TRIMERIC CORPORATION

Preliminary Technoeconomic Analysis for Power Generation from Coal Water Slurry Using Modular Staged-OMB Gasifier

DE-FE0031506

This document has been revised as indicated below and described in the revision record on the following page. Please destroy all previous revisions.

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1 Executive Summary

Trimeric performed a technoeconomic analysis for a proposed 5 MW_e gross modular gasification power generation facility. The cost estimate was used to determine the economic advantages of the proposed modular, staged-OMB gasifier in comparison to other modular gasifier designs. The project, Staged OMB for Modular Gasifier/Burner, is led by The University of Kentucky Center for Applied Energy Research (UK CAER) and supported by East China University of Science and Technology (ECUST) and Trimeric Corporation. A high-level summary of the findings from this project are presented below.

UK CAER Staged-OMB Gasifier Economics

Trimeric developed a fixed capital cost estimate, reported in Q4 2020 dollars, for the coal water slurry, staged-OMB gasifier unit to produce hydrogen-rich syngas to generate power. The capital costs are based on bare equipment costs with appropriate installation factors applied. Trimeric estimated a total purchased equipment cost of $21.0 MM with a total plant cost of $35.1 MM. The capital cost estimate does not include capital costs for coal preparation. Table 1-1 summarizes the purchased equipment costs and total plant costs by area.

Table 1-1. Purchased Equipment Costs and Total Plant Costs by Process Area.

Process Area	Purchased Equipment Costs ($MM)	Total Plant Costs ($MM)	% of TPC
Air Separation Unit	2.24	3.36	9.6
Gasification	2.43	8.21	23.4
Acid Gas Removal	8.18	10.42	29.7
Engines / Power Recovery	7.66	12.1	34.5
Balance of Plant	0.51	1.02	2.9
Total	**21.02**	**35.12**	

The total annual operating revenue, operating expenses, and indirect expenses are shown in Table 1-2. The gasifier facility produces power from coal feedstock as its only revenue stream. Electricity is produced via reciprocating internal combustion engines; thermal energy is not recovered from the gasification or power production process areas.

Table 1-2. Annual Revenue and Operating Costs for Gasifier Facility.

Revenue or Expense	Dollars per Year
Facility Revenue	$ 2,473,000
Facility Operating Expenses	$ 931,000
Facility Indirect Expenses	$ 2,339,000
Total Profit (Loss)	$ (796,000)

Note: 80% onstream factor used [1].

The primary variable operating expenses include: coal (fuel), water makeup, wastewater disposal, solids slag disposal, COS catalyst disposal, and LO-CAT® H_2S removal chemicals and solids disposal. Indirect, or fixed, operating expenses include staffing, maintenance, taxes, and insurance.

As shown in Table 1-2, the gasifier facility, as configured, will lose money when operating – independent of any upfront capital cost requirements. The most significant variable operating cost for the facility is the cost of fuel (54% of the total variable operating expenses). The ASU consumes nearly 19% of the gross power output of the reciprocating engines. The indirect expenses are substantial and are impacted by the low power production of the facility.

UK CAER Staged OMB-Gasifier Comparison to Other Gasifier Designs

The UK CAER staged OMB-gasifier was compared with four other gasification units generating power: two small-scale (5.3 and 18 MW_e gross) and two commercial-scale (763 and 738 MW_e gross). A sensitivity was performed on the process scale of the staged-OMB gasifier, increasing the total gross electrical output from 5.1 MW_e to 25 MW_e. Equipment was scaled using the ratio of total gross power raised to the exponent of 0.6. Variable costs were scaled linearly with gross power produced. The reciprocating engines were replaced by a combustion turbine assuming the technology is available at this scale for energy production from syngas. A summary of the facilities is provided in Table 1-3.

The syngas production performance was compared for the five facility designs as shown in Table 1-4. The staged-OMB gasifier outperforms the other gasifiers on total syngas production fraction (H_2+CO), syngas heating value, and H_2/CO ratio. Both commercial-scale gasifiers (DOE case B4A and S4A) have a moderately higher H_2/CO ratio [1, 2]. Higher heating value allows for more power generation and smaller equipment size per unit mass of feed. Higher H_2/CO ratio is more economical for conversion of syngas to chemicals should a polygeneration facility be desired [3]. The staged-OMB gasifier shows marginal differences for CO_2 content in the product syngas as well as oxygen demand.

A comparison of the facility costs using cost of electricity (COE) is summarized in Table 1-5. The contributions to the COE include the following: capital cost (assumed to be the total plant cost), fuel, variable operating costs, and fixed operating costs.

As shown in Table 1-5, the staged-OMB gasifier has a higher COE than most of the cases except the membrane-wall gasifier. Fuel and variable O&M are low in comparison with the other reference cases. The major contributors to the total COE ($281/MWh) are the capital cost (59%) and the fixed O&M (31%). While the staged-OMB gasifier decreased the gasification process area capital cost, the overall plant cost is largely defined by the other process areas. As shown previously in Table 1-1, gasification accounts for only 23% of the total plant cost with power production (34%) and acid gas removal (30%) being the most expensive process areas.

COE decreases from $281/MWh to $137/MWh when the facility scale is increased from 5.1 to 25 MW_e gross electrical output. Capital costs are significantly lower normalized to throughput because economies of scale favor larger facilities. Fixed O&M costs decrease because labor costs were assumed constant, but the facility output is higher.

Table 1-3. Facility Comparison for Technoeconomic Analysis.

Description	**UK CAER Staged-OMB Gasifier (Base Case)**	**UK CAER Staged-OMB Gasifier (Sensitivity Case)**	**UK CAER Membrane-Wall Gasifier** [4]	**UA Fairbanks HMI Gasifier** [5]	**DOE Case B4A (CB&I E-Gas™ Gasifier)** [1]	**DOE Case S4A (CoP E-Gas™ Gasifier)** [2]
Gross Power (MW_e)	5.1	25	5.3	18	763	738
Other Products	--	--	Steam Hydrocarbon Liquids and Waxes	Steam	--	--
Fuel Supply	Coal water slurry (North Dakota Lignite)	Coal water slurry (North Dakota Lignite)	Coal Fines (Impoundment Fines)	Coal + Biomass (Usibelli Sub-Bituminous/Wood Chips)	Coal water slurry (Illinois No.6 – Bituminous)	Coal water slurry (Powder River Basin – Subbituminous)
Oxidant Supply	Cryogenic ASU (99.6 vol% O_2)	Cryogenic ASU (99.6 vol% O_2)	Cryogenic ASU (>95 vol% O_2)	Air (21 vol% O_2)	Cryogenic ASU (>95 vol% O_2)	Cryogenic ASU (>95 vol% O_2)
Power Block	Reciprocating internal combustion engines	Combustion Turbine	Reciprocating internal combustion engines + turboexpander	Reciprocating internal combustion engines + diesel engine generator	Combustion turbine + steam turbine	Combustion turbine + steam turbine
Acid Gas Removal	COS Hydrolysis LO-CAT®	COS Hydrolysis LO-CAT®	MDEA Unit SulfaTreat	Short contact time caustic scrubber	COS Hydrolysis MDEA Unit Claus Unit	COS Hydrolysis MDEA Unit Claus Unit
Sulfur Load (LTPD)	1.55	7.60	0.58	0.10	55.23	51.18

Table 1-4. Syngas Production Performance Comparison.

Description	Units	UK CAER Staged-OMB Gasifier (Base Case)	UK CAER Staged-OMB Gasifier (Sensitivity Case)	UK CAER Membrane-Wall Gasifier [4]	UA Fairbanks HMI Gasifier [5]	DOE Case B4A (CB&I E-Gas™ Gasifier) [1]	DOE Case S4A (CoP E-Gas™ Gasifier) [2]
$O_2/(H_2+CO)$	Mol/Mol	0.39	Performance of the 25 MW facility assumed the same as the base case.	0.51	0.24	0.36	0.48
Carbon Conversion	%	98.0		97.2	N/A	99.2	99.1
Syngas Quality							
HHV @ Outlet	Btu/SCF	268		262	167	240	242
H_2+CO	Mole Frac	0.81		0.70	0.46	0.71	0.69
H_2/CO	--	0.82		0.64	0.70	0.91	0.95
CO/CO_2	--	2.58		2.99	3.70	1.94	1.32

Table 1-5. Cost of Electricity Comparison.

Description	Units	UK CAER Staged-OMB Gasifier (Base Case)	UK CAER Staged-OMB Gasifier (Sensitivity Case)	UK CAER Membrane-Wall Gasifier [4]	UA Fairbanks HMI Gasifier [5]	DOE Case B4A (CB&I E-Gas™ Gasifier) [1]	DOE Case S4A (CoP E-Gas™ Gasifier) [2]
Gross Power	MW_e	5.3	25	5.3	18	763	738
COE	$/MWh	281	137	355	156	99	74
Capital	$/MWh	164	78	175	93	58	45
Fuel	$/MWh	15	15	0		14	7
Variable O&M	$/MWh	13	13	90	63	10	7
Fixed O&M	$/MWh	88	32	91		17	15

Technical measures that can be considered to improve the economics for the proposed 5 MW staged-OMB gasifier facility are listed below.

- Alternate Coal Type: Lower sulfur coal could reduce the capital and operating expense for acid gas removal. If the sulfur load was the same as with the membrane-wall gasifier case (0.58 LTPD), the COE could be reduced by 15% to $239/MWh. However, input from UK CAER indicated that coal processing with sulfur removal prior to gasification would be more economically favorable for this scale.
- Larger Production Scales: The largest operating costs for the gasification facility are the indirect expenses comprising labor and overhead costs, property tax and insurance, and facility maintenance and upkeep. A larger production facility would improve the scaling of fixed operating costs against generated revenue.
- Replace Engines with Turbines: The reciprocating engines have high capital cost because of the engine efficiency and sizeable derate (50%) due to the low heating value of the syngas and limited application experience. Significant cost escalators are incurred for this special design. Combustion turbines are not available for small-scale processes. Advancements in turbine design at small scale, or increasing the facility throughput, would be required to switch from reciprocating engines to combustion turbines. Increasing the production scale from 5 to 25 MW, and shifting from reciprocating engines to a combustion turbine, yielded a 51% reduction in the COE from $281/MWh to $137/MWh.

2 Introduction

UK CAER was selected as a recipient for the U.S. Department of Energy (DOE) funding opportunity DE-FOA-0001719, titled "Small Scale Modularization of Gasification Technology Components for Radically Engineered Modular Systems (REMS)" under Area of Interest 1 (AOI1), "Modularization of Emerging Gasification Technologies." The main objective of AOI1 is "research into the development of one or more components having application on a REMS gasifier skid that can produce clean syngas from coal and produce power via conversion of syngas in a fuel cell, combustion turbine, or other heat engine" for an energy conversion system scaled to 1-5 MW_e. UK CAER's project, "Staged OMB for Modular Gasifier/Burner", proposes a staged-OMB (opposed multiburner) gasifier to utilize coal water slurry (CWS) for power production, designed to offer flexibility of fuel and load, improved fuel conversion and gasification efficiency, and prolonged wall/burner service life.

Trimeric was selected to perform a technoeconomic analysis as part of the staged-OMB gasification project. The purpose of the technoeconomic analysis was to determine the economic and performance advantages of the proposed modular, staged-OMB gasifier in comparison to other existing modular gasifier designs and full-scale state of the art commercial integrated gasification combined cycle (IGCC) power production facilities. This report contains the following sections of information regarding the technoeconomic analysis task:

- Process engineering design basis
- Purchased equipment and total plant costs
- Fixed and variable operating costs, and revenue streams
- A comparison of normalized CAPEX and OPEX as well as other performance metrics to other existing modular gasifier designs and for full-scale state of the art commercial IGCC power production facilities. Cost and production information was gathered (using publicly available information) for other existing modular gasifier designs funded by the U.S. DOE, and also for full-scale IGCC facilities discussed in reference reports published by the U.S. DOE.

3 Staged-OMB Gasifier Process Engineering Design Basis

A preliminary process design for power generation from coal water slurry using modular, staged-OMB gasification was prepared by ECUST. A process flow diagram (PFD) and a heat and material balance (H&MB) table generated by process simulation of the gasification process using Aspen HYSYS were developed by ECUST. ECUST's gasifier PFD and H&MB table can be found in Appendix A and B. Trimeric developed additional process areas including air separation, acid gas removal, power generation, and balance of plant from the materials provided by ECUST. Block flow diagrams (BFDs) produced by Trimeric can be found in Appendix C. The technoeconomic analysis in this report is based on material and energy balances for the overall facility. A description of the overall process is provided below.

3.1 Process Description

Coal water slurry and oxidant (high purity oxygen) are fed to staged burners along the refractory-lined reaction zone of the gasifier unit. Coal, water, and oxygen react to generate syngas composed mainly of carbon dioxide, carbon monoxide, and hydrogen. The exothermic reaction results in gas leaving the adiabatic reaction chamber at 1300°C. The hot syngas gas leaving the reaction zone is first cooled by a quench ring as it passes downward through a dip tube which guides the gas through the quench zone. The dip tube is submerged in the sump containing quench water. The gas reverses direction passing upwards through the sump where the gas continues to cool. The quenched syngas leaves the gasifier quench zone at 190°C out of a horizontal nozzle. Ash, unrecovered material, and other solids collect at the bottom of the sump where they are transferred to a lock hopper for handling in the gray/black water handling circuit. "Black water" is defined as process water that contains some amount of solid material, while "gray water" is defined as recycled process water that does not contain solids.

Oxygen for the gasifier is generated in a cryogenic air separation unit. In the air separation unit, air is compressed, cooled in an air cooler, and chilled in a refrigeration exchanger. The compression and cooling also condense and separate much of the water vapor in the inlet air stream. The remaining water and carbon dioxide are removed using molecular sieve beds. From the molecular sieve beds, the stream is pre-chilled by exchanging heat with the cryogenic product and waste gas streams from the downstream distillation columns. The exiting product streams are warmed to near-ambient temperature.

The raw oxygen and nitrogen stream is further chilled to cryogenic temperatures. The distillation system uses two distillation columns in series to make product oxygen. The first column operates at a higher pressure and separates nitrogen from oxygen, argon, and other impurities. The raw oxygen stream from the bottoms of the first column is fed to a second, lower pressure distillation column where it is further purified. The bottoms product from this second distillation column is 95 vol%+ purity oxygen at a pressure of approximately 4 MPaG.

Syngas leaving the gasifier enters a cyclonic separator which knocks out any remaining solids or entrained liquids. The separated syngas flows to a water scrubber which counter-currently contacts the syngas with condensate and black water using trayed internals. The scrubbed syngas exits the top of the tower. Black water from the cyclone and water scrubber bottoms is collected and piped to the evaporator tower in the gray/black water circuit.

Collected gray and black water are treated by an evaporator tower and a series of vacuum flash tanks to remove physically entrained hydrogen and acid gases. Evolved acid gases are flared or vented to atmosphere. Produced black water is combined with flocculant in the settling tank to agglomerate solids. Higher density, agglomerated solids settle by gravity to the bottom of the tank and the concentrated solids slurry is pumped to a filtration system. Solids are disposed of as waste, and gray water is recycled to the settling tank. Gray water from the overheads of the settling tank is combined with condensate from the evaporator tower overhead separator and vacuum flash overhead separators in the gray water tank. The gray water is reconditioned for use by the addition of dispersants to break down large hydrocarbon particles. A portion of the gray water recycle is sent to large on-site retention ponds where organic material settles by gravity at the bottom of the retention ponds. Clarified water from the retention ponds is softened to remove additional ions and recycled back to the process. The remaining gray water from the gray water tank is sent back to the lock hopper flush tank. Black water from the lock hopper is sent to a settling pool for slag removal and disposal.

Raw syngas from the water scrubber requires removal of sulfur-containing species to meet the fuel gas feed requirements of the downstream reciprocating internal combustion engines.

Raw syngas is fed to a hydrolysis reactor containing activated alumina catalyst which hydrolyzes COS to form H_2S. The reactor effluent is sent to a condenser to cool the gas and knock out water which is recycled internally. The cooled syngas exits the condenser knockout and enters the liquid redox unit for H_2S removal. The syngas enters an absorber tower where it comes into contact with an alkaline solution containing an oxidant, usually a chelated, multivalent metal. H_2S is oxidized into elemental sulfur and the metal is reduced. The resulting solution is then passed to a reaction tank where oxidation air is added to regenerate (re-oxidize) the metal. Elemental sulfur is filtered out of the liquid solution and dewatered. The regenerated redox solution is pumped back to the absorber tower. The treated syngas with reduced H_2S content exits the overhead of the absorber tower.

Treated syngas from liquid redox unit operation is passed through a letdown valve to meet the feed pressure specification of the reciprocating internal combustion engines. The syngas is fed to the power generation unit to produce electrical power via a bank of three reciprocating internal combustion engines.

Cooling water is supplied to the process by a packaged cooling water system. Treated water is used as makeup to the cooling water loop. Cooling water blowdown is sent to the water recycle tank for use in the gasification process area.

3.2 Review of ECUST Deliverables

ECUST conducted a preliminary process engineering design for a 5 MW_e gross modular gasification unit that utilizes a high-temperature, staged-OMB gasifier to produce syngas that is used to generate power. ECUST's scope included the CWS gasification unit to produce hydrogen-rich syngas. No other energy products (steam, hydrocarbon fuels, etc.) are produced in the system.

Trimeric received the following process engineering deliverables from ECUST to support Trimeric's technoeconomic analysis of the facility:

- Process Flow Diagram: ECUST provided a process flow diagram for the gasification unit. The process flow diagram identified major equipment in the process. The process flow diagram also included stream numbers that correlated to information in the heat and material balance table.
- Heat and Material Balance Table: ECUST modeled the gasification process with Aspen HYSYS®. ECUST provided Trimeric with a heat and material balance table in Excel format from the simulation. Trimeric requested the simulation files to aid in reviewing the process; however, ECUST could not provide the model due to concern of their proprietary data in the simulation.
- Cost Information: ECUST provided the equipment size and cost estimate for the gasifier, which included nozzles, flow controller, burners, and lock hopper for slag discharge if it is built in the china.

ECUST provided an initial process flow diagram and heat and material balance table to Trimeric on February 27, 2021. Based on Trimeric's review of that information, ECUST made several revisions and provided Trimeric with updated files on March 29, 2021, which are included in Appendix A and B.

Trimeric did not fully evaluate ECUST's process engineering design, but a cursory review of the deliverables was performed as follows:

- Trimeric compared a membrane-wall modular gasifier process design to gasify coal fines from a previous project completed by Trimeric for UK CAER in 2019 [4] with the current, staged-OMB gasifier design to better understand the differences and performance impacts of the two gasifier process designs. The major differences are summarized in

Table 3-1 below. The variations in designs will have an impact on the type, size, and cost of the equipment and operating expenses for the OMB gasifier facility as described later in the report.

Table 3-1. Comparison of Membrane-Wall and Staged-OMB Gasification Facility Designs.

Parameter	**2019 Membrane-Wall Gasification Facility Design** [4]	**2021 Staged-OMB Gasification Facility Design**
Products/Revenue Streams	- Electricity - Steam - Hydrocarbon fuels	- Electricity
Air Separation Unit		
Type	Cryogenic air separation	Cryogenic air separation
Gasification		
Gasifier	- Membrane-wall configuration	- Opposed multi-stage burner - Heavy refractory wall similar to E-Gas, Shell, and others [1]
Coal feed	Coal fines slurry	Coal water slurry
Steam production	Included	Eliminated
Gray/Black Water Handling	- Gray water streams from gray water tank flow to slag pool and water scrubber - Gray water from the gray water tank returned to water scrubber - No fresh water to evaporator water tower (formerly HP Flash) - Condensate from acid gas cooler/separator flows to settling tank with water from Vacuum Flash Separator - Vacuum pump skid condensate recycled internally to vacuum cooler/separator and ultimately to the gray water tank	- All gray water diverted to lock hopper - Black water from evaporator water tower returned to water scrubber - Fresh water fed to evaporator water tower - Condensate from acid gas cooler/separator flows to a makeup tank in the balance of plant area - Vacuum pump skid condensate to gray water tank

Parameter	2019 Membrane-Wall Gasification Facility Design [4]	2021 Staged-OMB Gasification Facility Design
Acid Gas Removal		
Sulfur recovery	- Amine (MDEA) unit separates acid gas from syngas - H_2S removed from amine unit acid gas in amine unit regenerator overhead using solid scavengers (SulfaTreat)	- COS removal from syngas using UNICAT catalyst - H_2S removal from syngas using LO-CAT®
Sulfur load	0.58 LTPD	1.55 LTPD
Fischer-Tropsch		
Hydrocarbon fuel production	Included	Eliminated
Steam production	Included	Eliminated
Engines / Power Recovery		
Turboexpander	Included	Eliminated
Engines	Reciprocating internal combustion engines	Reciprocating internal combustion engines
Balance of Plant		
Scope	- Water treatment system - Steam (deaerator, BFW pumps, blowdown equipment) - Cooling tower - Recycle water - Hydrocarbon fuels (tanks, pumps) - Flare stack	- Water treatment system - Cooling tower - Flare stack - Settling pond recycle water softener

- Trimeric identified additional equipment that was not shown or listed on the ECUST process flow diagram which were required to complete the staged-OMB modular gasification design. These equipment items are listed below:

  - Syngas cooler (E-1304) and knockout drum (V-1213A) to cool the gas prior to the LO-CAT® unit.
  - Filter press (M-1301) to dewater the slag/solids from the settling tank.
  - Filtrate tank (V-1312) and pump (P-1311) to collect filtrate from the filter press and recycle it to the process.
  - Filtrate separator (V-1313), vacuum pump separator (V-1314), and filter vacuum pump (P-1312) in the filter press system.
  - P-1307A vacuum condensate pump to transfer liquids under vacuum from V-1304 to V-1309.
  - P-1309A/B vacuum pump to transfer liquids under vacuum from V-1303A to V-1308.
  - Fuel gas tank (V-1211) to store natural gas on site.
  - Other various small items (A-1203 slurry tank agitator, A-1202 slag pool agitator, A-1302 settling tank rake, S-1203 natural gas filter, Z-1203 spark generator, X-1201 slag crusher, V-1210 oxygen tank, A-1205 flocculant agitator, and A-1204 filtrate tank agitator).
  - Trimeric also assigned tag IDs for new equipment not included on ECUST's process flow diagram.
- Trimeric identified several items related to water handling in the overall system including those discussed below:
  - The syngas passes through a cooler (E-1304) where water condenses and is separated in a knockout drum (V-1213A). Normally, the condensed water would be sent back to a water scrubber or the gasifier. However, ECUST only modeled the condenser and did not include the knockout or water recycle. Trimeric created their own simulation of the syngas cooler and water knockout. The recovered condensate from the water knockout is sent to the makeup water tank in the balance of plant area to offset the raw water makeup requirements. Sending the condensate back to the water scrubber would have required manual manipulation of ECUST's stream table outputs.

  - The ECUST process flow diagram shows the water in stream 1323 being purged to "waste water". However, the water actually flows to a large settling pond on site. Once the organic matter settles to the bottom of the pond, the water is treated in an ion exchange unit and pumped back to the main process as a recycle stream.
  - The slag water filtrate (stream 1334) is returned to the settling tank (V-1308), and does not exit the overall process as shown on the process flow diagram.
- Trimeric performed material and energy balances around selected key unit operations using information from the heat and material balance table provided by ECUST. Trimeric did not discover any significant issues with the heat and material balances other than the mass balance around V-1303A (vacuum flash evaporator). Stream 1329 had to be adjusted to close the mass balance around V-1303 and the water balance around the entire process. Trimeric did not have enough information to fully evaluate the enthalpy balance in the reaction zone of the gasifier (F-1201A) since it is a proprietary ECUST design.
- Trimeric reviewed the stream tables in the Excel spreadsheet and identified a few areas requiring clarification or correction:
  - Streams 1101, 1102, and 1103 reported mass flow rates for water when the entries were labeled as molar flow rates. Trimeric corrected this in the stream table included in Appendix B.
  - Streams 1205 and 1232 had incorrect molar flow rates for water. The mass flowrate divided by the molecular weight did not match. Trimeric corrected this in the stream table included in Appendix B.
  - Streams 1101, 1102, and 1103 do not include any sulfur species in the solid phase. Trimeric requested this information to perform a sulfur balance around the gasifier to see the partition of sulfur species present in the feed coal between H_2S and slag. However, ECUST was not able to provide the data but the total sulfur content in the syngas was provided.

- The combustion efficiency for the reciprocating internal combustion engines was increased from 33% to 40% based on information gained from the engine vendors from the past project [4].
- Trimeric cross-referenced streams on the process flow diagram with streams in the material balance tables. Several items were identified and corrected by ECUST:
  - Stream 1317 (flocculant addition) was added to the drawing.
  - Stream 1303 (between V-1301A and P-1301 A1/A2) was added to the drawing.
  - A description was included for Stream 1322 (dispersant).
  - The normal routing for the P-1301A pump discharge is to T-1201A via stream 1233. The pump can also discharge to stream 1229 or stream 1217 as shown on the process flow diagram, but these streams are normally no flow (NNF) during regular process operation.
- Trimeric requested background information on several components of the equipment in the system:
  - Since the staged-OMB gasifier does not produce steam as in the membrane-wall gasifier design [4], Trimeric inquired how heat would be removed from the gasifier to maintain temperature. ECUST indicated that the gasifier will operate adiabatically in the gasification chamber with heavy refractory wall design similar to the gasifiers from CB&I, Shell and others. A cooling water system is used for the burners in the staged-OMB gasifier design as a replacement to the steam system from the past project.
  - The sulfur content of the syngas increased from 0.5 LTPD in the membrane-wall gasifier design that uses coal fines [4] to 1.5 LTPD in the staged-OMB gasifier which uses coal water slurry. ECUST confirmed that coal fines typically have lower sulfur content than coal water slurry systems.

After performing this cursory review of equipment and process information within ECUST's scope, Trimeric developed a list of additional unit operations (outside of ECUST's scope) needed to complete the facility. These major unit operations are noted below:

- Air separation unit to produce oxygen for gasification
- Acid gas removal from the syngas to remove COS and H_2S to specifications needed for the downstream engines
- Reciprocating internal combustion engines for electricity generation
- Balance of plant for water makeup, waste water disposal, cooling tower, and flare

It should be noted that while ECUST provided the overall process flow scheme and heat and material balance table for the gasification unit, Trimeric sized the major equipment besides gasification island in the process to complete the design for this unit operation.

3.3 Process Areas Added by Trimeric

3.3.1 *Internal Combustion Engines*

Previous communications with established equipment vendors found electricity production from syngas using turbines at small-scale to be impractical [4]. High hydrogen content in syngas would require steam or water injection to reduce NO_x and maintain a wide range of stable combustion for "diffusion combustors" [6], and special mitigation to prevent flashback which may include dilution with natural gas or injection modifications [7]. For example, Siemens offers two gas turbine options that produce 5 MW of electricity: the SGT-A05 and the SGT-100. However, these gas turbines can only tolerate up to 5 vol% H_2 in the fuel to mitigate against flashback. The quantity of natural gas required to dilute syngas to 5 vol% H_2 to use the 5 MW gas turbines was determined to be economically unfavorable.

In a past project, Trimeric evaluated two alternative options with GE Distributed Power (now INNIO), Siemens Gas Engines and Dresser-Rand Environmental Industrial Solutions: reciprocating internal combustion engines (RICE), and an external combustion chamber coupled with a KG2-3G/EF gas turbine [4]. The Dresser-Rand Environmental Industrial Solutions group proposed the Ener-Core Power Oxidizer technology to oxidize the syngas and then feed this hot exhaust gas to the KG2-3G/EF gas turbine; however, the technology only had one commercial installation in operation [8]. Trimeric selected reciprocating internal combustion engines because they are a proven technology at the scale of interest for the previous work.

The current project is of similar scale and syngas composition relative to the previous work. Therefore, Siemens and INNIO were sent the composition and flow rate of syngas produced by the gasifier in this project to determine if new options had become available for this production scale. Siemens reconfirmed the previous recommendations and stated that at least 90% natural gas would be required to mix with the syngas. INNIO also reconfirmed their previous recommendations but noted that the higher CO_2 content may require a different J620 engine version; however, the effect on performance and costing would be negligible. J620 RICE gensets were selected for this project per recommendation by INNIO. Details regarding estimated costs for RICE gensets are covered in the Capital Cost Estimate and Operating Cost Estimate sections of the report.

The selected RICE gensets have a recommended fuel supply pressure of 60 to 70 psig (0.41 to 0.48 MPaG). The syngas leaving the acid gas removal unit is approximately 1.74 MPaG, so a turboexpander was investigated as an option to recover additional electricity from the syngas prior to combustion. Trimeric estimates that an additional 0.11 MW of electricity can be generated from the turboexpander. However, simulation results showed significant cooling of the expanded syngas and the presence of condensed liquid. Syngas reheat is necessary to meet fuel supply requirements of 10 to 40°C and less than 80% relative humidity [9]. A preliminary cost estimate for a gas-gas heat exchanger to reheat the syngas using the exhaust gas from a single RICE was developed, but the total investment of the turboexpander and exchanger were determined to be too high relative to the value of the additional recovered energy. The high-pressure syngas is passed through a regulator to achieve the necessary inlet conditions for the RICE gensets.

