Contingency	30%			\$ 13,620
Offsite & Utilities Investment				\$ 59,020
General Service Facilities	5%	of BLI & Utilities Investment		\$ 41,830
Waste Treatment	1%	of BLI Investment		\$ 7,780
Outside Battery Limits Investment				\$ 108,600
			<i>Total Fixed Capital (TFC) Investment</i>	\$ 886,100

Table B.5
Power Generation from UCG
Annual Estimated Operating Costs

Plant Net Capacity			485 MW
<i>Costs in 2011 thousand US\$</i>			
			<i>Costs</i>
Plant Investment, Battery Limits (BLI)			\$ 777,530
Plant Investment, Outside Battery Limits (OBLI)			<u>\$ 108,600</u>
Total Fixed Capital (TFC)			\$ 886,130
Operating Costs, Per Year			
Raw Material Costs			\$ 8,127
Utility Costs			<u>\$ 6,766</u>
	Variable Costs		\$ 14,893
Estimated Annual Drilling Costs			\$ 47,840
Operating Labor			
		Number/year	Unit Cost
Gas Cleanup and Power Generation Personnel (3 shifts)		18	\$ 62 wages/year
UCG Field Operations Personnel (3 shifts)		33	\$ 62 wages/year
Maintenance Labor	\$1.20 per GWh	3880 GWh	\$ 4,660
Control Laboratory Labor, 10% of Operating Labor	10% of Operating Labor		\$ 320
Direct Labor Costs			\$ 8,140
Maintenance Materials	\$1.80 per GWh	3880 GWh	\$ 6,980
Operating Supplies, 12% of Operating Labor	12% of Operating Labor		\$ 380
Total Direct Costs			\$ 86,380
Plant Overhead	80% of Labor Costs		\$ 6,820
Taxes and Insurance	1.60% of TFC		\$ 14,180
Cash Costs			\$ 107,380
Depreciation	14.3% of TFC		\$ 126,590
Gate Costs			\$ 233,970
General, Admin, Sales, Research	5% of Gate Costs		\$ 11,700
Production Costs			\$ 245,670
TFC + Estimated Working Capital			\$ 912,975
ROI	12.5% of Capital Investment		\$ 114,120
Production Cost + Cost of Capital			\$ 359,790
<i>Production Cost + Cost of Capital without Depreciation Charge</i>			\$ 233,200
Nominal Net Capacity, MW			485
Stream Factor			0.913
Estimated Annual Energy Output, MWh			3,880,000
Cost including Capital Return per MWh, \$			\$ 93
Cost excluding Depreciation but including Capital Return per MWh, \$			\$ 60

Appendix C

Table C.1
Synthetic Natural Gas from UCG
Stream Table

	BALSYN1	BALSYN2	BALSYNC	BALSYNH	BFW1
Temperature K	727	235	259	525	298
Pressure atm	50	45	45	45	45
Component Mole Flow					
<i>kmol/hr</i>					
H2	17,072	16,594	16,940	16,933	-
CO	5,234	4,946	5,122	5,116	-
H2O	27,242	Negl.	Negl.	Negl.	45,100
CO2	20,549	120	6,967	2,136	-
CH4	5,055	4,116	4,736	4,651	-
N2	901	866	889	889	-
C2H6	451	53	291	161	-
H2S	150	1	3	2	-
NH3	601	Negl.	Negl.	Negl.	-
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	-	3	21	8	-
Component Mass Flow					
<i>kg/hr</i>					
H2	34,415	33,451	34,149	34,134	-
CO	146,617	138,526	143,470	143,298	-
H2O	490,773	Negl.	Negl.	Negl.	812,489
CO2	904,367	5,280	306,595	94,021	-
CH4	81,089	66,034	75,971	74,610	-
N2	25,247	24,272	24,910	24,893	-
C2H6	13,550	1,584	8,735	4,852	-
H2S	5,119	33	115	77	-
NH3	10,233	Negl.	Negl.	Negl.	-
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	-	82	689	248	-
Total Flow kmol/hr	77,255	26,698	34,969	29,896	45,100
Total Flow kg/hr	1,711,410	269,261	594,634	376,133	812,489
Total Flow l/min	1,519,300	194,573	271,313	487,237	17,978
Vapor Frac	1.00	1.00	1.00	1.00	0.00
Liquid Frac	0.00	0.00	0.00	0.00	1.00

