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Coal-Based Power Plants of the Future: Electricity and Ammonia Polygeneration Final Pre-FEED Case Study Report

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1. Concept Background

Team AST developed a coal-based power system for application in the evolving bulk power system. Specifically, the design is a polygeneration plant for the co-production of electricity and ammonia from coal in a flexible system that can adapt to complex and shifting realities inherent in a modern electrical grid with significant renewable penetration. At a high level, the plant consists of two gasifier trains, a power island and two ammonia loops.

The general business philosophy of the polygeneration design centers on offering multiple potential revenue streams, including (1) commercial electricity available for sale to the grid, (2) salable ancillary services (e.g., capacity markets, frequency stability, voltage regulation, etc.), (3) and NH_3 for commercial delivery at or near retail (as opposed to wholesale) prices. By combining these three different revenue streams in a polygeneration facility that offers high operational flexibility, it is possible to modulate plant operations on a very short time scale to meet emerging market signals and opportunities. This ability to correctly match production to market demand will allow for optimization of plant profitability.

While the plant has the flexibility to operate at a multitude of operating points, the edges of the overall operating range are currently described by five specific operation modes, as seen in Table 1-1.

Table 1-1. Summary of Operating Modes

Operating Point	Net Export Power	Ammonia Production	Gasifier Operation	GT Operation	ST Operation	Ammonia Loop Operation
Balanced Generation, 3 GTs	48 MW	600 MTPD	100% of Capacity	Three Turbines @ 67% Capacity	Primary ST @ 86% load	Both Trains @ 100% Capacity
Balanced Generation, 2 GTs	51 MW	600 MTPD	100% of Capacity	Two Turbines @ 100% Capacity	Primary ST @ 91% Load	Both Trains @ 100% Capacity
Net Zero Power	0 MW	600 MTPD	66% of Capacity	One Turbine at 67% Capacity	Primary ST @ 40% Load	Both Trains @ 100% Capacity
High Electricity Production	82 MW	380 MTPD	100% of Capacity	Three Turbines @ 100% Capacity	Primary ST @ 88% Load	Both Trains @ 63% Capacity

Operating Point	Net Export Power	Ammonia Production	Gasifier Operation	GT Operation	ST Operation	Ammonia Loop Operation
Max Electricity Production	112 MW	59 MTPD	100% of Capacity	Three Turbines @ 100% Capacity	Primary ST @ 100% Load, Secondary ST @ 85% Load	Both Trains @ 10% Capacity

These operating modes define an operating window that provides the flexibility to modulate ammonia and net electricity production to meet market demand while enabling the two gasifier trains to operate at ~65% of capacity even in the absence of net electricity demand by the grid. This will allow the plant owner to choose operating points to maximize profitability while reducing the potential of being forced into outage by curtailed market demand.

The intent is to operate the polygeneration facility at a high service factor more typical of a chemical production facility rather than what would be normally expected from a pure, fossil fuel-based electricity generation facility that is subjected to forced curtailment. A number of design decisions have been made to support this goal. Multiple gasifier trains have been selected to provide the ability to run one train in conjunction with utilization of stored syngas (if required) while another train is shut down for maintenance. Additionally, if service is required to either the ammonia loop or power island, it can be performed at time when high demand is predicted for the alternative plant production capacity (i.e., if ammonia loop maintenance is required, it can be scheduled during a time of predicted high net energy demand, reducing the overall turndown for the plant as a whole).

The ability to perform opportunistic maintenance as described above, as well as the ability to match plant output to market demand, should support a service factor closer to the 96% metric achievable by chemical production facilities. However, it should be noted that the standard electrical generation service metric does not have as clear of a meaning for a polygeneration plant with multiple, viable operating points.

At the reference *Balanced Production, 3 GTs* operating point, ~71,000 kg/hour of as-received, Illinois #6 coal will be dried in a fluidized bed before passing to two SES U-Gas gasifiers, which will produce ~172,000 kg/hour of raw syngas. After passing through a water-gas shift reactor and various syngas cleaning and emission control technologies[1], the clean syngas will be nominally distributed to the ammonia train and power block. This *Balanced Production* syngas disposition will support net power generation of 48 MW and ammonia generation of 600 MTPD.

As detailed above in Table 1-1, the 600 MTPD represents the maximum ammonia production for this plant. By shifting to the *High Electricity Production* operating mode, it is possible to increase net power generation to 82 MW while reducing ammonia production to ~380 MTPD. This net power export can be further increased to 112 MW, as seen in the *Max Electricity Production*

[1] Details on ammonia removal, mercury removal, acid gas removal, CO_2 compression and drying, sulfur recovery, and tail gas treatment can be found in *Performance Results* (Section 3).

operating point. This 112 MW net power export relies on a deep turndown of the ammonia trains (both trains operating at 10% of maximum capacity).

To maximize cross-comparison against existing studies, and to maintain compliance with the site characteristics and conditions provided by the awarded contract, general siting characteristics and air composition will be adopted in accordance with those found in the June 2019 release of National Energy and Technology Laboratory's (NETL's) *Quality Guideline for Energy System Studies: Process Modeling Design Parameters*[2].

[2] These exhibits correspond with Site Conditions found in the June 2019 release of NETL's *Quality Guideline for Energy System Studies: Process Modeling Design Parameters*. However, some differences do exist. In these instances, this report has defaulted to the values in the latest QGESS document.

2. Process Description

The overall plant concept is an innovative application of largely established technology components to design and develop a coal-based, polygeneration system that contributes to the modern bulk power system. This coal-based system functions at a smaller scale than traditional baseload coal and natural gas power plants to provide both distributed, dispatchable power and ancillary services to power systems that are stressed due to lower inertia and a more complex, geographically disjointed topology.

To do so, the system's optimal scale must be centered on a design philosophy that values operational response, adaptability, and resiliency in addition to the standard concerns of availability and efficiency. Rather than relying on significant technological innovation that can be both risky and costly, the approach to meet the objectives of the Coal FIRST Initiative (CFI) is centered on intelligent and purposeful application of solid engineering and process development.

2.1 System Block Flow Diagram, Heat and Mass Balance, and Process Block Descriptions

At a high level, the conceptual design includes a coal gasifier to produce syngas that can be combusted in a conventional, combined cycle power block as well as used to produce ammonia for use as a chemical storage medium. The selected approach of creating a system based on established components and technology makes all of the major equipment of this design basis commercially available. A block flow diagram[3], with accompanying stream tables/heat and mass balance for the *Balanced Generation, 3 GTs* operating point, can be seen in Figure 2-1 and Table 2-1 followed by short process descriptions of each major subsystem. [4]

[3] The "Fluid Bed Dryer" that appears in the block flow diagram was previously referred to as the "Devolatilizer" in previous reports related to the polygeneration design effort. As the primary purpose of this vessel is drying, as opposed to devolatilization, this re-branding is appropriate

[4] Details for the other four operating points can be found in Appendix D.

Figure 2-1. Polygeneration Plant Block Flow Diagram

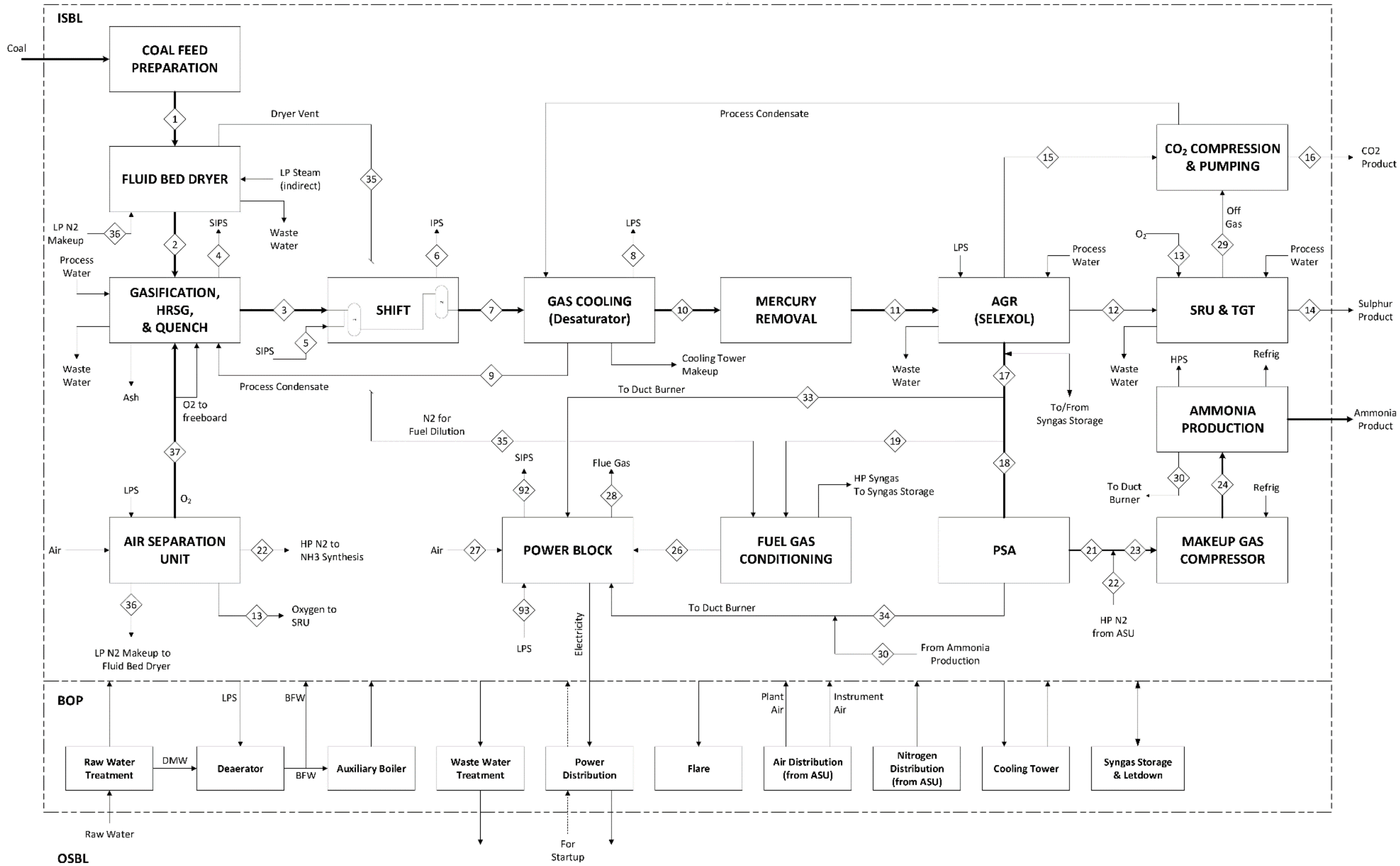

Table 2-1. Balanced Generation, 3 GTs Stream Table/Heat and Mass Balance

STREAM NUMBER		1		2		3		4		5		6	
STREAM NAME		AR coal feed		Dried Coal Feed		Scrubbed Syngas		Net Steam from Gasifier		Steam to Shift 1		Steam Raised in Shift	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	0.00	0.00	2552.07	40.23	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	28.013	0.00	0.00	0.00	0.00	24.50	0.39	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Monoxide	28.010	0.00	0.00	0.00	0.00	2596.96	40.94	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	44.010	0.00	0.00	0.00	0.00	998.24	15.74	0.00	0.00	0.00	0.00	0.00	0.00
Methane	16.042	0.00	0.00	0.00	0.00	91.95	1.45	0.00	0.00	0.00	0.00	0.00	0.00
Argon	39.948	0.00	0.00	0.00	0.00	7.67	0.12	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Sulfide	34.082	0.00	0.00	0.00	0.00	55.57	0.88	0.00	0.00	0.00	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	16.46	0.26	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h	62984		62984									
HHV / LHV (MW)		533.72	514.77	533.72	515.10	429.50	396.06	0.00	0.00	0.00	0.00	0.00	0.00
Temperature	°C	15.00		75.00		178.32		398.89		300.00		258.79	

STREAM NUMBER		1		2		3		4		5		6	
STREAM NAME		AR coal feed		Dried Coal Feed		Scrubbed Syngas		Net Steam from Gasifier		Steam to Shift 1		Steam Raised in Shift	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara	1.01		1.01		36.35		41.00		41.00		46.00	
Total Dry Molar Flow (kg.mol/h)			0.00		0.00	6343.43	100.00	0.00	0.00	0.00	0.00	0.00	0.00
Water	kg.mol/h	437.42		184.01		2511.81		2076.96		7003.44		2619.21	
Total Wet (kg.mol/h)		437.42		184.01		8855.24		2076.96		7003.44		2619.21	
Total Mass Flow (kg/h)			70,900		66,300		171,700		37,400		126,200		47,200
Molecular Weight							19.39		18.02		18.02		18.02

Table 2-1. Balanced Generation, 3 GTs Stream Table/Heat and Mass Balance

STREAM NUMBER		7		8		9		10		11		12	
STREAM NAME		Hot Syngas		LPS from Cooling Train		Process Cond rec'le to sc'ber		Cold Syngas		Syngas (Hg free)		Sour Gas to SRU	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	5050.10	57.12	0.00	0.00	0.11	3.08	5049.31	57.28	5049.31	57.28	16.11	6.68
Nitrogen	28.013	24.50	0.28	0.00	0.00	0.00	0.02	24.49	0.28	24.49	0.28	0.05	0.02
Carbon Monoxide	28.010	98.93	1.12	0.00	0.00	0.00	0.05	98.92	1.12	98.92	1.12	0.53	0.22
Carbon Dioxide	44.010	3496.14	39.54	0.00	0.00	1.04	29.72	3488.51	39.57	3488.51	39.57	169.19	70.12
Methane	16.042	91.95	1.04	0.00	0.00	0.01	0.23	91.89	1.04	91.89	1.04	0.75	0.31
Argon	39.948	7.67	0.09	0.00	0.00	0.00	0.02	7.67	0.09	7.67	0.09	0.03	0.01
Hydrogen Sulfide	34.082	55.45	0.63	0.00	0.00	0.11	3.20	54.63	0.62	54.63	0.62	54.48	22.58
Carbonyl Sulfide	60.076	0.12	0.00	0.00	0.00	0.00	0.00	0.12	0.00	0.12	0.00	0.04	0.02
Ammonia	17.031	16.46	0.19	0.00	0.00	2.23	63.68	0.12	0.00	0.12	0.00	0.12	0.05
Oxygen	31.999	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		431.44	367.48	0.00	0.00	0.01	0.01	431.36	367.41	431.36	367.41	1.51	1.29