3.3.2 *Air Separation Unit (ASU)*

There are two commercially available options for oxygen generation at the scale required for the power generation facility: cryogenic separation and vacuum swing adsorption (VSA). Cryogenic separation is more costly (approximately twice the capital cost of a VSA unit per one major industrial gas supplier) and requires more electricity when compared to VSA, but it also produces a higher purity oxygen product. Cryogenic separation units can produce oxygen with approximately 99 vol% (or greater) purity, while VSA units are limited to a maximum of approximately 93 vol% purity. Trimeric selected cryogenic air separation for this application to meet the feed oxygen specification of 99.6 vol%.

Trimeric learned in a previous project that all major North American industrial gas suppliers have phased out the cryogenic ASU product lines at smaller scales; their customers have all migrated to VSA systems at this scale. A budgetary quote was received from Kaikong in China for a cryogenic ASU [4]. The quote from Kaikong was used as the costing basis for the cryogenic ASU specified in this project.

3.3.3 *Acid Gas Removal*

The selected reciprocating engines that produce electricity have maximum COS and H_2S limits for the feed gas. COS is limited to <0.02 mol% (200 ppmv) and H_2S is limited to 38 ppmv as calculated according to fuel supply requirements for the engines as found in documentation provided by GE Power & Water [9]. The acid gas removal equipment was designed to supply feed gas at or below the recommended COS and H_2S levels.

The syngas COS will be removed in a hydrolysis reaction that converts the COS to H_2S and CO_2. UNICAT offers a catalyst (CHC-5) that can be used for this application. UNICAT stated that the hydrolysis reaction with the catalyst is more favorable at higher temperature (~149ºC), and therefore they recommended locating the catalyst bed after the water scrubber (~187ºC) before syngas cooling. The catalyst can achieve 99% COS removal at these conditions (~5 ppmv COS in the treated gas). The catalyst would need to be replaced every 2 to 4 years. UNICAT also stated that there are no pre-activation steps required for the catalyst. UNICAT does not expect there to be a significant temperature change in the gas across the bed. There will be a low pressure drop from passing through the catalyst bed (~5 psi was assumed). The gas leaving the COS removal unit will flow to a cooler (E-1304) to reduce the temperature of the syngas to 40ºC and a knockout vessel (V-1213A) to remove condensed water. The cooled syngas stream will then flow to a H_2S removal unit.

In the 2019 membrane-wall gasifier study [4], an amine unit was used to remove acid gas from the cooled syngas stream exiting V-2213 (coal gas separator). The acid gas was then treated with solid scavenger for H_2S removal. While the total plant cost for this system was reasonable ($1.99 MM), the operating expenses were substantial due to the need to replace the non-regenerable solid scavenger (SulfaTreat) material. The operating cost for SulfaTreat was estimated to be $9,031/day at 0.5 LTPD of sulfur.

For the staged-OMB gasifier project, a liquid redox technology was selected for H_2S removal instead of using a non-regenerable H_2S scavenger. The liquid redox process can treat the entire syngas stream and would be more a more economical process than solid scavengers at the sulfur load for this application (1.55 LTPD). Solid scavengers are typically used for sulfur loads <0.1 LTPD. Iron redox processes are normally considered best suited for ~0.1 to 15 LTPD of elemental sulfur production when used in typical natural gas applications, which was assumed to also be applicable for syngas applications.

Liquid redox technologies are wet scrubbing systems that employ chelated iron solutions that contact the gas stream and convert H_2S to elemental sulfur. The spent solution containing the elemental sulfur is transferred to an oxidation vessel, where the chemistry is regenerated by sparging air (or pure oxygen) through the solution. The sulfur slurry is filtered to remove the byproduct wet sulfur cake, and the regenerated chemistry is returned to the absorber vessel. The solid sulfur filter cake can be disposed of in a landfill (nonhazardous) or shipped to a fertilizer company. Compounds that dissolve in the LO-CAT® solution can flash off from the pressure reduction between the absorber and oxidizer; this gas is clean of H_2S and can be repressurized to join the treated gas or vented if environmentally reasonable.

Trimeric received a budgetary estimate from Merichem for a LO-CAT® sulfur recovery unit. Merichem has over 230 LO-CAT® licenses worldwide, and the process has been used on syngas (5% of licenses) [10]. The H_2S in the treated gas from the LO-CAT® unit will be in the range of 4-10 ppmv. Merichem was not concerned with the operating pressure (~1.84 MPa) of the syngas stream; Merichem indicated that the LO-CAT® chemistry works at all pressures and the pressure for this application also reduces the size of the absorber. Merichem provided an estimate of the electrical and chemical requirements for the process and noted that no substantial cooling water was required at these conditions. The total plant cost for the LO-CAT® unit is $10.1 MM, but the operating expenses and sulfur disposal costs are only $938/day.

Caustic scrubbing is another potential H_2S treating option that can be used to remove H_2S at relatively low sulfur tonnages such as this. However, there is a substantial amount of CO_2 relative to H_2S in the syngas (36 kmol/hr CO_2 and 2 kmol/hr H_2S). Since the caustic will also react with the CO_2 in the syngas, it would be necessary to couple caustic scrubbing with an upfront selective treating unit (for CO_2 removal) or possibly use a special caustic scrubber design to limit the CO_2 pickup. Less CO_2 pickup avoids operating issues including the potential for solid sodium carbonate (Na_2CO_3) formation and plugging, and minimizes unnecessary caustic usage while improving the NaHS quality, which is the ultimate end product when treating with caustic. If the NaHS product is of high enough purity, it could potentially be sold. However, if there is too much CO_2 pickup, then the NaHS product stream would be contaminated with carbonate salts and may need to be disposed of as a waste byproduct. In addition, at high CO_2 to H_2S ratios, annual caustic makeup operating costs make this option economically unfavorable.

It may be that a selective amine (such as ExxonMobil Flexsorb™ or BASF's OASE sulfexx™) could be used to selectively remove H_2S in the presence of CO_2 if the selectivity is high enough. Short contact time (SCT) caustic scrubber designs, that use static mixers or other devices to limit the contact time, preferentially absorb H_2S over CO_2 since H_2S has significantly faster reaction kinetics relative to CO_2. The SCT caustic scrubber requires careful design and operation for selective H_2S removal or the product can go off specification or plugging can occur in the system. The economics for caustic scrubbing will be highly dependent on the selectivity achievable by the system. These options could be considered in future phases of the project.

3.3.4 *Balance of Plant*

In addition to the process areas noted above, Trimeric developed a preliminary process design for balance of plant (BOP) areas not covered by ECUST. This scope and process design covers areas such as:

- Process water, including feedwater treatment and recycle water softening
- Cooling tower and cooling water distribution
- Flare systems

Trimeric created block flow diagrams for cooling water systems, process water distribution, and the flare stack – which can be found in Appendix C. Using this information, Trimeric developed an equipment list for all equipment and systems outside of ECUST scope; estimated capital costs for this equipment are covered in Section 4.2.2 of the report.

Coal is assumed "as-delivered"; additional costs for preparation are outside of Trimeric's scope. Additionally, large settling ponds are used to clarify gray water before returning to the gasification process. Evaporation or other water control strategies and any additional costs directly associated with the settling ponds were not included in Trimeric's scope.

4 Capital Cost Estimate

4.1 Summary

This section provides a high-level summary of the approach used to estimate the equipment sizing and cost estimates for the gasification facility. It also presents a description of the overall capital costs.

4.1.1 *Scope of Capital Cost Estimate*

The scope of Trimeric's capital cost estimate includes equipment in the following areas:

- Air Separation Unit
- Gasification
- Acid Gas Removal
- Engines / Power Recovery
- Balance of Plant

The capital cost estimate does not include coal preparation.

4.1.2 *Equipment Sizing*

ECUST provided a process flow diagram and heat and material balance table for the gasification area of the facility. Trimeric used this information to size the gasification equipment.

In order to compare costs for the staged-OMB gasifier and membrane-wall gasifier [4] on the same basis, similar equipment sizing criteria were used for this application. For example, the material of construction for the equipment was assumed to be the same as in the membrane-wall gasifier study [4]. In some cases, the size of equipment (e.g., separators, tanks, slag pool, etc.) was scaled based on equivalent equipment from the membrane-wall gasifier study [4]. For other equipment (e.g., towers and heat exchangers), new designs were developed. Lastly, the size for specific other equipment (e.g., gasifier) was obtained directly from ECUST or UK CAER. More details of the sizing of the equipment in the gasification area are presented in Section 4.2.2.

Trimeric developed the conceptual design for the other areas of the facility (air separation unit, acid gas removal, engines / power recovery, and balance of plant). More details regarding this equipment are given in the cost section for each process area.

4.1.3 *Equipment Cost Estimation*

Trimeric relied on four different methods for estimating the bare or skidded equipment costs:

- Application of cost information supplied directly by ECUST (gasifier only)
- Costs obtained directly from vendors
- Costs scaled from past project information
- Costs obtained from Aspen Capital Cost Estimator

Aspen Capital Cost Estimator (CCE) was used for all standard equipment including separators, columns, tanks, pumps, heat exchangers, agitators, and conveyors.

Trimeric's experience is that the Aspen CCE software does a reasonable job of estimating costs for steel fabricated equipment for which the material costs are the primary contributor to the overall cost – and for standard rotating equipment. Trimeric has Aspen CCE V11.1 with a cost database from Q1 2019. Costs from Aspen CCE were escalated to Q4 2020 using indices the Chemical Engineering Plant Cost Index (CEPCI) [11].

In some cases, the costs for certain tanks, pumps, and other equipment (slag pool, slag conveyor, etc.) were scaled from the Aspen CCE estimates in the membrane-wall gasifier study [4] that used a cost database from Q1 2016. These costs were also escalated to Q4 2020 using a CEPCI conversion as well.

Table 4-1 gives the key sizing criteria by equipment type used to cost the equipment in Aspen CCE.

Table 4-1. Key Sizing Criteria for Aspen CCE.

Equipment Type	Sizing Parameters for Aspen CCE
Heat Exchangers	Materials of construction (MOC), design pressure and temperature, heat transfer surface area, shell diameter, tube length
Vessels or Tanks	MOC, design pressure and temperature, diameter, and height (or vessel volume)
Columns	Vessel MOC, design pressure and temperature, vessel diameter and height, number of trays, tray MOC
Pumps	MOC, design pressure and temperature, total developed head, motor power, type of pump (e.g. centrifugal)

Tables with detailed equipment sizing and cost data are given in Appendix D.

If the material of construction was not available in Aspen CCE for the equipment type, then Trimeric selected an alloy with a composition as close as possible to the specified material. For example, for several pieces of equipment in the past study [4] Q235B was specified according to the Chinese standard. Trimeric estimated costs in Aspen for this equipment assuming ASME A36 mild steel. Also, in some equipment, the design pressure and temperature were adjusted for this application using the same ratio of operating to design pressure from the membrane-wall gasifier project [4].

ECUST provided a cost for the gasifier (including nozzles, flow controller, burners, and lock hopper for slag discharge) of approximately CYN 4 million (equivalent to $610,000 USD).

Details of the basis for the equipment costs (Aspen, vendor data, etc.) are given in the equipment cost section for each area.

4.1.4 *Results*

Trimeric developed a fixed capital cost estimate for the gasifier facility based on bare equipment costs with appropriate installation factors applied. The capital is reported as Q4 2020 dollars, with all source costs adjusted to a CEPCI of 595.9 [11]. Trimeric estimated a total purchased equipment cost of $21.0 MM with a total plant cost of $35.1 MM.

The overall installation factor is 1.7. This is a relatively low installation factor that is consistent with Trimeric's experience with highly skidded units; most of the high-cost equipment within the facility is skidded and expected to have a low installation cost. Some equipment (LO-CAT® unit, reciprocating engine gensets, air separation unit, cooling tower, etc.) will come directly from vendors as packaged units, while other loose equipment will be assembled and packaged modularly at a fabrication shop before being delivered to site.

Table 4-2 summarizes the purchased equipment costs and total plant costs by area.

Table 4-2. Purchased Equipment Cost and Total Plant Cost Summary.

Process Area	Purchased Equipment Costs ($MM)	Total Plant Costs ($MM)	% of TPC
ASU	2.24	3.36	9.56
Gasification	2.43	8.21	23.39
Acid Gas Removal	8.18	10.42	29.68
Engines / Power Recovery	7.66	12.1	34.47
Balance of Plant	0.51	1.02	2.90
Total	**21.02**	**35.12**	

4.2 Equipment Cost Estimates by Area (Details)

4.2.1 *Air Separation Unit*

The estimated equipment cost for the Air Separation Unit (ASU) is $2.24 MM with an installed cost of $3.36 MM (9% of the total plant cost). A low installation factor of 1.5 was selected because this is a skidded unit. Table 4-3 summarizes cost data for the Air Separation Unit, including the Purchased Equipment Cost (PEC) and the Total Plant Cost (TPC), which includes installation.

Table 4-3. Air Separation Unit Cost Summary.

Tag	Equipment Name	Equipment Type	PEC (per unit)	No. Units	PEC (Total)	TPC
n/a	Air Separation Unit	Other	$2,239,000	1	2,239,000	$3,359,000

A cryogenic ASU was selected for the current project scope due to the high oxygen purity requirement (99.6 vol%) for the gasifier feed. Trimeric had previous discussions with multiple U.S. vendors for similar applications; they stated that their typical offerings of cryogenic units are for much larger scales and, therefore, they would likely decline if requested to provide a quote. Per the vendors consulted, the cryogenic unit is considered an outdated technology at this smaller scale and most customers select a vacuum swing adsorption system (VSA), which typically has significantly lower capital and operating costs. However, VSA units have a lower maximum product oxygen purity of approximately 93 vol% that does not meet the current feed oxygen requirement.

A previous budgetary cost estimate for a cryogenic ASU from a vendor (Kaikong) in China was used [4]. The estimate from Kaikong (CNY 22.1 million) was approximately 40% lower than the concept level estimate provided by a U.S. vendor in early 2018. The capital cost estimate uses the budgetary quote from the Chinese supplier, but it is noted that there is significant uncertainty in the cost.

The purchased equipment cost was estimated by scaling according the oxygen demand of the current case relative to the quoted basis. The estimate purchased equipment cost was scaled from 2018 dollars to 2021 dollars using the CEPCI index [11].

4.2.2 *Gasification Area*

The estimated purchased equipment cost for the Gasification area is $2.43 MM with an installed cost of $8.21 MM (23% of the total plant cost). The unit comprises a mix of highly skidded equipment (such as the gasifier) and loose equipment (such as a number of pumps, vessels, and heat exchangers). An overall installation factor of 3.4 was applied to this area. Table 4-4 summarizes the cost data for the Gasification area.

Table 4-4. Gasification Area Cost Summary.

Tag	Equipment Name	Equipment Type	PEC (per unit)	No. Units	PEC (Total)	TPC
F-1201A	Gasifier	Other	$610,000	1	$610,000	$1,220,000
E-1201	Burner Cooling Water Heat Exchanger	Heat Exchanger	$72,000	1	$72,000	$288,000
E-1304	Syngas Cooler	Heat Exchanger	$73,000	1	$73,000	$292,000
E-1301A	Acid Gas Condenser	Heat Exchanger	$16,000	1	$16,000	$64,000
E-1303A/B	Waste Water Cooler	Heat Exchanger				See Table 4-5
E-1302A	Vacuum Flash Evaporative Condenser	Heat Exchanger	$22,000	1	$22,000	$88,000
T-1201A	Water Scrubber	Vessel / Tower	$55,000	1	$55,000	$220,000
V-1101	Charcoal Slurry Tank	Vessel / Tower	$310,000	1	$310,000	$1,240,000
V-1202	Medium Pressure Nitrogen Tank	Vessel / Tower				See Table 4-5
V-1211	Fuel Gas Tank	Vessel / Tower				See Table 4-5
V-1204	Burner Cooling Water Tank	Vessel / Tower	$49,000	1	$49,000	$196,000
V-1205A1/A2/A3/A4/A5	Burner Cooling Water & Gas Separator	Vessel / Tower	$5,000	5	$25,000	$100,000
V-1209	Accident Burner Cooling Water Tank	Vessel / Tower	$67,000	1	$67,000	$268,000
V-1203A	Water Sealed Tank	Vessel / Tower	$44,000	1	$44,000	$176,000
V-1206A	Lock Hopper	Other				See Table 4-5
V-1207A	Lock Hopper Flush Water Tank	Vessel / Tower	$16,000	1	$16,000	$64,000
L-1201A	Slag Chain Conveyor	Other	$38,000	1	$38,000	$76,000
V-1208A	Slag Pool	Other	$8,000	1	$8,000	$24,000
V-1213A	Raw Gas Separator	Vessel / Tower	$33,000	1	$33,000	$132,000
V-1201A/B	High Pressure Nitrogen Tank	Vessel / Tower				See Table 4-5
V-1302A	Acid Gas Separator	Vessel / Tower	$6,000	1	$6,000	$24,000
T-1301A	Evaporator Water Tower	Vessel / Tower	$34,000	1	$34,000	$136,000
V-1303A	Vacuum Flash Evaporator	Vessel / Tower	$31,000	1	$31,000	$124,000
V-1304A	Vacuum Flash Evaporative Separator	Vessel / Tower	$5,000	1	$5,000	$20,000
V-1305A	Vacuum Pump Separator	Vessel / Tower	$13,000	1	$13,000	$52,000

Tag	Equipment Name	Equipment Type	PEC (per unit)	No. Units	PEC (Total)	TPC
V-1307	Dispersant Tank	Vessel / Tower	$3,000	1	$3,000	$12,000
V-1306A/B	Flocculant Tank	Vessel / Tower	$35,000	1	$35,000	$140,000
V-1308	Settling Tank	Vessel / Tower	$202,000	1	$202,000	$808,000
V-1309	Gray Water Tank	Vessel / Tower	$57,000	1	$57,000	$228,000
V-1312	Filtrate Tank	Vessel / Tower	$72,000	1	$72,000	$288,000
V-1313	Filtrate Separator	Vessel / Tower				See Table 4-5
V-1314	Vacuum Pump Separator	Vessel / Tower				See Table 4-5
P-1101A1/A2	Charcoal Slurry Feed Pump	Pump	$14,000	2	$28,000	$112,000
P-1101A3	Charcoal Slurry Feed Pump	Pump	$3,000	1	$3,000	$12,000
P-1202A/B	Burner Cooling Water Pump	Pump	$34,000	2	$68,000	$272,000
P-1203A1/A2	Lock Hopper Recycling Pump	Pump	$4,000	2	$8,000	$32,000
P-1204A	Slag Pool Pump	Pump	$3,000	2	$6,000	$24,000
P-1201A1/A2	Black Water Recycling Pump	Pump	$6,000	2	$12,000	$48,000
P-1307A	Vacuum Condensate Pump	Pump	$3,000	2	$6,000	$24,000
P-1302A	Vacuum Pump	Vacuum Pump	$10,000	1	$10,000	$40,000
P-1303A/B	Low Pressure Gray Water Pump	Pump	$6,000	2	$12,000	$48,000
P-1304A/B	Settling Tank Substrate Pump	Pump	$9,000	2	$18,000	$72,000
P-1311	Filtrate Pump	Pump	$7,000	2	$14,000	$56,000
P-1312	Filter Vacuum Pump	Vacuum Pump				See Table 4-5
P-1306A/B	Flocculant Pump	Pump	$3,000	2	$6,000	$24,000
P-1305A/B	Dispersant Pump	Pump	$2,000	2	$4,000	$16,000
A-1203	Slurry Tank Agitator	Other				See Table 4-5
A-1202	Slag Pool Agitator	Other				See Table 4-5
A-1302	Settling Tank Rake	Other				See Table 4-5
Y-1201A1/A2/A3/A4/A5	Oxygen Silencer	Other				See Table 4-5
S-1203	Natural Gas Filter	Other	$6,000	1	$6,000	$12,000
Z-1201A1/A2/A3/A4/A5	Burner	Other				See Table 4-5
Z-1203	Spark Generator	Other	$7,000	1	$7,000	$14,000
X-1201A	Slag Grinding Mill	Other				See Table 4-5

Tag	Equipment Name	Equipment Type	PEC (per unit)	No. Units	PEC (Total)	TPC
A-1201A	Mixer	Other				See Table 4-5
S-1202A	Cyclone	Vessel / Tower	$29,000	1	$29,000	$116,000
A-1301	Static Mixer	Other				See Table 4-5
M-1301	Vacuum Belt Filter	Other	$154,000	1	$154,000	$462,000
Z-1202A	Preheat Burner	Other				See Table 4-5
S-1101	Hydraulic Cylinder Sieve	Other				See Table 4-5
V-1102	Flush Water Tank	Vessel / Tower				See Table 4-5
P-1102A/B	Flush Water Pump	Pump				See Table 4-5
E-1202A	Lock Hopper Flush Water Cooler	Heat Exchanger	$23,000	1	$23,000	$69,000
P-1205A	Preheated Water Pump	Pump				See Table 4-5
S-1201A1/A2	Black Water Filter	Other				See Table 4-5
J-1201A	Startup Ejector	Other				See Table 4-5
Y-1202A	Ejector Silencer	Other				See Table 4-5
V-1301A	High Temperature Water Tank	Vessel / Tower	$14,000	1	$14,000	$56,000
P-1301A1/A2	High Temperature Water Pump	Pump	$48,000	2	$96,000	$384,000
P-1309A/B	V-1303 discharge pump	Pump	$5,000	2	$10,000	$40,000
V-1310	Nitrogen Sealing Tank	Vessel / Tower				See Table 4-5
V-1210	Oxygen tank	Vessel / Tower				See Table 4-5
A-1205	Flocculant agitator	Other				See Table 4-5
A-1204	Filtrate tank agitator	Other				See Table 4-5
				TOTAL	**$2,430,000**	**$8,213,000**

The gasification equipment was sized following the general guidelines listed below. Detailed notes on the specific criteria for sizing and costing for each piece of equipment can be found in Appendix D.

- Gasifier – ECUST provided the size of the staged-OMB gasifier.
- Towers – The towers were sized using guidelines provided by UK CAER, which included using a superficial velocity of ~1.5 m/s and assuming 4 trayed stages in T-

1201A (water scrubber) and 4 trayed stages plus one chimney tray in T-1301A (evaporator water tower). A spacing of 2 feet (0.61 m) was assumed for the trays and at liquid and vapor feed points in the tower. Trimeric also assumed a two-minute residence time at 50% full in the sump. The sump diameter is wider than the main column diameter for both towers.

- Separators – The separators were sized generally in one of several methods including checking the minimum diameter required for vapor separation with simple Souders-Brown calculations [12]; scaling the size from the past study [4] for separators that appeared to be dominated by liquid holding requirements, and estimating sizes in commercial simulation software (Schlumberger Symmetry v2002.2). The length-to-diameter (L/D) from the past study [4] was used to approximate the separator dimensions.
- Tanks – The size of the tanks in the gasification area were scaled from the total liquid volume and L/D ratios for the tanks from the past study [4].
- Heat Exchangers – The heat exchangers were sized using the duty requirements from the heat and material balance table, and assuming the same overall heat transfer coefficient and type/designation of exchanger (e.g., BEM, BEU, etc.) as in the past study [4]. The bundle diameter and tube length were resized to produce typical L/D designs as necessary. A 10% overdesign was included in the size estimates.
- Pumps – The flow rates for the pumps were taken from the heat and material balance tables and an overdesign factor of 1.2 to 1.3 was applied. The power was estimated for the pumps using the same pump efficiencies for analogous pumps as in the past study [4]. It was also noted in the past study that the flow rates for the settling tank pump and filtrate pump were significantly higher than the simulation results (possibly because they operate intermittently). For this reason, the flow rates for the P-1304A/B (settling tank substrate pump) and P-1311 (filtrate pump) in this application were adjusted accordingly higher.
- Vacuum Pump – A factor of 1.2 was applied to the flow rate for P-1302A (vacuum pump) in the material balance table. The power requirement for the vacuum pump was estimated using general vendor guidelines for these specific types of pumps.

- Special Equipment – The size (and cost) of some of the specialty equipment in the process (e.g., slag pool, etc.) were scaled from the past study [4].

Table 4-5. Gasification Area – Key Notes on Cost and Size Estimate.

Tag	Equipment Name	Notes
E-1303A/B	Waste Water Cooler	Eliminated - Waste water is sent to a retention pond, cooling not required
V-1202	Medium Pressure Nitrogen Tank	Excluded. N_2 tank included in ASU supplier scope
V-1211	Fuel Gas Tank	Excluded: Assume NG stored onsite in bullet containers.
V-1206A	Lock Hopper	Size basis: not estimated Cost basis: included with gasifier
V-1201A/B	High Pressure Nitrogen Tank	Excluded. N_2 tank included in ASU supplier scope
V-1313	Filtrate Separator	Excluded from cost per 2019 project (vapor flow rate negligible).
V-1314	Vacuum Pump Separator	Excluded from cost per 2019 project (vapor flow rate negligible).
P-1312	Filter Vacuum Pump	Excluded: vapor flow not provided and assumed negligible per 2019 project.
A-1203	Slurry Tank Agitator	Integrated with V-1101.
A-1202	Slag Pool Agitator	Integrated with V-1208.
A-1302	Settling Tank Rake	Integrated with V-1308.
Y-1201A1/A2/A3/A4/A5	Oxygen Silencer	Minimal cost. Did not include in capital cost.
S-1203	Natural Gas Filter	Assumed same cost as in 2019 project.
Z-1201A1/A2/A3/A4/A5	Burner	Burners included with ECUST gasifier cost.
Z-1203	Spark Generator	Assumed same cost as in 2019 project.
X-1201A	Slag Grinding Mill	Assumed included in gasifier cost.
A-1201A	Mixer	Minimal cost. Did not include in capital cost.
A-1301	Static Mixer	Minimal cost. Did not include in capital cost.
Z-1202A	Preheat Burner	Minimal cost. Did not include in capital cost.
S-1101	Hydraulic Cylinder Sieve	Minimal cost. Did not include in capital cost.
V-1102	Flush Water Tank	Minimal cost. Did not include in capital cost.
P-1102A/B	Flush Water Pump	Minimal cost. Did not include in capital cost.
P-1205A	Preheated Water Pump	Not included in cost estimate (for startup only).
S-1201A1/A2	Black Water Filter	Did not include in cost estimate (no cost in 2019 project either).

Tag	Equipment Name	Notes
J-1201A	Startup Ejector	Not included in cost estimate (for startup only).
Y-1202A	Ejector Silencer	Not included in cost estimate (for startup only).
V-1310	Nitrogen Sealing Tank	Did not include in cost estimate (no cost in 2019 project either).
V-1210	Oxygen tank	Did not include in cost estimate (no cost in 2019 project either).
A-1205	Flocculant agitator	No cost in 2019 project; minimal cost and not included in 2021 project.
A-1204	Filtrate tank agitator	No cost in 2019 project; minimal cost and not included in 2021 project.

Tables that compare the equipment sizing and costs from the membrane-wall gasifier design [4] to the staged-OMB gasifier are included in Appendix E.

4.2.3 *Acid Gas Removal Area*

The estimated equipment cost for the Acid Gas Removal area is $8.18 MM with an installed cost of $10.42 MM (30% of the total plant cost). The overall installation factor for the COS and H_2S removal units is 1.3. Table 4-6 summarizes the cost data.

Table 4-6. Acid Gas Removal Area Cost Summary.

Tag	Equipment Name	Equipment Type	PEC (per unit)	No. Units	PEC (Total)	TPC
n/a	Liquid Redox Equipment	Other	$8,107,000	1	$8,107,000	$10,14,000
n/a	COS Removal	Other	$36,000	2	$72,000	$288,000
				TOTAL	**$8,179,000**	**$10,422,000**

UNICAT provided an estimate of the size of the vessel to hold the catalyst bed (155 ft^3 or 4.4 m^3). UNICAT also recommended that the vessel have an L/D ratio of 2. Given this information, the vessel was sized to be 1.45 m in diameter and 2.85 m in height. The vessel was assumed to be made of the same material as the water scrubber T-1201 (13MnNiMoR+316L (4 mm) Cladded). The design includes a spare vessel for maintenance or replacement. An installation factor of 4 was used for the vessels.

Merichem provided a budgetary estimate for a 1.84 LTPD unit based on the sulfur load from the first heat and material balance table that ECUST provided (with 33% engine efficiency). The sulfur load was reduced to 1.55 LTPD in the final version of the heat and material balance as a result of increasing the engine efficiency to 40%. The budgetary estimate was adjusted to the new sulfur load using a scaling exponent of 0.6. The budgetary cost included all the equipment to build the job as well as the engineering. Merichem indicated that the LO-CAT unit is mostly stick built and that the total project cost would also need to include installation, civil work, utilities, and other miscellaneous costs. Merichem stated that this usually ranges from 25% to 100% of the quoted budgetary cost, depending on the user and location. Merichem's preference is to only provide the proprietary equipment (absorber, oxidizer, filter, etc.). Trimeric assumed a factor of 1.25 to arrive at the total plant cost for the LO-CAT® unit. This represents a best-case estimate of total plant costs for the acid gas removal unit.

4.2.4 *Power Production Area*

The estimated purchased equipment cost for the engines is $7.66 MM with an installed cost of $12.10 MM (33% of the total plant cost). An installation factor of 1.6 was applied to this area based on reference data provided by the U.S. EPA [13]. Table 4-7 summarizes the cost data.

Table 4-7. Engines Cost Summary.