Table C.1
Synthetic Natural Gas from UCG
Stream Table

	BFW2	CH4COOL	CLAUSFD	CLAUSFD2	CLAUSOXY
Temperature K	298	313	285	243	298
Pressure atm	2	44.9	2	2	2
Component Mole Flow					
<i>kmol/hr</i>					
H2	-	98	132	132	-
CO	-	1	112	112	-
H2O	30,000	5,860	Negl.	Negl.	-
CO2	-	1,764	13,258	13,237	-
CH4	-	10,138	319	319	-
N2	-	889	12	12	-
C2H6	-	161	158	158	-
H2S	-	2	135	134	-
NH3	-	Negl.	Negl.	Negl.	-
O2	-	-	1	1	1
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	-	8	417	30	-
Component Mass Flow					
<i>kg/hr</i>					
H2	-	198	266	266	-
CO	-	22	3,148	3,148	-
H2O	540,458	105,563	9	Negl.	-
CO2	-	77,639	583,483	582,574	-
CH4	-	162,643	5,118	5,117	-
N2	-	24,893	337	337	-
C2H6	-	4,852	4,744	4,741	-
H2S	-	77	4,591	4,562	-
NH3	-	Negl.	7	1	-
O2	-	-	46	46	46
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	-	248	13,354	963	-
Total Flow kmol/hr	30,000	18,921	14,545	14,136	1
Total Flow kg/hr	540,458	376,133	615,102	601,755	46
Total Flow l/min	11,967	120,298	2,796,750	2,305,740	291
Vapor Frac	0.00	0.69	1.00	1.00	1.00
Liquid Frac	1.00	0.31	0.00	0.00	0.00

Table C.1
Synthetic Natural Gas from UCG
Stream Table

	CLRSYN	CO24CCS	COLDSYN	CW1	FRSNGRCT
Temperature K	250	245	277	293	264
Pressure atm	45	1	45	1	25
Component Mole Flow					
<i>kmol/hr</i>					
H2	17,091	8	17,091	-	1
CO	5,205	6	5,205	-	Negl.
H2O	-	Negl.	-	7,500,000	105
CO2	20,532	5,771	20,532	-	904
CH4	5,053	468	5,053	-	383
N2	901	8	901	-	7
C2H6	449	166	449	-	36
H2S	149	8	149	-	4
NH3	-	Negl.	-	-	5
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	-	32	-	-	16,947
Component Mass Flow					
<i>kg/hr</i>					
H2	34,453	16	34,453	-	1
CO	145,794	171	145,794	-	Negl.
H2O	-	Negl.	-	135,115,000	1,900
CO2	903,609	253,968	903,609	-	39,788
CH4	81,064	7,512	81,064	-	6,151
N2	25,240	223	25,240	-	205
C2H6	13,501	4,979	13,501	-	1,096
H2S	5,078	269	5,078	-	137
NH3	-	Negl.	-	-	86
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	-	1,038	-	-	543,031
Total Flow kmol/hr	49,380	6,467	49,380	7,500,000	18,394
Total Flow kg/hr	1,208,740	268,177	1,208,740	135,115,000	592,396
Total Flow l/min	337,989	2,148,080	388,911	2,981,560	15,634
Vapor Frac	1.00	1.00	1.00	0.00	0.00
Liquid Frac	0.00	0.00	0.00	1.00	1.00