STREAM NUMBER		7		8		9		10		11		12	
STREAM NAME		Hot Syngas		LPS from Cooling Train		Process Cond rec'le to sc'ber		Cold Syngas		Syngas (Hg free)		Sour Gas to SRU	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Temperature	°C	304.09		153.02		192.20		39.30		39.30		39.30	
Pressure	bara	34.95		5.16		44.35		34.05		34.05		34.05	
Total Dry Molar Flow (kg.mol/h)		8841.33	100.00	0.00	0.00	3.50	100.00	8815.67	100.00	8815.67	100.00	241.30	100.00
Water	kg.mol/h	7017.35		3219.35		1105.29		17.35		17.35		5.40	
Total Wet (kg.mol/h)		15858.68		3219.35		1108.78		8833.01		8833.01		246.70	
Total Mass Flow (kg/h)			297,900		58,000		20,000		171,100		171,100		9,500
Molecular Weight			18.78		18.02		18.04		19.37		19.37		38.37

Table 2.1. Balanced Generation, 3 GTs Stream Table/Heat and Mass Balance

STREAM NUMBER		13		14		15		16		17		18	
STREAM NAME		O2 to SRU		Sulphur Product		Feed to CO2 Comp		CO2 Product		total sweet syngas		syngas to PSA	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	0.00	0.00	15.75	0.50	26.44	0.79	5017.45	92.71	2560.36	92.71
Nitrogen	28.013	0.00	0.00	0.00	0.00	0.07	0.00	7.20	0.21	24.38	0.45	12.44	0.45
Carbon Monoxide	28.010	0.00	0.00	0.00	0.00	0.98	0.03	1.34	0.04	97.41	1.80	49.71	1.80
Carbon Dioxide	44.010	0.00	0.00	0.00	0.00	3143.40	99.40	3319.69	98.87	175.92	3.25	89.77	3.25
Methane	16.042	0.00	0.00	0.00	0.00	1.75	0.06	2.26	0.07	89.39	1.65	45.62	1.65
Argon	39.948	0.21	0.50	0.00	0.00	0.08	0.00	0.33	0.01	7.55	0.14	3.85	0.14
Hydrogen Sulfide	34.082	0.00	0.00	0.00	0.00	0.10	0.00	0.16	0.00	0.05	0.00	0.02	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.08	0.00	0.08	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	0.00	0.00	0.02	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	41.94	99.50	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	55.16	100.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		0.00	0.00	0.00	0.00	1.76	1.53	2.76	2.38	428.10	364.59	218.45	186.05
Temperature	°C	20.00		135.00		38.61		49.90		38.61		38.64	

STREAM NUMBER		13		14		15		16		17		18	
STREAM NAME		O2 to SRU		Sulphur Product		Feed to CO2 Comp		CO2 Product		total sweet syngas		syngas to PSA	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara	3.00		1.01		34.05		145.00		34.05		33.05	
Total Dry Molar Flow (kg.mol/h)		42.15	100.00	55.16	100.00	3162.21	100.00	3357.53	100.00	5412.15	100.00	2761.77	100.00
Water	kg.mol/h	0.00		0.00		5.49		0.00		0.00		0.00	
Total Wet (kg.mol/h)		42.15		55.16		3167.70		3357.53		5412.15		2761.77	
Total Mass Flow (kg/h)			1,400		1,800		138,500		146,500		23,000		11,700
Molecular Weight			32.04		32.07		43.73		43.62		4.25		4.25

Table 2-1. Balanced Generation, 3 GTs Stream Table/Heat and Mass Balance

STREAM NUMBER		19		20		21		22		23		24	
STREAM NAME		syngas to GT		Total Exhaust from GTs (x3)		PSA H2 to NH3 loop		N2 to NH3 loop		Feed to MUG Comp		Feed to NH3 loop	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	2457.09	92.71	0.00	0.00	2201.91	100.00	0.00	0.00	2201.91	75.00	2201.91	75.00
Nitrogen	28.013	11.94	0.45	20601.07	79.53	0.00	0.00	733.97	100.00	733.97	25.00	733.97	25.00
Carbon Monoxide	28.010	47.70	1.80	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	44.010	86.15	3.25	190.86	0.74	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methane	16.042	43.78	1.65	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Argon	39.948	3.70	0.14	343.69	1.33	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Sulfide	34.082	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	0.00	0.00	4768.06	18.41	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		209.64	178.54	0.00	0.00	174.81	147.89	0.00	0.00	174.81	147.89	174.81	147.89

STREAM NUMBER		19		20		21		22		23		24	
STREAM NAME		syngas to GT		Total Exhaust from GTs (x3)		PSA H2 to NH3 loop		N2 to NH3 loop		Feed to MUG Comp		Feed to NH3 loop	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Temperature	°C	38.64		422.40		38.64		40.00		37.93		123.30	
Pressure	bara	32.75		1.05		33.05		33.30		33.05		142.00	
Total Dry Molar Flow (kg.mol/h)		2650.38	100.00	25903.70	100.00	2201.91	100.00	733.97	100.00	2935.88	100.00	2935.88	100.00
Water	kg.mol/h	0.00		3658.76		0.00		0.00		0.00		0.00	
Total Wet (kg.mol/h)		2650.38		29562.46		2201.91		733.97		2935.88		2935.88	
Total Mass Flow (kg/h)		11,266		817,700		4,400		20,600		25,000		25,000	
Molecular Weight		4.25		27.66		2.02		28.01		8.52		8.52	

Table 2-1. Balanced Generation, 3 GTs Stream Table/Heat and Mass Balance

STREAM NUMBER		25		26		27		28		29		30	
STREAM NAME		PSA Tail Gas to recompression		Diluted Fuel to GT (x1)		Air to GGT (x1)		Flue Gas (total)		SRU Off Gas to CO2 Compressor		Ammonia Purge to duct burner	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016			818.95	74.31	0.00	0.00	0.00	0.00	10.69	5.47	4.40	50.03
Nitrogen	28.013			222.67	20.21	6644.35	75.52	20616.44	79.88	7.13	3.65	1.47	16.68
Carbon Monoxide	28.010			15.90	1.44	0.00	0.00	0.00	0.00	0.35	0.18	0.00	0.00
Carbon Dioxide	44.010			28.71	2.61	4.42	0.05	375.96	1.46	176.29	90.26	0.00	0.00
Methane	16.042			14.59	1.32	0.00	0.00	0.00	0.00	0.51	0.26	0.00	0.00
Argon	39.948			1.23	0.11	113.33	1.29	347.54	1.35	0.25	0.13	0.00	0.00
Hydrogen Sulfide	34.082			0.01	0.00	0.00	0.00	0.00	0.00	0.06	0.03	0.00	0.00
Carbonyl Sulfide	60.076			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031			0.00	0.00	0.00	0.00	0.00	0.00	0.02	0.01	2.93	33.29
Oxygen	31.999			0.00	0.00	2035.97	23.14	4468.31	17.31	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065			0.00	0.00	0.00	0.00	0.05	0.00	0.01	0.00	0.00	0.00
Sulphur	32.070			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)				69.87	59.51	0.00	0.00	0.00	0.00	1.00	0.86	0.35	0.30
Temperature	°C			121.00		15.00		101.00		39.79		6.00	

STREAM NUMBER		25		26		27		28		29		30	
STREAM NAME		PSA Tail Gas to recompression		Diluted Fuel to GT (x1)		Air to GGT (x1)		Flue Gas (total)		SRU Off Gas to CO2 Compressor		Ammonia Purge to duct burner	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara			45.00		1.01		1.01		1.20		20.00	
Total Dry Molar Flow (kg.mol/h)				1102.07	100.00	8798.06	100.00	25808.30	100.00	195.31	100.00	8.80	100.00
Water	kg.mol/h			0.46		35.16		4117.27		9.62		0.00	
Total Wet (kg.mol/h)				1102.53		8833.22		29925.56		204.93		8.80	
Total Mass Flow (kg/h)					9,900		256,600		825,100		8,184		100
Molecular Weight					8.97		29.05		27.57		39.94		11.35

Table 2-1. Balanced Generation, 3 GTs Stream Table/Heat and Mass Balance

STREAM NUMBER		32		33		34		35		36		37	
STREAM NAME		Duct Burner Exhaust		Syngas to duct burner		PSA tail to duct burner		HP N2 Diluent to GT Feed		sweep N2 to dryer		Total Oxygen Feed to Gasifier	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	0.01	92.71	358.45	64.02	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	28.013	20616.44	75.52	0.00	0.45	12.44	2.22	656.15	100.00	656.15	100.00	0.00	0.00
Carbon Monoxide	28.010	0.00	0.00	0.00	1.80	49.71	8.88	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	44.010	375.96	0.05	0.00	3.25	89.77	16.03	0.00	0.00	0.00	0.00	0.00	0.00
Methane	16.042	0.00	0.00	0.00	1.65	45.62	8.15	0.00	0.00	0.00	0.00	0.00	0.00
Argon	39.948	347.54	1.29	0.00	0.14	3.85	0.69	0.00	0.00	0.00	0.00	7.67	0.50
Hydrogen Sulfide	34.082	0.00	0.00	0.00	0.00	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	4468.31	23.14	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1526.47	99.50
Sulphur Dioxide	64.065	0.05	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		0.00	0.00	0.00	0.00	43.65	38.15	0.00	0.00	0.00	0.00	0.00	0.00
Temperature	°C	557.50		38.64		40.00		40.00		40.00		150.00	
Pressure	bara	1.04		33.05		1.30		32.90		2.30		45.00	

STREAM NUMBER		32		33		34		35		36		37	
STREAM NAME		Duct Burner Exhaust		Syngas to duct burner		PSA tail to duct burner		HP N2 Diluent to GT Feed		sweep N2 to dryer		Total Oxygen Feed to Gasifier	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Total Dry Molar Flow (kg.mol/h)		25808.30	100.00	0.01	100.00	559.86	100.00	656.15	100.00	656.15	100.00	1534.14	100.00
Water	kg.mol/h	4117.27		0.00		0.00		1.39		0.00		0.00	
Total Wet (kg.mol/h)		29925.56		0.01		559.86		657.54		656.15		1534.14	
Total Mass Flow (kg/h)			825,100		0		7,300		18,400		18,400		49,200
Molecular Weight			27.57		4.25		13.04		27.99		28.01		32.04

2.1.1 *Coal Receiving and Handling*

This operating section consists of two (2) primary unit operations:

- Handling systems designed to unload Illinois #6 coal and pile in yard stockpiles
- A storage area with active and inactive storage piles to service the plant

In the standard plant configuration, 8 cm x 0 (3" x 0) coal will be delivered to the site by 100-car trains comprised of 100-ton rail cars. Coal will be unloaded through the trestle bottom dumpster into two receiving hoppers and be subsequently transported by a vibratory feeder and belt conveyor to either the long-term storage pile or the reclaim area. Iron will be removed by passing the coal under a magnetic plate separator prior to delivery to the reclaim pile.

Vibratory feeders, located in the reclaim hopper, and a belt conveyor transfer the coal to the coal surge bin located in the crusher tower. The coal is reduced to 3 cm x 0 ($1^1/_4$" x 0) before a conveyor delivers it to the transfer tower and onto the tripper before being sent to the storage silos.

2.1.2 *Coal Preparation and Feed Systems*

The Coal Receiving and Handling subsystem ends at the coal silo. The Coal Preparation and Feed subsystem takes coal from the silo and performs two primary unit operations:

- Crushing the coal to a size suitable for use in the fluid bed dryer
- Transporting the coal from the coal silo to the fluid bed dryer

The crushed coal (roughly 0.125" x 0) is delivered to a surge bin before being transported to the fluid bed dryer through use of a lock hopper utilizing captured CO_2 as the transport gas.

2.1.3 *Coal Fluid Bed Drying System*

The primary purpose of the fluid bed dryer is to facilitate drying of the coal and releasing any hydrocarbons that are adsorbed in the pores of the crushed coal.[5] Additionally, while not examined in-depth in this report, the fluid bed drying system can serve to increase the overall system adaptability by facilitating a wider range of acceptable coal feedstocks and mitigating concerns of coal caking and swelling of the fuel feedstock prior to gasification.

The fluid bed dryer meets these objectives by:

- Reducing the moisture content of the coal prior to delivery to the gasifier
- Reducing the amount of light hydrocarbons adsorbed in the pores of the coal[6]

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[5] In the current design basis, the coal is heated only to temperatures sufficient to drive of adsorbed water. As such operating temperatures above 200 ºC are not anticipated. The significantly higher Design Temperature specification in the equipment list reflects the desire for the vessel to be specified such that higher temperature operations may be considered (following proper management of change) without a full vessel replacement, this higher temperature does not reflect operations in the current design basis.

[6] The coal selected for this study, as defined by DOE, is assumed to be "adsorbed hydrocarbon free." However, it is believed that the potential exists for trace amounts of adsorbed hydrocarbons in real-world feedstocks. It is anticipated that any adsorbed

Through these functions, the fluid bed dryer assures a more consistent feedstock for the gasifier. Specifically, the wet coal (11.12% moisture content by weight) is dried within the fluid bed dryer to a 5% moisture content by weight through indirect heating supplied by excess low-pressure steam that is generated in other plant processes. Nitrogen supplied by the air separation unit (ASU) will be introduced as a stripping gas into the fluid bed dryer to aid in stripping of the removed moisture and absorbed light hydrocarbons from the system. In addition to serving as the stripping gas, this nitrogen forms the bulk of the diluent that will be required to ensure that the syngas composition meets the requirements of the selected turbines (additional discussion can be found in Section 2.1.11.3).

The resulting overhead stream from this drying and desorption process contains the stripping gas, the moisture driven off of the as-received coal, and any desorbed hydrocarbons.[7] Water is knocked-out from the overhead stream by condensation through a transfer line exchanger prior to re-integration of the overhead stream with the post-water gas shifted (WGS) syngas stream. This re-integration occurs after the acid gas removal (AGR) system and before fuel gas conditioning.