Tag	Equipment Name	Equipment Type	PEC (per unit)	No. Units	PEC (Total)	TPC
n/a	Engines	Other	$2,554,000	3	$7,662,000	$12,105,000
				TOTAL	**$7,662,000**	**$12,105,000**

In the previous project [4], several vendors recommended estimating engine costs for a similar application using an existing EPA study [13] that published normalized costs ($/kW) for gas engine generators using natural gas in combined heat and power applications. GE Distributed Power, now INNIO, provided an estimation of the engine performance for the delivered syngas composition and provided a list of comments and recommended modifications to the costing and performance tables in the referenced EPA document. GE Distributed Power's comments include:

- The hydrogen content in the syngas requires an approximate 50% derate in gross electric output over an equivalent natural gas fired engine.
- Very few special gas engines are produced; therefore, this application would be considered a "one-off" build, which results in additional incurred costs associated with special system requirements of the genset.
- There is potential for toxic tar formation which creates special EHS requirements and further increase the maintenance costs beyond the effect of the reduced power output. Additional industrial hygiene monitoring may be required. However, UK CAER has indicated that this concern should be negligible.
- The significant power derate leads to higher costs in the exhaust gas treatment system(s), heat recovery steam generator, and auxiliary engine systems on a normalized basis. A 50% derate results in a doubling of the system costs.

The provided comments and recommendations result in a capital cost escalation factor between two to three for the syngas application relative to the reference natural gas application.

Trimeric confirmed the same performance derate and cost escalation factors with INNIO for this project. The current project's syngas is above the upper limit for CO_2 for the selected engine version, per commentary from INNIO. A different J620 engine version might be better suited for the current syngas composition; however, only minor changes to heat balances would be expected.

4.2.5 *Balance of Plant*

The estimated equipment cost for the Balance of Plant equipment is $0.51 MM with an installed cost of $1.02 MM (3% of the total plant cost). An installation factor of 2.0 was applied to this area. A lower installation factor was used because the Water Treatment System, Cooling Tower, and Flare System can be purchased "off the shelf" from vendors as skidded units. Table 4-8 summarizes the cost data.

Table 4-8. Balance of Plant Cost Summary.

Tag	Equipment Name	Equipment Type	PEC (per unit)	No. Units	PEC (Total)	TPC
n/a	Water Treatment System	Other	$218,000	1	$218,000	$436,000
n/a	Cooling Tower	Other	$224,000	1	$224,000	$448,000
n/a	Water Softener	Other	$9,000	2	$18,000	$36,000
n/a	Flare Stack	Other	$50,000	1	$50,000	$100,000
				TOTAL	**$510,000**	**$1,020,000**

Trimeric sized the equipment using the corresponding heat and material balance. The costs for the Water Treatment System, Cooling Tower, and Flare Stack were scaled based on vendor data from other projects. The cost of the Water Softener used for gray water recycle was based on industrial water softener prices located during this project. Trimeric assumes that the Water Treatment System will include coarse filtration, multimedia filtration, and a reverse osmosis (RO) module.

5 Operating Expenses and Revenue

5.1 Summary

The gasifier facility produces power from coal feedstock as its sole revenue stream. Electricity is produced by three reciprocating internal combustion engines operating in parallel; thermal energy is not recovered from the gasification or power production process areas. The expected net power production from the facility at full production rates is 3,778 kW; this includes power produced from three reciprocating engines less the facility parasitic load. The expected revenue for power production by the gasifier facility is $8,470 per day.

Operating costs for the facility include the fuel to the gasifier as well as utilities and other material costs. Fuel to the gasifier is a slurry composed of prepared coal, process water, and additives. The price of additives or other coal preparation steps are not included in the cost basis. Details for utility and material costs are provided later in this section. Rates and resulting operating costs are shown in Table 5-1 and Table 5-2, respectively.

Table 5-1. Gasification Facility Materials and Utilities Required.

Facility Material/Utility	Quantity	Details
Coal	58.4 tons/day	Coal is received containing moisture
Water	30,290 lb/hr	Water for Cooling Tower and Gasification Unit
Waste Water	5,540 lb/hr	Water from Gasifier Unit
Natural Gas	2 MMBtu/day	Enrichment gas required for flaring events
Slag Solids (Coarse and Fine)	522 lb/hr	From gasifier unit
LO-CAT® Sludge	240 lb/hr	Sulfur cake at 40% moisture from H_2S removal
LO-CAT® Chemicals/Other	$780/day	Estimated from approximate total operating cost less electricity demand
COS Catalyst	$26/day	UNICAT CHC-5 hydrolysis catalyst changed out every 3 years

Table 5-2. Gasification Facility Operating Expenses per Day.

Facility Operating Expense	Cost per Day
Coal	$ 1,713
Water	$ 215
Waste Water	$ 53
Natural Gas	$ 4
Slag (Coarse and Fine) Disposal	$ 238
LO-CAT® Sludge Disposal	$ 159
LO-CAT® Chemical/Other Makeup	$ 780
COS Catalyst	$ 26
Total Operating Expenses per Day	**$ 3,187**

The gasifier facility indirect expenses are for staffing (operating and maintenance labor), maintenance, taxes, and insurance. Details regarding indirect expense bases are given in Section 5.4. These additional expenses are shown in Table 5-3.

Table 5-3. Indirect Expenses for Gasification Facility.

Indirect Expense	Cost per Day
Staffing	$ 2,558
Maintenance	$ 1,924
Taxes and Insurance	$ 1,924
Total Indirect Expenses per Day	**$ 6,407**

The total annual operating revenue, operating expenses, and indirect expenses are shown in Table 5-4. The facility's annual revenue and operating expenses consider the gasifier facility online 80% of the time during a calendar year which is consistent with the DOE reference report [1]. Indirect expenses are not impacted by the fraction of time the facility is online.

Table 5-4. Annual Revenue and Operating Costs for Gasification Facility.

Revenue or Expense	Dollars per Year
Revenue	$ 2,473,000
Operating Expenses	$ (931,000)
Indirect Expenses	$ (2,339,000)
Total Profit (Loss)	**$ (796,000)**

Table 5-4 shows that the gasification facility will lose money when operating – independent of capital expenditure.

The major variable operating cost for the facility is the cost of fuel. The fuel price was assumed to be the levelized price of North Dakota lignite to maintain consistency with the assumed price of electricity, which was assumed to be the COE for an IGCC plant using North Dakota lignite coal [14, 15]. A cheaper or local low sulfur coal would reduce fuel costs, but the lower heating value and higher moisture content would increase the coal drying costs and the size and operating costs of the overall facility to meet the same net production. Lower rank coal is typically more reactive and contains more volatile matter than higher rank coals, leading to faster reaction and more converted gas relative to the amount of char produced [16]. Lower rank coal also does not have caking properties, so it can be used in a wider variety of gasifiers without addition pre-treatment [17]. However, lower rank coals may have a higher tendency to form tar or other potential byproducts caused by less efficient combustion which can lead to additional post-combustion scrubbing requirements.

The parasitic load on the gasification unit could be reduced by changes to the feed oxygen specification. The cryogenic ASU consumes nearly 19% of the gross power output of the reciprocating engines. Lowering the required feed oxygen concentration would allow switching from a cryogenic ASU to VSA that has the potential to decrease capital cost and increase the amount of electricity that can be sold to the grid.

The largest costs of the gasification facility are fixed operating costs comprising labor and overhead costs, property tax and insurance, and facility maintenance and upkeep. A larger production facility would improve the scaling of fixed operating costs against generated revenue.

5.2 Revenue (Details)

5.2.1 *Electricity*

The gasifier facility burns syngas to produce electrical power using three reciprocating internal combustion engines in parallel. Each reciprocating engine produces approximately

1,700 kW of electricity, for a total gross output of 5,100 kW. Thermal energy of the engine exhaust is not utilized in this project.

A portion of the generated power is consumed by equipment in the gasifier facility. The largest parasitic load is the ASU, which contains two series of compressors necessary to compress ambient air to separation pressure and the oxygen product up to gasifier feed pressure. The remainder of the parasitic load is energy supplied to the LO-CAT® unit and other small users throughout the facility including pumps, mixers/agitators, solids handling, and other miscellaneous equipment. Table 5-5 shows the expected electricity producers and users in the gasifier facility.

Table 5-5. Electricity Generation and Consumption in Gasifier Facility.

Electricity Generator/Consumer	**Quantity**	**Details**
Reciprocating Engines	5,100 kW	Three reciprocating engines at 1,700 kW_e
ASU Compressors	-961 kW	Main Air Compressor and Oxygen Compressor
LO-CAT® AGR	-221 kW	Merichem estimate of the total power requirements for this unit
Miscellaneous Users	-140 kW	Summation of equipment list power requirements
Total Electricity Generated	**3,778 kW**	**Net Power to Sell to Grid**

The sales price of power is based upon the COE for a commercial-scale IGCC facility using North Dakota lignite coal, scaled from 2007 dollars to 2021 dollars using CEPCI [14]. At the assumed $0.093 per kWh, the total revenue generation from power production is $8,470 per day of operation.

5.3 Operating Expense (Details)

Operating expenses for the gasifier facility are expenses directly related to production rates, and not related to the purchased cost of the facility or the personnel required to operate the facility.

5.3.1 *Coal (Fuel)*

The gasifier facility is fed coal water slurry to staged burner nozzles which combust in the presence of oxygen to produce syngas for power production. The price of coal used was $29 per ton [15], which results in a daily fuel cost of $1,713. North Dakota lignite coal was selected for the costing basis to maintain consistency with the selected price of electricity [14]. The reported price is the levelized coal price assuming the facility is located at the mine site (minemouth) [15].

5.3.2 *Water*

The gasifier facility requires fresh water for different unit operations. Where possible, water is recycled or used multiple times in an effort to minimize the quantity of water flowing to or from the facility. BFDs for the water system at the gasification facility are in Appendix C.

A fresh water cost of $2.46 per 1,000 gallons was used [18], which results in a daily operating cost of $215. Trimeric notes that these costs are highly variable dependent upon geographic location, and Trimeric selected a cost for an Owensboro, KY, public utility – which was the closest location to Eastern Kentucky (excluding major metropolitan areas). The fresh water cost, reported as $1.86 per 1,000 gallons in 2016, was escalated to the year 2021 using the reported annual water price escalation rate of 5.72% [18].

5.3.3 *Waste Water*

The gasifier facility uses water to produce the coal water slurry feed and to scrub solids and other water-soluble impurities from the product syngas. The scrubber water effluent is treated for solids removal and most dissolved gases are flashed off and sent to flare or atmosphere. However, dissolved solids (ionic species) accumulate in the process water such that discharge is required.

The process material balance provided by ECUST assumes that the majority of water is recycled within the gasification facility. The main water stream leaving the gasification process is gray water from the gray water tank that is sent to on-site settling ponds which allow for settling of organic material. The clarified water is passed through a water softener before it is reused to supply fresh water users located in the gasification area. Blowdown from the cooling water loop is also used as a source of fresh water supply to the gasification area to minimize fresh water demand for the facility. Due to the large flow rate of blowdown and water recycle strategy of the gasifier, excess water is available to supply the fresh water demand of the gasification unit. Excess water is sent to waste to avoid accumulation of contaminants, along with the water contained in the slag and fines waste streams.

This analysis assumes that the water discharges to a Publicly Owned Treatment Works (POTW) without any additional treatment beyond treatment for solids removal shown on the gasification PFD in Appendix A. The cost for waste water sent to the POTW is \$3.29 per 1,000 gallons [18], and results in a daily operating cost of \$53. Trimeric notes that these costs are also highly variable dependent upon geographic location, and Trimeric selected a cost for a location in Chesterfield County, VA – which was the closest location to Eastern Kentucky (excluding major metropolitan areas). The waste water cost, reported as \$2.25 per 1,000 gallons in 2016, was escalated to the year 2021 using the reported annual waste water price escalation rate of 4.05% [18].

5.3.4 *Natural Gas (Flare Enrichment)*

Syngas or other gases with low heating values sent to flare require enrichment gas to achieve the minimum heating value required for combustion. Based on past vendor quotes received by Trimeric, a minimum lower heating value (LHV) of 300 Btu/SCF was used to estimate the required enrichment gas flow rate assuming the full syngas flow is sent to flare. Enrichment gas was assumed to be natural gas with a LHV of 983 Btu/scf priced at \$2.59 per MMBtu. The natural gas price was taken to be the spot price for the Midwest, according to the data reported by the U.S. Energy Information Administration (EIA) [19].

Daily natural gas usage was estimated assuming two flaring events per year at 24 hours per event, normalized by annual online production time. Pilot, assist, and purge gas rates were assumed negligible. The daily operating cost of natural gas is \$4.

5.3.5 *Slag (Coarse and Fine) Disposal*

Solid byproducts are formed during coal gasification. These products, including slag, ash, and coal fines, are separated from the syngas using a combination of scrubbing water and physical separation devices. Solids removed from the effluent water are collected in roll-off bins which are moved by truck to a landfill for disposal. Details on the water treatment system of the gasification unit for the removal of coarse and fine slag from black and gray water are in the gasification PFD, found in Appendix A.

This analysis assumes that all solids discharged from the gasification unit will be disposed of in this manner. The cost for disposal is $38 per ton of slag material [1]. The gasification unit produces approximately 6.26 tons of slag per day of operation for a total daily operating cost of $238.

5.3.6 *Acid Gas Removal*

According to UNICAT, the COS hydrolysis catalyst (CHC-5) will need to be replaced every two to four years. For the purposes of this work, Trimeric assumed the catalyst changeout frequency would be every three years. UNICAT stated that the cost of the catalyst for this application is $28,000 (per vessel). This equates to a cost of $26/day. More information on the catalyst material is provided in Appendix F of this report.

Merichem provided an estimate of the LO-CAT® chemical and electrical costs for a 1.84 LTPD unit (based on the heat and material balance table with an engine efficiency of 33%). Merichem stated that the annual operating costs would be $500,000/yr (assuming 365 day/yr operation). Merichem indicated that the electricity usage would be 263 kW, which gives an annual electrical cost of $161,272/yr. The chemical costs were estimated by difference to be $338,728/yr or $928/day. The operating expenses were scaled by the ratio of the sulfur loads of the project relative to the original budgetary quote (~1.55 LTPD/1.84 LTPD). The adjusted total operating cost is $420,106/yr and the electrical load is 221 kW, giving a cost for chemicals/other consumables of $780/day.

LO-CAT® produces a sulfur cake that will need to be disposed of to a landfill as a non-hazardous waste. The sulfur cake is about 60% sulfur and 40% LO-CAT® solution. Trimeric assumed a disposal cost of $55/ton based on landfill disposal costs in Texas for solid waste material [20]. This equates to a disposal cost of $159/day. The sulfur could also be used as fertilizer but this disposal option was not considered for this work.

The total operating cost for acid gas removal for this application is $964/day.

5.4 Indirect Expense (Details)

The gasifier facility costs that are not directly related to power production include:

- Staffing for the facility, which includes plant operators, plant management personnel, and maintenance technicians directly employed by the facility.
- Maintenance and upkeep for the facility, based upon the total plant capital cost. This would include materials and outside labor required to keep the facility online and operating reliably, but not maintenance personnel employed directly by the facility.
- Property taxes and insurance for the facility, based upon the total plant capital cost.

5.4.1 *Staffing for Facility*

The gasifier facility requires appropriate staffing to run reliably and safely. Operations personnel are at the site 24 hours per day, 7 days per week with management and maintenance staff working a typical 40-hour work week. Table 5-6 shows the expected personnel required to staff the gasifier facility, which is based upon Trimeric's experience operating a facility of this complexity.

Table 5-6. Gasifier Facility Staffing Requirements.

Staff Category	Quantity Required	Cost to Facility per Unit	Total Cost to Facility per Year
Operator	8 (2 per shift, 4 shifts)	$ 79,500	$ 636,000
Maintenance	2	$ 55,500	$ 111,000
Management and Admin	N/A	25% of Operator and Maintenance Staff	$ 187,000
		Total Staffing Cost	**$ 934,000**

Annual salaries for operators and maintenance personnel are based off of May 2020 wage estimates reported by the U.S. Bureau of Labor and Statistics (BLS) [21]. Trimeric used annual mean wages for the State of Kentucky for "Chemical Plant and System Operators" (Occupation Code: 51-8091) for operators, and "Welders, Cutters, Solderers and Brazers" (Occupation Code: 51-4121) for maintenance personnel. Base salaries are escalated by 30% to account for benefits paid to the employee by the facility, consistent with the DOE reference report [1]. Management and administration costs were assumed to be 25% of the annual cost for the total operator and maintenance staff, consistent with the DOE reference report [1].

5.4.2 *Maintenance & Upkeep and Taxes & Insurance for the Facility*

Indirect expenses for maintenance and upkeep and for taxes and insurance for the facility are both estimated as 2% (total of 4%) of the total capital cost of the facility to maintain consistency with the DOE reference report [1]. The total capital cost for the facility is estimated to be $35.1 MM; more details on the capital cost for the facility are in Section 0 of this report. Total maintenance and upkeep costs are $702,000 per year, and taxes and insurance for the facility are also $702,000 per year.

6 Performance and Cost Comparison

6.1 Summary and Conclusions

The UK CAER staged-OMB gasifier performance was compared with four other gasification designs. Facility economics for the staged-OMB gasifier design were also studied as a function of process scale.

6.1.1 *Comparison to Other Gasifiers*

The UK CAER staged-OMB gasifier was compared with two small-scale (5.3 and 18 MW_e gross) and two commercial-scale (763 and 738 MW_e gross) gasification designs that generate power for electricity [4, 1, 5, 2]. A sensitivity was performed on the process scale of the staged-OMB gasifier design, increasing the total gross electrical output from 5.1 MW_e to 25 MW_e. Equipment was scaled by the ratio of the gross power output raised to the exponent of 0.6 and variable costs were scaled linearly with throughput. The reciprocating engines were replaced by a combustion turbine assuming that the technology is available for power production from syngas at that production scale without requiring cost escalation for additional special equipment or derate for the lower heating value of the syngas.

All cases have common process areas including oxidant preparation and feed, gasification and syngas scrubbing, acid gas removal, power generation and balance of plant. Cryogenic air separation is used with the exception of the HMI-designed gasifier which is supplied by blown ambient air. Power is generated by reciprocating internal combustion engines at small production scales; combustion turbines are used in conjunction with the commercial scale gasifier. The UK CAER staged-OMB gasifier design case does not include heat recovery steam generator (HRSG), which is a common addition in all other cases. The UK CAER membrane-wall gasifier design case converts a portion of the generated syngas (derived from the gasification of waste coal fines) to hydrocarbon liquids using a Fischer-Tropsch unit.

A comparison of syngas production performance is summarized in Table 6-1. The influence of coal rank is not accounted for in these results outside of the effects included in the reported heat and material balances. The staged-OMB gasifier outperforms the other gasifiers on total syngas production fraction (H_2+CO), syngas heating value, and H_2/CO ratio. The commercial-scale gasifiers (DOE cases B4A and S4A) have a moderately higher H_2/CO ratio. Higher heating value allows for more power generation and smaller equipment size per unit mass of feed. Higher H_2/CO ratio is more economical for conversion of syngas to chemicals should a polygeneration facility to produce both electricity and chemicals be desired [3]. The staged-OMB gasifier shows marginal differences for CO_2 content in the product syngas as well as oxygen demand.

A comparison of the facility costs using Cost of Electricity (COE) is summarized in Table 6-2. The contributions to the COE are listed below: capital cost (assumed to be the total plant cost), fuel, variable O&M costs, and fixed O&M costs. The staged-OMB gasifier has a higher COE compared to most of the cases considered. Capital costs account for 59% of the COE, followed by fixed O&M at 31%. While the staged-OMB gasifier decreased the gasification process area capital cost at the 5 MW_e gross power production scale compared to the membrane-wall gasifier, the overall plant cost is largely defined by the other process areas. Gasification accounts for only 23% of the total plant cost. Capital intensive areas ordered by the percentage of the total plant cost are power production (34%), acid gas removal (30%), air separation (10%), and balance of plant (3%).

Reductions in capital for power production could be made by technology advancements. Reciprocating engines could be replaced by more cost-effective combustion turbines if small-scale turbine design can be adapted to accommodate higher hydrogen syngas concentrations. In addition, reducing the sulfur load to the facility by switching to a lower rank coal would reduce capital and operating expenses associated with the acid gas removal area. Lower rank coal would also decrease fuel costs, but potentially increase capital costs in other process areas due to a lower syngas heating value.

Table 6-1. Syngas Production Performance Comparison of UK CAER Staged-OMB Gasification with Other Small and Commercial Scale Power Generating Gasification Facilities.

Description	Units	UK CAER Staged-OMB Gasifier (Base Case)	UK CAER Staged-OMB Gasifier (Sensitivity Case)	UK CAER Membrane-Wall Gasifier [4]	UA Fairbanks HMI Gasifier [5]	DOE Case B4A (CB&I E-Gas™ Gasifier) [1]	DOE Case S4A (CoP E-Gas™ Gasifier) [2]
$O_2/(H_2+CO)$	Mol/Mol	0.39	Performance of the 25 MW facility assumed the same as the base case.	0.51	0.24	0.36	0.48
Carbon Conversion	%	98.0		97.2	N/A	99.2	99.1
Syngas Quality							
HHV @ Outlet	Btu/SCF	268		262	167	240	242
H_2+CO	Mole Frac	0.81		0.70	0.46	0.71	0.69
H_2/CO	--	0.82		0.64	0.70	0.91	0.95
CO/CO_2	--	2.58		2.99	3.70	1.94	1.32

Table 6-2. Cost Comparison of UK CAER Staged-OMB Gasification with Other Small and Commercial Scale Power Generating Gasification Facilities.

Description	Units	UK CAER Staged-OMB Gasifier (Base Case)	UK CAER Staged-OMB Gasifier (Sensitivity Case)	UK CAER Membrane-Wall Gasifier [4]	UA Fairbanks HMI Gasifier [5]	DOE Case B4A (CB&I E-Gas™ Gasifier) [1]	DOE Case S4A (CoP E-Gas™ Gasifier) [2]
Gross Power	MWe	5.1	25	5.3	18	763	738
COE	$/MWh	281	137	355	156	99	74
Capital	$/MWh	164	78	175	93	58	45
Fuel	$/MWh	15	15	0		14	7
Variable O&M	$/MWh	13	13	90	63	10	7
Fixed O&M	$/MWh	88	32	91		17	15

The influence of coal rank is reflected by comparison of Case B4A and Case S4A. Oxygen demand is lower for Case S4A that uses subbituminous coal compared to bituminous coal used in Case B4A. The lower price of subbituminous coal is directly reflected in the fuel cost; however, Case B4A contains additional units for raw syngas cleaning (HCl and ammonia scrubbing) which increase the capital and operating expenses relative to Case S4A. Additionally, Case B4A includes significantly more capital expenditure in feedwater preparation relative to Case S4A. Higher rank coals typically permit smaller equipment sizes because the heating values of the produced syngas are higher; however, the capital cost contribution is higher for the bituminous gasifier (Case B4A) compared to the subbituminous gasifier (Case S4A).

COE decreases from $281/MWh to $137/MWh when the facility scale is increased from 5.1 to 25 MW_e gross electrical output. Capital costs are significantly lower normalized to throughput because economies of scale favor larger facilities. Fixed O&M costs decrease because labor costs were assumed constant, but the facility output is higher.

6.2 <u>Comparison Cases (Details)</u>

6.2.1 *Case Descriptions*

<u>UK CAER Staged-OMB Gasifier (5.1 MW)</u>

Details regarding the preliminary design, performance, and costing of the proposed modular, staged-OMB gasifier are contained in this report. The staged-OMB gasifier converts coal water slurry to syngas that is used to produce 5.1 MW gross electrical power by a bank of three reciprocating engines. Oxygen is fed to the gasifier by a cryogenic ASU due to the feed oxygen specification of 99.6 vol%. Sulfur is removed from the raw syngas by a two-step process involving COS hydrolysis followed by liquid redox using the LO-CAT® process.

<u>UK CAER Staged-OMB Gasifier (25 MW)</u>

The UK CAER staged-OMB gasifier design was scaled from 5.1 MW gross to 25 MW gross to study the economic impacts of a larger production facility. The reciprocating engines were replaced by a combustion turbine. Siemens supplied an estimated purchased equipment cost for the SGT-600 packed system of $12.4 MM and recommended an installation factor of 2. Siemens recommended value is consistent with other publicly available costs for similar scale combustion turbines [22]. The cost of the SGT-600 turbine was used without escalation or derating to account for the lower heating value of syngas; however, advancements in turbine design are still required for syngas operation. In this technoeconomic analysis, Trimeric assumed that a comparable technology was available to represent a best-case scenario for the staged-OMB gasifier at the larger process scale.

All other equipment selections remained the same,but were scaled according to the gross power output of the facility raised to the exponent of 0.6.

UK CAER Membrane-Wall Gasifier

A previous project report [4] from UK CAER, Trimeric, and others details the preliminary design, performance, and costing of a modular, membrane-wall gasifier that produces syngas from waste coal fines. Membrane-wall gasifiers include a lining of steam generating tubing between the refractory wall and the gasifier shell. The gasifier facility is a combined heat and power (CHP), polygeneration facility. The gasifier converts coal fines from impoundment ponds to syngas and steam. The CHP unit produces 5.3 MW gross electrical power as well as steam from reciprocating engine exhaust gas, steam from the gasifier steam drum, and hydrocarbon (HC) fuels, HC waxes, and steam from a Fischer-Tropsch (F-T) unit. Oxygen is fed to the gasifier by a cryogenic ASU due to the feed oxygen specification of 95 vol%. Acid gases are removed from the raw syngas first by separation using an aqueous methyldiethanolamine (MDEA) unit; sulfur is then removed from the concentrated acid gas stream using a non-regenerable solid scavenger (SulfaTreat).

UA Fairbanks Gasifier

A second small-scale modular gasification reference case is presented in the report by the University of Alaska Fairbanks (UAF) [5]. The gasifier in that report is a refractory lined, air blown, atmospheric, moving bed, up-draft type gasifier supplied by Hamilton Maurer International (HMI). The gasifier is based on the Wellman-Galusha design. The gasifier facility is a CHP facility. The gasifier converts coal or a mixture of coal and biomass to syngas and pyrolysis liquids (a mixture of tars and oils). The CHP unit produces electricity and steam from syngas using reciprocating engines with an HSRG. Pyrolysis liquids from the gasifier are mixed with ultra-low sulfur diesel as fuel to a diesel engine generator to produce additional power and steam. The facility has a gross power output of 18 MW combined between the reciprocating engines and the diesel engine. Oxygen is fed to the gasifier by an air blower. Raw syngas is cleaned by a wet electrostatic precipitator to remove pyrolysis liquids followed by a short contact time caustic scrubber to remove sulfur.

DOE Reference Case B4A (CB&I E-Gas^{TM} Gasifier)

The first commercial reference gasifier is Case B4A from a DOE baseline report that is an IGCC facility fed by bituminous coal [1]. The gasifier is a Chicago Bridge & Iron Company (CB&I) E-Gas^{TM} design. The gasifier is a pressurized, upflow, entrained, slagging gasifier. Coal water slurry is converted to syngas that is used to produce power in a combustion turbine. Hot exhaust is sent to a HRSG which generates steam. The steam is let down through a steam turbine to produce additional power. The overall gross electrical output of the facility is 763 MW. Oxygen is fed to the gasifier by a cryogenic ASU due to the feed oxygen specification of 95 vol%. Contaminants are separated from the raw syngas using an HCl scrubber, an ammonia scrubber, and a mercury removal unit. Acid gases are removed from the raw syngas by COS hydrolysis followed by acid gas enrichment in an MDEA unit. The concentrated acid gas stream from the MDEA unit is sent to a Claus unit for sulfur recovery.

DOE Reference Case S4A (CoP E-Gas^{TM} Gasifier)

The final reference gasifier is Case S4A from a DOE baseline report that is an IGCC facility fed by subbituminous coal [2]. The gasifier is a Conoco Phillips (CoP) E-Gas^{TM} design. The gasifier is a pressurized, upflow, entrained, slagging gasifier. Coal water slurry is converted to syngas that is used to produce power in a combustion turbine. Hot exhaust is sent to a HRSG which generates steam. The steam is let down through a steam turbine to produce additional power. The overall gross electrical output of the facility is 738 MW. Oxygen is fed to the gasifier by a cryogenic ASU due to the feed oxygen specification of 95 vol%. Mercury is removed from the raw syngas by a mercury removal unit. Acid gases are removed from the raw syngas by COS hydrolysis followed by acid gas enrichment in an MDEA unit. The concentrated acid gas stream from the MDEA unit is sent to a Claus unit for sulfur recovery.

6.2.2 *Comparison of Cases*

The cases used for the cost and performance comparison are similar in that they all produce power from coal-derived syngas; however, the cases have many differences including their production scale, diversity of product streams, fuel source, oxidant supply, power block units, and acid gas load and acid gas removal design (Table 6-3).