Table C.1
Synthetic Natural Gas from UCG
Stream Table

	HPPRDSNG	L1CO2RCH	L1COLD	L1CR	L1HL
Temperature K	244	248	233	241	294
Pressure atm	25	2	45	45	3
Component Mole Flow					
<i>kmol/hr</i>					
H2	98	8	-	132	Negl.
CO	1	6	-	112	Negl.
H2O	Negl.	393	430	72	71
CO2	898	5,936	302	13,633	375
CH4	9,751	468	-	319	Negl.
N2	881	8	-	12	Negl.
C2H6	124	166	-	160	2
H2S	1	17	17	150	15
NH3	Negl.	26	31	5	5
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	4	86,439	103,798	17,348	16,931
Component Mass Flow					
<i>kg/hr</i>					
H2	197	16	-	266	Negl.
CO	22	171	-	3,148	Negl.
H2O	Negl.	7,086	7,747	1,296	1,287
CO2	39,516	261,258	13,291	599,996	16,513
CH4	156,439	7,512	-	5,118	1
N2	24,683	223	-	337	Negl.
C2H6	3,739	4,979	-	4,815	71
H2S	33	563	579	5,102	511
NH3	Negl.	440	528	88	81
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	120	2,769,690	3,325,910	555,867	542,513
Total Flow kmol/hr	11,758	93,467	104,578	31,944	17,400
Total Flow kg/hr	224,748	3,051,940	3,348,060	1,176,030	560,976
Total Flow l/min	143,786	604,271	88,021	24,168	15,423
Vapor Frac	1.00	0.03	0.00	0.00	0.00
Liquid Frac	0.00	0.97	1.00	1.00	1.00

Table C.1
Synthetic Natural Gas from UCG
Stream Table

	L1HR	L1MAKEUP	L1RECY1	L1RECY2	L1TOCABS
Temperature K	298	298	308	308	233
Pressure atm	5	50	5	50	45
Component Mole Flow					
<i>kmol/hr</i>					
H2	132	-	Negl.	-	-
CO	112	-	Negl.	-	-
H2O	72	-	465	430	288
CO2	13,633	-	541	302	202
CH4	319	-	Negl.	-	-
N2	12	-	Negl.	-	-
C2H6	160	-	2	-	-
H2S	150	-	24	17	11
NH3	5	-	31	31	21
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	17,348	58	103,338	103,740	69,478
Component Mass Flow					
<i>kg/hr</i>					
H2	266	-	Negl.	-	-
CO	3,148	-	Negl.	-	-
H2O	1,296	-	8,372	7,747	5,185
CO2	599,996	-	23,804	13,291	8,896
CH4	5,118	-	1	-	-
N2	337	-	Negl.	-	-
C2H6	4,815	-	71	-	-
H2S	5,102	-	805	579	388
NH3	88	-	520	528	353
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	555,867	1,858	3,311,170	3,324,050	2,226,220
Total Flow kmol/hr	31,944	58	104,400	104,520	70,000
Total Flow kg/hr	1,176,030	1,858	3,344,740	3,346,200	2,241,050
Total Flow l/min	1,162,050	52	94,412	94,227	58,917
Vapor Frac	0.45	0.00	0.00	0.00	0.00
Liquid Frac	0.55	1.00	1.00	1.00	1.00