The above description includes five significant process updates (relative to the process presented in the Conceptual Design Basis report) intended to better meet program objectives:[8]

1. The target moisture level of the coal existing the fluid bed dryer has been changed from 0% to 5% as this is the moisture content that is specified by the SES U-Gas gasifier for Illinois #6 coal. The advantage of this update is that reduction in the required drying of the coal represents a reduction in the amount of energy required to operate the fluid bed drying process.

2. Previously, the primary energy to drive the fluid bed dryer was obtained by a partial oxidation of the coal. While this was effective, it resulted in lower usable energy for other system processes, resulting in a reduced overall plant efficiency. In contrast, the current process provides the advantage of leveraging sensible heat integration to drive the system with excess process heat made elsewhere in the plant. Particular focus on this heat integration process during the Performance Modeling phase will help to ensure that these gains are maximized.

3. The fluid bed dryer is no longer supplied with an oxygen-rich stream from the ASU. In the previous Conceptual Design Basis, the oxygen was supplied primarily to drive the partial oxidation of the coal. Since this partial oxidation is no longer required, there is no longer a need for oxygen delivery to the fluid bed dryer.

4. CO_2 is no longer used as the stripping gas. While effective, this approach essentially reintroduced CO_2 that was already removed from the system resulting in removal of the

hydrocarbons that exist in a real-world feedstock would be a negligible amount in the overhead stream, that is ultimately routed through the fuel gas conditioning and will not significantly impact the plant's combustion and emission characteristics.

[7] It is the intention and belief that the overhead stream will only contain minimal amounts of desorbed hydrocarbons with pilot plant testing to quantify and characterize hydrocarbons that wind up in the fluid bed dryer overhead stream (most likely desorbed hydrocarbons from the pore volume of the coal, but possibly generated but unintended chemical transformation of the coal in "hots spots" or other poor operation transients).

[8] This was previously detailed and accepted in the Design Basis Report.

same captured CO_2 multiple times. This increased the overall size of the Selexol system and lowered overall plant efficiency. In the current system, the CO_2 has been replaced with a nitrogen-rich stream from the ASU which not only acts as the stripping gas but also serves as the diluent required for proper operation of the combustion turbine.

5. The overhead vapor stream will now be reintegrated with the shifted syngas stream at the directly before fuel gas conditioning, bypassing the mercury removal bed and AGR system.[9]

The core product of the fluid bed dryer is the sufficiently dried coal stream. This solid stream is delivered to the gasifier for conversion to syngas via a typical dry coal injection system. The solid effluent from the dryer is fed to this system which accomplishes the pressurization required to enter the gasifier. This intermediate system reduces the need to couple operating details of these unit operations at this time. It should be noted that the design pressure of the fluid bed dryer was set to the same value as the design pressure of the gasifier so future adaptions of a built system can consider more direct communication and interaction between these operating sections although this is explicitly not part of the current design basis.

While not formally part of our current design basis or Pre-FEED objectives, it is important to note that this specific technology (i.e., a bubbling fluid bed) was selected for coal drying out of a desire to ensure that deployed capital equipment would allow for increased operational flexibility and additional option value opportunities throughout the plant's lifecycle. Specifically, the inclusion of this fluid bed vessel and system offers the opportunity to handle coals with sulfur content beyond that of the design basis coal while minimizing the need for future plant modifications and capital outlay. To this end, the specified vessel is designed such that it could accommodate limestone injection for sulfur scavenging if the plant operator determines that this is a desired process implementation. This additional sulfur mitigation opportunity can enable the use of high sulfur coal sources at some point in the plant's lifecycle without the need to expand the fixed capacity of the acid gas removal system beyond the size of the originally installed system. Similar to the ability of refineries to accept various qualities of crude oil feedstocks, this unit operation increases overall plant flexibility and supports potential future arbitrage opportunities among different available coal feedstocks[10].

2.1.4 *Air Separation*

An oxygen rich stream (99.5 vol% O_2) for use in the gasifier and the Sulfur Recovery Unit (SRU), as well as a nearly pure nitrogen-rich stream for use throughout the facility, are separated in a cryogenic ASU. It is intended for this unit is to be provided as a complete vendor package.

In the ASU, atmospheric air is compressed and dried. A portion of the dry air stream is sent to a booster compressor before being passed to the "cold box." The remainder is fed directly to the

[9] It is believed that the fluid bed drying process will not produce enough organic-mercury compounds in the overhead stream to make mercury scrubbing of the overhead stream necessary, but this is something that should be confirmed through pilot plant testing.

[10] While current efforts have focused on the use of Illinois #6 as the primary fuel feedstock, initial analysis in the Conceptual Design phase suggests that this approach could support the use of additional coal feedstocks, including waste coal streams. However, it should be noted that this analysis is preliminary in nature and would require plant modifications as well as a full hazard and operability study.

ASU cold box. In the cold box, the dry air is cooled against the low temperature product streams. The cold air leaving the main heat exchanger is sent to a distillation column arrangement typically consisting of a high pressure (HP) and low pressure (LP) column.

Liquid O_2 from the sump of the LP column is pumped up to the gasifier operating pressure and passed back to the main heat exchanger where it is vaporized, cooling the incoming air. The gaseous O_2 product stream is of 99.5% purity and is at approximately 45 bar(g). Gaseous N_2 leaves the top of the LP column and also passes back through the main heat exchanger cooling the incoming air. Oxygen and nitrogen storage are provided to maintain plant operations during short outages of the ASU.

The ASU is typically provided as a vendor package. The following description is not specific to any ASU vendor. The air separation process begins by compressing ambient air in the main air compressor. The main air compressor has inter-stage and discharge cooling provided by cooling water. The cooled, compressed air then passes through a temperature swing adsorption system where the water, carbon dioxide, and organic material are removed.

The dry air stream is then split, and a portion of the air is sent to a booster compressor. Expansion of the air sent through the booster compressor supplies additional refrigeration to the process to make up for heat gained in the cold box during operation.

The "cold box" is a large structure containing all of the major cryogenic process equipment. Voids in the cold box are filled with perlite to provide insulation and reduce ambient heat gain.

Both the main compressor air stream and the air sent through the booster compressor flow into the ASU cold box. On entering the cold box, dry air is passed through a brazed aluminum heat exchanger where it is cooled against low temperature product streams. Cold air leaving the main heat exchanger enters a distillation column arrangement typically consisting of a high pressure (HP) and low pressure (LP) column. Reducing the pressure of the chilled air in a cryogenic turbo expander provides additional cooling. Nitrogen vapor from the top of the HP column is used to re-boil the LP column. A small portion of the condensed liquid nitrogen is extracted from the HP column, pumped to ~35 bar, and vaporized in the main heat exchanger. This stream is used in the ammonia synthesis loop and for fuel dilution in the power block. Additional nitrogen is vaporized and used to provide N_2 for the fluid bed dryer stripping gas, purge gas to the sulfur recovery unit, transport gas for coal milling and drying, and lock hopper pressurization for the gasifier.

An ASU will be included to create both oxygen-rich and nitrogen-rich streams for use in other system processes. Specifically, the oxygen-rich stream will supply the oxidation reactions driving the core process in the gasifier while the nitrogen-rich stream will be used to supply (1) the ammonia synthesis loop, (2) stripping gas to the fluid bed dryer, (3) fuel diluent for the combustion turbine, and (4) product tank blanketing.

The sizing of the ASU is set by the oxygen requirements and must support a demand of ~39,000 kg/hour of nitrogen for system processes and ~50,000 kg/hour of oxygen. The ASU represents significant parasitic loads on the system with the ASU package (i.e., ASU main compressor, ASU auxiliaries, and oxygen and nitrogen stream compressors) accounting for over 30% of the plant total.

2.1.5 *Gasifier*

The gasifier follows an SES U-Gas design with dimensions limited by the ability to shop fabricate and transport over-land to the site to ensure that modularity is maintained. The represents a significant update relative to the Conceptual Design report. Whereas the previous Conceptual Design focused on a KRW-style gasifier, the Pre-FEED process has focused on the SES U-Gas style gasifier. Initially, the KRW gasifier was selected because it offered a number of positive characteristics in terms of package size and aspect ratio, which resulted in perceived advantages in shop fabricability and modularity. While a KRW gasifier has not been recently manufactured, it was believed that this was more of an issue of resurrecting a sufficiently mature, if abandoned, technology. However, in discussions with teaming-partner experts in the field of commercial gasification technology, it is now believed that adopting the KRW gasifier represents unnecessary risks in the areas of manufacturability and commercialization to meet the aggressive deployment timeline of the Coal First Initiative.

In order to help reduce the risk of manufacturability and commercialization, the SES U-Gas gasifier has been selected. This risk reduction is driven both by the fact that this style of gasifier is supported by an existing and willing vendor and the fact that there are a number of existing commercial operations, helping to ensure a flow of active and fresh operating knowledge. Additionally, both the vendor and selected gasifier design have demonstrated experience operating with the selected Illinois #6 feedstock. These factors combine to lower the technological risk associated with piloting and commercialization of the overall plant design.

The devolatilized and dried coal is conveyed to the top of the lock hopper system where it is pressurized using N_2 before being fed to bottom of the fluidized bed gasifier. In the gasifier, the coal reacts with a sub-stoichiometric quantity of oxygen and steam to convert to a synthesis gas which contains primarily CO, H_2, CO_2, steam (H_2O), lesser amounts of N_2, CH_4 and a small amount of Ar. As this gasifier operates at about 1000°C, the syngas exiting the fluidized bed in standard SES U-gas operations contains roughly 7% methane. Methane content at this level can significantly reduce the effectiveness of pre-combustion carbon capture efforts. To address this concern, the design basis utilizes partial oxidation occurring in the freeboard of the gasifier to reduce methane content to roughly 1%. The WGS (Eq. 2.1) and steam methane reforming[11] (Eq. 2-2) reactions operate according to the following equations:

$$CO + H_2O \Leftrightarrow CO_2 + H_2 \qquad \textit{Eq. 2-1}$$

$$CO + 3H_2 \Leftrightarrow CH_4 + H_2O \qquad \textit{Eq. 2-2}$$

It is important to minimize the operating pressure of the gasifier in order to achieve this large methane reduction as lower pressures promote the steam methane reforming reaction.

The sulfur in the coal is converted primarily to H_2S with the remainder converting to COS. The small amount of chlorine present in the coal is converted to HCl. Small amounts of HCN and NH_3 are also produced in the gasifier. The operating conditions of the gasifier are selected to eliminate

[11] Note that this refers to steam methane reforming occurring within the gasifier through the partial oxidation in the freeboard as a means of reducing overall methane content in the raw syngas. This is opposed to operating a separate steam methane reformer elsewhere in the plant.

the production of tars, phenols, and other condensable organic materials from the produced syngas. The gasifier is non-slagging and the inorganic material in the feed is discharged as a fly ash and coarse char material from the overhead cyclones and gasifier bottom discharge hopper. This material is cooled, discharged from the gasifier, collected and disposed of offsite.

Hot syngas exits from the top of the gasifier and is cooled in a gasifier heat recovery steam generator (HRSG.) The HRSG generates HP superheated steam which is used in the process. The syngas discharges from the gasifier HRSG at ~300°C and ~40 bar and enters a scrubber column which removes the residual particulates in the raw syngas and any HCl. The scrubber column also saturates the syngas with water. The blowdown water from the scrubber column is sent to the waste water treatment plant which purifies the water so that it can be used within the plant or discharged offsite.

It is anticipated that the gasifier will produce ~172,000 kg/hour of scrubbed syngas from the coal feedstock. Parasitic loads are relatively light for the gasifier, accounting for ~1% of the total for the plant. Additionally, the gasifier allows for recovery of a significant amount of process heat that can be used to meet other plant thermal loads.

The temperature and pressure of the coarse ash from the gasifier is reduced as ash flows out through the ash classifier and bottom ash handling system. Fine ash and carbon particles leave the gasifier fluidized bed with the syngas. The primary fines recovery and recycle system consists of two cyclones in series, which collect nearly all fines from the gas stream leaving the gasifier. The fines collected in the cyclones are returned to the gasifier by means of a dip-leg. The syngas from the primary cyclones is cooled in the syngas cooler and then passes to the third cyclone and ceramic/metal filters for further removal of dust. The additional fines that are collected from the third cyclone and filters are routed to a fines silo through a lockhopper system, where they are collected in the baghouse and returned to the gasifier for further conversion. The bottom ash, upon leaving the ash classifier, is cooled and removed from the plant via an ash cooler, lockhopper system, and screw coolers before being transported outside by belt conveyors for truck unloading. In the initial ash cooler, steam is generated through direct contact with the ash and directed through the annulus into the gasifier.

2.1.6 *Water Gas Shift*[12]

Water gas shift forms a central part of the plant's emissions strategy by serving as a mechanism to maximize the amount of pre-combustion CO_2 capture. This approach is synergistic to ammonia production as WGS increases the hydrogen content within the syngas stream. This shift is accomplished by reacting the raw syngas in the presence of steam and a catalyst in a fixed-bed reactor. Required cooling in this process will remove sensible heat that is generated in the shift reaction for use in other system processes.

To accomplish this process, additional steam is added to the raw syngas stream from the scrubbers to increase the steam content of the syngas to ~60% by volume. This level of steam content both

[12] As the process described in this section represents a sulfur-tolerant water gas shift that includes the CO shift converter upstream of the acid gas removal, this process can be more accurately described as a "sour gas shift." However, the term "water gas shift" has been selected instead to match the process naming convention observed in *Cost and Performance Baseline for Fossil Energy Plants Volume 1b: Bituminous Coal (IGCC) to Electricity Revision 2b – Year Dollar Update* report.

facilitates the shift reaction and prevents damage to the catalyst. All of the syngas is preheated to 300°C in a feed-product interchanger and passed through a single WGS train consisting of two WGS reactors in series, where the carbon monoxide (CO) in the gas reacts with water vapor (H_2O) to produce hydrogen (H_2) and (CO_2) according to the WGS reaction (Eq. 2-1, as seen above):

Other reactions also occur in the WGS reactors. Carbonyl sulfide (COS) is hydrolyzed to hydrogen sulfide (H_2S) (Eq. 2-3) and hydrogen cyanide (HCN) to ammonia (NH_3) (Eq. 2-4) as seen below:

$$COS + H_2O \Leftrightarrow CO_2 + H_2S \qquad \text{Eq. 2-3}$$

$$HCN + H_2O \Leftrightarrow NH_3 + CO \qquad \text{Eq. 2-4}$$

The shift reaction is exothermic with a temperature rise across the first reactor of approximately 150°C.