The differences between cases have the following effects on facility cost and performance:

- Process Scale: Purchased equipment costs scale non-linearly with capacity; the price of equipment is typically more favorable per unit capacity as production scale increases. Labor costs scale favorably with production scale as well.
- Fuel Supply: The price of coal varies significantly with coal rank and transportation cost and represents a major operating cost to gasification facilities. Using lower rank coals or locating the gasification facility closer to the fuel source improve fuel costs. The UK CAER Membrane-Wall Gasifier study used coal fines recovered from impoundment ponds as a "free" source of fuel; excavation and other preparation costs were not considered as part of the Trimeric scope [4].
- Oxidant Supply: Air separation units are a significant cost contributor to gasification plants. Different ASU technologies can save considerably on capital expenses and parasitic load on the power generation facility, freeing up additional power to sell to the grid. High oxygen purity at small scales provided by cryogenic ASUs is very capital and energy intensive, and most North American industrial gas suppliers no longer provide designs for modular ASUs. VSAs can significantly reduce costs for the oxygen supply process area, but there are limitations on oxygen purity.

- Power Block: Combustion turbines operating on syngas at small scale are currently not available unless significant quantities of natural gas are blended with the syngas to reduce hydrogen concentration; reciprocating internal combustion engines must be used. The low heating value of syngas affects the cost and performance of reciprocating engines; the power output is roughly half of combustion with natural gas per input from vendors. Because of the significant derate in performance, reciprocating engines for these applications are still considered "one-off" builds which require additional special systems and cost escalators. The use of combustion turbines at larger scale are expected to reduce capital cost per unit of power generated. Syngas turboexpanders and steam turbines may be used to convert pressure and heat into additional electrical output.
- Acid Gas Removal: Acid gas removal is a significant cost center for coal gasification. Non-regenerable methods such as solid scavengers can only be used at small sulfur loads (<0.1 LTPD) without leading to excessive operating costs. Regenerable methods incur additional capital costs but will typically lower operating costs considerably. For the reference cases considered, there is a substantial variation in sulfur load. Lower rank coals would reduce the sulfur load in addition to increasing gasification efficiency (higher reactivity) at reduced cost. However, lower rank coals have lower heating values which require more throughput for equivalent power generation. Higher throughput could increase capital costs in other process areas.

Table 6-3. Case Information Used for Technoeconomic Comparison.

Description	UK CAER Staged-OMB Gasifier (Base Case)	UK CAER Staged-OMB Gasifier (Sensitivity Case)	UK CAER Membrane-Wall Gasifier [4]	UA Fairbanks HMI Gasifier [5]	DOE Case B4A (CB&I E-Gas™ Gasifier) [1]	DOE Case S4A (CoP E-Gas™ Gasifier) [2]
Gross Power (MW)	5.1	25	5.3	18	763	738
Other Products	--	--	Steam Hydrocarbon Liquids and Waxes	Steam	--	--
Fuel Supply	Coal water slurry (North Dakota Lignite)	Coal water slurry (North Dakota Lignite)	Coal Fines (Impoundment Fines)	Coal + Biomass (Usibelli Sub-Bituminous/Wood Chips)	Coal water slurry (Illinois No.6 – Bituminous)	Coal water slurry (Powder River Basin – Subbituminous)
Oxidant Supply	Cryogenic ASU (99.6 vol% O_2)	Cryogenic ASU (99.6 vol% O_2)	Cryogenic ASU (>95 vol% O_2)	Air (21 vol% O_2)	Cryogenic ASU (>95 vol% O_2)	Cryogenic ASU (>95 vol% O_2)
Power Block	Reciprocating internal combustion engines	Combustion Turbine	Reciprocating internal combustion engines + turboexpander	Reciprocating internal combustion engines + diesel engine generator	Combustion turbine + steam turbine	Combustion turbine + steam turbine
Acid Gas Removal	COS Hydrolysis LO-CAT®	COS Hydrolysis LO-CAT®	MDEA Unit SulfaTreat	Short contact time caustic scrubber	COS Hydrolysis MDEA Unit Claus Unit	COS Hydrolysis MDEA Unit Claus Unit
Sulfur Load (LTPD)	1.55	7.60	0.58	0.10	55.23	51.18

6.2.3 *Normalization of Cases*

The technoeconomic analysis results are presented by comparing normalized total capital and operating costs on gross and net power bases, and incorporating these costs into a cost of electricity (COE). "*The COE is the revenue received by the generator per net megawatt-hour during the power plant's first year of operation, assuming that the COE escalates thereafter at a nominal annual rate equal to the general inflation rate, i.e., that it remains constant in real terms over the operational period of the power plant*" [23]. COE incorporates the capacity factor of the facility as well as the capital charge factor. The capacity factor was assumed to be 0.8 for all cases to maintain consistency with the DOE reference case [1]. A capital charge factor of 0.124 was selected for all cases which assumes a high-risk investor-owned utility (IOU) finance structure with a capital expenditure period of five years [23].

6.2.4 *Results*

A summary of the cost and performance results are presented in Table 6-4.

<u>Gasifier Performance</u>

Gasifier performance is similar for the staged-OMB gasifier relative to the comparison cases. All cases have high carbon conversion and produce syngas with similar HHV. Carbon conversion cannot be calculated from the UA Fairbanks report because neither a proximate nor ultimate analysis are presented. A significant portion of the carbon is converted to pyrolysis liquids – exclusion of the pyrolysis liquids would yield an artificially low carbon conversion though the carbon is still used in that process.

Table 6-4. Technoeconomic Analysis Results Summary.

Description	Units	UK CAER Staged-OMB Gasifier (Base Case)	UK CAER Staged-OMB Gasifier (Sensitivity Case)	UK CAER Membrane-Wall Gasifier [4]	UA Fairbanks HMI Gasifier [5]	DOE Case B4A (CB&I E-Gas™ Gasifier) [1]	DOE Case S4A (CoP E-Gas™ Gasifier) [2]
Gross Power	MW	5.1	25	5.3	18	763	738
Net Power	MW	3.8	19.2	3.5	N/A	641	605
$O_2/(H_2+CO)$	Mol/Mol	0.39	Same as Base Case	0.51	0.24	0.36	0.48
Carbon Conversion	%	98.0		97.2	N/A	99.2	99.1
Syngas Quality							
HHV @ Outlet	Btu/SCF	268	Same as Base Case	262	167	240	240
H_2+CO	Mole Frac	0.81		0.70	0.46	0.71	0.69
H_2/CO	--	0.82		0.64	0.70	0.91	0.95
CO/CO_2	--	2.58		2.99	3.70	1.94	1.32
Purchased Equipment Cost							
Gross Output	$/MWe	4,122,000	1,883,000	3,051,000	N/A	1,240,000	1,074,000
Net Output	$/MWe	5,564,000	2,452,000	4,597,000	N/A	1,477,000	1,310,000
Total Plant Cost							
Gross Output	$/MWe	6,886,000	3,381,000	6,550,000	N/A	2,756,000	2,076,000
Net Output	$/MWe	9,295,000	4,403,000	9,867,000	N/A	3,280,000	2,534,000
Operating Cost							
Variable	$MM	0.43	2.11	2.74	N/A	54.5	39.6
Fuel	$MM	0.50	2.45	0.00	N/A	80.3	39.1
Fixed	$MM	2.34	3.96	2.22	N/A	78.1	61.8
Total O&M (Gross)	$/MWe	641,000	355,000	945,000	556,000	279,000	190,000
Total O&M (Net)	$/MWe	865,000	462,000	1,424,000	N/A	332,000	232,000
COE	$/MWh	281	137	355	156	99	74
Capital	$/MWh	164	78	175	93	58	45
Fuel	$/MWh	15	15	0	63	14	7
Variable O&M	$/MWh	13	13	90		10	7
Fixed O&M	$/MWh	88	32	91		17	15

The staged-OMB gasifier shows a higher H_2/CO ratio than the membrane-wall gasifier which makes the syngas easier to convert to chemicals or substitute natural gas. H_2/CO ratio is a weak function of coal rank; temperature and pressure more strongly influence the water gas shift reaction [17]. The optimum H_2/CO ratio for methanol and substitute natural gas is two and three, respectively [3]. CO_2 is in higher proportion relative to the membrane-wall gasifier but lower than the DOE reference case. CO_2 removal would improve Fischer-Tropsch reactions because CO_2 is a product of those reactions; lower CO_2 in the feed increases total conversion [24]. CO_2 removal would also lessen potential poisoning effects in other conversion catalysts, for example, catalytic production of ammonia from N_2 and H_2 by Haber-Bosch [3, 25].

The total syngas fraction produced (H_2+CO), as well as oxygen demand, is a strong function of the coal rank. Lower rank coals have smaller crystalline sizes and larger pore volumes which increase the reactivity of coal with steam. Steam reactions with surface carbon are reaction rate limited and surface reactions between product gases and surface carbon/carbon-oxides may be limited by the resistance of pore diffusion [17, 24]. Higher reactivity also lowers steam consumption as steam is reacted more efficiently [17]. The total syngas fraction and syngas HHV are lower in the UA Fairbanks gasifier case because the oxidant used is blown ambient air. Components in air other than oxygen dilute the resulting syngas.

The staged-OMB gasifier outperforms the comparison gasifiers on total syngas production, syngas heating value, and H_2/CO ratio. Differences in CO_2 content in the product syngas and oxygen demand of the gasifier are marginal.

Facility Costs

COE for the 5 MW_e staged-OMB gasifier is high; however, fuel and variable O&M are low in comparison with the other reference cases. Fixed O&M and capital costs are the main contributors to the high COE.

North Dakota lignite was used as the fuel cost basis. Utilizing a lower sulfur coal could reduce the operating and capital cost of the acid gas removal unit. Assuming the sulfur load to the acid gas removal unit could be lowered to the sulfur load of the membrane-wall gasifier case (0.58 LTPD), the total plant cost would decrease by $4.5 MM and variable O&M would decrease by $0.24 MM per year, assuming there are no other changes to costs in other unit areas. The parasitic load of the acid gas removal unit would decrease by 0.14 MW which would increase the power sold to the grid. COE under these assumptions decreases by 15% to $239/MWh.

Fixed O&M and capital costs are high relative to the facility output. Economies of scale favor larger production facilities for both capital costs as well as labor. Other fixed operating costs including taxes, insurance, and maintenance materials are estimated as a percentage of the total plant cost, so they scale proportionally. Increasing the facility gross electrical output from 5.1 to 25 MW, and shifting from RICE gensets to a combustion turbine, decreases the COE from $281/MWh to $137/MWh.

While the staged-OMB gasifier does provide moderate cost savings in the gasification process area, the other process areas account for nearly 77% of the total plant cost. Significant capital savings in other process areas, specifically power production, air separation, and acid gas removal, would have the biggest impact on the economic feasibility of this process.

The reciprocating engines in the current facility design represent nearly 34% of the total plant cost. The reciprocating engines have high capital cost because of the substantial derate caused by the low heating value of the syngas and limited application experience. Significant cost escalators are incurred due to the "one-off" design and associated special equipment needs of the gensets. Combustion turbines for syngas (without significant natural gas blending) are currently unavailable at smaller scales. Advancements in turbine design at small scale, or an increase in throughput, would be required to switch from reciprocating engines to combustion turbines.

The ASU accounts for approximately 10% of the total plant cost. Due to the high inlet oxygen purity of the gasifier, cryogenic ASU was the only feasible technology selection.

Finally, the acid gas removal unit comprises 30% of the total plant cost. The capital and operating costs of the acid gas removal unit scales with the sulfur load to the unit. While regenerable sulfur removal by LO-CAT® significantly improved operating expenses compared to the membrane-wall gasifier design with non-regenerable adsorbent, additional process equipment is required. Reducing the sulfur load improves costs considerably. Switching to a lower sulfur coal would reduce the sulfur load; however, other process equipment could increase in size and cost to account for changes in heating value and changes to throughput required for equivalent production.

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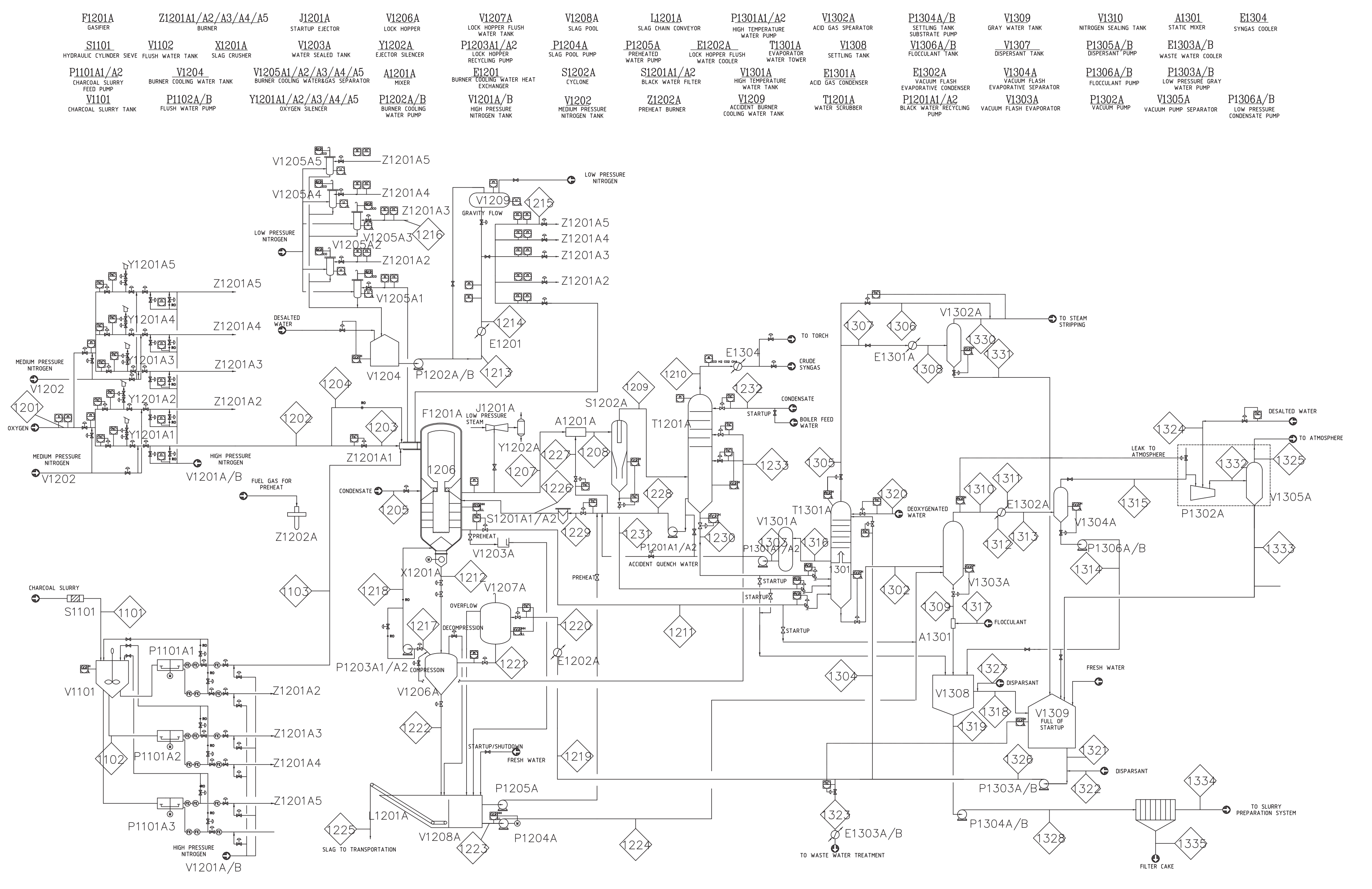
7 Appendices

- Appendix A – ECUST PFD
- Appendix B – ECUST H&MB
- Appendix C – Trimeric BFDs
- Appendix D – Detailed Size and Cost Tables
- Appendix E – Comparison to Membrane-Wall Gasifier
- Appendix F – Vendor Information

Appendix A

ECUST Process Flow Diagram



UNIVERSITY OF KENTUCKY
53TPD STAGED OPPOSED MULTI-BURNERS
COAL WATER SLURRY GASIFICATION UNIT
PROCESS FLOW DIAGRAM

East China University of Science and Technology	Page 1 of 1	Total 1 of 1	PFD

Appendix B

ECUST Heat and Material Balance Table

Gasification Heat and Material Balance

UK-CAER Staged OMB Gasifier

Stream No.		**1101**	**1102**	**1103**	**1201**	**1202**	**1203**	**1204**	**1205**	**1206**	**1207**	**1208**
Temperature	°C	50.00	50.00	50.00	25.00	25.00	25.00	25.00	138.30	1300.00	190.21	188.71
Pressure	Mpa(G)	0.00	0.10	3.00	3.50	2.50	2.50	2.50	3.00	2.00	1.98	1.91
Vapor Fraction	--				1.00	1.00	1.00	1.00	0.00	1.00	1.00	0.79
Liquid Fraction	--				0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.21
Vapor + Liquid Substream												
Molar Flow Rate												
H_2	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0017	75.6268	75.6280	75.6348
CO	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0020	92.2027	92.2045	92.2100
CO_2	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1305	36.0913	36.1156	36.1438
H_2S	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0306	2.0208	2.0304	2.0351
COS	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0934	0.0935	0.0935
CH_4	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.1000	0.1000	0.1000
N_2	kmol/h	0.0000	0.0000	0.0000	0.0199	0.0050	0.0009	0.0041	0.0001	0.8897	0.8897	0.8898
Ar	kmol/h	0.0000	0.0000	0.0000	0.2449	0.0612	0.0110	0.0502	0.0000	0.2449	0.2449	0.2449
NH_3	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0271	0.6250	0.6537	0.6977
HCN	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0002	0.0002
HCOOH	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0002	0.0000	0.0000
HCl	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	kmol/h	0.0000	0.0000	0.0000	65.9166	13.1833	2.3730	10.8103	0.0000	0.0000	0.0000	0.0000
H_2O	kmol/h	81.6993	32.6797	16.3399	0.0000	0.0000	0.0000	0.0000	257.5129	62.0907	344.7967	500.0956
Total flowrate	kmol/h	81.6993	32.6797	16.3399	66.1813	13.2363	2.3825	10.8537	257.7053	269.9857	552.7571	708.1705
Total flowrate	kg/h	1472.22	588.89	294.44	2119.59	423.92	76.31	347.61	4643.33	5563.50	10658.46	13459.42
Total flowrate	m^3/h				45.55	12.61	2.27	10.34	5.00	1685.98	981.96	1017.71
Mass Density	kg/m^3				46.53	33.61	33.61	33.61	928.67	3.30	10.85	13.23
Average molecular weight	g/gmol	18.02	18.02	18.02	32.02	32.02	32.02	32.02	18.02	20.61	19.28	19.01
Gas Phase												
Viscosity	cP				0.0206	0.0206	0.0206	0.0206	0.0000	0.0493	0.0199	0.0198
Thermal conductivity	kcal/m·h·°C				0.0227	0.0227	0.0227	0.0227	0.0000	0.1671	0.0428	0.0426
Liquid Phase												
Viscosity/cP	cP								0.1958	0.0000	0.0000	0.1417
Thermal conductivity	kcal/m·h·°C								0.5818	0.0000	0.0000	0.5593
Solid phase												
Ash	kg/h									236.65	11.84	11.88
Coal	kg/h	2208.33	883.33	441.67						0.00	0.00	0.00
Limestone/Additives	kg/h											
Total Solid flowrate	kg/h	2208.33	883.33	441.67						236.65	11.84	11.88
Overall												
Total mass flowrate	kg/h	3680.56	1472.22	736.11	2119.59	423.92	76.31	347.61	4643.33	5800.15	10670.29	13471.29
Total Enthalpy	W				3.42E-11	8.54E-12	1.54E-12	7.01E-12	-1.99E+07	-7.12E+06	-2.92E+07	-4.10E+07

Gasification Heat and Material Balance

UK-CAER Staged OMB Gasifier

Stream No.		**1209**	**1210**	**1211**	**1212**	**1213**	**1214**	**1215**	**1216**	**1217**	**1218**	**1219**
Temperature	°C	188.71	184.96	183.10	95.00	49.00	43.00	43.00	49.00	65.01	65.01	72.38
Pressure	Mpa(G)	1.91	1.86	1.98	1.98	1.40	1.35	0.95	0.00	1.98	2.51	0.60
Vapor Fraction	--	1.00	1.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Liquid Fraction	--	0.00	0.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Vapor + Liquid Substream												
Molar Flow Rate												
H_2	kmol/h	75.6278	75.6083	0.0129	0.0002	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.0001
CO	kmol/h	92.2043	92.1897	0.0103	0.0001	0.0000	0.0000	0.0000	0.0000	0.0001	0.0001	0.0000
CO_2	kmol/h	36.1150	36.0292	0.0500	0.0019	0.0000	0.0000	0.0000	0.0000	0.0011	0.0003	0.0021
H_2S	kmol/h	2.0304	2.0122	0.0080	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0013
COS	kmol/h	0.0935	0.0936	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	kmol/h	0.1000	0.1000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	kmol/h	0.8897	0.8896	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Ar	kmol/h	0.2449	0.2448	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NH_3	kmol/h	0.6535	0.5101	0.0759	0.0071	0.0000	0.0000	0.0000	0.0000	0.0205	0.0203	0.0330
HCN	kmol/h	0.0002	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCOOH	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCl	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	kmol/h	345.1476	298.4599	274.2415	72.0636	4399.5560	4399.5560	549.9445	549.9445	88.6861	88.6852	83.1273
Total flowrate	kmol/h	553.1069	506.1374	274.4495	72.1084	4399.5560	4399.5560	549.9445	549.9445	88.7605	88.7596	83.2127
Total flowrate	kg/h	10664.74	9816.36	4946.44	1299.79	79280.00	79280.00	9910.00	9910.00	1600.00	1600.00	1500.00
Total flowrate	m^3/h	1014.52	947.66	5.60	1.35	80.20	80.00	10.00	10.03	1.63	1.63	1.54
Mass Density	kg/m^3	10.51	10.36	883.02	961.88	988.50	991.00	991.00	988.50	981.42	981.42	976.99
Average molecular weight	g/gmol	19.28	19.39	18.02	18.03	18.02	18.02	18.02	18.02	18.03	18.03	18.03
Gas Phase												
Viscosity	cP	0.0198	0.0200	0.0000	0.0000						0.0000	0.0000
Thermal conductivity	kcal/m·h·°C	0.0426	0.0437	0.0000	0.0000						0.0000	0.0000
Liquid Phase												
Viscosity/cP	cP	0.0000	0.0000	0.1460	0.2973	0.5600	0.6220	0.6220	0.5600	0.4388	0.4388	0.3911
Thermal conductivity	kcal/m·h·°C	0.0000	0.0000	0.5618	0.6773	0.5550	0.5480	0.5480	0.5550	0.5596	0.5596	0.5711
Solid phase												
Ash	kg/h	1.13	0.00	22.49	202.40					0.00	0.00	0.00
Coal	kg/h	0.00	0.00	0.00	0.00					0.00	0.00	0.00
Limestone/Additives	kg/h											
Total Solid flowrate	kg/h	1.13	0.00	22.49	202.40					0.00	0.00	0.00
Overall												
Total mass flowrate	kg/h	10665.87	9816.36	4968.93	1502.18	79280.00	79280.00	9910.00	9910.00	1600.00	1600.00	1500.00
Total Enthalpy	W	-2.92E+07	-2.62E+07	-2.08E+07	-5.38E+06					-6.98E+06	-6.98E+06	-6.52E+06

Gasification Heat and Material Balance

UK-CAER Staged OMB Gasifier

Stream No.		1220	1221	1222	1223	1224	1225	1226	1227	1228	1229	1230
Temperature	°C	45.00	45.00	65.01	65.01	65.01	65.01	188.01	188.01	188.01	188.01	188.01
Pressure	Mpa(G)	0.55	0.00	0.00	0.00	0.40	0.00	2.01	2.54	2.54	2.54	1.86
Vapor Fraction	--	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Liquid Fraction	--	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Vapor + Liquid Substream												
Molar Flow Rate												
H_2	kmol/h	0.0001	0.0001	0.0001	0.0001	0.0001	0.0000	0.0125	0.0068	0.0193	0.0125	0.0025
CO	kmol/h	0.0000	0.0000	0.0001	0.0001	0.0001	0.0000	0.0100	0.0055	0.0155	0.0100	0.0020
CO_2	kmol/h	0.0021	0.0021	0.0008	0.0002	0.0002	0.0000	0.0513	0.0282	0.0795	0.0513	0.0101
H_2S	kmol/h	0.0013	0.0013	0.0000	0.0001	0.0001	0.0000	0.0084	0.0046	0.0130	0.0084	0.0017
COS	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0000	0.0001	0.0001	0.0000
Ar	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NH_3	kmol/h	0.0330	0.0330	0.0154	0.0144	0.0144	0.0008	0.0799	0.0439	0.1237	0.0799	0.0158
HCN	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001	0.0000	0.0000
HCOOH	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCl	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	kmol/h	83.1273	83.1273	66.5028	63.1216	63.1216	3.3805	282.6501	155.2988	437.9489	282.6501	55.9993
Total flowrate	kmol/h	83.2127	83.2127	66.5586	63.1746	63.1746	3.3834	282.8585	155.4133	438.2718	282.8585	56.0406
Total flowrate	kg/h	1500.00	1500.00	1199.79	1138.80	1138.80	60.99	5097.86	2800.96	7898.82	5097.86	1010.00
Total flowrate	m^3/h	1.51	1.51	1.22	1.16	1.16	0.06	5.81	3.19	9.00	5.81	1.15
Mass Density	kg/m^3	992.80	992.80	981.42	981.42	981.42	981.42	878.02	878.02	878.02	878.02	877.93
Average molecular weight	g/gmol	18.03	18.03	18.03	18.03	18.03	18.03	18.02	18.02	18.02	18.02	18.02
Gas Phase												
Viscosity	cP	0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Thermal conductivity	kcal/m·h·°C	0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Liquid Phase												
Viscosity/cP	cP	0.6532	0.6532	0.4388	0.4388	0.4388	0.4388	0.1423	0.1423	0.1423	0.1423	0.1423
Thermal conductivity	kcal/m·h·°C	0.5418	0.5418	0.5596	0.5596	0.5596	0.5596	0.5600	0.5600	0.5600	0.5600	0.5600
Solid phase												
Ash	kg/h	0.00	0.00	202.40	19.43	19.43	182.97	0.07	0.04	0.11	0.07	1.02
Coal	kg/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Limestone/Additives	kg/h											
Total Solid flowrate	kg/h	0.00	0.00	202.40	19.43	19.43	182.97	0.07	0.04	0.11	0.07	1.02
Overall												
Total mass flowrate	kg/h	1500.00	1500.00	1402.18	1158.23	1158.23	243.96	5097.93	2801.00	7898.93	5097.93	1011.02
Total Enthalpy	W	-6.58E+06	-6.58E+06	-4.98E+06	-4.94E+06	-4.94E+06	-4.25E+04	-2.15E+07	-1.18E+07	-3.33E+07	-2.15E+07	-4.26E+06

Gasification Heat and Material Balance

UK-CAER Staged OMB Gasifier

Stream No.		**1231**	**1232**	**1233**	**1301**	**1302**	**1303**	**1304**	**1305**	**1306**	**1307**	**1308**
Temperature	°C	188.71	138.30	131.14	133.55	133.55	131.14	72.38	131.14	131.14	131.14	75.00
Pressure	Mpa(G)	1.91	3.00	2.00	0.20	0.20	0.18	0.60	0.18	0.18	0.18	0.16
Vapor Fraction	--	0.00	0.00	0.00	1.00	0.00	0.00	0.00	1.00		1.00	0.00
Liquid Fraction	--	1.00	1.00	1.00	0.00	1.00	1.00	1.00	0.00	1.00	0.00	1.00
Vapor + Liquid Substream												
Molar Flow Rate												
H_2	kmol/h	0.0070	0.0023	0.0000	0.0224	0.0000	0.0000	0.0001	0.0224	0.0000	0.0224	0.0224
CO	kmol/h	0.0056	0.0028	0.0000	0.0179	0.0000	0.0000	0.0000	0.0179	0.0000	0.0179	0.0179
CO_2	kmol/h	0.0288	0.1782	0.0001	0.1131	0.0004	0.0001	0.0022	0.1316	0.0000	0.1316	0.0426
H_2S	kmol/h	0.0047	0.0418	0.0001	0.0216	0.0003	0.0001	0.0013	0.0279	0.0000	0.0279	0.0076
COS	kmol/h	0.0000	0.0005	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	kmol/h	0.0000	0.0001	0.0000	0.0001	0.0000	0.0000	0.0000	0.0001	0.0000	0.0001	0.0001
Ar	kmol/h	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NH_3	kmol/h	0.0441	0.0369	0.0265	0.0920	0.0755	0.0265	0.0342	0.1220	0.0000	0.1220	0.0113
HCN	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCOOH	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCl	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	kmol/h	154.9480	351.4236	95.6442	50.2163	434.9965	95.6442	86.2307	40.8187	0.0000	40.8187	40.7307
Total flowrate	kmol/h	155.0635	351.6862	95.6786	50.4835	435.0947	95.6786	86.3193	41.1407	0.0000	41.1407	41.0516
Total flowrate	kg/h	2794.67	6336.67	1723.76	912.50	7838.62	1723.76	1556.00	744.73	0.00	744.73	744.73
Total flowrate	m^3/h	3.19	6.82	1.85	556.92	8.42	1.85	1.59	483.68		483.68	1.76
Mass Density	kg/m^3	876.77	928.67	933.07	1.64	931.18	933.07	976.99	1.54		1.54	422.16
Average molecular weight	g/gmol	18.02	18.02	18.02	18.08	18.02	18.02	18.03	18.10		18.10	18.14
Gas Phase												
Viscosity	cP	0.0000	0.0000	0.0000	0.0139	0.0000		0.0000	0.0138		0.0138	0.0174
Thermal conductivity	kcal/m·h·°C	0.0000	0.0000	0.0000	0.0237	0.0000		0.0000	0.0235		0.0235	0.0396
Liquid Phase												
Viscosity/cP	cP	0.1417	0.1958	0.2074	0.0000	0.2034	0.2074	0.3911	0.0000	0.0000	0.0000	0.3803
Thermal conductivity	kcal/m·h·°C	0.5593	0.5818	0.5885	0.0000	0.5889	0.5885	0.5711	0.0000	0.0000	0.0000	0.5474
Solid phase												
Ash	kg/h	10.75		0.00	0.00	34.25		0.00	0.00		0.00	0.00
Coal	kg/h	0.00		0.00	0.00	0.00		0.00	0.00		0.00	0.00
Limestone/Additives	kg/h											
Total Solid flowrate	kg/h	10.75		0.00	0.00	34.25	0.00	0.00	0.00	0.00	0.00	0.00
Overall												
Total mass flowrate	kg/h	2805.42	6336.67	1723.76	912.50	7872.87	1723.76	1556.00	744.73	0.00	744.73	744.73
Total Enthalpy	W	-1.18E+07	-2.71E+07	-7.39E+06	-3.34E+06	-3.35E+07	-7.39E+06	-6.77E+06	-2.72E+06		-2.72E+06	-3.22E+06