Table C.1
Synthetic Natural Gas from UCG
Stream Table

	L1TOCST2	L1TOSABS	L1TOSNGR	L1W2CRYO	L1WL
Temperature K	250	233	233	308	308
Pressure atm	5	45	45	50	2
Component Mole Flow					
<i>kmol/hr</i>					
H2	7	-	-	-	Negl.
CO	6	-	-	-	Negl.
H2O	288	72	70	430	71
CO2	5,032	51	49	302	375
CH4	85	-	-	-	Negl.
N2	1	-	-	-	Negl.
C2H6	129	-	-	-	2
H2S	12	3	3	17	15
NH3	21	5	5	31	5
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	69,492	17,369	16,951	103,798	16,931
Component Mass Flow					
<i>kg/hr</i>					
H2	14	-	-	-	Negl.
CO	171	-	-	-	Negl.
H2O	5,186	1,296	1,265	7,747	1,287
CO2	221,470	2,224	2,170	13,291	16,513
CH4	1,361	-	-	-	1
N2	18	-	-	-	Negl.
C2H6	3,883	-	-	-	71
H2S	426	97	95	579	511
NH3	354	88	86	528	81
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	2,226,660	556,556	543,135	3,325,910	542,513
Total Flow kmol/hr	75,073	17,500	17,078	104,578	17,400
Total Flow kg/hr	2,459,550	560,261	546,751	3,348,060	560,976
Total Flow l/min	63,092	14,729	14,374	94,279	66,300
Vapor Frac	0.00	0.00	0.00	0.00	0.01
Liquid Frac	1.00	1.00	1.00	1.00	0.99

Table C.1
Synthetic Natural Gas from UCG
Stream Table

	L12CO2ST	LEAN1COL	LEANL1	LTRFLSH	METHH2O
Temperature K	253	308	325	250	313
Pressure atm	45	1	1	5	45
Component Mole Flow					
<i>kmol/hr</i>					
H2	346	Negl.	Negl.	339	Negl.
CO	176	Negl.	Negl.	170	Negl.
H2O	288	393	393	Negl.	5,824
CO2	7,049	166	166	2,016	11
CH4	619	Negl.	Negl.	535	3
N2	23	Negl.	Negl.	22	Negl.
C2H6	238	Negl.	Negl.	109	1
H2S	14	9	9	1	Negl.
NH3	21	26	26	Negl.	Negl.
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	69,497	86,407	86,407	5	7
Component Mass Flow					
<i>kg/hr</i>					
H2	698	Negl.	Negl.	683	Negl.
CO	4,944	Negl.	Negl.	4,773	Negl.
H2O	5,186	7,085	7,085	Negl.	104,927
CO2	310,211	7,290	7,290	88,741	506
CH4	9,937	Negl.	Negl.	8,576	53
N2	638	Negl.	Negl.	621	4
C2H6	7,152	Negl.	Negl.	3,269	17
H2S	470	294	294	44	1
NH3	354	439	439	Negl.	Negl.
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	2,226,830	2,768,660	2,768,660	166	232
Total Flow kmol/hr	78,271	87,000	87,000	3,198	5,847
Total Flow kg/hr	2,566,420	2,783,760	2,783,760	106,872	105,740
Total Flow l/min	65,224	78,724	80,349	212,998	2,363
Vapor Frac	0.00	0.00	0.00	1.00	0.00
Liquid Frac	1.00	1.00	1.00	0.00	1.00

Table C.1
Synthetic Natural Gas from UCG
Stream Table

	METHREC	MPSTEAM0	MPSTEAM1	MPSTEAM2	MPSTEAM3	RAWACIDG
Temperature K	243	398	398	398	398	286
Pressure atm	2	45	2	45	45	3
Component Mole Flow						
<i>kmol/hr</i>						
H2	Negl.	-	-	-	-	132
CO	Negl.	-	-	-	-	112
H2O	Negl.	45,100	30,000	42,845	2,255	Negl.
CO2	21	-	-	-	-	13,258
CH4	Negl.	-	-	-	-	319
N2	Negl.	-	-	-	-	12
C2H6	Negl.	-	-	-	-	158
H2S	1	-	-	-	-	135
NH3	Negl.	-	-	-	-	Negl.
O2	Negl.	-	-	-	-	-
SO2	-	-	-	-	-	-
S2	-	-	-	-	-	-
S8	-	-	-	-	-	-
CH3OH	387	-	-	-	-	417
Component Mass Flow						
<i>kg/hr</i>						
H2	Negl.	-	-	-	-	266
CO	Negl.	-	-	-	-	3,148
H2O	9	812,489	540,458	771,865	40,624	9
CO2	909	-	-	-	-	583,483
CH4	Negl.	-	-	-	-	5,118
N2	Negl.	-	-	-	-	337
C2H6	3	-	-	-	-	4,744
H2S	29	-	-	-	-	4,591
NH3	6	-	-	-	-	7
O2	Negl.	-	-	-	-	-
SO2	-	-	-	-	-	-
S2	-	-	-	-	-	-
S8	-	-	-	-	-	-
CH3OH	12,391	-	-	-	-	13,354
Total Flow kmol/hr	409	45,100	30,000	42,845	2,255	14,544
Total Flow kg/hr	13,348	812,489	540,458	771,865	40,624	615,057
Total Flow l/min	344	19,575	8,052,400	18,596	979	1,861,310
Vapor Frac	0.00	0.00	1.00	0.00	0.00	1.00
Liquid Frac	1.00	1.00	0.00	1.00	1.00	0.00