The syngas leaving the first shift reactor is cooled by raising HP steam in a boiler. The syngas then enters the second shift reactor at approximately 290°C. The syngas leaving the second shift reactor is cooled by heating up the feed to the first shift reactor in the interchanger. The remaining fraction of CO ("slippage") after the shift reactor is less than 2.0% by volume on a dry basis. The syngas is cooled to approximately 190°C by transferring heat to HP boiler feed water (BFW), and then enters the bottom of the desaturator column where it is cooled by circulating process water fed to the top of the column.

The effluent of WGS operating section, neglecting the water that will be knocked out in the syngas cooling process, is ~172,000 kg/hour comprised primarily of CO_2 (~154,000 kg/hour) and H_2 (~10,000 kg/hour).

2.1.7 *Syngas Cooling*

Final cooling of the syngas prior to cleaning occurs in the desaturator, a direct contact cooler which uses multiple beds of random packing in a tower. Most of the water present in the syngas from the WGS reactor condenses in the desaturator. The syngas leaves the top of the desaturator column at ~40°C, containing only a small fraction of the water vapor that entered with the gas at the bottom of the column.

Hot syngas exits the HP BFW preheater at ~190°C and ~34.3 bar(a), enters the bottom of the desaturator, and contacts hot water flowing down through the packing in the column. The process water leaving the bottom of the desaturator at ~181°C is split into several streams as part of the overall plant's heat integration. A portion of the hot process water (~20 MTPH) is pumped back to the gasifier scrubbers as described above. The majority (~1,100 MTPH) of the process water leaving the bottom of the desaturator column if fed to a second HP BFW preheater, where it preheats the BFW to 170°C. Additional heat is extracted from this stream in the LP boiler by raising ~35 MTPH of steam at 5.16 bar(a).

After passing through the LP boiler, the process water (now at ~163°C) is split into three streams:

1. About 20 MTPH is fed to the gas turbine (GT) feed preheater. This exchanger preheats the fuel to the gas turbines to 121°C after the fuel has been compressed in the GT fuel

compressor. The outlet from the GT feed preheater is fed to the ammonia stripper.

2. About 170 MTPH of process water from the LP boiler is used to preheat LP BFW to 150°C. Part of the preheated BFW is fed to the LP boiler while the majority is pumped to 55 bar(a) and fed to the syngas cooler in the gasifier island and the HP boiler downstream of the first shift reactor.

3. The balance of the process water is used to re-boil the Selexol stripper column in the AGR. The hot water exiting the stripper reboiler is split into two streams.

    a. ~760 MTPH is returned to the top of the lower section of the desaturator at 149°C.

    b. The balance is used to produce low-low pressure steam (LLPS) at 2 bar(a) in the LLPS boiler, which is used exclusively as stripping steam in the deaerator.

The process water from the outlet of the LP BFW preheater (Stream 2, above) and the process water from the outlet of the LLPS boiler (Stream 3.b., above) are combined and used to preheat demineralized makeup water (DMW). The process water stream is split at the outlet of the DMW preheater into two streams: (1) ~250 MTPH of process water is cooled to 40°C (accomplished by initially cooling to 65°C using an air cooler, with an exchange against cooling water providing the remaining cooling duty) before being fed to the top of the desaturator and (2) the balance fed to the ammonia stripper column to remove any excess ammonia that may be present. Process condensate from the ammonia stripper can then be used as make-up for the cooling tower.

The desaturator and most of the associated exchangers are located adjacent to the shift reactors. The GT feed preheater, LLPS boiler and DMW preheater are all located in the power block. The Selexol reboiler and ammonia stripper column are located in the AGR.

The syngas exits the top of the desaturator at 40°C and 34 bar(a).

A key feature of the desaturator is that most of the water is recycled to the middle of the desaturator at 149°C. This increases the quantity of 181°C water available at the bottom of the desaturator and improves overall heat recovery.

Using a desaturator column in the configuration described enables optimal integration of heat from the raw syngas with rest of the plant. Any heat that is not required for process heating duties is used to preheat LP or HP boiler feed water or provide duty for LP steam generation. An additional critical advantage of using the desaturator is that this complex heat recovery can be accomplished while maximizing efficiency and minimizing pressure drop (~0.3 bar drop rather than a 1.5-2 bar drop commensurate with a series of exchangers and knock-out pots) through the system. This reduction in pressure drop through the cooling train allows for the gasifier to be operated at a lower pressure which, as stated above, promotes a reduction in methane formation in the gasifier.

2.1.8 *Syngas Clean Up*

The purpose of the syngas clean-up operation is to remove impurities from the shifted syngas stream (e.g., CO_2, sulfur, and mercury) to provide a hydrogen-rich, "pure" stream suitable for both power and chemical storage generation. The approach to syngas clean-up is as follows:

2.1.8.1 Ammonia Removal

Ammonia is separated from the syngas and process water streams through the use of an ammonia stripper fed by a side stream of process water drawn from the water circulating around the desaturator column. The moisture in the overhead from the column is mostly condensed in the overhead condenser of the ammonia stripper. Condensate from the overhead condenser is returned to the top of the column. The remaining, ammonia-rich vapor stream from the overhead condenser is sent to the Claus plant furnace in the SRU where the ammonia is destroyed by combustion. Stripped water from the bottom of the ammonia stripper column is used as make-up water for the cooling tower.

2.1.8.2 Mercury Removal

Mercury removal will be accomplished through the inclusion of a sulfur-impregnated, activated carbon bed. A representative system is described in the *Cost and Performance Baseline for Fossil Energy Plants Volume 1b: Bituminous Coal (IGCC) to Electricity Revision 2b – Year Dollar Update* report. Syngas leaving the desaturator will pass through these mercury guard beds before passing to the H_2S absorber in the AGR unit. This will serve to remove traces of mercury that may be in the syngas. Typically, carbon replacement is needed after 18 – 24 months of operations.

2.1.8.3 Acid Gas Removal

The objective of the AGR is to remove the sulfur compounds and carbon dioxide from the syngas. Sulfur is present primarily as H_2S which is removed to achieve a maximum total sulfur concentration in the syngas to the gas turbine of <10 ppmv (dry basis). Sizing and operation of the AGR system is selected to ensure that sufficient carbon dioxide is captured to support a 90% carbon removal rate for the plant as a whole.

The technology selected for the AGR is Selexol licensed by Honeywell UOP.

Major equipment in the acid gas removal unit includes the H_2S absorber, CO_2 absorber, H_2S concentrator, Selexol stripper, flash gas compressor, stripping gas compressor, CO_2 recycle compressor, flash vessels, pumps, and heat exchangers.

Shifted, cooled syngas from the mercury guard beds enters the AGR unit where it is blended with a cooled stream of recycle gas from the H_2S concentrator. The gas blend is fed into the H_2S absorber where it is contacted with cooled, loaded, Selexol solution. "Loaded solution" is defined as Selexol solution that has been through the CO_2 absorber and, consequently, is loaded with CO_2. H_2S, COS, some CO_2, and small quantities of other gases (primarily hydrogen) are absorbed into the solution.

The syngas, now free of sulfur but containing most of the original incoming CO_2, exits the top of the H_2S absorber and is fed to the bottom of the CO_2 absorber where it is first contacted with semi-lean solution. The "semi-lean solution" is so named because it is regenerated by pressure flash, rather than steam stripping. The CO_2 is recovered from the Selexol solution in a series of three vessels where the solution is flashed at progressively lower pressures. The semi-lean solution is then cooled and pumped back to the center of the CO_2 absorber. This is an energy efficient method for recovering the bulk of the CO_2 from the syngas, resulting in most of the CO_2 being absorbed from the syngas. In the top section of the CO_2 absorber, the gas stream comes into contact with lean solution (solution regenerated by steam stripping in the Selexol stripper vs. pressure flash regeneration for "semi-lean solution") and finally exits the CO_2 absorber at approximately ~33 bar and containing ~4% CO_2.

The solvent leaving the bottom of the H_2S Absorber, called "rich liquor", enters the lean-rich exchanger, where the temperature of the stream is increased by heat exchange with the lean solvent from the Selexol stripper. The stream is then fed to the H_2S concentrator which increases the proportion of H_2S in the rich liquor by stripping most of the CO_2, CO, and H_2 from the rich liquor through the use of nitrogen, part of which is sourced from the overhead of the fluid bed dryer. The overhead stream from the H_2S concentrator is cooled and fed back to the inlet of the H_2S absorber.

Rich liquor from the bottom of the H_2S concentrator is sent to the Selexol stripper, where the solution is stripped with steam to remove the H_2S. Stripping steam is generated from the Selexol solution in the Selexol stripper reboilers, which are heated by recycled water from the desaturator and LP steam. The overhead acid gas product from the Selexol stripper is sent to the SRU. The lean solution is pumped to the lean-rich interchanger and then cooled further before being sent to the top of the CO_2 absorber.

The solvent exiting the CO_2 absorber is termed "loaded solvent," as it contains high level of CO_2 but very little sulfur. A portion of the loaded solvent is sent to the H_2S absorber, to absorb the sulfur compounds. The majority of the loaded solvent is fed into the HP CO_2 flash drum where a portion of the absorbed gases are flashed off. The overheads from this drum (primarily H_2 and CO_2) are compressed in the CO_2 recycle compressor and recycled to the CO_2 absorber syngas inlet to recover the H_2.

The solvent stream leaving the HP flash drum is flashed further through use of both an IP flash drum and an LP flash drum. The overhead of the IP and LP flash drums is the CO_2 product gas and is sent to the CO_2 product compressor. The semi-lean solvent exiting the LP flash drum is cooled in a semi-lean cooler and returned to the CO_2 absorber via the semi-lean pump.

The sweet syngas stream is split with additional details appearing in Section 2.1.9.

2.1.8.4 CO_2 Compression and Drying

Flashed gas containing CO_2 and water vapor is compressed to ~90 bar(g) and dried during the compression process.

Flashed gas from the AGR IP and LP CO_2 flash drums is fed into the CO_2 compressor package to compress the product CO_2. The gas from the LP flash drum is fed to the first stage of the compressor, while the gas from the IP flash drum is fed to the second stage. The majority of the water present is knocked-out after the first and second compression stages. The remaining water is separated from the product CO_2 in the CO_2 drying package. The condensed water is returned to the desaturator as makeup.

The CO_2 stream is ~90 bar(g) at the compressor discharge. This stream is condensed to liquid in the compressor after-cooler, then pumped to the export pressure of 145 bar(g) for eventual routing to a CO_2 pipeline and storage.

2.1.8.5 Sulfur Recovery

The acid gas from the H_2S stripper, along with sulfur containing streams from the ammonia stripper and flash gas from the gasifier scrubber blowdown, is sent to a Claus-based SRU to recover the sulfur as elemental sulfur. The Claus technology consists of a thermal oxidation stage where part of the H_2S is reacted with pure oxygen from the ASU to form SO_2 followed by three catalytic stages (each utilizing the standard Claus catalyst) where SO_2 is reacted with H_2S to produce elemental sulfur. Condensers present between each catalytic stage are used to remove elemental sulfur at each point along the series of catalytic reactors. After passing through each condenser, the gas is reheated before entering the next reactor.

In the thermal oxidation stage, about one third of the H_2S in the acid gas is burned in an oxygen-deficient environment to form SO_2. The quantity of acid gas oxidized is adjusted to achieve third stage tail gas concentrations of H_2S between 0.8-1.0 vol%. LP steam is produced in the sulfur condensers and fed to the LP steam header.

The tail gas from the final sulfur condenser goes to the tail gas treatment (TGT) unit where sulfur compounds in the tail gas are removed before the gas is fed to the inlet of the CO_2 compressor.

Condensed molten sulfur from the Claus plant SRU contains H_2S which must be removed before storage or shipment. The liquid sulfur product from the SRU is degassed by stripping with nitrogen. The sulfur product off-gas is routed to the Shell Claus Off-Gas (SCOT) absorber (01-T-0602) in the TGT unit.

The plant is expected to produce 1,776 kg/hour of sulfur, which will be sent through the solids handling system with the anticipation this byproduct will be sold to generate an ancillary revenue stream for the plant.

2.1.8.6 Tail Gas Treatment Unit

The Claus plant tail gas is processed in a TGT unit to remove the residual sulfur compounds so that the stream can be safely vented to atmosphere utilizing a SCOT absorber.

The tail gas from the final stage of the SRU is hydrogenated in a fixed catalytic bed. If required, a small stream of syngas from the desaturator may be used as a supplemental source of hydrogen. The hydrogenation process reduces the sulfur compounds in the tail gas, primarily COS in this application, to H_2S. The hydrogenated tail gas is then quenched in a wash tower. In the wash tower, most of the water in the hydrogenated tail gas stream is condensed. The wash tower uses circulating water for washing the gas feed. The circulating water is cooled before entering the top of the wash tower. Any net production of water is sent to water treatment.

The washed gas is combined with the off-gas from sulfur de-gassing and sent to the packed column SCOT absorber. Lean amine solvent is used to absorb most of the H_2S from the tail gas, while minimizing removal of CO_2. The rich solvent is pumped to the regeneration column to recover the H_2S. Desulfurized gas leaving the top of the absorber is incinerated and discharged to atmosphere.

The rich solvent flows through a lean-rich exchanger to the SCOT regeneration column. The lean-rich exchanger heats the rich solvent feed by cooling the hot lean solvent leaving the regenerator. The rich solvent then enters the regenerator where the solvent is stripped by steam produced in the regenerator reboiler. The stripped solvent is cooled by the lean-rich exchanger before returning to the SCOT absorber. The acid gas stripped from the rich solvent is cooled and sent to the regenerator knock-out drum. From the regenerator knock-out drum, the acid gas returns to the feed section of the Claus unit. Condensed water is used to scrub the acid gas at the top of the regenerator to remove trace solvent from the acid gas.

2.1.9 *Syngas Management*

The purpose of the syngas management operation is to monitor and regulate the distribution of syngas (as well as relevant ancillary streams such as nitrogen, steam, etc.) between the various operating sections. This includes managing storage capacity to respond to changes in electrical load and extraction of hydrogen for ammonia synthesis. Primarily, this involves routing clean syngas between one of three possible dispositions: (1) a tank for temporary storage[13], (2) the gas turbines, and (3) the hydrogen recovery pressure swing adsorption (PSA) unit.