Gasification Heat and Material Balance

UK-CAER Staged OMB Gasifier

Stream No.		1309	1310	1311	1312	1313	1314	1315	1316	1317	1318	1319
Temperature	°C	78.69	78.69	33.00	43.00	75.00	69.37	69.37	131.14	30.00	78.14	78.14
Pressure	Mpa(G)	-0.06	-0.06	0.40	0.25	-0.06	-0.07	-0.07	0.18	0.11	0.00	0.00
Vapor Fraction	--	0.00	1.00	0.00	0.00	0.01	0.00	1.00	0.00	0.00	0.00	0.00
Liquid Fraction	--	1.00	0.00	1.00	1.00	0.99	1.00	0.00	1.00	1.00	1.00	1.00
Vapor + Liquid Substream												
Molar Flow Rate												
H_2	kmol/h	0.0000	0.0009	0.0000	0.0000	0.0009	0.0000	0.0009	0.0000	0.0000	0.0000	0.0000
CO	kmol/h	0.0000	0.0008	0.0000	0.0000	0.0008	0.0000	0.0008	0.0000	0.0000	0.0000	0.0000
CO_2	kmol/h	0.0002	0.0874	0.0000	0.0000	0.0384	0.0002	0.0468	0.0001	0.0000	0.0002	0.0000
H_2S	kmol/h	0.0002	0.0367	0.0000	0.0000	0.0118	0.0002	0.0160	0.0001	0.0000	0.0002	0.0000
COS	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Ar	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NH_3	kmol/h	0.2415	0.2114	0.0000	0.0000	0.1310	0.1063	0.0382	0.0265	0.0000	0.2407	0.0109
HCN	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCOOH	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCl	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	kmol/h	757.4183	39.7764	2221.0314	2221.0314	39.7324	38.9826	0.7580	95.6442	1.2212	756.3139	34.2481
Total flowrate	kmol/h	757.7748	40.1136	2221.0314	2221.0314	40.0646	39.2125	0.8607	95.6786	1.2212	756.6692	34.2642
Total flowrate	kg/h	13652.97	725.30	40012.50	40012.50	725.30	708.37	16.94	1723.76	22.00	13633.05	617.34
Total flowrate	m^3/h	14.04	2579.37	40.21	40.35	30.15	0.73	71.21	1.85	0.02	14.01	0.63
Mass Density	kg/m^3	972.16	0.28	995.21	991.55	24.05	975.83	0.24	933.07	995.70	973.07	973.07
Average molecular weight	g/gmol	18.02	18.08	18.02	18.02	18.10	18.06	19.68	18.02	18.02	18.02	18.02
Gas Phase												
Viscosity	cP	0.0000	0.0118			0.0127	0.0000	0.0122	0.0000	0.0000	0.0000	0.0000
Thermal conductivity	kcal/m·h·°C	0.0000	0.0193			0.0194	0.0000	0.0189	0.0000	0.0000	0.0000	0.0000
Liquid Phase												
Viscosity/cP	cP	0.3615	0.0000	0.7645	0.6292	0.3787	0.4100	0.0000	0.2074		0.3628	0.3628
Thermal conductivity	kcal/m·h·°C	0.5702	0.0000	0.5329	0.5470	0.5634	0.5606	0.0000	0.5885		0.5745	0.5745
Solid phase												
Ash	kg/h	53.68	0.00	0.00	0.00	0.00	0.00	0.00	0.00		0.00	53.68
Coal	kg/h	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00		0.00	0.00
Limestone/Additives	kg/h											
Total Solid flowrate	kg/h	53.68	0.00	0.00	0.00	0.00	0.00	0.00	0.00		0.00	53.68
Overall												
Total mass flowrate	kg/h	13706.66	725.30	40012.50	40012.50	725.30	708.37	16.94	1723.76	22.00	13633.05	671.03
Total Enthalpy	W	-5.93E+07	-2.67E+06			-3.13E+06	-3.07E+06	-5.63E+04	-7.39E+06	-1.07E+04	-5.93E+07	-2.62E+06

Gasification Heat and Material Balance

UK-CAER Staged OMB Gasifier

Stream No.		**1320**	**1321**	**1322**	**1323**	**1324**	**1325**	**1326**	**1327**	**1328**	**1329 (Est.)**	**1330**
Temperature	°C	0.00	72.38	30.00	72.38	40.00	31.72	72.38	30.00	78.14	--	75.00
Pressure	Mpa(G)	2.00	0.00	0.11	0.60	0.60	0.00	0.60	0.11	0.35	--	0.16
Vapor Fraction	--	0.00	0.00	0.00	0.00	0.00	1.00	0.00	0.00	0.00		1.00
Liquid Fraction	--	1.00	1.00	1.00	1.00	1.00	0.00	1.00	1.00	1.00		0.00
Vapor + Liquid Substream												
Molar Flow Rate												
H_2	kmol/h	0.0000	0.0006	0.0000	0.0005	0.0000	0.0006	0.0006	0.0000	0.0000	0.0000	0.0221
CO	kmol/h	0.0000	0.0005	0.0000	0.0004	0.0000	0.0006	0.0005	0.0000	0.0000	0.0000	0.0177
CO_2	kmol/h	0.0000	0.0231	0.0000	0.0188	0.0000	0.0010	0.0231	0.0000	0.0000	0.0000	0.0335
H_2S	kmol/h	0.0000	0.0139	0.0000	0.0113	0.0000	0.0002	0.0139	0.0000	0.0000	0.0000	0.0039
COS	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0001
Ar	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NH_3	kmol/h	0.0000	0.3584	0.0000	0.2912	0.0000	0.0000	0.3584	0.0000	0.0109	0.0000	0.0001
HCN	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCOOH	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCl	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	kmol/h	0.0000	903.3494	0.5551	734.5465	66.6101	0.0001	903.9045	0.1110	34.2481	299.7933	0.0136
Total flowrate	kmol/h	0.0000	904.2784	0.5551	735.3015	66.6101	0.0026	904.8335	0.1110	34.2642	299.7933	0.0910
Total flowrate	kg/h	0.00	16300.61	10.00	13254.61	1200.00	0.07	16310.61	2.00	617.34	5400.86	2.40
Total flowrate	m^3/h	0.00	16.68	0.01	13.57	1.21	0.07	16.69	0.00	0.63		1.00
Mass Density	kg/m^3	0.00	976.99	995.70	976.99	995.21	1.12	976.99	995.81	973.07		2.39
Average molecular weight	g/gmol	0.00	18.02	18.02	18.03	18.02	28.01	18.03	18.02	18.02	18.02	26.37
Gas Phase												
Viscosity	cP	0.0000	0.0000		0.0000		0.0157		0.0000	0.0000		0.0174
Thermal conductivity	kcal/m·h·°C	0.0000	0.0000		0.0000		0.0344		0.0000	0.0000		0.0396
Liquid Phase												
Viscosity/cP	cP	0.0000	0.3911		0.3911	0.7645		0.3911		0.3628		0.0000
Thermal conductivity	kcal/m·h·°C	0.0000	0.5711		0.5711	0.5329		0.5711		0.5745		0.0000
Solid phase												
Ash	kg/h	0.00	0.00		0.00	0.00	0.00	0.00		53.68		0.00
Coal	kg/h	0.00	0.00		0.00	0.00	0.00	0.00		0.00		0.00
Limestone/Additives	kg/h											
Total Solid flowrate	kg/h	0.00	0.00		0.00	0.00	0.00	0.00	0.00	53.68		0.00
Overall												
Total mass flowrate	kg/h	0.00	16300.61	10.00	13254.61	1200.00	0.07	16310.61	2.00	671.03	5400.86	2.40
Total Enthalpy	W	0.00E+00	-6.55E+07	-1.47E+04	-5.76E+07	-5.29E+06	-1.39E+02	-7.09E+07	-8.81E+03	-2.62E+06	-2.35E+07	-5.10E+03

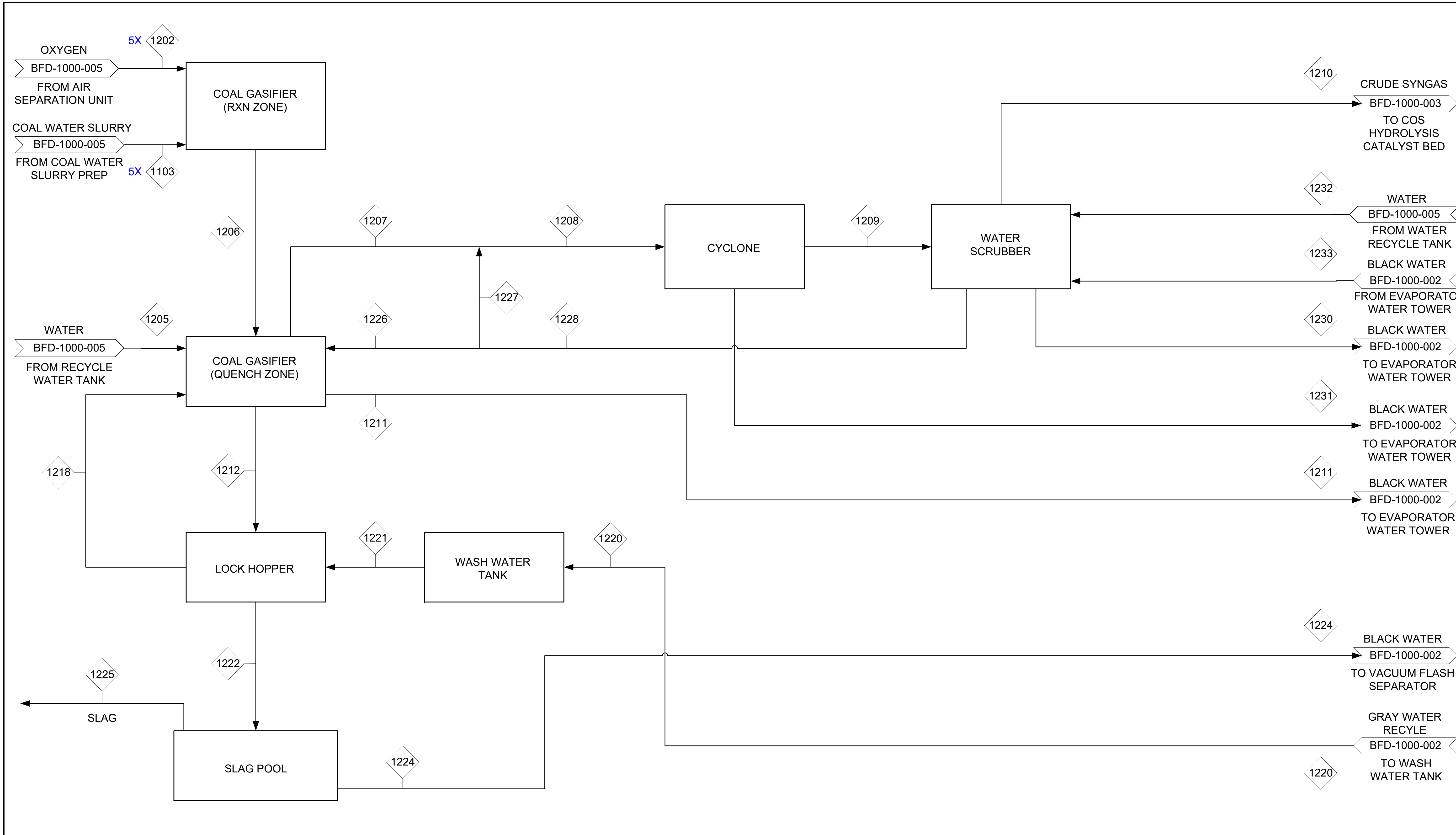
Gasification Heat and Material Balance

UK-CAER Staged OMB Gasifier

Stream No.		**1331**	**1332**	**1333**	**1334**	**1335**
Temperature	°C	75.00	31.72	31.72	78.88	78.88
Pressure	Mpa(G)	0.16	0.00	0.00	0.00	0.00
Vapor Fraction	--	0.00	0.00	0.00	0.00	0.00
Liquid Fraction	--	1.00	1.00	1.00	1.00	1.00
Vapor + Liquid Substream						
Molar Flow Rate						
H_2	kmol/h	0.0003	0.0009	0.0002	0.0000	0.0000
CO	kmol/h	0.0002	0.0008	0.0002	0.0000	0.0000
CO_2	kmol/h	0.0091	0.0147	0.0137	0.0000	0.0000
H_2S	kmol/h	0.0036	0.0100	0.0098	0.0000	0.0000
COS	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000
Ar	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000
NH_3	kmol/h	0.0113	0.0001	0.0001	0.0101	0.0008
HCN	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000
HCOOH	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000
HCl	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	kmol/h	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	kmol/h	40.7171	67.3360	67.3358	31.8114	2.4366
Total flowrate	kmol/h	40.9606	67.4387	67.4361	31.8264	2.4378
Total flowrate	kg/h	742.33	1216.94	1216.86	573.42	43.92
Total flowrate	m^3/h	0.76	1.29	1.22	0.59	0.05
Mass Density	kg/m^3	976.28	945.08	995.68	972.07	972.07
Average molecular weight	g/gmol	18.12	18.05	18.04	18.02	18.02
Gas Phase						
Viscosity	cP	0.0000	0.0157			
Thermal conductivity	kcal/m·h·°C	0.0000	0.0344			
Liquid Phase						
Viscosity/cP	cP	0.3803	0.7902	0.7902	0.3606	0.3606
Thermal conductivity	kcal/m·h·°C	0.5474	0.5169	0.5169	0.5703	0.5703
Solid phase						
Ash	kg/h	0.00	0.00	0.00	0.00	53.68
Coal	kg/h	0.00	0.00	0.00	0.00	0.00
Limestone/Additives	kg/h					
Total Solid flowrate	kg/h	0.00	0.00	0.00	0.00	53.68
Overall						
Total mass flowrate	kg/h	742.33	1216.94	1216.86	573.42	97.60
Total Enthalpy	W	-3.21E+06	-5.35E+06	-5.35E+06	-2.49E+06	-1.25E+05

Appendix C

Block Flow Diagrams



CONFIDENTIAL

FILENAME: BFD UKCAER OMB REV2.VSDX | DATE: 03/11/2021 | DRAWN BY: STEVEN FULK

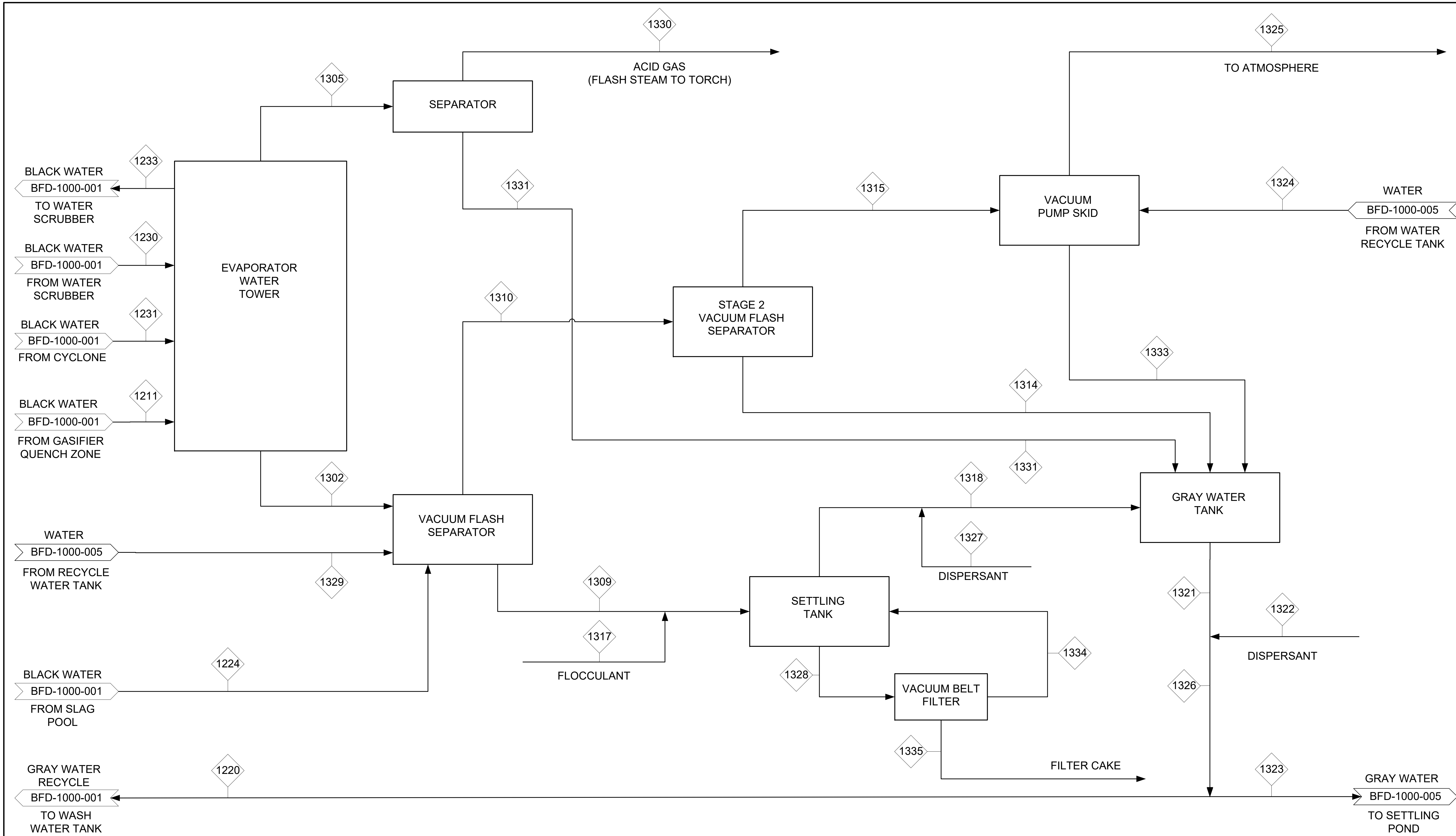
REVISIONS

REV.	DATE	DESCRIPTION	BY	CHECKED	APPROVED	APPROVED
0	03/11/2021	DRAFT BFD	SMF	AJS		
1	04/27/2021	REV0 COMMENTS – ADD BOP	SMF			

TRIMERIC CORPORATION
P.O. Box 826
Buda, Texas 78610

UK CAER – OMB GASIFIER
GASIFICATION
BLOCK FLOW DIAGRAM

CLIENT/SITE: UK CAER | JOB NUMBER: 50157.02
DRAWING NUMBER: BFD-1000-001 | SCALE: NONE



CONFIDENTIAL

FILENAME	DATE		DRAWN BY
BFD UKCAER OMB REV2.VSDX	03/11/2021		STEVEN FULK

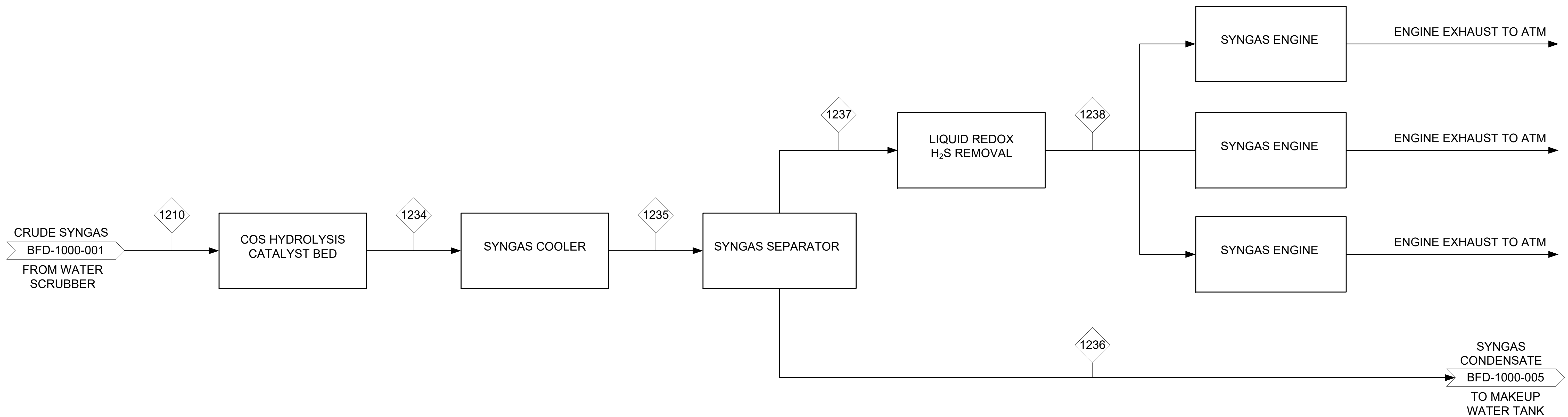
REVISIONS

REV.	DATE	DESCRIPTION	BY	CHECKED	APPROVED	APPROVED
0	03/11/2021	DRAFT BFD	SMF	AJS		
1	04/27/2021	REV0 COMMENTS – ADD BOP	SMF			

TRIMERIC CORPORATION
P.O. Box 826
Buda, Texas 78610

UK CAER – OMB GASIFIER
GASIFICATION
BLOCK FLOW DIAGRAM

CLIENT/SITE	JOB NUMBER
UK CAER	50157.02
DRAWING NUMBER	SCALE
BFD-1000-002	NONE



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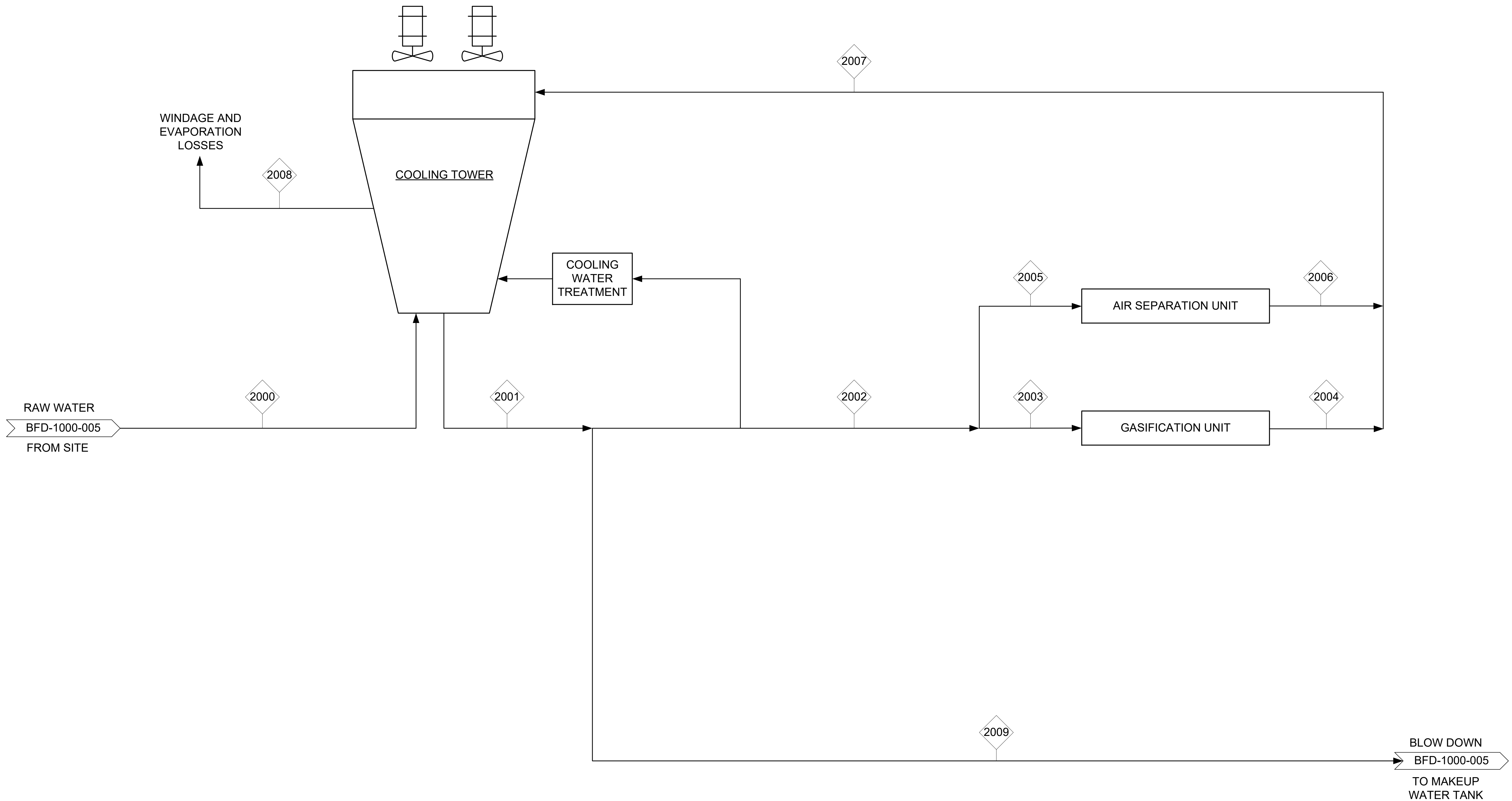
REVISIONS						
REV.	DATE	DESCRIPTION	BY	CHECKED	APPROVED	APPROVED
0	04/27/2021	DRAFT BFD	SMF			

FILENAME	DATE		DRAWN BY
BFD UKCAER OMB REV2.VSDX	04/27/2021		STEVEN FULK

TRIMERIC CORPORATION
P.O. Box 826
Buda, Texas 78610

UK CAER – OMB GASIFIER
BLOCK FLOW DIAGRAM

CLIENT/SITE	JOB NUMBER
UK CAER	50157.02
DRAWING NUMBER	SCALE
BFD-1000-003	NONE



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FILENAME	DATE		DRAWN BY
BFD UKCAER OMB REV2.VSDX	04/27/2021		STEVEN FULK

REVISIONS

REV.	DATE	DESCRIPTION	BY	CHECKED	APPROVED	APPROVED
0	04/27/2021	DRAFT BFD	SMF			

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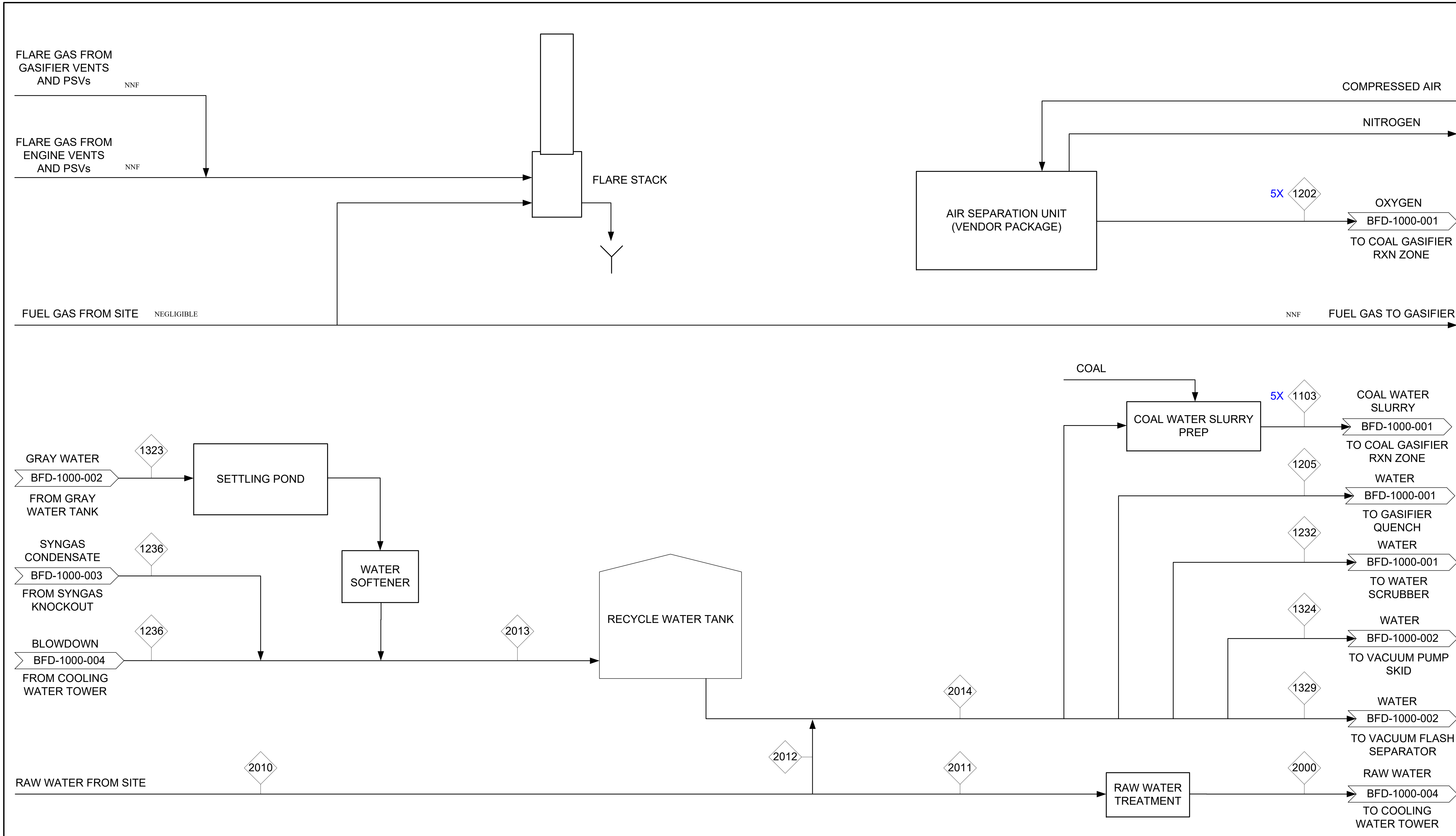
Buda, Texas 78610