Table C.1
Synthetic Natural Gas from UCG
Stream Table

	RAWC1	RAWPRD	RAWSYN	SYN2SABS	SYNCOND1
Temperature K	575	313	873	233	
Pressure atm	44.9	45	50	45	50
Component Mole Flow					
<i>kmol/hr</i>					
H2	98	98	17,049	17,072	-
CO	1	1	5,257	5,234	-
H2O	5,860	35	25,010	-	-
CO2	1,764	1,753	20,526	20,549	-
CH4	10,138	10,135	5,055	5,055	-
N2	889	888	901	901	-
C2H6	161	161	451	451	-
H2S	2	2	150	150	-
NH3	Negl.	Negl.	601	-	-
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	8	Negl.	-	-	-
Component Mass Flow					
<i>kg/hr</i>					
H2	198	198	34,368	34,415	-
CO	22	22	147,261	146,617	-
H2O	105,563	635	450,562	-	-
CO2	77,639	77,133	903,356	904,367	-
CH4	162,643	162,590	81,089	81,089	-
N2	24,893	24,889	25,247	25,247	-
C2H6	4,852	4,835	13,550	13,550	-
H2S	77	76	5,119	5,119	-
NH3	Negl.	Negl.	10,233	-	-
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	248	16	-	-	-
Total Flow kmol/hr	18,921	13,074	75,000	49,412	-
Total Flow kg/hr	376,133	270,393	1,670,790	1,210,410	-
Total Flow l/min	328,661	117,671	1,774,160	264,514	-
Vapor Frac	1.00	1.00	1.00	0.79	
Liquid Frac	0.00	0.00	0.00	0.21	

Table C.1
Synthetic Natural Gas from UCG
Stream Table

	SYNCOND2	SYNGAS2	SYNMHOT	SYNMHT1	SYNNOASH
Temperature K	298	727	760	760	873
Pressure atm	50	45	50	50	50
Component Mole Flow					
<i>kmol/hr</i>					
H2	-	17,049	17,049	17,049	17,049
CO	-	5,257	5,257	5,257	5,257
H2O	27,242	27,265	25,010	25,010	25,010
CO2	-	20,526	20,526	20,526	20,526
CH4	-	5,055	5,055	5,055	5,055
N2	-	901	901	901	901
C2H6	-	451	451	451	451
H2S	-	150	150	150	150
NH3	601	601	601	601	601
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	-	-	-	-	-
Component Mass Flow					
<i>kg/hr</i>					
H2	-	34,368	34,368	34,368	34,368
CO	-	147,261	147,261	147,261	147,261
H2O	490,773	491,187	450,562	450,562	450,562
CO2	-	903,356	903,356	903,356	903,356
CH4	-	81,089	81,089	81,089	81,089
N2	-	25,247	25,247	25,247	25,247
C2H6	-	13,550	13,550	13,550	13,550
H2S	-	5,119	5,119	5,119	5,119
NH3	10,233	10,233	10,233	10,233	10,233
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	-	-	-	-	-
Total Flow kmol/hr	27,843	77,255	75,000	75,000	75,000
Total Flow kg/hr	501,006	1,711,410	1,670,790	1,670,790	1,670,790
Total Flow l/min	11,162	1,688,230	1,543,640	1,543,640	1,774,160
Vapor Frac	0.00	1.00	1.00	1.00	1.00
Liquid Frac	1.00	0.00	0.00	0.00	0.00