Estimates related to syngas storage capacity used a syngas storage capacity of 1,000 m^3. The design basis for the storage capacity was motivated by the desire to ease transitions between plant operating points, as well as assisting in handling process upsets (i.e. syngas to be diverted to storage while the gasifier is backdown in event of an issue with the PSA or ammonia train). These transition needs set the capacity requirement, primarily by evaluating the lag in the transition time of the ammonia loop relative to the gasifier trains and the power island. The capacity selected will provide 40 minutes of storage which is sufficient to handle the most drastic operating point transition, and this storage time can be extended to 60 – 80 minutes by performing other operational adjustments during the transition period.

In the *Balanced Production, 3 GTs* operating mode, the syngas flowrate to the combustion turbine is ~11,300 kg/hour with the balance (11,700 kg/hr) going to the PSA. Of the ~11,700 kg/hour to the PSA, ~4,400 kg/hour of pure hydrogen is sent to the ammonia loop, with the remainder sent to the power island for combustion in the turbines and duct burners

[13] The intended use of the storage tanks is to dampen the impacts of lagging system components during the transitions between operating modes. They can accomplish this by (1) storing excess syngas created while the syngas production system turns down at a slower rate than the combustion turbine or by (2) supplying surge syngas to the gas turbines while the syngas production system ramps up at a slower rate than the combustion turbine. Based on this intended equipment usage, the storage tanks will accommodate the bi-directional flow of syngas.

Estimates related to syngas storage capacity used a syngas storage capacity of 1,000 m3. The design basis for the storage capacity was motivated by the desire to ease transitions between plant operating points, as well as assisting in handling process upsets (i.e. syngas to be diverted to storage while the gasifier is backdown in event of an issue with the PSA or ammonia train). These transition needs set the capacity requirement, primarily by evaluating the lag in the transition time of the ammonia loop relative to the gasifier trains and the power island. The capacity selected will provide 40 minutes of storage which is sufficient to handle the most drastic operating point transition, and this storage time can be extended to 60 – 80 minutes by performing other operational adjustments during the transition period.

As the plant is designed with syngas storage, flaring is not standard operating procedure, and is only used in start-up, shutdown and during upset conditions for safety purposes. During normal operation, including transitions, flaring is not carried out if for no other reason the flare is burning valuable product. If, during transitions, excess syngas is being produced (e.g. the power island has reduced capacity rapidly and the ammonia loop and / or the gasifier island has not responded as quickly as expected) the excess syngas is sent to syngas storage either directly from the AGR, or via the GT feed gas compressor. Once stable operation is achieved, the syngas storage unit is depressurized by feeding the GT and / or the duct burners.

Waste gas containing 33% (dry) ammonia is being fed to the duct burner in very small quantities. The ammonia purge from the ammonia loop (stream 30) is fed at a rate of 5.6 kmol/h, where it is combined with stream 34 at 355 kmol/h and stream 33 which varies in flow depending on operation. The ammonia composition in the overall duct burner feed is low. Although, the amount of NOx generation has not been detailed, it is expected that the downstream SCR catalyst will be able to handle the NOx due to ammonia combustion.

2.1.10 *Ammonia Generation*

2.1.10.1 Hydrogen Purification

Hydrogen is recovered from the sweet syngas using pressure swing adsorption with the resulting high purity hydrogen fed to the ammonia synthesis unit. Depending on the operating scenario, the off-gas from the PSA can have two final dispositions: (1) compression for use as fuel in the gas turbine and (2) fuel for the duct burners in the HRSG.

2.1.10.2 Ammonia Synthesis and Refrigeration

The primary goal of the ammonia synthesis train is to provide a chemical storage medium to support overall system reliability, availability, and modularity with the additional opportunity to provide a supplemental value stream for the polygeneration plant. Based on the nominal amount of hydrogen available in the plant, a scale-down of the conventional, existing Haber-Bosch approaches is believed to be most applicable.

Nitrogen from the ASU is compressed to 33 bar(a) (utilizing the same compressor used for nitrogen dilution of the GT fuel) and then mixed with hydrogen from the PSA. The mixed stream is chilled to ~7 °C (using excess refrigeration capacity from the ammonia recovery unit) and compressed to 135 bar(a) in a two-stage, intercooled compressor. The fresh feed to the loop is mixed with recycle gas from the knock-out pot and compressed further in the circulator compressor.

The syngas enters the loop at 145 bar(a), preheated occurring against the ammonia product stream, and fed to a three-bed converter with intercooling. The ammonia product from the reactor is at ~400 °C and cooled through multiple process, including:

1. Raising steam at 105 bar(a)
2. Heat exchange to the syngas feed in the feed/product interchanger
3. Heat exchange against cooling water

4. Heat exchange against the recycle gas from the knock out pot
5. A refrigeration unit

The syngas and product ammonia streams enter the knock-out pot at ~4 °C with the overhead from the knock-out pot being reheated against the incoming product stream and fed to the inlet of the recycle compressor.

Liquid ammonia is recovered from the knock-out pot and flashed to remove the bulk of the dissolved and entrained gases. The flash stream is routed to the SRU and used as fuel gas. The liquid ammonia enters the refrigeration unit, is chilled, and then passed to the product tanks.

The 105 bar(a) steam raised in the ammonia synthesis loop is depressurized to 68 bar(a) and fed to the HP steam superheaters in the power block HRSG's.

2.1.11 *Power Block*

The overall power block follows a combined cycle design. There are three LM2500+ gas turbines, modified for the combustion of high H_2 syngas. Associated with each gas turbine is a HRSG configured to produce two levels of superheated steam. Steam generated in the each of the three HRSG's is combined with surplus steam generated in the process blocks and can be fed to a combination of two steam turbines: a primary steam turbine rated for 47 MWe and a secondary steam turbine rated for 25 MWe.

The desire for rapid, frequent turndown and ramping, while maintaining high overall plant efficiency, has influenced a number of decisions throughout the design process. For example, aeroderivative turbine designs were selected as they have the ability to rapidly ramp up in response to changes in grid demand faster than a single, large frame turbine. By selecting a three-turbine configuration, it is possible to achieve higher net power production for export while still allowing for high levels of overall plant turndown. For example, the *Net Zero Power* case, which is essentially full turndown from a power export standpoint, can be achieved with a single turbine operating at 68% of maximum capacity).

Additionally, the use of three turbines allows for greater options in both meeting demand at a given point within the operating window. Specifically, the *Balanced Production* operating point can be met through either three turbines as 67% capacity or two turbines at 100% capacity. This flexibility in reaching different points within the operating window the plant operator with more tools at his/her disposal to quickly transition to meet rapidly changing market demands and conditions.

The use of the three turbines also helps to ensure emissions compliance across a wide range of operating conditions as there should never be a case when a single turbine is forced to turn down so significantly as to operate outside the advertised operational range with full emissions compliance. If a situation arose where a turbine did need turned down below the emission compliant range, the plant operator could simply choose to completely shut a turbine down while increasing the load(s) on the remaining operational turbine(s) to make up for the reduced power output.

2.1.11.1 Fuel Gas Conditioning

The fuel to the gas turbine needs to be conditioned to meet the GE's specifications for high hydrogen fuel for LM2500+ gas turbines. This includes compression to the required inlet pressure (33 bar), dilution to meet the composition specification (primarily through the use of nitrogen) and preheating to 121°C against circulating process water from the desaturator. While most of the fuel gas is fed directly from the AGR, a portion of the PSA off-gas is compressed and fed to inlet of the GT under some operating scenarios.

2.1.11.2 LM2500+ Gas Turbine

The LM2500+ is an advanced gas turbine designed to fire high H_2 syngas in its combustors. The key metric for high hydrogen syngas service used by GE is "H2 + ½CO". This is defined as the mole fraction of H_2 plus half the mole fraction of CO, with the maximum molar fraction limit of the LM 2500+ set at a 0.75. It is noted that the sweet syngas produced by the plant has a "H_2 + ½ CO" of 0.94. In order to create a turbine fuel that conforms to GE's requirements, the syngas fuel is diluted with nitrogen.

Water is injected to the combustors to reduce the production of thermal NOx, resulting in the gas turbine exhaust containing 25 ppmvd of NOx when adjusted to 15 vol% O_2 (dry basis). Because there is so much less carbon in this high hydrogen fuel than is found in typical hydrocarbon or syngas fuels due to the pre-combustion capture methods employed, the CO in the turbine exhaust is expected to be less than 10 ppmvd (adjusted to 15 vol% O_2 on a dry basis).

2.1.11.3 Heat Recovery and Steam Generation

Heat from each gas turbine exhaust raises steam in the associated two-pressure level HRSG. The exhaust temperature from the LM2500+ operating on high hydrogen syngas is only 450°C, which serves to limit the pressure and superheat temperature of the steam generated in the HRSG to below what is required for the steam feed to the shift reactor. To alleviate this concern, each HRSG is fitted with a duct burner configured to combust high hydrogen syngas. In addition to raising the exhaust temperature from the gas turbines, the duct burners additionally serve as an opportunity to utilize any fuel that has not already been employed to produce ammonia or to supply the gas turbines directly.

HP steam is raised in the HRSG's at 64 bar and 487 °C with the combined steam raised by the three HRSG's driving one steam turbine generator. The total main-steam flow is limited to 160 MTPH although this can be produced by two of the three trains together. IP steam is fed from a pass out in the extraction steam turbine to the shift unit to supplement the steam feed to the shift reactors at 43 bar and 430°C. LP steam, in excess of that required by the process units, is blended with steam raised in the HRSG's and fed to the IP/LP crossover in the steam turbine which is at 4.9 bar. Stack gas is discharged to the atmosphere at 110°C via the stacks associated with each HRSG. Additional information is provided in Appendix E and Appendix F.

The steam system is designed to allow steam export to the plant for start-up and to heat the fuel gas and nitrogen diluent for the gas turbine.

The steam turbine last stage exhaust quality is approximately 88% in normal operation. The steam turbine condenses the remaining water vapor in the exhaust steam by rejecting the heat to cooling water. Steam condensate is transferred to the vacuum deaerator package which operates at 70 mbar(a). Condensate is de-aerated using LLP steam generated by a side stream from the desaturator.

Condensate pumps distribute the de-aerated BFW to all steam generators in the plant.

2.1.11.4 Selective Catalytic Reduction (SCR)

This facility has been designed to reduce the concentration of NOx in the HRSG stack gas to a maximum of 5 ppmvd adjusted to 15% O_2 (dry basis) during normal operation.

The concentration of NOx in the gas turbine exhaust is 25 ppmvd adjusted to 15% O_2 (dry basis). Selective catalytic reduction (SCR) units installed in the HRSG's reduce the NOx in the flue gas from 25 to 5 ppmvd adjusted to 15% O_2 (dry basis) through the reduction of NOx to N_2 and H_2O by the reaction with ammonia on the catalyst. This ammonia is injected into the flue gas in the HRSG's upstream of the SCR catalyst beds. The ammonia serves to activate the SCR catalyst as the flue gas passes through the catalyst beds. The addition of ammonia is controlled to limit the ammonia slip (i.e., the concentration in the stack gas) to 5 ppmvd. The SCR design specification for NOx inlet and flue gas are presented in the equipment list. The inlet specification is 25 ppmv and the outlet specification is 5 ppmv. Typically, NOx generation is expected to be trace amounts in this stream, thus not specified in the HMB.

2.2 Key System Assumptions

System assumptions for the polygeneration plant design are compiled in Table 2-2.

Table 2-2. Key System Assumptions

Metric	Value/Notes
Combustion Turbine	3x GE LM2500+ (30.2 MW output each)
Ammonia Synthesis Loop	2x 300 MTPD Capacity Ammonia Loops
Gasifier Tech	SES U-Gas
Oxidant	95% vol% O_2
Coal	Illinois No. 6
Coal Feed Moisture Content %	5%
COS Hydrolysis Reactor	Occurs in WGS
Water Gas Shift	Yes
H_2S Sep	Selexol 1st Stage
Sulfur Removal %	~100.0
Sulfur Recovery	Claus Plant with Tail Gas Treatment (SCOT); Recovered as Elemental Sulfur
Mercury Control	Dual Carbon Bed in Series
NOx Control	N_2 Dilution, Humidification, and SCR
CO_2 Sep	Selexol 2nd Stage
Overall Carbon Capture	90%

2.3 Five Operating Points for Insight into Operational Performance and Flexibility

It is envisioned that the plant will provide the flexibility to operate efficiently across a wide operational window in order to respond to changing demands of the bulk electric grid, both in the short term (e.g., changes to instantaneous and day ahead electricity demand) and long term (e.g., changes to the overall renewable penetration rate).

While it would be impractical to attempt to fully define operations across the full envisioned operating window of the proposed plant, it is prudent to define general operations at a number of key operating points. These points help to both define the bounds of the logical, intended operating window as well as provide relevant understanding of the advantages and trade-offs of operating the plant at different points.

2.3.1 *Balanced Ammonia and Electricity Generation, Three Turbines*

In support of the overall polygeneration design, it is important to investigate operating characteristics when the plant is producing a balance between a moderate to high level of production of both electricity for export and ammonia.

In this mode, ammonia production of 600 MTPD is achieved by operating two, 300 MTPD ammonia trains at full capacity. The power island delivers 48 MW of net power for export (101 MW gross), generated by three LM2500+ turbines operating at 67% of maximum capacity and running the primary steam turbine at 86% load. The LM2500+ turbines will be fueled by nitrogen-diluted syngas. PSA off-gas provides fuel to fire duct burners to support greater power generation in the steam turbine.

2.3.2 *Balanced Ammonia and Electricity Generation, Two Turbines*

One major advantage of the three-turbine design is the ability to utilize different combinations of equipment and operating conditions to achieve similar plant results. For example, it is possible to achieve roughly the same output of the *Balanced Ammonia and Electricity Production, Three Turbines* by using two turbines operating at a higher load.

Specifically, while ammonia production stays at 600 MTPD, the turbine operation shifts from three turbines at 67% capacity to two turbines at 100% capacity. Combined with a slightly higher utilization of the primary steam turbine (91% capacity, up from 86% capacity), the net power for export increases slightly to 51 MW (103 MW gross).