UK CAER – OMB GASIFIER
COOLING WATER SYSTEM
BLOCK FLOW DIAGRAM

CLIENT/SITE	JOB NUMBER
UK CAER	50157.02
DRAWING NUMBER	SCALE
BFD-1000-004	NONE



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FILENAME	DATE		DRAWN BY
BFD UKCAER OMB REV2.VSDX	04/22/2021		STEVEN FULK

REVISIONS

REV.	DATE	DESCRIPTION	BY	CHECKED	APPROVED	APPROVED
0	04/27/2021	DRAFT BFD	SMF			

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Buda, Texas 78610

UK CAER – OMB GASIFIER GENERAL UTILITIES BLOCK FLOW DIAGRAM

CLIENT/SITE	JOB NUMBER
UK CAER	50157.02
DRAWING NUMBER	SCALE
BFD-1000-005	NONE

Appendix D

Equipment Size and Cost

Detailed Equipment Sizing - Heat Exchangers

UK-CAER Staged OMB Gasiifer

Tag	Equipment Name	MOC (Shell for HX)	MOC (Tubes for HX)	Design temp, C (Shell for HX)	Design temp, C (Tubes for HX)	Design Pressure (shell for HX), MPaG	Design Pressure (tubes for HX), MPaG	Cold temp in, C	Cold temp out, C	Hot temp in, C	Hot temp out, C	LMTD	Duty, kW	U-Value Btu/hr-ft2-F	U-Value W/m2-C	Area, m2 (Spec'd)	Type	Shell Diameter, mm	Tube Length, m	Purchased Cost (per unit)	Basis	Basis - CEPCI Index	Project CEPCI	Project Cost (per unit)	Number of Units	Purchased Cost (total)	Installed Capital Cost Multiplier	Installed Cost	Notes
E-1201	Burner Cooling Water Heat Exchanger	CS	304 SS	60	80	2.0	0.70	25	35	49	43	15.9	553	56	319	109	BEU	700	4.50	$74,500	Q1 2019	617.3	595.9	$72,000	1	$72,000	4	$288,000	Size Basis: Calculated duty from stream table, assume same overall U, resize bundle diameter and tube length for appropriate L/D with 10% overdesign Cost Basis: Aspen for Q1, 2019
E-1304	Syngas Cooler	CS	304 SS	310	80	3.50	0.70	25	35	185	40	58.6	4224	80	454	159	BEM	700	6.0	$75,100	Q1 2019	617.3	595.9	$73,000	1	$73,000	4	$292,000	Size Basis: Calculated duty from stream table, assume same overall U, resize bundle diameter and tube length for appropriate L/D with 10% overdesign Cost Basis: Aspen for Q1, 2019
E-1301A	Acid Gas Condenser	CS	304 SS	80	170	0.70	0.50	25	35	131	75	70.5	499	141	798	9	BEM	300	2.44	$16,100	Q1 2019	617.3	595.9	$16,000	1	$16,000	4	$64,000	Size Basis: Calculated duty from stream table, assume same overall U, resize bundle diameter and tube length for appropriate L/D with 10% overdesign Cost Basis: Aspen for Q1, 2019
E-1303A/B	Cooler	CS	304 SS	80	80	0.70	0.70	25	35	72	40	24.4	1158	45	256	186	BEM	850	4.50										Eliminated - Waste water is sent to a retention pond, cooling not required
E-1302A	Vacuum Flash Evaporative Condenser	CS	304 SS	80	120	0.70	-0.1 / 0.4	33	43	79	75	38.9	463	109	617	19	BEM	425	2.44	$22,300	Q1 2019	617.3	595.9	$22,000	1	$22,000	4	$88,000	Size Basis: Calculated duty from stream table, assume same overall U, resize bundle diameter and tube length for appropriate L/D with 10% overdesign Cost Basis: Aspen for Q1, 2019
E-1202A	Lock Hopper Flush Water Cooler	CS	304 SS	80	80	0.70	0.70	25	35	72	45	27.8	56	45	256	8	BEU	300	2.44	$24,200	Q1 2019	617.3	595.9	$23,000	1	$23,000	3	$69,000	Size Basis: Calculated duty from stream table, assume same overall U, resize bundle diameter and tube length for appropriate L/D with 10% overdesign Cost Basis: Aspen for Q1, 2019

Detailed Equipment Sizing - Vessel / Tower

UK-CAER Staged OMB Gasifier

Tag	Equipment Name	MOC	Design temp, C	Design Pressure, MPaG	Type	Vap Volumetric flow rate, m3/hr	Liq Volumetric flow rate, m3/hr	Volume, m3	Diameter, mm	Height, mm	Internals	Basis - Purchased Cost (per unit)	Basis	Basis - CEPCI Index	Cost Scale exponent	Purchased Cost (per unit) If Scaled from Past Project	Project CEPCI	Project Cost (per unit)	Number of Units	Purchased Cost (total)	Installed Capital Cost Multiplier	Installed Cost	Notes
T-1201A	Water Scrubber	13MnNiMoR+316L (4 mm) Cladded	250	3	Vertical	948	10	1.6	500-700	6050	4 trays	$56,700	Q1 2019	617.3			595.9	$55,000	1	$55,000	4	$220,000	Size basis: ~1.5 m/s, 2 minute liquid residence time at 50% holdup, 4 stages (2-ft tray spacing), sump diameter wider than column. Cost basis: from Aspen for Q1 2019.
V-1101	Charcoal Slurry Tank	Q235B	100	0.5	Vertical		3.7	58	4250	4100		$306,200	Q1 2016	536.4	0.57	$279,200	595.9	$310,000	1	$310,000	4	$1,240,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 slurry flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent; cost includes agitator, A-2201.
V-1202	Medium Pressure Nitrogen Tank	CS	70	7.2	Vertical			3.2	1000	4000													Excluded. N2 tank included in ASU supplier scope
V-1211	Fuel Gas Tank	304 SS	70	2.5	Vertical				600	1900													Excluded: Assume NG stored onsite in bullett containers.
V-1204	Burner Cooling Water Tank	CS	70	0.5	Vertical		80.2	10.6	1750	4400		$16,900	Q1 2016	536.4	0.57	$44,100	595.9	$49,000	1	$49,000	4	$196,000	Size basis: 2019 tank volume times the proportioin of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
V-1205A1/A2/A3/A4/A5	Burner Cooling Water & Gas Separator	CS	70	0.5	Vertical		10	0.16	450	1150		$5,200	Q1 2019	617.3			595.9	$5,000	5	$25,000	4	$100,000	Size basis: no vapor flow provided so assumed size based on liquid holdup; 2019 separator volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the separator dimensions. Cost basis: from Aspen for Q1 2019.
V-1209	Accident Burner Cooling Water Tank	CS	70	0.5	Horizontal		80	34	2800	5600		$23,300	Q1 2016	536.4	0.57	$60,700	595.9	$67,000	1	$67,000	4	$268,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
V-1203A	Water Sealed Tank	SS	250	4	Vertical			0.35	650	1050		$39,300	Q1 2016	536.4	0.57	$39,300	595.9	$44,000	1	$44,000	4	$176,000	Size basis: assumed same as in past project. Cost basis: same Q1 2016 cost as in past project.
V-1207A	Lock Hopper Flush Water Tank	CS	100	0.8	Vertical		1.5	1.41	950	2100		$17,300	Q1 2016	536.4	0.57	$14,700	595.9	$16,000	1	$16,000	4	$64,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
V-1213A	Raw Gas Separator	304 SS	70	2.50	Vertical	281	5.4	0.91	750	2400		$34,500	Q1 2019	617.3			595.9	$33,000	1	$33,000	4	$132,000	Size basis: Checked size with 4 methods - vapor velocity and L/D, ratioing volume based on liquid rates, Souders-Brown equation, and Symmetry software; 2019 separator volume times the proportion of the 2021 and 2019 liquid rates gave a reasonably sized vessel. Cost basis: from Aspen Q1 2019.
V-1201A/B	High Presure Nitrogen Tank	CS	70	7.2	Vertical			2	1000	2500													Excluded. N2 tank included in ASU supplier scope
V-1302A	Acid Gas Separator	CS	140	0.5	Vertical	1.0	0.8	0.32	550	1500		$6,100	Q1 2019	617.3			595.9	$6,000	1	$6,000	4	$24,000	Size basis: the vapor velocity in the 2019 project was very low, indicating sizing dominated by liquid rate; 2019 separator volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the separator dimensions. Cost basis: from Aspen for Q1 2019.
T-1301A	Evaporator Water Tower	CS	170	0.5	Vertical	484	10	1.5	350-700	7100	4 trays + 1 chimney	$34,900	Q1 2019	617.3			595.9	$34,000	1	$34,000	4	$136,000	Size basis: ~1.5 m/s, 2 minute liquid residence time at 50% holdup, 4 stages + 1 chimney tray (2-ft tray spacing), sump diameter wider than column. Cost basis: from Aspen for Q1 2019.
V-1303A	Vacuum Flash Evaporator	CS	100	-0.1 / 0.5	Vertical	2579	14	8.02	1800	3150		$33,700	Q1 2016	536.4	0.57	$28,100	595.9	$31,000	1	$31,000	4	$124,000	Size basis: Checked size with 3 methods - vapor velocity and L/D, ratioing volume based on liquid rates, and Souders-Brown equation; 2019 separator volume times the proportion of the 2021 and 2019 liquid rates gave a reasonably sized vessel. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
V-1304A	Vacuum Flash Evaporative Separator	CS	130	-0.1 / 0.5	Vertical		0.7	0.14	400	1200		$5,000	Q1 2019	617.3			595.9	$5,000	1	$5,000	4	$20,000	Size basis: the vapor velocity in the 2019 project was very low, indicating sizing dominated by liquid rate; 2019 separator volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the separator dimensions. Cost basis: from Aspen for Q1 2019.
V-1305A	Vacuum Pump Separator	CS	100	0.5	Vertical	0.07	1.22	0.65	650	1950		$12,800	Q1 2019	617.3			595.9	$13,000	1	$13,000	4	$52,000	Size basis: evaluated size using 30-minute liquid residence time and Souders-Brown equation for minimum vapor diameter; residence time and liquid flow rate gave reasonable size; assumed an L/D of 3. Cost basis: from Aspen for Q1 2019.
V-1307	Dispersant Tank	SS	70	0.5	Vertical		0.002	0.03	300	450		$6,100	Q1 2016	536.4	0.57	$2,600	595.9	$3,000	1	$3,000	4	$12,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
V-1306A/B	Flocculant Tank	SS	70	0.5	Vertical		0.02	3.68	1550	1950		$48,600	Q1 2016	536.4	0.57	$31,600	595.9	$35,000	1	$35,000	4	$140,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
V-1308	Settling Tank	Q235B	100	0.1	Vertical		14.6	29.3				$228,200	Q1 2016	536.4	0.57	$182,100	595.9	$202,000	1	$202,000	4	$808,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent; includes settling tank rake, A-2301 (cost as agitator).
V-1309	Gray Water Tank	CS	100	0.5	Vertical		16.7	23	3450	2500		$60,300	Q1 2016	536.4	0.57	$51,700	595.9	$57,000	1	$57,000	4	$228,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
V-1312	Filtrate Tank	CS	100	0.5	Vertical		0.59	47.56	3400	5400		$35,700	Q1 2016	536.4	0.57	$65,100	595.9	$72,000	1	$72,000	4	$288,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
V-1313	Filtrate Separator				Vertical																		Excluded from cost per 2019 project (vapor flow rate negligible).
V-1314	Vacuum Pump Separator	CS			Vertical																		Excluded from cost per 2019 project (vapor flow rate negligible).
S-1202A	Cyclone	CS + 316 (4mm)		2.5	Cyclone	1015	3.2	1.28	750	2900		$29,500	Q12019	617.3			595.9	$29,000	1	$29,000	4	$116,000	Size basis: used 2019 vapor velocity and 2021 vapor flow to estimate diameter; used 2019 L/D to estimate separator length. Cost basis: from Aspen for Q1 2019.
V-1102	Flush Water Tank																						Minimal cost. Did not include in capital cost.
V-1301A	High Temperature Water Tank			0.5			1.85	0.62	750	2250		$13,700	Q1 2019	617.3			595.9	$14,000	1	$14,000	4	$56,000	Size basis: assumed liquid residence time of 20-minutes (same as emergency cooling water tank) and L/D of 3. Cost Basis: from Aspen for Q1 2019.
V-1310	Nitrogen Sealing Tank																						Did not include in cost estimate (no cost in 2019 project either).
V-1210	Oxygen tank																						Did not include in cost estimate (no cost in 2019 project either).

Detailed Equipment Sizing - Pumps
UK CAER Staged OMB Gasifier

Tag	Equipment Name	MOC	Op temp, C (Shell for HX)	Pump Type	Sizing Flow Rate, m3/hr	Sizing Flow Rate, L/s	P1, MPaG	P2, MPaG	dP, Mpa	density, kg/m3	Calc Head, m	Calc Power, kW	Basis - Purchased Cost (per unit)	Basis	Basis - CEPCI Index	Cost Scale exponent	Purchased Cost (per unit) If Scaled from Past Project	Project CEPCI	Project Cost (per unit)	Number of Units	Purchased Cost (total)	Installed Capital Cost Multiplier	Installed Cost	Notes
P-1101A1/A2	Charcoal Slurry Feed Pump	316 SS	50	Diaphragm	1.8	0.5	0.1	3	2.9	1000	318	7.3	$18,900	Q1 2016	536.4	0.34	$13,000	595.9	$14,000	2	$28,000	4	$112,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (20%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
P-1101A3	Charcoal Slurry Feed Pump	316 SS	50	Diaphragm	0.9	0.3	0.1	3	2.9	1000	318	3.7	$4,100	Q1 2016	536.4	0.34	$2,700	595.9	$3,000	1	$3,000	4	$12,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (20%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
P-1202A/B	Burner Cooling Water Pump	CS	49	Centrifugal	96.2	26.7	0	1.4	1.4	989	155	81.4	$35,700	Q1 2019	617.3			595.9	$34,000	2	$68,000	4	$272,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (46%) and power calculations as in 2019 project. Cost basis: from Aspen Q1 2019
P-1203A1/A2	Lock Hopper Recycling Pump	CS	65	Centrifugal	2.1	0.6	1.98	2.51	0.53	981	59	0.7	$4,100	Q1 2016	536.4	0.34	$3,200	595.9	$4,000	2	$8,000	4	$32,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (46%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
P-1204A	Slag Pool Pump	CS	65	Centrifugal	1.4	0.4	0	0.4	0.4	981	45	0.5	$4,500	Q1 2016	536.4	0.34	$2,900	595.9	$3,000	2	$6,000	4	$24,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (30%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
P-1201A1/A2	Black Water Recycling Pump	Body:06Cr13Ni4Mo; Impeller:06Cr13Ni4Mo	188	Centrifugal	12.0	3.3	1.86	2.54	0.68	878	85	4.9	$6,400	Q1 2016	536.4	0.34	$5,400	595.9	$6,000	2	$12,000	4	$48,000	Size basis: simulation flow with factor of ~1.3 over design; same pump efficiency (46%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
P-1307A	Vaccum Condensate Pump	CS	69	Centrifugal	0.9	0.2	-0.067	0.3	0.367	976	41	0.3	$4,500	Q1 2016	536.4	0.34	$2,400	595.9	$3,000	2	$6,000	4	$24,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (30%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
P-1303A/B	Low Pressure Gray Water Pump	CS	72	Centrifugal	20.0	5.6	0	0.6	0.6	977	67	10.1	$4,900	Q1 2016	536.4	0.34	$5,800	595.9	$6,000	2	$12,000	4	$48,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (33%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
P-1304A/B	Settling Tank Substrate Pump	CS	78	Centrifugal	77.5	21.5	0	0.35	0.35	973	39	15.1	$9,300	Q1 2019	617.3			595.9	$9,000	2	$18,000	4	$72,000	Size basis: 2019 work showed actual pump flow significantly higher than simulation flow; scaled the pump flow for this application on same basis as past project and included factor of ~1.2 over design; same pump efficiency (50%) and power calculations as in 2019 project. Cost basis: from Aspen Q1 2019
P-1311	Filtrate Pump	CS	79	Centrifugal	41.7	11.6	0	0.3	0.3	972	34	7.0	$6,900	Q1 2019	617.3			595.9	$7,000	2	$14,000	4	$56,000	Size basis: 2019 work showed actual pump flow significantly higher than simulation flow; scaled the pump flow for this application on same basis as past project and included factor of ~1.3 over design; same pump efficiency (50%) and power calculations as in 2019 project. Cost basis: from Aspen Q1 2019
P-1306A/B	Flocculant Pump	CS	30	Metering Pump	0.027	0.01	0	0.11	0.11	996	12	0.002	$3,700	Q1 2016	536.4	0.34	$2,800	595.9	$3,000	2	$6,000	4	$24,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (50%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
P-1305A/B	Dispersant Pump	CS	30	Metering Pump	0.002	0.001	0	0.11	0.11	996	12	0.000	$2,600	Q1 2016	536.4	0.34	$1,500	595.9	$2,000	2	$4,000	4	$16,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (50%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
P-1102A/B	Flush Water Pump																							Minimal cost. Did not include in capital cost.
P-1205A	Preheated Water Pump																							Not included in cost estimate (for startup only).
P-1301A1/A2	High Temperature Water Pump	CS		Centrifugal	2.2	0.6	0.18	2.54	2.36	933	277	3.2	$48,700	Q1 2019	617.3			595.9	$48,000	2	$96,000	4	$384,000	Size basis: simulation flow with factor of ~1.2 overdesign; same pump efficiency (46%) as quench water recycle pump Cost basis: from Aspen for Q1 2019
P-1309A/B	V-1303 discharge pump	CS	79	Centrifugal	16.9	4.7	-0.056	0.2	0.256	972	29	4.0	$4,900	Q1 2019	617.3			595.9	$5,000	2	$10,000	4	$40,000	Size basis: simulation flow with factor of ~1.2 overdesign; same pump efficiency (30%) as vacuum condensate pump Cost basis: from Aspen Q1 2019

Detailed Equipment Sizing - Vacuum Pumps
UK-CAER Staged OMB Gasifier

Tag	Equipment Name	Equipment Type	MOC	Op temp, C	Pump Type	Sizing Flow Rate, m3/hr	Sizing Flow Rate, L/s	P1, MPaG	P2, MPaG	dP, Mpa	density, kg/m3	Calc Head, m	Calc Power, kW	Project Cost (per unit)	Number of Units	Purchased Cost (total)	Installed Capital Cost Multiplier	Installed Cost	Notes
P-1302A	Vacuum Pump	Vacuum Pump	CS	69	Liquid ring	84.7	23.5	-0.067	0	0.067	0.24		2.0	$10,000	1	$10,000	4	$40,000	Size basis: power estimated from correlations specifically for vacuum pumps. Cost basis: scaled from February 2020 vacuum pump quote on a $/HP basis.
P-1312	Filter Vacuum Pump	Vacuum Pump																	Excluded: vapor flow not provided and assumed negligible per 2019 project.

Tag	Equipment Name	Equipment Type	MOC	Design temp, C	Design Pressure, MPaG	Type	Vapor Volumetric flow rate, m3/hr	Liq Volumetric flow rate, m3/hr	Liq Residence Time, hr	Volume, m3	Diameter, mm	Height, mm	Internals	Basis - Purchased Cost (per unit)	Basis	Basis - CEPCI Index	Cost Scale exponent	Purchased Cost (per unit) If Scaled from Past Project	Project CEPCI	Project Cost (per unit)	Number of Units	Purchased Cost (total)	Installed Capital Cost Multiplier	Installed Cost	Notes
	Air Separation Unit	Other					9090							$2,316,171	2018	616.5			595.9	$2,239,000	1	$2,239,000	1.5	$3,358,500	ASU - Scaled on O2 demand from 2019. Same quote used which was in 2018 RMB. 2018 RMB to 2018 dollars. CEPCI to current index
F-1201A	Gasifier	Other									820-1820	4500								$610,000	1	$610,000	2	$1,220,000	Gasifier size and cost from ECUST. Size basis: 1.82 m outside metal shell diameter; 0.82 m inner diameter; 4.5 m straight height of gasification chamber (excluding dome and slag hole). Cost basis: includes nozzles, flow controller and lock hopper for slag discharge, and burners.
V-1206A	Lock Hopper	Other	CS + 316L	180	4	Vertical		2.9		0.29				$0	Q1 2016	536.4	0.00	$0	595.9	$0	1	$0	3	$0	Size basis: not estimated Cost basis: included with gasifier
L-1201A	Slag Chain Conveyor	Other						0.17						$65,400	Q1 2016	536.4	0.60	$34,000	595.9	$38,000	1	$38,000	2	$76,000	Size basis: not estimated Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.60 exponent.
V-1208A	Slag Pool	Other	CS	100	0.1	Horizontal		1.4	2	2.9				$14,500	Q1 2016	536.4	0.57	$6,900	595.9	$8,000	1	$8,000	3	$24,000	Size basis: 2019 slag pool volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the slag pool dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 slag pool volumes and a 0.57 exponent; cost includes agitator, A-2202.
A-1203	Slurry Tank Agitator	Other																							Integrated with V-1101.
A-1202	Slag Pool Agitator	Other																							Integrated with V-1208.
A-1302	Settling Tank Rake	Other																							Integrated with V-1308.
Y-1201A1/A2/A3/A4/A5	Oxygen Silencer	Other	316 SS		DN150																				Minimal cost. Did not include in capital cost.
S-1203	Natural Gas Filter	Other		15	100									$5,200	Q1 2016	536.4			595.9	$6,000	1	$6,000	2	$12,000	Assumed same cost as in 2019 project.
Z-1201A1/A2/A3/A4/A5	Burner	Other																			1	$0	2	$0	Burners included with ECUST gasifier cost.
Z-1203	Spark Generator	Other												$7,000	Q4 2018	616.5			595.9	$7,000	1	$7,000	2	$14,000	Assumed same cost as in 2019 project.
X-1201A	Slag Grinding Mill	Other		280	4																				Assumed included in gasifier cost.
A-1201A	Mixer	Other																							Minimal cost. Did not include in capital cost.
A-1301	Static Mixer	Other	316 SS	64		Mixer					100	1000													Minimal cost. Did not include in capital cost.
M-1301	Vacuum Belt Filter	Other						0.63						$75,500	Q1 2016	536.4	0.60	$138,200	595.9	$154,000	1	$154,000	3	$462,000	Size basis: No physical size. Cost basis: scaled Q1 2016 cost with 2021 and 2019 liquid flow rates and a 0.6 exponent. Q1 2016 cost based on a rotary drum filter estimate in Aspen.
Z-1202A	Preheat Burner	Other																							Minimal cost. Did not include in capital cost.
S-1101	Hydraulic Cylinder Sieve	Other																							Minimal cost. Did not include in capital cost.
S-1201A1/A2	Black Water Filter	Other																							Did not include in cost estimate (no cost in 2019 project either).
J-1201A	Startup Ejector	Other																							Not included in cost estimate (for startup only).
Y-1202A	Ejector Silencer	Other																							Not included in cost estimate (for startup only).
A-1205	Flocculant agitator	Other																							No cost in 2019 project; minimal cost and not included in 2021 project.
A-1204	Filtrate tank agitator	Other																							No cost in 2019 project; minimal cost and not included in 2021 project.
	Liquid Redox Equipment	Other																				$8,107,000	1.25	$10,133,750	Budget estimate from Merichem for a 1.8 LTPD liquid redox unit scaled to a 1.5 LTPD unit with a 0.6 exponent.
	COS Removal Equipment	Other												$37,200	Q1 2019	617.3			595.9	$36,000	2	$72,000	4	$288,000	Catalyst and vessel requirements from UNICAT vendor.
	Engines	Other												$2,431,850	2013	567.3			595.9	$2,554,000	3	$7,662,000	1.58	$12,104,616	recommendations, derated 50%.
	Water Treatment System	Other												$196,579	Q1 2016	536.4			595.9	$218,000	1	$218,000	2	$436,000	Sized based on 2% evaporation/windage loss and 0.4% blowdown of cooling water requirement
	Cooling Tower	Other												$213,256	2013	567.3			595.9	$224,000	1	$224,000	2	$448,000	Scaled based on cooling water rate from 4 cell vendor quote. Adjusted for temperature change.
	Water Softener	Other												$9,300	2021	595.9			595.9	$9,000	2	$18,000	2	$36,000	Sized based on 30% overdesign of full recycle flow to select model + spare.
	Flare Stack	Other												$51,803	2018	616.5			595.9	$50,000	1	$50,000	2	$100,000	Scaled on total flow rate based on vendor quote from 2018.