Table C.1
Synthetic Natural Gas from UCG
Stream Table

	SYNTOCRY	TOC1RXR	TORXR	WGHPRDC1	WGHPRDC2
Temperature K	233	240	293	298	298
Pressure atm	45	45	45	50	50
Component Mole Flow					
<i>kmol/hr</i>					
H2	17,091	16,933	16,860	17,072	17,072
CO	5,205	5,116	5,049	5,234	5,234
H2O	-	Negl.	-	27,242	-
CO2	20,532	2,136	831	20,549	20,549
CH4	5,053	4,651	4,460	5,055	5,055
N2	901	889	883	901	901
C2H6	449	161	96	451	451
H2S	149	2	2	150	150
NH3	-	Negl.	-	601	-
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	-	8	4	-	-
Component Mass Flow					
<i>kg/hr</i>					
H2	34,453	34,134	33,988	34,415	34,415
CO	145,794	143,298	141,425	146,617	146,617
H2O	-	Negl.	-	490,773	-
CO2	903,609	94,021	36,572	904,367	904,367
CH4	81,064	74,610	71,551	81,089	81,089
N2	25,240	24,893	24,736	25,247	25,247
C2H6	13,501	4,852	2,887	13,550	13,550
H2S	5,078	77	68	5,119	5,119
NH3	-	Negl.	-	10,233	-
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	-	248	128	-	-
Total Flow kmol/hr	49,380	29,896	28,185	77,255	49,412
Total Flow kg/hr	1,208,740	376,133	311,354	1,711,410	1,210,410
Total Flow l/min	263,827	219,170	258,037	392,356	382,645
Vapor Frac	0.79	1.00	1.00	0.64	1.00
Liquid Frac	0.21	0.00	0.00	0.36	0.00

Table C.1
Synthetic Natural Gas from UCG
Stream Table

	WGSBYPAS	WGSFEED	WGSPRDH	WGSPRDM	WW1
Temperature K	727	727	753	446	295
Pressure atm	45	45	50	50	1
Component Mole Flow					
<i>kmol/hr</i>					
H2	16,878	170	193	17,072	-
CO	5,205	53	30	5,234	-
H2O	26,992	273	250	27,242	7,500,000
CO2	20,321	205	228	20,549	-
CH4	5,004	51	51	5,055	-
N2	892	9	9	901	-
C2H6	446	5	5	451	-
H2S	149	2	2	150	-
NH3	595	6	6	601	-
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	-	-	-	-	-
Component Mass Flow					
<i>kg/hr</i>					
H2	34,025	344	390	34,415	-
CO	145,788	1,473	829	146,617	-
H2O	486,275	4,912	4,498	490,773	135,115,000
CO2	894,322	9,034	10,045	904,367	-
CH4	80,278	811	811	81,089	-
N2	24,995	252	252	25,247	-
C2H6	13,415	136	136	13,550	-
H2S	5,068	51	51	5,119	-
NH3	10,130	102	102	10,233	-
O2	-	-	-	-	-
SO2	-	-	-	-	-
S2	-	-	-	-	-
S8	-	-	-	-	-
CH3OH	-	-	-	-	-
Total Flow kmol/hr	76,482	773	773	77,255	7,500,000
Total Flow kg/hr	1,694,300	17,114	17,114	1,711,410	135,115,000
Total Flow l/min	1,671,350	16,882	15,750	727,843	2,985,260
Vapor Frac	1.00	1.00	1.00	0.81	0.00
Liquid Frac	0.00	0.00	0.00	0.19	1.00