This ability to achieve roughly the same net plant outputs from different combinations of operating equipment characteristics allows for greater flexibility for the plant operator to efficiently and intelligently meet real-world demands. For example, if two turbines are already on-line, it is possible to quickly ramp up to the *Balanced Generation* point without the need to start the third turbine. If it is anticipated that no additional grid demand beyond the 51 MW of export will be requested in the near future, the plant can continue to operate on just the two turbines[14]. In contrast, if it is expected that grid demand for net export electricity will increase, the operator can begin the process of bringing the third turbine online. As it ramps up, the other two turbines can be turned down until all reach a steady state of 67% of capacity. While the net power export will still be similar to the *Balanced, Two Turbine* point, the plant will now be better positioned to quickly ramp up in response to future expected grid demands.

2.3.3 *Zero Net Power*

It is envisioned that there are times when the Independent System Operator (ISO) or Regional Transmission Organization (RTO) would require the polygeneration facility to fully curtail the electricity exported to the grid (i.e., the net electricity production will be set to zero). In this scenario, the proposed plant will need to significantly ramp down electrical generation such that only enough electricity is generated to meet internal demands and parasitic loads.

Fortunately, this polygeneration-based system offers a number of inherent advantages to limit the negative impacts of this turndown relative to the overall plant subsystems. First, even in scenarios where there is no net power export requested by the grid, it is anticipated that the ammonia train will still largely be operating at full capacity. This is not a small operation, relatively speaking, requiring that many of the other plant subsystems operate towards the upper one third of the operating ranges. Specifically, it is anticipated that the overall plant parasitic loads to maintain the ammonia trains at full capacity will be 40 MW (this compares to ~52 MW of parasitic loads in the *Balanced Generation* operating points). To supply enough syngas to generate 40 MW of power and provide sufficient feed to operate the ammonia trains at full capacity, the gasifier will be

[14] It is possible that a developer of this plant may assess the modeled financial performance of the plant and determine that the plant may not operate in a mode utilizing three generators often enough to justify the capital cost of the third generator. We defer that to be a project by project decision.

required to operate at 66% of its nameplate capacity. By limiting the overall turndown required by the majority of the plant subsystems, it is anticipated that the proposed design will reduce wear and tear on capital equipment, maintain reasonable efficiency across the projected operating ranges, and offer good transient response and capabilities.

The plant subsystem that will see the largest turndown will be the power block. While there will still be 40 MW of parasitic load that must be met, this can be accomplished using just one of the selected LM2500+ turbines operating at 67% of capacity paired with the steam turbine operating at 40% of capacity. This turbine will fire using nitrogen-diluted syngas while the PSA off-gas will be fired in the duct burners to increase output of the steam turbine.

2.3.4 *High Electricity Production*

In the *High Electricity Production* mode, the plant will have all three turbines in the power block operating at full capacity and the primary steam turbine operating at 88% capacity to provide a net export of 82 MW to the grid. This represents an increase of ~30 MW relative to the *Balanced Generation* operating points.

To achieve this higher next power export, significant amounts of syngas will need to be diverted to the power island from the ammonia production trains. As a result, the ammonia production will reduce from 600 MTPD to 380 MTPD, which is achieved by running both trains at 63% of capacity.

As the ammonia train is inherently a "recycle process" due to equilibrium limitations, it is anticipated to be able to handle this increase in recycle rates to accommodate the turndown without significant issue. The majority of the operational and control system design challenge will be assuring the heat integration between operating sections adapts smoothly during these turndown scenarios. The impact of transitioning through the operating window on utilities and heat integration have been considered, Appendix E and F provides relevant details of the integration. Additionally, since this scenario is essentially just shifting the overall syngas disposition to ensure that more syngas reaches the power block, there is no turndown required from any operating sections other than those directly involved in the ammonia production (e.g., the ammonia trains, ammonia compressors, syngas PSA to supply hydrogen to the ammonia train, etc.), reducing system transients and stresses on capital equipment.

In this scenario, all three LM2500+ combustion turbines will be operating at their full rated capacity, fueled entirely by nitrogen-diluted syngas. Additionally, the PSA off-gas will be the sole source of fuel used to fire the duct burners to increase the temperature of the turbine exhaust to support steam generation in the HRSG. As previously stated, the LM2500+ turbines in combined cycle configurations have ramp rates of over 60% per minute, relative to full load, once they have been started. This ensures that transitioning to this operating mode can occur in only a handful of minutes from any point on the operating window[15].

[15] Transitions to operating points assume the plant is running within the warm operating point window; cold start information is provided in Section 2.4.7

2.3.5 *Maximum Electricity Production*

It was also of interest to examine what the impacts and trade-offs would be of diverting even more syngas to the power island beyond what is seen in the *High Electricity Production* case. As the turbines are already operating at maximum capacity and the primary steam turbine is already at 88% of capacity, there is little room for additional net electricity generation without adding additional capital equipment.

Rather than adding a fourth combustion turbine, an additional, secondary steam turbine was selected instead as it represented the most efficient choice for increasing power production capabilities. By adding a secondary steam turbine with 25 MW of capacity, it is possible to operate the both ammonia trains at 10% of capacity (59 MTPD total) will producing 112 MW of power for export.

It should be noted that it is not intended for the plant to operate at this point for significant periods of time as it is fairly inefficient relative to the other described operating points. The primary reason for its inclusion is that it does provide greater operational flexibility by offering an increase of net power of export of nearly 40% relative to the *High Electricity Production* operating point with relatively low increase in capital expenditures. As flexibility is a key component of the Coal FIRST program, it is believed that a 40% increase in net export power available provides a legitimate value opportunity. However, individual plant operators will need to be judicious in how they leverage this greater flexibility to ensure that the benefits outweigh the costs associated with the much lower HHV efficiency.

2.3.6 *Summary of Operating Points*

A narrative summary of the described operating points can be seen in Figure 2-2, with a tabular representation in Table 2-3.

Figure 2-2. Summary Description of Defined Operating Points

Balanced Generation, 3 GT's	***Represents operating conditions when there is neither the need for surge net power generation to meet increased grid demand nor the need for power generation curtailment to respond to reductions in grid demand*** • Ammonia Train: Two, 300 MTPD trains operating at full capacity to provide 600 MTPD of ammonia product for export • Power Island: Three GE LM2500+ turbines operating at 67% capacity, primary steam turbine operating at 86% capacity, and secondary steam turbine operating at 0% capacity
Balanced Generation, 2 GT's	***Alternative means to provide output of "Balanced Generation, 3 GT's" Case to provide greater transient flexibility*** • Ammonia Train: Two, 300 MTPD trains operating at full capacity to provide 600 MTPD of ammonia product for export • Power Island: Two GE LM2500+ turbines operating at full capacity, primary steam turbine operating at 91% capacity, and secondary steam turbine operating at 0% capacity
Zero Net Power	***Represents operating conditions when net power production is zero in response to reductions in grid demand*** • Ammonia Train: Two, 300 MTPD trains operating at full capacity to provide 600 MTPD of ammonia product for export • Power Island: One GE LM2500+ turbine operating at 67% capacity, primary steam turbine operating at 40% capacity, and secondary steam turbine operating at 0% capacity
High Electricity Generation	***Represents operating conditions when ammonia generation reduced to support increased electrical grid demand*** • Ammonia Train: Two, 300 MTPD trains operating at 63% of overall capacity to provide ~380 MTPD of ammonia product • Power Island: Three GE LM2500+ turbines operating at 100% capacity, primary steam turbine operating at 88% capacity, and secondary steam turbine operating at 0% capacity
Max Electricity Generation	***Represents more extreme version of "High Electricity Generation" Case*** • Ammonia Train: Two, 300 MTPD trains operating at 10% of overall capacity to provide ~60 MTPD of ammonia product • Power Island: Three GE LM2500+ turbines operating at 100% capacity, primary steam turbine operating at 100% capacity, and secondary steam turbine operating at 85% capacity

Table 2-3. Summary Table of Defined Operating Points

Operating Point	Net Export Power	Ammonia Production	Gasifier Operation	GT Operation	ST Operation	Ammonia Train Operation
Balanced Generation, 3 GTs	48 MW	600 MTPD	100% of Capacity	Three Turbines @ 67% Capacity	Primary ST @ 86% load	Both Trains @ 100% Capacity
Balanced Generation, 2 GTs	51 MW	600 MTPD	100% of Capacity	Two Turbines @ 100% Capacity	Primary ST @ 91% Load	Both Trains @ 100% Capacity
Net Zero Power	0 MW	600 MTPD	66% of Capacity	One Turbine at 67% Capacity	Primary ST @ 40% Load	Both Trains @ 100% Capacity
High Electricity Production	82 MW	380 MTPD	100% of Capacity	Three Turbines @ 100% Capacity	Primary ST @ 88% Load	Both Trains @ 63% Capacity
Max Electricity Production	112 MW	59 MTPD	100% of Capacity	Three Turbines @ 100% Capacity	Primary ST @ 100% Load, Secondary ST @ 85% Load	Both Trains @ 10% Capacity

2.4 System Transients

In general, process plants are generally not designed for rapid turndown, although it is always an option to vent to flare in an emergency. As such, the most relevant transient cases discussed herein are those that require a turndown of the process equipment which, in this case, refers chiefly to equipment in the syngas production train (e.g., gasifier trains, ASU, shift reactors, AGR, SRU, CO_2 compressors, etc.) and in the ammonia loop. The impact of transitioning through the operating window on utilities and heat integration have been considered, Appendix E and Appendix F provide relevant details of the integration. The largest turndowns for these two process equipment groups are:

1. Ammonia Loop Train - *Balanced* operating mode to *High Electricity Production* operating mode, in which the ammonia loop reduces from 100% load down to 63% load.

2. Syngas Production Train - *High Electricity Production* Operating Mode to *Zero Net Power* Operating Mode where the syngas production train transitions from 100% load to 66% load. This is a particularly interesting transition to examine as it also represents a ramping of the ammonia loop from 63% load to 100% load.

Five transition cases are considered:

1. *Balanced Generation, Three Turbines* to *High Electricity Production*
2. *High Electricity Production* to *Zero Net Power*
3. *High Electricity Production* to *Max Electricity Production*
4. *Max Electricity Production* to *Balanced Generation, Two Turbines*
5. *Balanced Generation, Two Turbines* to *High Electricity Production*

2.4.1 *Balanced Generation, 3 GTs to High Electricity Production*

Starting Point - The gasifier, ASU, shift, AGR, SRU and CO_2 compression at 100% load, the ammonia loop is at 100% load, the three LM2500+ turbines are at 67% load, and primary steam turbine is at 86% load. Ammonia production is at 600 MTPD. Net power production 48 MW.

Finishing Point - The gasifier, ASU, shift, AGR, SRU and CO2 compression at 100% load, ammonia loop is at 63% load, the three LM2500+ turbines at 100% load, and the steam turbine at 88% load. Net power production 82 MW.

Table 2-4. Transient Case Study - Balanced Generation, Three Turbines to High Electricity Production

Operating Point	Net Export Power	Ammonia Production	Gasifier Operation	GT Operation	ST Operation	Ammonia Train Operation	Comments
Balanced Generation, 3 GTs	48 MW	600 MTPD	100% of Capacity	Three Turbines @ 67% Capacity	Primary ST @ 86% Load	Both Trains @ 100% Capacity	- Ramp in ~1 minute with NG firing (LM2500+ ramps @ 20MW/min)[16]
High Electricity Production	82 MW	380 MTPD	100% of Capacity	Three Turbines @ 100% Capacity	Primary ST @ 88% Load	Both Trains @ 65% Capacity	- Ammonia turndown and NG back-out in ~40-50 minutes
Delta	+34 MW	-220 MTPD	No Change	Three Turbines @ 33% Ramp	Primary ST @ 2% Ramp	Both Trains @ 35% Turndown	- Parasitic load stabilization in ~20 min

Transition Details - The gasifier, ASU, shift, AGR, SRU and CO_2 compression all remain at 100% operation and can essentially remain in steady state operation for this transition case. Based on the excellent ramp rates of the LM2500+ turbines (20MW/minute), it is anticipated that ramping them to full load for the turbines and steam generation should take only 1-2 minutes, assuming that there is adequate fuel supply available. This fuel supply can be met through the use of syngas stored on-

[16] Natural gas supply will be 80 MMscfd, based on a constraining scenario where the gasifiers fail while operating, natural gas can be used to both maintain output as well as restart the facility.

site or, if needed, by use of natural gas to supplement the produced syngas. Bringing the additional steam turbine capacity will take an additional 5-10 minutes.

At the same time, the turndown of the ammonia loop and reduction in associated parasitic power loads will begin. At a turndown rate of ~1% of full load per minute, the 35% ammonia train turndown required here will take 40-50 minutes with an additional 20 minutes required to stabilize refrigeration loads and other, ancillary parasitic loads. As the ammonia loop is turned down, syngas can be shifted to the power island, allowing for the use of stored syngas or natural gas to be gradually reduced until a steady state is reached.

As part of the energy integration strategy, the heat produced by the ammonia loop is used for considerable heat integration. As the ammonia loops are turned down, less feed to the ammonia loops is required to be pre-heated, lowering the overall heat integration needs during the transient. To address this, excess heat of reaction from the ammonia train is rejected to the air coolers by partial by-passing the hot side of the main feed/product interchanger. Additional detail is provided in Appendix E and Appendix F.

2.4.2 *High Electricity Production Operating Mode to Zero Net Power Operating Mode*

Starting Point - The gasifier, ASU, shift, AGR, SRU and CO_2 compression at 100% load, ammonia loop at 63% load, three LM2500+ turbines at 100% load, and primary steam turbine at 88% load. Net power production 82 MW.

Finishing Point - The gasifier, ASU, shift, AGR, SRU and CO_2 compression at 66% load, ammonia loop at 100% load, one LM2500+ turbine at 67% load, and primary steam turbine at 40% load. Net power production 0 MW.