Appendix E

Comparison to Membrane-Wall Gasifier

Equipment Size and Cost Comparison
UK-CAER Staged OMB Gasifier (2021)
Membrane-Wall Gasifier (2019)

Project Year	Tag	Equipment Name	MOC (Shell for HX)	MOC (Tubes for HX)	Design temp, C (Shell for HX)	Design temp, C (Tubes for HX)	Design Pressure (shell for HX), MPaG	Design Pressure (tubes for HX), MPaG	Cold temp in, C	Cold temp out, C	Hot temp in, C	Hot temp out, C	LMTD	Duty, kW	U-Value Btu/hr-ft2-F	U-Value W/m2-C	Area, m2 (Spec'd)	Type	Shell Diameter, mm	Tube Length, m	Project Cost (per unit)	Number of Units	Purchased Cost (total)	Installed Capital Cost Multiplier	Installed Cost	Notes
2021	E-1201	Burner Cooling Water Heat Exchanger	CS	304 SS	60	80	2.0	0.70	25	35	49	43	15.9	553	56	319	109	BEU	700	4.50	$72,000	1	$72,000	4	$288,000	Size Basis: Calculated duty from stream table, assume same overall U, resize bundle diameter and tube length for appropriate L/D with 10% overdesign Cost Basis: Aspen for Q1, 2019
2019	E-2201	Burner cooling water heat exchanger	CS	304 SS	60	80	2.0	0.7									38	BEU	450	4.5	$41,000	1	$41,000	4	$164,000	
2021	E-1304	Syngas Cooler	CS	304 SS	310	80	3.50	0.70	25	35	185	40	58.6	4224	80	454	159	BEM	700	6.0	$73,000	1	$73,000	4	$292,000	Size Basis: Calculated duty from stream table, assume same overall U, resize bundle diameter and tube length for appropriate L/D with 10% overdesign Cost Basis: Aspen for Q1, 2019
2019	E-2203	Raw Gas Cooler	CS	304 SS	310	80	3.5	0.7									96	BEM	650	6.0	$56,000	1	$56,000	4	$224,000	
2021	E-1301A	Acid Gas Condenser	CS	304 SS	80	170	0.70	0.50	25	35	131	75	70.5	499	141	798	9	BEM	300	2.44	$16,000	1	$16,000	4	$64,000	Size Basis: Calculated duty from stream table, assume same overall U, resize bundle diameter and tube length for appropriate L/D with 10% overdesign Cost Basis: Aspen for Q1, 2019
2019	E-2301	HP Flash Gas Cooler	CS	304 SS	80	170	0.7	0.5									90	BEM	650	4.5	$61,000	1	$61,000	4	$244,000	
2021	E-1303A/B	Waste Water Cooler	CS	304 SS	80	80	0.70	0.70	25	35	72	40	24.4	1158	45	256	186	BEM	850	4.50						Eliminated - Waste water is sent to a retention pond, cooling not required
2019	E-2304	Waste Water Cooler	CS	304 SS	80	80	0.7	0.7									33	BEM	400	4.5						
2021	E-1302A	Vacuum Flash Evaporative Condenser	CS	304 SS	80	120	0.70	-0.1 / 0.4	33	43	79	75	38.9	463	109	617	19	BEM	425	2.44	$22,000	1	$22,000	4	$88,000	Size Basis: Calculated duty from stream table, assume same overall U, resize bundle diameter and tube length for appropriate L/D with 10% overdesign Cost Basis: Aspen for Q1, 2019
2019	E-2302	Vacuum Cooler	CS	304 SS	80	120	0.7	- 0.1 / 0.4									109	BEM	700	4.5	$53,000	1	$53,000	4	$212,000	
2021	E-1202A	Lock Hopper Flush Water Cooler	CS	304 SS	80	80	0.70	0.70	25	35	72	45	27.8	56	45	256	8	BEU	300	2.44	$23,000	1	$23,000	3	$69,000	Size Basis: Calculated duty from stream table, assume same overall U, resize bundle diameter and tube length for appropriate L/D with 10% overdesign Cost Basis: Aspen for Q1, 2019
2021		**TOTAL**																							**$801,000**	

Equipment Size and Cost Comparison
UK-CAER Staged OMB Gasifier (2021)
Membrane-Wall Gasifier (2019)

Project Year	Tag	Equipment Name	MOC	Design temp, C	Design Pressure, MPaG	Type	Vap flow rate, m3/hr	Liq flow rate, m3/hr	Volume, m3	Diameter, mm	Height, mm	Internals	Project Cost (per unit)	Number of Units	Purchased Cost (total)	Installed Capital Cost Multiplier	Installed Cost	Notes
2021	T-1201A	Water Scrubber	13MnNiMoR+316L (4 mm) Cladded	250	3	Vertical	948	10	1.6	500-700	6050	4 trays	$55,000	1	$55,000	4	$220,000	Size basis: ~1.5 m/s, 2 minute liquid residence time at 50% holdup, 4 stages (2-ft tray spacing), sump diameter wider than column. Cost basis: from Aspen for Q1 2019.
2019	T-2201	Water Scrubber	13MnNiMoR+316L (4 mm) Cladded	280	4	Vertical	584	12	4.8	1200	4200	none	$277,000	1	$277,000	4	$1,108,000	Past vessel cost estimated with diameter and height reversed.
2021	V-1101	Charcoal Slurry Tank	Q235B	100	0.5	Vertical		3.7	58	4250	4100		$310,000	1	$310,000	4	$1,240,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 slurry flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent; cost includes agitator, A-2201.
2019	V-2201	Slurry Tank	Q235B	100	0.5	Vertical		4.5	68	4500	4300		$352,000	1	$352,000	4	$1,408,000	
2021	V-1202	Medium Pressure Nitrogen Tank	CS	70	7.2	Vertical			3.2	1000	4000							Excluded. N2 tank included in ASU supplier scope
2019	V-2202	HP Nitrogen Gas Tank	CS	70	7.2	Vertical				1000	4000							
2021	V-1211	Fuel Gas Tank	304 SS	70	2.5	Vertical				600	1900							Excluded: Assume NG stored onsite in bullett containers.
2019	V-2204	Fuel Gas Tank	304 SS	70	2.5					600	1900							
2021	V-1204	Burner Cooling Water Tank	CS	70	0.5	Vertical		80.2	10.6	1750	4400		$49,000	1	$49,000	4	$196,000	Size basis: 2019 tank volume times the proportioin of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
2019	V-2205	Burner Cooling Water Tank	CS	70	0.5	Vertical		15	2.0	1000	2500		$19,000	1	$19,000	4	$76,000	
2021	V-1205A1/A2/A3/A4/A5	Burner Cooling Water & Gas Separator	CS	70	0.5	Vertical		10	0.16	450	1150		$5,000	5	$25,000	4	$100,000	Size basis: no vapor flow provided so assumed size based on liquid holdup; 2019 separator volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the separator dimensions. Cost basis: from Aspen for Q1 2019.
2019	V-2206	Burner Cooling Water Gas Separator	CS	70	0.5			15	0.25	500	1250		$6,000	1	$6,000	4	$24,000	
2021	V-1209	Accident Burner Cooling Water Tank	CS	70	0.5	Horizontal		80	34	2800	5600		$67,000	1	$67,000	4	$268,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
2019	V-2207	Emergency Burner Cooling Water Tank	CS	70	0.5	Horizontal		15	6	1600	3200		$27,000	1	$27,000	4	$108,000	
2021	V-1203A	Water Sealed Tank	SS	250	4	Vertical			0.35	650	1050		$44,000	1	$44,000	4	$176,000	Size basis: assumed same as in past project. Cost basis: same Q1 2016 cost as in past project.
2019	V-2211	Water Seal	SS	250	4	Vertical			0.35	650	1050		$45,000	1	$45,000	4	$180,000	
2021	V-1207A	Lock Hopper Flush Water Tank	CS	100	0.8	Vertical		1.5	1.41	950	2100		$16,000	1	$16,000	4	$64,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
2019	V-2209	Lock Hopper Wash Water Tank	CS	100	0.8	Vertical		2.1	2.0	1050	2300		$20,000	1	$20,000	4	$80,000	
2021	V-1213A	Raw Gas Separator	304 SS	70	2.5	Vertical	281	5.4	0.91	750	2400		$33,000	1	$33,000	4	$132,000	Size basis: Checked size with 4 methods - vapor velocity and L/D, ratioing volume based on liquid rates, Souders-Brown equation, and Symmetry software; 2019 separator volume times the proportion of the 2021 and 2019 liquid rates gave a reasonably sized vessel. Cost basis: from Aspen Q1 2019.
2019	V-2213	Raw Gas Separator	304 SS	70	3.5	Vertical	256	3.2	0.54	600	1900		$27,000	1	$27,000	4	$108,000	
2021	V-1201A/B	High Presure Nitrogen Tank	CS	70	7.2	Vertical			2	1000	2500							Excluded. N2 tank included in ASU supplier scope
2019	V-2216	HP Nitrogen Gas Tank	CS	70	7.2	Vertical				1000	2500							
2021	V-1302A	Acid Gas Separator	CS	140	0.5	Vertical	1.0	0.8	0.32	550	1500		$6,000	1	$6,000	4	$24,000	Size basis: the vapor velocity in the 2019 project was very low, indicating sizing dominated by liquid rate; 2019 separator volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the separator dimensions. Cost basis: from Aspen for Q1 2019.
2019	V-2301	HP Flash Separator	CS	70	3.5	Vertical	2.9	1.7	0.73	700	1900		$19,000	1	$19,000	4	$76,000	
2021	T-1301A	Evaporator Water Tower	CS	170	0.5	Vertical	484	10.3	1.5	350-700	7100	4 trays + 1 chimney	$34,000	1	$34,000	4	$136,000	Size basis: ~1.5 m/s, 2 minute liquid residence time at 50% holdup, 4 stages + 1 chimney tray (2-ft tray spacing), sump diameter wider than column. Cost basis: from Aspen for Q1 2019.
2019	V-2305	HP Flash Tank	CS	170	0.5	Vertical	1050	18	0.3	500	1500		$6,000	1	$6,000	4	$24,000	
2021	V-1303A	Vacuum Flash Evaporator	CS	100	-0.1 / 0.5	Vertical	2579	14	8.02	1800	3150		$31,000	1	$31,000	4	$124,000	Size basis: Checked size with 3 methods - vapor velocity and L/D, ratioing volume based on liquid rates, and Souders-Brown equation; 2019 separator volume times the proportion of the 2021 and 2019 liquid rates gave a reasonably sized vessel. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
2019	V-2306	Vacuum Flash Tank	CS	100	-0.1 / 0.5	Vertical	4915	20	11.00	2000	3500		$39,000	1	$39,000	4	$156,000	
2021	V-1304A	Vacuum Flash Evaporative Separator	CS	130	-0.1 / 0.5	Vertical		0.7	0.14	400	1200		$5,000	1	$5,000	4	$20,000	Size basis: the vapor velocity in the 2019 project was very low, indicating sizing dominated by liquid rate; 2019 separator volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the separator dimensions. Cost basis: from Aspen for Q1 2019.
2019	V-2303	Vacuum Flash Separator	CS	70	-0.1 / 0.5	Vertical	0.37	1.5	0.29	500	1500		$6,000	1	$6,000	4	$24,000	

Equipment Size and Cost Comparison
UK-CAER Staged OMB Gasifier (2021)
Membrane-Wall Gasifier (2019)

Project Year	Tag	Equipment Name	MOC	Design temp, C	Design Pressure, MPaG	Type	Vap flow rate, m3/hr	Liq flow rate, m3/hr	Volume, m3	Diameter, mm	Height, mm	Internals	Project Cost (per unit)	Number of Units	Purchased Cost (total)	Installed Capital Cost Multiplier	Installed Cost	Notes
2021	V-1305A	Vacuum Pump Separator	CS	100	0.5	Vertical	0.07	1.22	0.65	650	1950		$13,000	1	$13,000	4	$52,000	Size basis: evaluated size using 30-minute liquid residence time and Souders-Brown equation for minimum vapor diameter; residence time and liquid flow rate gave reasonable size; assumed an L/D of 3. Cost basis: from Aspen for Q1 2019.
2019	V-2304	Vacuum Separator																
2021	V-1307	Dispersant Tank	SS	70	0.5	Vertical		0.002	0.03	300	450		$3,000	1	$3,000	4	$12,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
2019	V-2308	Dispersant Tank	SS	70	0.5	Vertical		0.010	0.15	500	750		$7,000	1	$7,000	4	$28,000	
2021	V-1306A/B	Flocculant Tank	SS	70	0.5	Vertical		0.02	3.68	1550	1950		$35,000	1	$35,000	4	$140,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
2019	V-2311	Flocculant Tank	SS	70	0.5	Vertical		0.05	7.85	2000	2500		$56,000	1	$56,000	4	$224,000	
2021	V-1308	Settling Tank	Q235B	100	0.1	Vertical		14.6	29.3				$202,000	1	$202,000	4	$808,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent; includes settling tank rake, A-2301 (cost as agitator).
2019	V-2307	Settling Tank	Q235B	100	0.1	Vertical		21.8	43.5				$262,000	1	$262,000	4	$1,048,000	
2021	V-1309	Gray Water Tank	CS	100	0.5	Vertical		16.7	23	3450	2500		$57,000	1	$57,000	4	$228,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
2019	V-2310	Grey Water Tank	CS	100	0.5	Vertical		22.4	31	3800	2700		$69,000	1	$69,000	4	$276,000	
2021	V-1312	Filtrate Tank	CS	100	0.5	Vertical		0.59	47.56	3400	5400		$72,000	1	$72,000	4	$288,000	Size basis: 2019 tank volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the tank dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 tank volumes and a 0.57 exponent.
2019	V-2312	Filtrate Tank	CS	100	0.5	Vertical		0.21	17.19	2400	3800		$41,000	1	$41,000	4	$164,000	
2021	V-1313	Filtrate Separator				Vertical												Excluded from cost per 2019 project (vapor flow rate negligible).
2019	V-2313	Filtrate Separator	CS			Vertical												
2021	V-1314	Vacuum Pump Separator	CS			Vertical												Excluded from cost per 2019 project (vapor flow rate negligible).
2019	V-2314	Vacuum Pump Separator	CS			Vertical												
2021	S-1202A	Cyclone	CS + 316 (4mm)		2.5	Cyclone	1015	3.2	1.28	750	2900		$29,000	1	$29,000	4	$116,000	Size basis: used 2019 vapor velocity and 2021 vapor flow to estimate diameter; used 2019 L/D to estimate separator length. Cost basis: from Aspen for Q1 2019.
2019	S-2203	Cyclone Separator	CS + 316 (4mm)		2.85		1098		1.33	750	3000		$31,000	1	$31,000	4	$124,000	
2021	V-1102	Flush Water Tank																Minimal cost. Did not include in capital cost.
2021	V-1301A	High Temperature Water Tank			0.5	Vertical		1.85	0.62	750	2250		$14,000	1	$14,000	4	$56,000	Size basis: assumed liquid residence time of 20-minutes (same as emergency cooling water tank) and L/D of 3. Cost Basis: from Aspen for Q1 2019.
2021	V-1310	Nitrogen Sealing Tank																Did not include in cost estimate (no cost in 2019 project either).
2021	V-1210	Oxygen tank																Did not include in cost estimate (no cost in 2019 project either).
2019	V-2203	Oxygen tank																
2021		**TOTAL**															**$4,400,000**	

Project Year	Tag	Equipment Name	MOC	Op temp, C	Pump Type	Sizing Flow Rate, m3/hr	Sizing Flow Rate, L/s	P1, MPaG	P2, MPaG	dP, Mpa	density, kg/m3	Calc Head, m	Calc Power, kW	Project Cost (per unit)	Number of Units	Purchased Cost (total)	Installed Capital Cost Multiplier	Installed Cost	Notes
2021	P-1101A1/A2	Charcoal Slurry Feed Pump	316 SS	50	Diaphragm	1.8	0.5	0.1	3	2.9	1000	318	7.3	$14,000	2	$28,000	4	$112,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (20%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
2019	P-2201A	HP Coal Slurry Pump	316SS	30	Diaphragm	5.5	1.5	0	3.5	3.5	1000	384	26.7	$22,000	2	$44,000	4	$176,000	
2021	P-1101A3	Charcoal Slurry Feed Pump	316 SS	50	Diaphragm	0.9	0.3	0.1	3	2.9	1000	318	3.7	$3,000	1	$3,000	4	$12,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (20%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
2019	P-2201B	HP Coal Slurry Circulate Pump	CS	30	Centrifugal	3	0.8	0	0.3	0.3	1000	33	1.3	$5,000	2	$10,000	4	$40,000	
2021	P-1202A/B	Burner Cooling Water Pump	CS	49	Centrifugal	96.2	26.7	0	1.4	1.4	989	155	81.4	$34,000	2	$68,000	4	$272,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (46%) and power calculations as in 2019 project. Cost basis: from Aspen Q1 2019
2019	P-2202A/B	Burner Cooling Water Pump	CS	45	Centrifugal	15	4.2	0	1.5	1.5	1000	165	13.6	$20,000	2	$40,000	4	$160,000	
2021	P-1203A1/A2	Lock Hopper Recycling Pump	CS	65	Centrifugal	2.1	0.6	1.98	2.51	0.53	981	59	0.7	$4,000	2	$8,000	4	$32,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (46%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
2019	P-2203A/B	Lock Hopper Recycle Pump	CS	150	Centrifugal	4.5	1.3	3	3.5	0.5	1000	55	1.4	$5,000	2	$10,000	4	$40,000	
2021	P-1204A	Slag Pool Pump	CS	65	Centrifugal	1.4	0.4	0	0.4	0.4	981	45	0.5	$3,000	2	$6,000	4	$24,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (30%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
2019	P-2204	Slag Pool Pump	CS	55	Centrifugal	5.5	1.5	0	0.4	0.4	1000	44	2.0	$5,000	2	$10,000	4	$40,000	
2021	P-1201A1/A2	Black Water Recycling Pump	Body:06Cr13Ni4Mo; Impeller:06Cr13Ni4Mo	188	Centrifugal	12.0	3.3	1.86	2.54	0.68	878	85	4.9	$6,000	2	$12,000	4	$48,000	Size basis: simulation flow with factor of ~1.3 over design; same pump efficiency (46%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
2019	P-2205A/B	Quench Water Recycle Pump	Body:06Cr13Ni4Mo; Impeller:06Cr13Ni4Mo	206	Centrifugal	20	5.6	2.85	3.1	0.25	1000	27	3.0	$7,000	2	$14,000	4	$56,000	
2021	P-1307A	Vaccum Condensate Pump	CS	69	Centrifugal	0.9	0.2	-0.067	0.3	0.37	976	41	0.3	$3,000	2	$6,000	4	$24,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (30%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
2019	P-2303A/B	Vacuum Condensate Pump	CS	40	Centrifugal	5.5	1.5	-0.08	0.3	0.38	1000	42	1.9	$5,000	2	$10,000	4	$40,000	
2021	P-1303A/B	Low Pressure Gray Water Pump	CS	72	Centrifugal	20.0	5.6	0	0.6	0.6	977	67	10.1	$6,000	2	$12,000	4	$48,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (33%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
2019	P-2306A/B	LP Grey Water Pump	CS	61	Centrifugal	12	3.3	0	0.5	0.5	1000	55	5.1	$6,000	2	$12,000	4	$48,000	
2021	P-1304A/B	Settling Tank Substrate Pump	CS	78	Centrifugal	77.5	21.5	0	0.35	0.35	973	39	15.1	$9,000	2	$18,000	4	$72,000	Size basis: 2019 work showed actual pump flow significantly higher than simulation flow; scaled the pump flow for this application on same basis as past project and included factor of ~1.2 over design; same pump efficiency (50%) and power calculations as in 2019 project. Cost basis: from Aspen Q1 2019
2019	P-2308A/B	Settling Tank Pump	CS	61	Centrifugal	28	7.8	0	0.5	0.5	1000	55	7.8	$7,000	2	$14,000	4	$56,000	
2021	P-1311	Filtrate Pump	CS	79	Centrifugal	41.7	11.6	0	0.3	0.3	972	34	7.0	$7,000	2	$14,000	4	$56,000	Size basis: 2019 work showed actual pump flow significantly higher than simulation flow; scaled the pump flow for this application on same basis as past project and included factor of ~1.3 over design; same pump efficiency (50%) and power calculations as in 2019 project. Cost basis: from Aspen Q1 2019
2019	P-2311A/B	Filtrate Pump	CS	65	Centrifugal	15	4.2	0	0.3	0.3	1000	33	2.5	5000.0	2	$10,000	4	$40,000	
2021	P-1306A/B	Flocculant Pump	CS	30	Metering Pump	0.027	0.01	0	0.11	0.11	996	12	0.002	$3,000	2	$6,000	4	$24,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (50%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
2019	P-2310A/B	Flocculant Pump	CS	30	Metering	0.06	0.02	0	0.15	0.15	1000	16	0.005	$4,000	2	$8,000	4	$32,000	
2021	P-1305A/B	Dispersant Pump	CS	30	Metering Pump	0.002	0.001	0	0.11	0.11	996	12	0.000	$2,000	2	$4,000	4	$16,000	Size basis: simulation flow with factor of ~1.2 over design; same pump efficiency (50%) and power calculations as in 2019 project. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.34 exponent.
2019	P-2305A/B	Dispersant Pump	CS	30	Metering	0.012	0.003	0	0.15	0.15	1000	16	0.001	$3,000	2	$6,000	4	$24,000	
2021	P-1102A/B	Flush Water Pump																	Minimal cost. Did not include in capital cost.
2021	P-1205A	Preheated Water Pump																	Not included in cost estimate (for startup only).
2021	P-1301A1/A2	High Temperature Water Pump	CS		Centrifugal	2.2	0.6	0.18	2.54	2.36	933	277	3.2	$48,000	2	$96,000	4	$384,000	Size basis: simulation flow with factor of ~1.2 overdesign; same pump efficiency (46%) as quench water recycle pump Cost basis: from Aspen for Q1 2019
2021	P-1309A/B	V-1303 discharge pump	CS	79	Centrifugal	16.9	4.7	-0.056	0.2	0.256	972	29	4.0	$5,000	2	$10,000	4	$40,000	Size basis: simulation flow with factor of ~1.2 overdesign; same pump efficiency (30%) as vacuum condensate pump Cost basis: from Aspen Q1 2019
2021		TOTAL																$1,164,000	

Equipment Size and Cost Comparison
UK-CAER Staged OMB Gasifier (2021)
Membrane-Wall Gasifier (2019)

Project Year	Tag	Equipment Name	MOC	Op temp, C	Pump Type	Sizing Flow Rate, m3/hr	Sizing Flow Rate, L/s	P1, MPaG	P2, MPaG	dP, Mpa	density, kg/m3	Calc Head, m	Calc Power, kW	Project Cost (per unit)	Number of Units	Purchased Cost (total)	Installed Capital Cost Multiplier	Installed Cost	Notes
2021	P-1302A	Vacuum Pump	CS	69	Liquid ring	84.7	23.5	-0.067	0	0.067	0.24		2.0	$10,000	1	$10,000	4	$40,000	Size basis: power estimated from correlations specifically for vacuum pumps. Cost basis: scaled from February 2020 vacuum pump quote on a $/HP basis.
2019	P-2302A/B	Vacuum Pump																	Excluded P-2302A/B, E-2203, V-2304 from scope. Vapor flow rate negligible.
2021	P-1312	Filter Vacuum Pump																	Excluded: vapor flow not provided and assumed negligible per 2019 project.
2019	P-2312A/B	Filter Vacuum Pump																	Excluded P-2312A/B,V-2313,and V-2314. Vapor flow rate negligible.
2012		TOTAL																$40,000	

Equipment Size and Cost Comparison
UK-CAER Staged OMB Gasifier (2021)
Membrane-Wall Gasifier (2019)

Project Year	Tag	Equipment Name	MOC	Design temp, C	Design Pressure, MPaG	Type	Vapor Volumetric Flow Rate, m3/hr	Liq Volumetric Flow Fate, m3/hr	Volume, m3	Diameter, mm	Height, mm	Project Cost (per unit)	Number of Units	Purchased Cost (total)	Installed Capital Cost Multiplier	Installed Cost	Notes
2021		Air Separation Unit					9090					$2,239,000	1	$2,239,000	1.5	$3,358,500	RMB to 2018 dollars. CEPCI to current index
2019		Air Separation Unit					14673					$3,087,000	1	**$3,087,000**	**1.5**	**$4,630,500**	
2021	F-1201A	Gasifier								820-1820	4500	$610,000	1	$610,000	2	$1,220,000	Gasifier size and cost from ECUST. Size basis: 1.82 m outside metal shell diameter; 0.82 m inner diameter; 4.5 m straight height of gasification chamber (excluding dome and slag hole). Cost basis: includes nozzles, flow controller and lock hopper for slag discharge, and burners.
2019	F-2201	Gasifier										$840,000	1	$840,000	2	$1,680,000	
2021	V-1206A	Lock Hopper	CS + 316L	180	4	Vertical		2.9	0.29			$0	1	$0	3	$0	Size basis: not estimated Cost basis: included with gasifier
2019	V-2208	Lock Hopper	CS + 316L	180	4	Vertical		5.2	0.6			$17,000	1	$17,000	3	$51,000	
2021	L-1201A	Slag Chain Conveyor						0.17				$38,000	1	$38,000	2	$76,000	Size basis: not estimated Cost basis: scaled Q1 2016 cost with the 2021 and 2019 liquid flow rates and a 0.60 exponent.
2019	L-2201	Slag Conveyor						0.51				$75,000	1	$75,000	2	$150,000	
2021	V-1208A	Slag Pool	CS	100	0.1	Horizontal		1.4	2.9			$8,000	1	$8,000	3	$24,000	Size basis: 2019 slag pool volume times the proportion of the 2021 and 2019 liquid flow rates; used the 2019 L/D to estimate the slag pool dimensions. Cost basis: scaled Q1 2016 cost with the 2021 and 2019 slag pool volumes and a 0.57
2019	V-2210	Slag Pool	CS	100	0.1	Horizontal		5.2	10.4			$17,000	1	$17,000	3	$51,000	
2021	A-1203	Slurry Tank Agitator															Integrated with V-1101.
2019	A-2202	Slurry Tank Agitator															
2021	A-1202	Slag Pool Agitator															Integrated with V-1208.
2019	A-2202	Slag Pool Agitator															
2021	A-1302	Settling Tank Rake															Integrated with V-1308.
2019	A-2301	Settling Tank Rake															
2021	Y-1201A1/A2/A3/A4/A5	Oxygen Silencer	316 SS		DN150												Minimal cost. Did not include in capital cost.
2019	Z-2201	Silencer															
2021	S-1203	Natural Gas Filter		15	100							$6,000	1	$6,000	2	$12,000	Assumed same cost as in 2019 project.
2019	S-2201	Natural Gas Filter		15	100							$6,000	1	$6,000	2	$12,000	
2021	Z-1201A1/A2/A3/A4/A5	Burner											1	$0	2	$0	Burners included with ECUST gasifier cost.
2019	Z-2202	Burner										$70,000	1	$70,000	2	$140,000	
2021	Z-1203	Spark Generator										$7,000	1	$7,000	2	$14,000	Assumed same cost as in 2019 project.
2019	Z-2203	Spark Generator										$7,000	1	$7,000	2	$14,000	
2021	X-1201A	Slag Grinding Mill		280	4												Assumed included in gasifier cost.
2019	H-2201	Slag Grinding Mill		280	4												
2021	A-1201A	Mixer															Minimal cost. Did not include in capital cost.
2019	S-2202	Venturi Tube															
2021	A-1301	Static Mixer	316 SS	64		Mixer				100	1000						Minimal cost. Did not include in capital cost.
2019	H-2301	Pipeline Mixer	316 SS	64		Mixer				100	1000						
2021	M-1301	Vacuum Belt Filter						0.63				$154,000	1	$154,000	3	$462,000	Size basis: No physical size. Cost basis: scaled Q1 2016 cost with 2021 and 2019 liquid flow rates and a 0.6 exponent. Q1 2016 cost based on a rotary drum filter estimate in Aspen.
2019	M-2301	Vacuum Belt Filter						0.23				$87,000	1	$87,000	3	$261,000	
2021	Z-1202A	Preheat Burner															Minimal cost. Did not include in capital cost.
2021	S-1101	Hydraulic Cylinder Sieve															Minimal cost. Did not include in capital cost.
2021	S-1201A1/A2	Black Water Filter															Did not include in cost estimate (no cost in 2019 project either).
2019	S-2203	Black Water Filter															
2021	J-1201A	Startup Ejector															Not included in cost estimate (for startup only).
2019	J-2201	Startup Ejector															
2021	Y-1202A	Ejector Silencer															Not included in cost estimate (for startup only).
2019	V-2212	Ejector Silencer															
2021	A-1205	Flocculant agitator															No cost in 2019 project; minimal cost and not included in 2021 project.
2019	A-2303	Flocculant agitator															
2021	A-1204	Filtrate tank agitator															No cost in 2019 project; minimal cost and not included in 2021 project.
2019	A-2304	Filtrate tank agitator															
2021		Liquid Redox Equipment												$8,107,000	1.25	$10,133,750	Budget estimate from Merichem for a 1.8 LTPD liquid redox unit scaled to a 1.5 LTPD unit with a 0.6 exponent.
2021		COS Removal Equipment										$36,000	2	$72,000	4	$288,000	Catalyst and vessel requirements from UNICAT vendor.
2021		Engines										$2,554,000	3	$7,662,000	1.58	$12,104,616	From vendor quote. Output basis remains the same. Based on performance recommendations, derated 50%.
2019		Engines										$2,643,000	3	$7,929,000	1.58	$12,527,820	
2021		Water Treatment System										$218,000	1	$218,000	2	$436,000	Sized based on 2% evaporation/windage loss and 0.4% blowdown of cooling water requirement

Equipment Size and Cost Comparison
UK-CAER Staged OMB Gasifier (2021)
Membrane-Wall Gasifier (2019)

Project Year	Tag	Equipment Name	MOC	Design temp, C	Design Pressure, MPaG	Type	Vapor Volumetric Flow Rate, m3/hr	Liq Volumetric Flow Fate, m3/hr	Volume, m3	Diameter, mm	Height, mm	Project Cost (per unit)	Number of Units	Purchased Cost (total)	Installed Capital Cost Multiplier	Installed Cost	Notes
2019	PG-7005	Water Treatment System										$205,000	1	$205,000	2	$410,000	
2021		Cooling Tower										$224,000	1	$224,000	2	$448,000	Scaled based on cooling water rate from 4 cell vendor quote. Adjusted for temperature change.
2019	T-7006	Cooling Tower										$283,000	1	$283,000	2	$566,000	
2021		Water Softener										$9,000	2	$18,000	2	$36,000	Sized based on 30% overdesign of full recycle flow to select model + spare.
2021		Flare Stack										$50,000	1	$50,000	2	$100,000	Scaled on total flow rate based on vendor quote from 2018.
2019	F-7014	Flare Stack												$5,000	3	$15,000	
2021		**TOTAL**														**$28,712,866**	

Equipment Cost Comparison - 2019 Equipment Removed from 2021 Scope

UK-CAER Staged OMB Gasifier (2021)
Membrane-Wall Gasifier (2019)