Table C.2
Synthetic Natural Gas (SNG) from UCG
Annual Estimated Variable Costs

Costs in 2011 US\$

<i>Materials Consumed</i>	<i>Number/year</i>	<i>Unit Cost</i>	<i>Costs</i>
Coal Royalty Costs	6,100,000 tonnes	\$3 per tonne	\$ 18,300,000
Oxygen Purchase	3,215,100 tonnes	\$30 per tonne	\$ 96,453,000
Estimated Tar By-product	45,000 tonnes	\$ (40.00) per tonne	\$ (1,800,000)
Catalyst Losses			\$ 700,000
Methanol Losses	23,990 tonnes	\$ 300 per tonne	<u>\$ 7,197,000</u>
		Raw Material Costs	\$ 120,850,000
	<i>Subtotal, Raw Material Costs</i>		\$ 120,850,000
<i>Utilities</i>			
Steam Export	1.35E+07 tonnes	\$ (2.00) per tonne	\$ (26,991,051)
Cooling Water	1.56E+08 cu.m	\$ 0.02 per cu.m	\$ 3,123,400
Electricity	4.80E+05 MWh	\$ 50.00 per MWh	\$ 23,993,640
Boiler Feed Water	1.59E+07 tonnes	\$ 0.20 per tonne	<u>\$ 3,173,152</u>
		Utility Costs	\$ 3,299,142
		Total Variable costs	\$ 124,149,142

Table C.3
Synthetic Natural Gas (SNG) from UCG
Annual Estimated Drilling Costs

Costs in 2011 thousand US\$

	<i>Number/run</i>	<i>Number/year</i>	<i>Unit Cost</i>	<i>Costs</i>
Production Wells	1	124 wells	\$ 374 per well	\$ 46,341
Injection Wells	3	372 wells	\$ 173 per well	\$ 64,480
Instrumentation Wells	4	496 wells	\$ 25 per well	\$ 12,400
Instrument Costs	4	496 wells	\$ 10 per well	\$ 4,960
Drill Waste Disposal	325	40300 tonne	\$ 0.05 per tonne	<u>\$ 2,015</u>
<i>Subtotal, Drilling Contractor Turnkey Costs</i>				\$ 130,196
Drilling Program Contingency	30%			\$ 39,059
Direct Employees for Oversight of Drilling Contract		5 employees	\$ 61 each	<u>\$ 310</u>
Total drilling costs				\$ 169,564
 Site Preparation Costs				
	<i>Number/run</i>	<i>Number/year</i>	<i>Unit cost</i>	
Land Lease Costs for Extraction	0.1	12.4 hectares	1.75 per hectare	\$ 22
Site Clearing and Preparation	0.1	12.4 hectares	4.5 per hectare	\$ 56
Utility Road Construction	0.4	49.6 km	8 per km	\$ 397
Field Piping & Installation	0.6	74.4 km	125 per km	<u>\$ 9,300</u>
Site Preparation Costs				\$ 9,774
 UCG Field Operation and Maintenance				
		<i>Number/year</i>		
Decommissioning of spent wells		124	10 each	\$ 1,240
Field Piping Maintenance				\$ 1,860
Monitoring Well Sampling		80	1.5 per sample	\$ 120
Environmental Reporting		2	20 each	<u>\$ 40</u>
Field Operation Costs				\$ 3,260
Total Annual UCG Field Operation Costs				\$ 182,600