Table 2-5. Transient Case Study - High Electricity Production to Net Zero Power

Operating Point	Net Export Power	Ammonia Production	Gasifier Operation	GT Operation	ST Operation	Ammonia Train Operation	Comments
High Electricity Production	82 MW	380 MTPD	100% of Capacity	Three Turbines @ 100% Capacity	Primary ST @ 88% Load	Both Trains @ 63% Capacity	- Syngas train turndown in 20-30 min (~1-2% per min)
Net Zero Power	0 MW	600 MTPD	66% of Capacity	One Turbine @ 67% Capacity	Primary ST @ 40% Load	Both Trains @ 100% Capacity	- Ammonia ramp in 40-50 minutes (0.8-1% per min)
Delta	-82 MW	+220 MTPD	34% Turndown	Two @ 100% Turndown, One @ 33% Turndown	Primary ST @ 48% Turndown	Both Trains @ 37% Ramp	- Power Island turndown in ~5-10 minutes - Excess syngas to storage - Utilize aux boiler for feed preheat

Transition Details – As with the previous transition scenario, the power block can generally respond much more rapidly than the other system components, with full turndown expected in under 10 minutes. In contrast, the ASU and gasifier (and the CO_2 compressor and AGR, to a lesser extent) can only change load by ~1-2% per minute (that holds in both turndown and ramping scenarios). Additionally, the ammonia loop can ramp at ~2% per minute. Overall, this results in ~20-30 minutes to turn down the syngas production loop, 40-50 minutes to reach 100% capacity on the ammonia loop, and an additional ~10-20 minutes to stabilize the refrigeration equipment associated with the ammonia loop.

For energy efficiency, the ammonia loop has considerable heat integration. As the ammonia loops ramp from 63% to 100% capacity, more ammonia loop feed preheating is required than can be recovered from the effluent from the reactor. The additional feed preheat is provided by a start-up feed preheater using HP steam (about 150 bar(a)) from the auxiliary boiler.

This disconnect between the time it takes the power block to transition compared to the rest of the plant can be primarily be addressed through the use of on-site storage. The lagging reduction in syngas production relative to the reduction in syngas demand by the system will result in excess syngas over that ~40-minute period. During that time, excess syngas can be sent to the syngas storage tanks.

2.4.3 *High Electricity Production to Max Electricity Production*

Starting Point - The gasifier, ASU, shift, AGR, SRU and CO_2 compression at 100% load, ammonia loop at 63% load, three LM2500+ turbines at 100% load, and primary steam turbine at 88% load. Net power production 82 MW.

Finishing Point - The gasifier, ASU, shift, AGR, SRU and CO_2 compression at 100% load, ammonia loop at 10% load, three LM2500+ turbines at 100% load, primary steam turbine at 100% load, and secondary steam turbine at 85% load. Net power production 112 MW.

Table 2-6. Transient Case Study - High Electricity Production to Max Electricity Production

Operating Point	Net Export Power	Ammonia Production	Gasifier Operation	GT Operation	ST Operation	Ammonia Train Operation	Comments
High Electricity Production	82 MW	380 MTPD	100% of Capacity	Three Turbines @ 100% Capacity	Primary ST @ 88% Load	Both Trains @ 63% Capacity	- Ammonia turndown in ~60 minutes - Parasitic load stabilization in ~20 minutes - Power island ramp in ~5-10 minutes
Max Electricity Production	112 MW	59 MTPD	100% of Capacity	Three Turbines @ 100% Capacity	Primary ST @ 100% Load, Secondary @ 85% Load	Both Trains @ 10% Capacity	
Delta	+30 MW	-321 MTPD	No Change	No Change	Primary ST @ 12% Ramp, Secondary ST @ 85% Ramp	Both Trains @ 53% Turndown	- Utilize NG or stored syngas for transition - Consider shut down of one ammonia train

Transition Details – The gasifier, ASU, shift, AGR, SRU and CO_2 compression all remain at 100% operation. The ammonia loop is ramped down at ~1% of full load per minute. The power island can change capacity much more rapidly, fully ramping in ~5-10 minutes. While waiting on additional syngas to be backed out of the lagging ammonia train, either stored syngas or supplemental natural gas can be utilized to meet the increased fuel demand of the power island.

Turndown of the ammonia loop will take ~50-60 minutes and a further ~20 minutes for the refrigeration system to stabilize. To maintain circulation within the ammonia train during this time, due to the higher recycle present under turndown, additional nitrogen (i.e., above the stoichiometric requirement for ammonia production) will be sent to the loop. This excess nitrogen will either be purged or reacted with hydrogen when ammonia production is ramped back up at some future point. As with the first transition example, as the ammonia loops' capacity is reduced, less feed is available for preheat, and excess heat of reaction is rejected to air coolers by partial by-passing of the main feed / product interchanger. It is assumed that the fixed heat loses from the synthesis loop mean that steam is not exported to the power block from the synthesis loop. Additional information is provided in Appendix E and Appendix F.

As stressed previously, overall plant efficiency at the *Max Electricity Production* point is relatively poor so it is not anticipated that the plant will operate there for extended periods of time. However, if the plant operator *does* expect that the plant will operate at this *Max Electricity Production* point for a considerable period of time (a week or more, for example), shutting down one ammonia train should be considered.

2.4.4 *Max Electricity Production to Balanced Generation, 2 GTs*

Starting Point - The gasifier, ASU, shift, AGR, SRU and CO_2 compression at 100% load, ammonia loop at 10% load, three LM2500+ turbines at 100% load, primary steam turbine at 100% load, and secondary steam turbine at 85% load. Net power production 112 MW.

Finishing Point - The gasifier, ASU, shift, AGR, SRU and CO_2 compression at 100% load, the ammonia loop is at 100% load, two LM2500+ turbines are at 100% load, and primary steam turbine is at 91% load. Duct burning is fired only from PSA off gas and ammonia loop flash gas, no syngas. Ammonia production is at 600 MTPD. Net power production 51 MW.

Table 2-7. Transient Case Study - Max Electricity Production to Balanced Generation, 2 GTs

Operating Point	Net Export Power	Ammonia Production	Gasifier	GT Operation	ST Operation	Ammonia Train Operation	Comments
Max Electricity Production	112 MW	59 MTPD	100% of Capacity	Three Turbines @ 100% Capacity	Primary ST @ 100% Load, Secondary @ 85% Load	Both Trains @ 10% Capacity	- Ammonia ramp in 80-90 minutes - Power island turndown in ~5-10 minutes - Excess syngas to storage
Balanced Generation, 2 GTs	51 MW	600 MTPD	100% of Capacity	Two Turbines @ 100% Capacity	Primary ST @ 91% Load	Both Trains @ 100% Capacity	
Delta	-61 MW	+541 MTPD	No Change	One @ 100% Turndown	Primary ST @ 9% Turndown, Secondary ST @ 85% Turndown	Both Trains @ 90% Ramp	- Utilize aux boiler for feed preheat

Transition Details – The gasifier, ASU, shift, AGR, SRU and CO_2 compression all remain at 100% operation. The ammonia loop is ramped up at 0.8-1% of full load per minute, taking 80-90 minutes to reach full capacity and for additional nitrogen to be purged from the loop. The power island can change capacity much more rapidly (about 5 to 10 minutes), to turn off one turbine and reduce the ST capacity. The additional syngas available while the increasing demand of the ammonia train lags the reduction in demand of the power island can be sent to on-site storage.

As noted previously in the second *System Transients* example, for energy efficiency, the ammonia loop has considerable heat integration. As the ammonia loops ramp from 10% to 100% capacity, more ammonia loop feed preheating is required than can be recovered from the effluent from the reactor. The additional feed preheat is provided by a start-up feed preheater using HP steam (about 150 bar(a)) from the auxiliary boiler. Additional information is provided in Appendix E and Appendix F.

While plant output is similar in both the *Balanced Generation, 2 GTs* and *Balanced Generation, 3 GTs,* operating points the *2 GT* solution should be used if the plant is expected to be at the *Balanced Generation* point for a considerable amount of time without the expectation of additional power demand. The advantage of the *2 GT* solution is that it is slightly more efficient with reduced maintenance costs deriving from the need to only operate two GTs (and supporting ancillary equipment) rather than three. This approach also allows for maintenance on the GT that is not in use.

In contrast, if it is expected that grid demand will increase in the near term, it is preferable to operate using the *3 GT* approach as it allows greater ramping and response since it will avoid the 30 minutes required to bring the shutdown GT up to full operating output.

2.4.5 *Balanced Generation, 2 GTs to High Electricity Production*

Starting Point - The gasifier, ASU, shift, AGR, SRU and CO_2 compression at 100% load, the ammonia loop is at 100% load, two LM2500+ turbines are at 100% load, and primary steam turbine is at 91% load. Ammonia production is at 600 MTPD. Net power production 51 MW.

Finishing Point - The gasifier, ASU, shift, AGR, SRU and CO2 compression at 100% load, ammonia loop is at 63% load, the three LM2500+ turbines at 100% load, and the steam turbine at 88% load. Net power production 82 MW.

Table 2-8. Transient Case Study - Balanced Generation, 2 GTs to High Electricity Production

Operating Point	Net Export Power	Ammonia Production	Gasifier	GT Operation	ST Operation	Ammonia Train Operation	Comments
Balanced Generation, 2 GTs	51 MW	600 MTPD	100% of Capacity	Two Turbines @ 100% Capacity	Primary ST @ 91% Load	Both Trains @ 100% Capacity	- Power island ramp in ~20-30 minutes - Ammonia turndown and in ~40-50 minutes - Parasitic load stabilization in additional ~20 min - Excess syngas to storage - Utilize aux boiler for feed preheat
High Electricity Production	82 MW	380 MTPD	100% of Capacity	Three Turbines @ 100% Capacity	Primary ST @ 88% Load	Both Trains @ 63% Capacity	
Delta	+31 MW	-220 MTPD	No Change	One @ 100% Ramp	Primary ST @ 3% Turndown,	Both Trains @ 37% Turndown	

Transition Details – The gasifier, ASU, shift, AGR, SRU and CO_2 compression all remain at 100% operation. Additional natural gas is fed to the power block to start-up the third, currently shutdown LM2500+ gas turbine and HRSG. This unit takes 20-30 minutes to reach 100% of capacity.

At the same time, turn down of the ammonia loop is started. To maintain stable operation of the ammonia loop, the turndown rate is limited to 0.8-1% of full load per minute, requiring ~40 minutes for the full 35% turndown to be completed with an additional ~20 minutes required to stabilize refrigeration and parasitic power loads.

In the early stages of start-up, the third GT (previously shut down) will be unable to utilize the full amount of syngas made available by the ammonia loop turndown. This excess syngas can be used to fire either the duct burners of the third HRSG or fed to syngas storage. As the ramp rate of the LM2500+ is fairly high after initial start-up (roughly 20 MW/min), it will eventually overtake the turndown of the ammonia loop, requiring continued burning of supplemental natural gas until the ammonia loop reaches steady state. As a consequence, total natural gas backout will not be completed for ~40-50 minutes after the start of the transition.

As part of the energy integration strategy, the heat produced by the ammonia loop is used for considerable heat integration. As the ammonia loops are turned down, less feed to the ammonia loops is required to be pre-heated, lowering the overall heat integration needs during the transient. To address this, excess heat of reaction from the ammonia train is rejected to the air coolers by partial by-passing the hot side of the main feed/product interchanger. Additional information is provided in Appendix E and Appendix F.

2.4.6 *Summary of System Transient Cases*

Table 2-9 summarizes the results of the five transient case studies presented.

Table 2-9. Summary of Transient Case Studies

Initial State	Final State	Delta			Transition Time		
		Net Power	Ammonia Product	Syngas Production	Power Island	Ammonia Train	Syngas Production
Balanced Generation, 3 GTs	High Electricity Production	+34 MW	-220 MTPD	No Change	~1 minute	~40-50 minutes for ammonia loop; additional 20 minutes for parasitic loads	N/A
High Electricity Production	Net Zero Power	-82 MW	+220 MTPD	-34% Capacity	~5-10 minutes	~40-50 minutes for ammonia loop; additional 20 minutes for parasitic loads	~20-30 minutes
High Electricity Production	Max Electricity Production	+30 MW	-321 MTPD	No Change	~5-10 minutes	~60 minutes for ammonia loop; additional 20 minutes for parasitic loads	N/A
Max Electricity Production	Balanced Generation , 2 GTs	-61 MW	+541 MTPD	No Change	~5-10 minutes	~80-90 minutes for ammonia loop; additional 20 minutes for parasitic loads	N/A
Balanced Generation, 2 GTs	High Electricity Production	+31 MW	-220 MTPD	No Change	~20-30 minutes	~40-50 minutes for ammonia loop; additional 20 minutes for parasitic loads	N/A

Transitions that are not covered above have been investigated, but do not warrant significant discussion. For example, transitions from Net Zero Power to other operating points occur with the ammonia loop fully operating, and basically involve ramping up of the power block, which has a fairly quick response. While we did not have an explicit description of the Balanced Generation 3 GTs or Balanced Generation 2 GTs operating points to the Net Zero Power operating point, these transitions are essentially milder and easier to implement variants of the High Electricity Production to Net Zero Power transition discussed in detail above.

2.4.7 *Initial Start-up*

The primary steps in a cold start of the syngas production train includes bringing the ASU online, heating up the gasifier and ASU, start-up of the AGR, introduction of coal into the gasifier and monitoring operational characteristics of components and product streams to ensure proper operation. The lagging variable in this process is the start-up of the ASU, which can take up to 48 hours to reach full product quality streams, although earlier operation can at times produce useable product quality. The GE LM2500+ has the capability to ramp from a cold start to full power in approximately 30 minutes with natural gas co-firing or stored syngas. Backing out of the supplemental natural gas or syngas from storage is driven by the ability to ramp up the syngas production train. The ammonia train is the unit with the longest start-up time, and hence is the limiting factor deciding the duration of a cold start. Start-up of the ammonia train is 24 – 48 hours and is largely driven by thermal management requirements related to the heat produced by this exothermic process.

3. Performance Results

At the *Balanced Production* operating points (obtained with either 3 GTs or 2 GTs), the plant produces ~50 MW of net power for export and 600 MTPD of ammonia with 90% carbon capture at a net plant efficiency of over 38% HHV.[17]

It is important to note that this 38% HHV efficiency is with carbon capture. As the CO_2 compressor alone represents over 10 MW of load (equivalent to 1.99% HHV efficiency), it should be clear that the plant is capable of achieving the Coal FIRST target of 40% HHV efficiency for non-capture cases.

[17] The net HHV efficiency for this plant is calculated as the combination of net power for export and the energy chemically stored as NH_3 divided by the total input energy of the input coal feed. While this approach is consistent with the approach found in other NETL reports, it is difficult to make a direct and equivalent comparison between this efficiency metric and the efficiency calculated for a traditional IGCC plant that is only producing electricity.