Time Basis		November 2018		November 2018		November 2018	October 2020
2019 Tag	**2019 Equipment Name**	**2019 Project Cost (per unit)**	**2019 Number of Units**	**2019 Purchased Cost (total)**	**2019 Installed Capital Cost Multiplier**	**2019 Installed Cost**	**2019 Project Installed Cost Brought to Same Time Basis as 2021 Project**
Gasification Equipment							
E-2204	Lock Hopper Circulating Water Cooler	18,000	1	18,000	4	72,000	69,590
E-2303	Vacuum Cooler (Condensate Cooler P&ID)	0	0	0	0	0	0
D-2201	Steam Drum	0	0	0	0	0	0
P-2206A/B	BFW Recycle Pump	10,000	2	20,000	4	80,000	77,330
P-2307A/B	HP Grey Water Pump	61,000	2	122,000	4	488,000	471,690
Total		**89,000**		**160,000**		**640,000**	**618,610**
Acid Gas Recovery Equipment							
T-3001	Absorber	50,000	1	50,000	4	200,000	193,320
T-3002	Regeneration Tower	56,000	1	56,000	4	224,000	216,520
E-3002	Lean/ Rich Heat Exchanger	9,000	1	9,000	4	36,000	34,800
E-3001	Lean Amine Cooler	11,000	1	11,000	4	44,000	42,530
E-3004	Reboiler	15,000	1	15,000	4	60,000	58,000
E-3003	Acid Gas Cooler	16,000	1	16,000	4	64,000	61,860
V-3005	Purified Gas Separator	24,000	1	24,000	4	96,000	92,790
V-3001	Defoamer Tank	10,000	1	10,000	4	40,000	38,660
V-3004	Acid Gas Separator	28,000	1	28,000	4	112,000	108,260
V-3002	The Low Storage Tank	24,000	1	24,000	4	96,000	92,790
V-3003	Amine Tank	43,000	1	43,000	4	172,000	166,250
P-3001A/B	Circulating Pump	81,000	2	162,000	4	648,000	626,350
P-3301	Defoamer Pump	5,000	2	10,000	4	40,000	38,660
P-3002A/B	Reflux Pump	7,000	2	14,000	4	56,000	54,130
P-3003	Submerged Pump	3,000	2	6,000	4	24,000	23,200
	SulfaTreat Vessels	0	0	20,000	4	80,000	77,330
Total		**382,000**		**498,000**		**1,992,000**	**1,925,450**
Fischer-Tropsch Equipment							
R-6002	F-T Reactor	515,000	1	515,000	3	1,545,000	1,493,370
E-6001	Syngas Preheater	16,000	1	16,000	4	64,000	61,860
E-6002	Product Cooler	35,000	1	35,000	4	140,000	135,320
E-6003	Higher Pressure Gas Cooler	16,000	1	16,000	4	64,000	61,860
E-6004	Washing Oil Heater	55,000	1	55,000	4	220,000	212,650
E-6005	Washing Oil Cooler	25,000	1	25,000	4	100,000	96,660
V-6001	Syngas Knock-out Drum	33,000	1	33,000	4	132,000	127,590
D-6001	Steam Drum	23,000	1	23,000	4	92,000	88,930
V-6002	Hot High Pressure Separator	28,000	1	28,000	4	112,000	108,260
V-6003	Cold High Pressure Separator	26,000	1	26,000	4	104,000	100,520
V-6004	Cold Low Pressure Separator	19,000	1	19,000	4	76,000	73,460
V-6005	Hot Low Pressure Separator	8,000	1	8,000	4	32,000	30,930
V-6006	Synthetic Water Separator	9,000	1	9,000	4	36,000	34,800
V-6007	C5-C12 Tank	0	0	0	0	0	0

Equipment Cost Comparison - 2019 Equipment Removed from 2021 Scope
UK-CAER Staged OMB Gasifier (2021)
Membrane-Wall Gasifier (2019)

Time Basis		November 2018		November 2018		November 2018	October 2020
2019 Tag	2019 Equipment Name	2019 Project Cost (per unit)	2019 Number of Units	2019 Purchased Cost (total)	2019 Installed Capital Cost Multiplier	2019 Installed Cost	2019 Project Installed Cost Brought to Same Time Basis as 2021 Project
V-6008	C13+ Tank	0	0	0	0	0	0
V-6009	Blowdown Flash Tank	0	0	0	0	0	0
V-6010	C13+ Polluted Oil Tank	41,000	1	41,000	4	164,000	158,520
V-6011	C5-C12 Polluted Oil Tank	0	0	0	0	0	0
V-6012	Washing Oil Buffer Tank	0	0	0	0	0	0
R-6001A/B	Guard Beds	17,000	2	34,000	3	102,000	98,590
P-6001A/B	BFW Pump	8,000	2	16,000	4	64,000	61,860
P-6002A/B	C13+ Pump	0	0	0	0	0	0
P-6003A/B	Synthetic Water Pump	5,000	2	10,000	4	40,000	38,660
P-6004A/B	C5-C12 Pump	0	0	0	0	0	0
P-6005	C5-C12 Polluted Oil Pump	0	0	0	0	0	0
P-6006	C13+ Polluted Oil Pump	3,000	2	6,000	4	24,000	23,200
P-6007	Washing Oil Recycle Pump	0	0	0	0	0	0
Total		**882,000**		**915,000**		**3,111,000**	**3,007,040**
Engine Equipment							
	Turboexpander	160,000	1	160,000	3	480,000	463,960
Total		**160,000**		**160,000**		**480,000**	**463,960**
Balance of Plant							
V-7000	Deaerator	41,000	1	41,000	4	164,000	158,520
P-7001	Boiler Feed Water Pumps	64,000	2	128,000	4	512,000	494,890
V-7002	Blowdown Drum	9,000	1	9,000	4	36,000	34,800
E-7003	Blowdown Cooler	10,000	1	10,000	4	40,000	38,660
P-7004	Blowdown Transfer Pump	5,000	2	10,000	4	40,000	38,660
TK-7008	Recycle Water Tank	29,000	1	29,000	4	116,000	112,120
P-7011	Recycle Water Pump	5,000	2	10,000	4	40,000	38,660
TK-7009	C5-C12 Product Storage Tank	69,000	1	69,000	4	276,000	266,780
P-7012	Light HC Loading Pump	5,000	2	10,000	4	40,000	38,660
TK-7010	C13+ Product Storage Tank	69,000	1	69,000	4	276,000	266,780
P-7013	Heavy HC Loading Pump	5,000	2	10,000	4	40,000	38,660
P-7015	Feed Prep Booster Pump	52,000	2	104,000	4	416,000	402,100
Total		**363,000**		**499,000**		**1,996,000**	**1,929,290**
Total Equipment Removed		**1,876,000**		**2,232,000**		**8,219,000**	**7,944,350**

Equipment Cost Comparison - Equipment Added to 2021 Scope

UK-CAER Staged OMB Gasifier (2021)
Membrane-Wall Gasifier (2019)

Time Basis		October 2020		October 2020	October 2020	October 2020
2021 Tag	**2021 Equipment Name**	**2021 Project Cost (per unit)**	**2021 Number of Units**	**2021 Purchased Cost (total)**	**2021 Installed Capital Cost Multiplier**	**2021 Installed Cost**
Gasification Equipment						
Z-1202A	Preheat Burner	0	0	0	0	0
S-1101	Hydraulic Cylinder Sieve	0	0	0	0	0
V-1102	Flush Water Tank	0	0	0	0	0
P-1102A/B	Flush Water Pump	0	0	0	0	0
E-1202A	Lock Hopper Flush Water Cooler	23,000	1	23,000	3	69,000
P-1205A	Preheated Water Pump	0	0	0	0	0
V-1301A	High Temperature Water Tank	14,000	1	14,000	4	56,000
P-1301A1/A2	High Temperature Water Pump	48,000	2	96,000	4	384,000
P-1309A/B	V-1303 discharge pump	5,000	2	10,000	4	40,000
V-1310	Nitrogen Sealing Tank	0	0	0	0	0
Total		**90,000**		**143,000**		**549,000**
Sulfur Recovery Equipment						
	Liquid Redox Equipment	0	0	8,107,000	1.25	10,133,750
	COS Removal	36,000	2	72,000	4	288,000
Total		**36,000**		**8,179,000**		**10,421,750**
Balance of Plant Equipment						
	Water Softener	9,000	2	18,000	2	36,000
Total		**9,000**		**18,000**		**36,000**
Total Equipment Added		**135,000**		**8,340,000**		**11,006,750**

Equipment Cost Comparison - Cost of Equipment Used in Both Projects

UK-CAER Staged OMB Gasifier (2021)
Membrane Wall Gasifier (2019)

Time Basis		November 2018		November 2018	November 2018		October 2020
2019 Tag	2019 Equipment Name	2019 Project Cost (per unit)	2019 Number of Units	2019 Purchased Cost (total)	2019 Installed Capital Cost Multiplier	2019 Installed Cost	2019 Project Installed Cost Brought to Same Time Basis as 2021 Project
Air Separation Equipment							
	Air Separation Unit	3,087,000	1	3,087,000	1.5	4,630,500	4,475,770
Total		**3,087,000**		**3,087,000**		**4,630,500**	**4,475,770**
Gasification Equipment							
F-2201	Gasifier	840,000	1	840,000	2	1,680,000	1,623,860
E-2201	Burner cooling water heat exchanger	41,000	1	41,000	4	164,000	158,520
E-2203	Raw Gas Cooler	56,000	1	56,000	4	224,000	216,520
E-2301	HP Flash Gas Cooler	61,000	1	61,000	4	244,000	235,850
E-2304	Waste Water Cooler	0	0	0	0	0	0
E-2302	Vacuum Cooler	53,000	1	53,000	4	212,000	204,920
T-2201	Water Scrubber	277,000	1	277,000	4	1,108,000	1,070,980
V-2201	Slurry Tank	352,000	1	352,000	4	1,408,000	1,360,950
V-2202	HP Nitrogen Gas Tank	0	0	0	0	0	0
V-2204	Fuel Gas Tank	0	0	0	0	0	0
V-2205	Burner Cooling Water Tank	19,000	1	19,000	4	76,000	73,460
V-2206	Burner Cooling Water Gas Separator	6,000	1	6,000	4	24,000	23,200
V-2207	Emergency Burner Cooling Water Tank	27,000	1	27,000	4	108,000	104,390
V-2211	Water Seal	45,000	1	45,000	4	180,000	173,990
V-2208	Lock Hopper	22,000	1	22,000	3	66,000	63,790
V-2209	Lock Hopper Wash Water Tank	20,000	1	20,000	4	80,000	77,330
L-2201	Slag Conveyor	75,000	1	75,000	2	150,000	144,990
V-2210	Slag Pool	17,000	1	17,000	3	51,000	49,300
V-2213	Raw Gas Separator	27,000	1	27,000	4	108,000	104,390
V-2216	HP Nitrogen Gas Tank	0	0	0	0	0	0
V-2301	HP Flash Separator	19,000	1	19,000	4	76,000	73,460
V-2305	HP Flash Tank	6,000	1	6,000	4	24,000	23,200
V-2306	Vacuum Flash Tank	39,000	1	39,000	4	156,000	150,790
V-2303	Vacuum Flash Separator	6,000	1	6,000	4	24,000	23,200
V-2304	Vacuum Separator	0	0	0	0	0	0
V-2308	Dispersant Tank	7,000	1	7,000	4	28,000	27,060
V-2311	Flocculant Tank	56,000	1	56,000	4	224,000	216,520
V-2307	Settling Tank	262,000	1	262,000	4	1,048,000	1,012,980
V-2310	Grey Water Tank	69,000	1	69,000	4	276,000	266,780
V-2312	Filtrate Tank	41,000	1	41,000	4	164,000	158,520
V-2313	Filtrate Separator	0	0	0	0	0	0
V-2314	Vacuum Pump Separator	0	0	0	0	0	0
P-2201A	HP Coal Slurry Pump	22,000	2	44,000	4	176,000	170,120
P-2201B	HP Coal Slurry Circulate Pump	5,000	2	10,000	4	40,000	38,660
P-2202A/B	Burner Cooling Water Pump	20,000	2	40,000	4	160,000	154,650
P-2203A/B	Lock Hopper Recycle Pump	5,000	2	10,000	4	40,000	38,660
P-2204	Slag Pool Pump	5,000	2	10,000	4	40,000	38,660
P-2205A/B	Quench Water Recycle Pump	7,000	2	14,000	4	56,000	54,130
P-2303A/B	Vacuum Condensate Pump	5,000	2	10,000	4	40,000	38,660
P-2302A/B	Vacuum Pump	0	0	0	0	0	0
P-2306A/B	LP Grey Water Pump	6,000	2	12,000	4	48,000	46,400
P-2308A/B	Settling Tank Pump	7,000	2	14,000	4	56,000	54,130
P-2311A/B	Filtrate Pump	5,000	2	10,000	4	40,000	38,660
P-2312A/B	Filter Vacuum Pump	0	0	0	0	0	0
P-2310A/B	Flocculant Pump	4,000	2	8,000	4	32,000	30,930
P-2305A/B	Dispersant Pump	3,000	2	6,000	4	24,000	23,200
A-2201	Slurry Tank Agitator	0	0	0	0	0	0
A-2202	Slag Pool Agitator	0	0	0	0	0	0
A-2301	Settling Tank Rake	0	0	0	0	0	0
Z-2201	Silencer	0	0	0	0	0	0
S-2201	Natural Gas Filter	6,000	1	6,000	2	12,000	11,600
Z-2202	Burner	70,000	1	70,000	2	140,000	135,320
Z-2203	Spark Generator	7,000	1	7,000	2	14,000	13,530
H-2201	Slag Grinding Mill	0	0	0	0	0	0
S-2202	Venturi Tube	0	0	0	0	0	0
S-2203	Cyclone Separator	31,000	1	31,000	4	124,000	119,860
H-2301	Pipeline Mixer	0	0	0	0	0	0
M-2301	Vacuum Belt Filter	87,000	1	87,000	3	261,000	252,280
S-2203	Black water filter	0	0	0	0	0	0
J-2201	Startup ejector (not in 2019 list but on 2019 PFD)	0	0	0	0	0	0
V-2212	Ejector silencer (not in 2019 list but on 2019 PFD)	0	0	0	0	0	0
V-2203	Oxygen tank	0	0	0	0	0	0
A-2303	Flocculant agitator	0	0	0	0	0	0
A-2304	Filtrate tank agitator	0	0	0	0	0	0
Total		**2,738,000**		**2,832,000**		**9,206,000**	**8,898,400**
Engine Equipment							
	Engines	2,643,000	3	7,929,000	1.58	12,527,820	12,109,210
Total		**2,643,000**		**7,929,000**		**12,527,820**	**12,109,210**
Balance of Plant Equipment							
PG-7005	Water Treatment System	0	0	205,000	2	410,000	396,300
T-7006	Cooling Tower	283,000	1	283,000	2	566,000	547,090
F-7014	Flare Stack	0	1	5,000	3	15,000	14,500
Total		**283,000**		**493,000**		**991,000**	**957,890**
Total of Similar Equipment		**8,751,000**		**14,341,000**		**27,355,320**	**26,441,270**

Time Basis		October 2020		October 2020		October 2020
2021 Tag	2021 Equipment Name	2021 Project Cost (per unit)	2021 Number of Units	2021 Purchased Cost (total)	2021 Installed Capital Cost Multiplier	2021 Installed Cost
Air Separation Equipment						
	Air Separation Unit	2,239,000	1	2,239,000	1.5	3,358,500
Total		**2,239,000**		**2,239,000**		**3,358,500**
Gasification Equipment						
F-1201A	Gasifier	610,000	1	610,000	2	1,220,000
E-1201	Burner Cooling Water Heat Exchanger	72,000	1	72,000	4	288,000
E-1304	Syngas Cooler	73,000	1	73,000	4	292,000
E-1301A	Acid Gas Condenser	16,000	1	16,000	4	64,000
E-1303A/B	Waste Water Cooler	0	0	0	0	0
E-1302A	Vacuum Flash Evaporative Condenser	22,000	1	22,000	4	88,000
T-1201A	Water Scrubber	55,000	1	55,000	4	220,000
V-1101	Charcoal Slurry Tank	310,000	1	310,000	4	1,240,000
V-1202?	Medium Pressure Nitrogen Tank	0	0	0	0	0
V-1211	Fuel Gas Tank	0	0	0	0	0
V-1204	Burner Cooling Water Tank	49,000	1	49,000	4	196,000
V-1205A1/A2/A3	Burner Cooling Water & Gas Separator	5,000	5	25,000	4	100,000
V-1209	Accident Burner Cooling Water Tank	67,000	1	67,000	4	268,000
V-1203A	Water Sealed Tank	44,000	1	44,000	4	176,000
V-1206A	Lock Hopper	0	1	0	3	0
V-1207A	Lock Hopper Flush Water Tank	16,000	1	16,000	4	64,000
L-1201A	Slag Chain Conveyor	38,000	1	38,000	2	76,000
V-1208A	Slag Pool	8,000	1	8,000	3	24,000
V-1213A	Raw Gas Separator	33,000	1	33,000	4	132,000
V-1201A/B?	High Presure Nitrogen Tank	0	0	0	0	0
V-1302A	Acid Gas Separator	6,000	1	6,000	4	24,000
T-1301A	Evaporator Water Tower	34,000	1	34,000	4	136,000
V-1303A	Vacuum Flash Evaporator	31,000	1	31,000	4	124,000
V-1304A	Vacuum Flash Evaporative Separator	5,000	1	5,000	4	20,000
V-1305A	Vacuum Pump Separator	13,000	1	13,000	4	52,000
V-1307	Dispersant Tank	3,000	1	3,000	4	12,000
V-1306A/B	Flocculant Tank	35,000	1	35,000	4	140,000
V-1308	Settling Tank	202,000	1	202,000	4	808,000
V-1309	Gray Water Tank	57,000	1	57,000	4	228,000
V-1312	Filtrate Tank	72,000	1	72,000	4	288,000
V-1313	Filtrate Separator	0	0	0	0	0
V-1314	Vacuum Pump Separator	0	0	0	0	0
P-1101A1/A2	Charcoal Slurry Feed Pump	14,000	2	28,000	4	112,000
P-1101A3	Charcoal Slurry Feed Pump	3,000	1	3,000	4	12,000
P-1202A/B	Burner Cooling Water Pump	34,000	2	68,000	4	272,000
P-1203A1/A2	Lock Hopper Recycling Pump	4,000	2	8,000	4	32,000
P-1204A	Slag Pool Pump	3,000	2	6,000	4	24,000
P-1201A1/A2	Black Water Recycling Pump	6,000	2	12,000	4	48,000
P-1307A	Vaccum Condensate Pump	3,000	2	6,000	4	24,000
P-1302A	Vacuum Pump	10,000	1	10,000	4	40,000
P-1303A/B	Low Pressure Gray Water Pump	6,000	2	12,000	4	48,000
P-1304A/B	Settling Tank Substrate Pump	9,000	2	18,000	4	72,000
P-1311	Filtrate Pump	7,000	2	14,000	4	56,000
P-1312	Filter Vacuum Pump	0	0	0	0	0
P-1306A/B	Flocculant Pump	3,000	2	6,000	4	24,000
P-1305A/B	Dispersant Pump	2,000	2	4,000	4	16,000
A-1203	Slurry Tank Agitator	0	0	0	0	0
A-1202	Slag Pool Agitator	0	0	0	0	0
A-1302	Settling Tank Rake	0	0	0	0	0
Y-1201A1/A2/A3	Oxygen Silencer	0	0	0	0	0
S-1203	Natural Gas Filter	6,000	1	6,000	2	12,000
Z-1201A1/A2/A3	Burner	0	1	0	2	0
Z-1203	Spark Generator	7,000	1	7,000	2	14,000
X-1201A	Slag Grinding Mill	0	0	0	0	0
A-1201A	Mixer	0	0	0	0	0
S-1202A	Cyclone	29,000	1	29,000	4	116,000
A-1301	Static Mixer	0	0	0	0	0
M-1301	Vacuum Belt Filter	154,000	1	154,000	3	462,000
S-1201A1/A2	Black Water Filter	0	0	0	0	0
J-1201A	Startup Ejector	0	0	0	0	0
Y-1202A	Ejector Silencer	0	0	0	0	0
V-1210	Oxygen tank	0	0	0	0	0
A-1205	Flocculant agitator	0	0	0	0	0
A-1204	Filtrate tank agitator	0	0	0	0	0
Total		**2,176,000**		**2,287,000**		**7,664,000**
Engine Equipment						
	Engines	2,554,000	3	7,662,000	1.58	12,104,616
Total		**2,554,000**		**7,662,000**		**12,104,616**
Balance of Plant Equipment						
	Water Treatment System	218,000	1	218,000	2	436,000
	Cooling Tower	224,000	1	224,000	2	448,000
	Flare Stack	50,000	1	50,000	2	100,000
Total		**492,000**		**492,000**		**984,000**
Total of Similar Equipment		**7,461,000**		**12,680,000**		**24,111,116**

Equipment Cost Comparison - Highest to Lowest

UK-CAER Staged OMB Gasifier (2021)

Time Basis			October 2020		October 2020		October 2020
Tag	Equipment Name	Equipment Type	Project Cost (per unit)	Number of Units	Purchased Cost (total)	Installed Capital Cost Multiplier	Installed Cost
To be included	Engines	Other	$2,554,000	3	$7,662,000	1.58	$12,104,616
	Liquid Redox Equipment	Other			$8,107,000	1.25	$10,133,750
	Air Separation Unit	Other	$2,239,000	1	$2,239,000	1.5	$3,358,500
V-1101	Charcoal Slurry Tank	Vessel / Tower	$310,000	1	$310,000	4	$1,240,000
F-1201A	Gasifier	Other	$610,000	1	$610,000	2	$1,220,000
V-1308	Settling Tank	Vessel / Tower	$202,000	1	$202,000	4	$808,000
M-1301	Vacuum Belt Filter	Other	$154,000	1	$154,000	3	$462,000
	Cooling Tower	Other	$224,000	1	$224,000	2	$448,000
	Water Treatment System	Other	$218,000	1	$218,000	2	$436,000
P-1301A1/A2	High Temperature Water Pump	Pump	$48,000	2	$96,000	4	$384,000
E-1304	Syngas Cooler	Heat Exchanger	$73,000	1	$73,000	4	$292,000
E-1201	Burner Cooling Water Heat Exchanger	Heat Exchanger	$72,000	1	$72,000	4	$288,000
V-1312	Filtrate Tank	Vessel / Tower	$72,000	1	$72,000	4	$288,000
P-1202A/B	Burner Cooling Water Pump	Pump	$34,000	2	$68,000	4	$272,000
	COS Removal Equipment	Other	$36,000	2	$72,000	4	$288,000
V-1209	Accident Burner Cooling Water Tank	Vessel / Tower	$67,000	1	$67,000	4	$268,000
V-1309	Gray Water Tank	Vessel / Tower	$57,000	1	$57,000	4	$228,000
T-1201A	Water Scrubber	Vessel / Tower	$55,000	1	$55,000	4	$220,000
V-1204	Burner Cooling Water Tank	Vessel / Tower	$49,000	1	$49,000	4	$196,000
V-1203A	Water Sealed Tank	Vessel / Tower	$44,000	1	$44,000	4	$176,000
V-1306A/B	Flocculant Tank	Vessel / Tower	$35,000	1	$35,000	4	$140,000
T-1301A	Evaporator Water Tower	Vessel / Tower	$34,000	1	$34,000	4	$136,000
V-1213A	Raw Gas Separator	Vessel / Tower	$33,000	1	$33,000	4	$132,000
V-1303A	Vacuum Flash Evaporator	Vessel / Tower	$31,000	1	$31,000	4	$124,000
S-1202A	Cyclone	Vessel / Tower	$29,000	1	$29,000	4	$116,000
P-1101A1/A2	Charcoal Slurry Feed Pump	Pump	$14,000	2	$28,000	4	$112,000
V-1205A1/A2/A3/A4/A5	Burner Cooling Water & Gas Separator	Vessel / Tower	$5,000	5	$25,000	4	$100,000
	Flare Stack	Other	$50,000	1	$50,000	2	$100,000
E-1302A	Vacuum Flash Evaporative Condenser	Heat Exchanger	$22,000	1	$22,000	4	$88,000
P-1304A/B	Settling Tank Substrate Pump	Pump	$9,000	2	$18,000	4	$72,000
L-1201A	Slag Chain Conveyor	Other	$38,000	1	$38,000	2	$76,000
E-1202A	Lock Hopper Flush Water Cooler	Heat Exchanger	$23,000	1	$23,000	3	$69,000
E-1301A	Acid Gas Condenser	Heat Exchanger	$16,000	1	$16,000	4	$64,000
V-1207A	Lock Hopper Flush Water Tank	Vessel / Tower	$16,000	1	$16,000	4	$64,000
P-1311	Filtrate Pump	Pump	$7,000	2	$14,000	4	$56,000
V-1301A	High Temperature Water Tank	Vessel / Tower	$14,000	1	$14,000	4	$56,000
V-1305A	Vacuum Pump Separator	Vessel / Tower	$13,000	1	$13,000	4	$52,000
P-1201A1/A2	Black Water Recycling Pump	Pump	$6,000	2	$12,000	4	$48,000
P-1303A/B	Low Pressure Gray Water Pump	Pump	$6,000	2	$12,000	4	$48,000
P-1302A	Vacuum Pump	Vacuum Pump	$10,000	1	$10,000	4	$40,000
P-1309A/B	V-1303 discharge pump	Pump	$5,000	2	$10,000	4	$40,000
	Water Softener	Other	$9,000	2	$18,000	2	$36,000

Equipment Cost Comparison - Highest to Lowest

UK-CAER Staged OMB Gasifier (2021)

Tag	Equipment Name	Equipment Type	Project Cost (per unit)	Number of Units	Purchased Cost (total)	Installed Capital Cost Multiplier	Installed Cost
P-1203A1/A2	Lock Hopper Recycling Pump	Pump	$4,000	2	$8,000	4	$32,000
V-1208A	Slag Pool	Other	$8,000	1	$8,000	3	$24,000
V-1302A	Acid Gas Separator	Vessel / Tower	$6,000	1	$6,000	4	$24,000
P-1204A	Slag Pool Pump	Pump	$3,000	2	$6,000	4	$24,000
P-1307A	Vaccum Condensate Pump	Pump	$3,000	2	$6,000	4	$24,000
P-1306A/B	Flocculant Pump	Pump	$3,000	2	$6,000	4	$24,000
V-1304A	Vacuum Flash Evaporative Separator	Vessel / Tower	$5,000	1	$5,000	4	$20,000
P-1305A/B	Dispersant Pump	Pump	$2,000	2	$4,000	4	$16,000
Z-1203	Spark Generator	Other	$7,000	1	$7,000	2	$14,000
V-1307	Dispersant Tank	Vessel / Tower	$3,000	1	$3,000	4	$12,000
P-1101A3	Charcoal Slurry Feed Pump	Pump	$3,000	1	$3,000	4	$12,000
S-1203	Natural Gas Filter	Other	$6,000	1	$6,000	2	$12,000
V-1206A	Lock Hopper	Other	$0	1	$0	3	$0
Z-1201A1/A2/A3/A4/A5	Burner	Other		1	$0	2	$0
E-1303A/B	Waste Water Cooler	Heat Exchanger					
V-1202	Medium Pressure Nitrogen Tank	Vessel / Tower					
V-1211	Fuel Gas Tank	Vessel / Tower					
V-1201A/B	High Presure Nitrogen Tank	Vessel / Tower					
V-1313	Filtrate Separator	Vessel / Tower					
V-1314	Vacuum Pump Separator	Vessel / Tower					
P-1312	Filter Vacuum Pump	Vacuum Pump					
A-1203	Slurry Tank Agitator	Other					
A-1202	Slag Pool Agitator	Other					
A-1302	Settling Tank Rake	Other					
Y-1201A1/A2/A3/A4/A5	Oxygen Silencer	Other					
X-1201A	Slag Grinding Mill	Other					
A-1201A	Mixer	Other					
A-1301	Static Mixer	Other					
Z-1202A	Preheat Burner	Other					
S-1101	Hydraulic Cylinder Sieve	Other					
V-1102	Flush Water Tank	Vessel / Tower					
P-1102A/B	Flush Water Pump	Pump					
P-1205A	Preheated Water Pump	Pump					
S-1201A1/A2	Black Water Filter	Other					
J-1201A	Startup Ejector	Other					
Y-1202A	Ejector Silencer	Other					
V-1310	Nitrogen Sealing Tank	Vessel / Tower					
V-1210	Oxygen tank	Vessel / Tower					
A-1205	Flocculant agitator	Other					
A-1204	Filtrate tank agitator	Other					
Total			**7,596,000**		**21,020,000**		**35,117,866**

Appendix F

Vendor Data

CHC - 5

COS Hydrolysis

Activated Alumina Catalyst

UNICAT's CHC-5 is an activated alumina catalyst with special promotion for high activity in the selective hydrolysis of COS to H_2S at ambient or low temperatures in liquid hydrocarbon streams such as propylene, propane, butane or similar light hydrocarbons. CHC-5 is a spherically shaped product that is also capable of removing H_2S, CO_2, and H_2O in ethylene or hydrogen plants. Whatever the application, Unicat's broad catalyst and adsorbent product line has the right product to optimize a unit's operational performance and provide the best overall value.

	GENERAL SPECIFICATIONS
Shape	Sphere
Sizes	2.0 - 4.0 mm, 3.0 - 5.0 mm, 4.0 - 6.0 mm
Material Composition	> 93 % Al_2O_3, 7.0 % LOI < 2 % proprietary modifier, 0.2 % NaO, 0.05 % SiO_2, 0.05 % Fe_2O_3
Density (Sock Loaded)	> 47 - 49 lb/ft^3 (0.75 - 0.78 kg/l)
Crush Strength	> 30 - 40 lb-f (14.1 - 18.1 kg-f (Size Dependent)
Optimum Operating Temperature	0 - 250 ° F (-17.8 - 121.1 ° C)
COS/CO_2 Removal Efficiency	99 % conversion
Surface Area	> 300 m^2/g
Porosity	< 0.45 - 0.5 ml/g

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