Table C.4
Synthetic Natural Gas (SNG) from UCG
Estimated Fixed Capital Costs

Plant Net Capacity		66 Trillion BTU/yr (34,500 BoPD Equivalent)	
<i>Costs in 2011 thousand US\$</i>			
Gas Cleanup and Power Plant			
Battery Limits Investment (BLI)	Equipment Cost	Installation Cost	Total Cost
Reactors	\$ 3,630	\$ 9,076	\$ 12,706
Catalyst Cost			\$ 2,100
Columns	\$ 14,870	\$ 23,800	\$ 38,670
Pressure Vessels	\$ 1,200	\$ 720	\$ 1,920
Heat Exchangers	\$ 41,260	\$ 45,400	\$ 86,660
Claus Package Unit			\$ 17,420
Particulate Removal	\$ 480	\$ 960	\$ 1,440
		<i>Subtotal</i>	\$ 160,916
BLI Contingency	30%	of Installed Equipment Costs	\$ 48,270
Battery Limits Investment			\$ 209,186
Tankage			
Methanol Storage Tanks			\$ 15,840
Methanol Surge Tanks			\$ 3,300
			\$ 19,140
Utilities	Purchased Cost	Installation Cost	Investment
Refrigeration	\$ 20,300	\$ 8,120	\$ 28,420
Boiler Feed Water	\$ 3,260	\$ 1,300	\$ 4,560
Cooling Water	\$ 2,900	\$ 1,160	\$ 4,060
		<i>Utilities Investment Subtotal</i>	\$ 37,000
Offsites & Utility Investment Contingency	30%		\$ 16,842
Offsite & Utilities Investment			\$ 72,982
General Service Facilities	25%	of BLI & Utilities Investment	\$ 70,540
Waste Treatment	5%	of BLI Investment	\$ 10,460
Outside Battery Limits Investment			\$ 154,000
		<i>Total Fixed Capital (TFC) Investment</i>	\$ 363,200

Table C.5
Synthetic Natural Gas (SNG) from UCG
Annual Estimated Operating Costs

Plant Net Capacity		66 Trillion BTU/yr (34,500 BoPD Equivalent)	
<i>Costs in 2011 thousand US\$</i>			
			<i>Costs</i>
Plant Investment, Battery Limits (BLI)			\$ 209,200
Plant Investment, Outside Battery Limits (OBLI)			\$ 154,000
Total Fixed Capital (TFM)			\$ 363,200
Operating Costs, Per Year			
Raw Material Costs (net)			\$ 120,850
Utility Costs (net)			\$ 3,299
Variable Costs			\$ 124,149
Estimated Annual Drilling Costs			\$ 182,600
Labor Costs			
Operating Labor, Gas Cleanup Personnel (3 shifts)	27	\$ 62 wages/year	\$ 1,674
Operating Labor, UCG Field Operations Personnel (3 shifts)	124	\$ 62 wages/year	\$ 7,688
Maintenance Labor	2.40% of BLI		\$ 5,021
Control Laboratory Labor, 10% of Operating Labor	10% of Operating Labor		\$ 940
Direct Labor Costs			\$ 15,320
Maintenance Materials	1.60% of BLI		\$ 3,347
Operating Supplies, 12% of Operating Labor	12% of Operating Labor		\$ 1,120
Total Direct Costs			\$ 341,860
Plant Overhead	80% of Direct Labor Costs		\$ 12,260
Taxes and Insurance	1.60% of TFC		\$ 5,810
Cash Costs			\$ 359,930
Depreciation	14.3% of TFC		\$ 51,890
Gate Costs			\$ 411,820
General, Admin, Sales, Research	5% of Gate Costs		\$ 20,590
Production Costs			\$ 432,410
TFC + Estimated Working Capital			\$ 453,183
ROI	12.5% of Capital Investment		\$ 56,650
Production Cost + Cost of Capital			\$ 489,060
<i>Production Cost + Cost of Capital without Depreciation Charge</i>			<i>\$ 437,170</i>
<i>Stream Factor</i>			<i>0.913</i>
Estimated Natural Gas Output, moles methane			7.78E+10
Estimated Natural Gas Output, MJ HHV			6.92E+10
Estimated Natural Gas Output, MMBTU HHV			6.55E+07
Production Cost including Capital Return per GJ, \$			\$ 7.1
Production Cost including Capital Return per MMBTU, \$			\$ 7.5
Production Cost including Capital Return per Barrel Oil Equivalent Energy, \$			\$ 43.1
Cost excluding Depreciation but including Capital Return per MMBTU, \$			\$ 6.7