Table 3-1. Polygeneration Plant Performance Summary

Performance Summary	Balanced Production, 3 Turbine	Balanced Production, 2 Turbines	Zero Net Power	High Electricity Production	Maximum Electricity Production
Combustion Turbine Power, MWe	61	60	20	91	91
Steam Turbine Power, MWe	40	43	19	41	68
Total Gross Power, MWe	101	103	39	132	159
Total Energy Chemically Stored as NH_3, MW	156	156	156	99	15
Combined Gross Power and Chemical Storage	257	259	195	231	174
ASU Package, kWe	14,400	14,400	9,500	14,400	14,400
Gasifier, kWe	50	50	50	50	50
Acid Gas Removal, kWe	4,400	4,400	4,400	4,400	4,400
CO_2 Compression, kWe	10,700	10,700	7,100	10,700	10,700
Cooling Tower Fans, kWe	1,400	1,400	1,000	1,400	1,400
Steam System, kWe	800	800	700	900	800
Drier Vent Compressors, kWe	2,800	2,600	1,500	3,400	3,400
N_2 Diluent Compressor, kWe	4,000	4,000	4,000	3,100	2,000
GT Fuel Feed Compressor, kWe	1,000	900	300	1,400	1,400

Performance Summary	Balanced Production, 3 Turbine	Balanced Production, 2 Turbines	Zero Net Power	High Electricity Production	Maximum Electricity Production
Make-up Gas Compressors, kWe	4,800	4,800	4,800	3,100	500
Ammonia Plant Loop (Compressors, chillers, etc.) kWe	6,000	5,900	4,800	6,000	5,000
Miscellaneous Balance of Plant, kWe	4,900	4,600	3,500	4,500	5,300
Total Parasitic Load, MWe	52	52	40	50	46
Combined Net Power and Chemical Storage	205	207	155	181	128
HHV Net Plant Efficiency	38.3%[18]	38.8%	44.0%	33.8%	23.9%
As-Received Coal Feed, kg/hr (lb/hr)	70,900 (156,300)	70,900 (156,300)	46,900 (103,400)	70,900 (156,300)	70,900 (156,300)
HHV Thermal Input, MWt	534	534	352	534	534
LHV Thermal Input, MWt	515	515	340	515	515
CO_2 Emissions, lb/MMBtu	19.8	19.7	17.1	19.9	19.9

[18] This efficiency represents non-capture cases. The least efficient way to operate this plant in a non-capture mode would be to simply vent the CO_2 once captured, eliminating the need for the CO_2 compressors. This elimination of 10.7 MW of parasitic load adds 2.00% to overall HHV efficiency, resulting in a 40% HHV efficiency for non-capture cases at the *Zero Net Power* operating point, as well as at both *Balanced Generation* operating points.

3.1 Environmental Performance

The summary of plant air emissions is presented in Table 3-2.

Table 3-2. Polygeneration Plant Emissions Summary Across Defined Operating Points

Performance Summary	Metric	Balanced Production, 3 GTs	Balanced Production, 2 GTs	Zero Net Power	High Electricity Production	Maximum Electricity Production
SO_2, lb/MWh-gross	Power Island Only	0.00	0.00	0.00	0.00	0.00
	Plant Total	0.00	0.00	0.00	0.00	0.00
NO_x, lb/MWh-gross	Power Island Only	0.30	0.29	0.26	0.34	0.29
	Plant Total	0.12	0.12	0.05	0.20	0.26
Particulates, lb/MWh-gross	Power Island Only	.035	.034	.090	.027	.022
	Plant Total	.014	.014	.018	.015	.020
Hg, lb/MWh-gross	Power Island Only	2.2E-6	2.2E-06	4.0E-06	1.7E-06	1.42E-06
	Plant Total	8.8E-7	8.7E-07	8.1E-07	9.8E-07	1.30E-06
CO_2, lb/MWh-gross	Power Island Only	357	348	556	274	227
	Plant Total	140	138	112	157	208
HCl, lb/MWh-gross	Power Island Only	0.000	0.000	0.000	0.000	0.000
	Plant Total	0.000	0.000	0.000	0.000	0.000

For IGCC plants, criteria emissions are typically calculated based on the MWh-gross of the power island. To be consistent with this approach, emissions in Table 3-2 are reported on a "Power Island Only" basis. However, applying this standard to a polygeneration plant can serve to distort the actual emissions performance as it does not take into account the high energy content stored in the cogeneration product (in this case, ammonia). In order to try and provide a more complete picture of the emissions performance of the polygeneration plant, emissions are also reported relative to MWh-gross on a "Plant Total" basis, which consists of the sum of the gross MWh from the power island and the energy stored in the cogeneration product.

The two-stage Selexol AGR process is the primary means of controlling SO_2 emissions in the polygeneration plant. The intensity of Selexol process is driven by the 90% carbon-capture goal, resulting in sulfur removal from the syngas beyond the emissions targets. A Claus plant is used to convert the H_2S-rich stream from the AGR system is to elemental sulfur.

This facility has been designed to reduce the concentration of NOx in the HRSG stack gas to a maximum of 5 ppmvd adjusted to 15% O_2 (dry basis) during normal operation. The plant utilizes N_2 dilution to limit the concentration of NOx in the gas turbine exhaust to 25 ppmvd adjusted to 15% O_2 (dry basis). SCR units installed in the HRSG's further reduce the NOx in the flue gas from 25 to 5 ppmvd adjusted to 15% O2 (dry basis). This is accomplished through the reduction of NOx to N_2 and H_2O by the reaction with ammonia on the catalyst. This ammonia is injected into the flue gas in the HRSG's upstream of the SCR catalyst beds. The ammonia serves to activate the SCR catalyst as the flue gas passes through the catalyst beds. The addition of ammonia is controlled to limit the ammonia slip (i.e., the concentration in the stack gas) to 5 ppmvd.[19]

Particulate emissions from normal operation of the LM2500+ turbines have an expected value of 3.5 lb/hr. While it is unclear exactly how duct burning with an SCR would increase or reduce these emissions, preliminary estimates by Worley suggests that the net impact will be largely negligible.

An Hg removal efficiency of just under 99% is required to ensure that the Hg emissions limit is met in all cases. A sulfur-impregnated bed system consisting of two beds in series is capable of achieving Hg removal in excess of 99%.[20]

The AGR system is able to capture 90% of the carbon contained in the syngas, at which point is it is compressed prior to sequestration.

All HCl will be removed in the syngas scrubber and will not enter the syngas stream.

The carbon balance for the plant for the reference *Balanced Generation, 3 GTs* operating point can be seen in Table 3-3. The carbon input to the plant includes both the carbon in the coal feedstock as well as the carbon contained in the air that supplies both the ASU and the GTs in the power island. Carbon leaves the plant as carbon in the form gasifier waste[21], the captured CO_2 product,

[19] While waste gas from the ammonia process is fed to the duct burners, it occurs in relatively small quantities and is not expected to impact NOx emission performance. Please refer to Section 2.1.9 for additional discussion.

[20] This matches the claim which appears in NETL's *Cost and Performance Baseline for Fossil Energy Plants Volume 1b: Bituminous Coal (IGCC) to Electricity Revision 2b – Year Dollar Update* National Energy Technology Laboratory, "Cost and Performance Baseline for Fossil Energy Plants Volume 1b: Bituminous Coal (IGCC) to Electricity Revision 2b – Year Dollar Update," U.S. Department of Energy, Pittsburgh, PA, 2019.)

[21] The *Baseline* reports refer to this generally as "slag", but the intention seems to be to ensure that the carbon capture balance/carbon capture performance calculations does not penalize a plant for unburnt carbon that exits the system as waste in

and CO_2 emitted to the atmosphere (this includes the stack gas from the power island as well as any vent gases from the various plant processes and equipment).

Table 3-3. Carbon Balance[22]

Carbon In		**Carbon Out**	
	kg/hr (lb/hr)		**kg/hr (lb/hr)**
Coal	45,172 (99,588)	Emitted to Atmosphere	4,487 (9,892)
Air (CO_2)	134 (295)[23]	CO_2 Product	39,916 (87,999)
		Gasifier Waste	903 (1,991)
Total	45,306 (99,883)	***Total***	45,306 (99,883)

$$\left(1-\left(\frac{Carbon\ to\ Atmosphere}{(Total\ Carbon\ In)-(Carbon\ in\ Gasifer\ Waste)}\right)\right)*100 \qquad \textit{Eq. 3-1}$$

$$\left(1-\left(\frac{4{,}487}{(45{,}306-903)}\right)\right)*100=90\%$$

The sulfur balance of the plant can be seen in Table 3-4 for the *Balanced Generation, 3 GTs* reference operating point. Sulfur input comes solely from the sulfur content in the coal feed. Sulfur outputs include both the elemental sulfur recovered in the Claus Plant as well as any sulfur content in the CO_2 product.

the gasification step. While the polygeneration plant use a "non-slagging" gasifier, there is ~2% of the carbon content of the coal feed that is lost in the gasification process. To account for this, the "slag" component of the carbon balance has been replaced with the more generic "Gasifier Waste" component.

[22] Additional Carbon Balance tables and calculations for the additional operating points can be seen in Appendix G.

[23] This represents the value of carbon contained in the air supplied to the GTs as well as the air supplied to the ASU using the air composition found in Appendix B.

Table 3-4. Sulfur Balance[24]

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	1,776 (3,916)	Emitted to Atmosphere	-
		CO_2 Product	8 (18)
		Elemental Sulfur	1,768(3,898)
Total	1,776 (3,916)	***Total***	1,776 (3,916)

Table 3-5 provides a water balance for the *Balanced Generation*, *3 GTs* operating point. The only defined operating point with a water balance that differs from the *Balanced Generation*, *3 GTs* operating point in Table 3-5 (allowing for round-off error) is the *Zero Net Power* operating point. The water balance at this operating point can be seen in Table 3-6.

[24] The sulfur balance for the *Balanced Generation, 2 GTs* operating point, the *High Electricity* operating point, and the *Max Electricity* operating point matches what is presented in Table 3-4 as there are no operational changes until after the AGR between these operating modes and the reference *Balanced Generation, 3 GTs* operating point. There are some differences in the *Zero Net Power* operating point (due primarily to the reduced feedstock flow) that can be seen in Appendix G.

Table 3-5. Balanced Generation, 3 GTs Water Balance

Water Makeup Area	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Discharge, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Consumption, m^3/min (gpm)
Coal Water in Feed	0	0	0	0.13 (35)	-0.13 (-35)
Raw Water to AGR	0.01 (3)	0	0.01 (3)	0.01 (3)	0
Raw Water SRU and TGT	0.08 (22)	0	0.08 (22)	0.08 (22)	0
Water Reaction Gasification	0	0	0	-0.43 (-114)	0.43 (114)
Water Reaction Shift	0	0	0	-0.75 (-199)	0.75 (199)
Water Reaction SRU and TGT	0	0	0	0.02 (7)	-0.02 (-4)
Cooling Tower	**5.22 (1378)**	**1.79 (472)**	**3.43 (907)**	**1.27 (335)**	**2.17 (572)**
Cooling Tower Blowdown	1.27 (335)	0	1.27 (335)	1.27 (335)	0
Cooling Tower Drift/Evaporation[1]	3.95 (1044)	0	3.95 (1044)	0	3.95 (1044)
ASU Knockout to CT Make-up	0	0.02 (6)	-0.02 (-6)	0	-0.02 (-6)
Desaturator (SWS Bottoms) to CT Make-up	0	1.76 (466)	-1.76 (-466)	0	-1.76 (-466)

Water Makeup Area	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Discharge, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Consumption, m^3/min (gpm)
Desaturator Make-up	**2.96 (681)**	**0**	**2.96 (781)**	**2.96 (781)**	**0**
IP Superheated Steam to Gasification	0.79 (210)	0	0.79 (210)	0.79 (210)	0
IP Superheated Steam to Shift	2.11 (557)	0	2.11 (557)	2.11 (557)	0
Steam Drum Blowdown and Makeup Requirement	0.06 (15)	0	0.06 (15)	0.06 (15)	0
Total:	**8.27 (2185)**	**1.79 (472)**	**6.48 (1713)**	**3.29 (870)**	**3.19 (843)**

Table 3-6. Zero Net Power Water Balance

Water Makeup Area	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Discharge, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Consumption, m^3/min (gpm)
Coal Water in Feed	0	0	0	0.13 (35)	-0.13 (-35)
Raw Water to AGR	0.01 (3)	0	0.01 (3)	0.01 (3)	0
Raw Water SRU and TGT	0.08 (22)	0	0.08 (22)	0.08 (22)	0
Water Reaction Gasification	0	0	0	-0.43 (-114)	0.43 (114)
Water Reaction Shift	0	0	0	-0.47 (-124)	0.47 (124)
Water Reaction SRU and TGT	0	0	0	0.02 (4)	-0.02 (-4)
Cooling Tower	**5.22 (1378)**	**1.17 (309)**	**4.05 (1069)**	**1.27 (335)**	**2.78 (735)**
Cooling Tower Blowdown	1.27 (335)	0	1.27 (335)	1.27 (335)	0
Cooling Tower Drift/Evaporation[1]	3.95 (1044)	0	3.95 (1044)	0	3.95 (1044)
ASU Knockout to CT Make-up	0	0.02 (6)	-0.02 (-6)	0	-0.02 (-6)
Desaturator (SWS Bottoms) to CT Make-up	0	1.15 (303)	-1.15 (-303)	0	-1.15 (-303)
Desaturator Make-up	**2.17 (6519)**	**0**	**2.17 (575)**	**2.17 (575)**	**0**

Water Makeup Area	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Discharge, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Consumption, m^3/min (gpm)
IP Superheated Steam to Gasification	0.79 (210)	0	0.79 (210)	0.79 (210)	0
IP Superheated Steam to Shift	1.35 (356)	0	1.35 (356)	1.35 (356)	0
Steel Drum Blowdown and Makeup Requirement	0.03 (9)	0	0.03 (9)	0.03 (9)	0
Total:	**7.49 (1978)**	**1.17 (309)**	**6.32 (1669)**	**2.79 (736)**	**3.53 (933)**

A water block flow diagram, with accompanying stream tables/heat and mass balance for the five defined operating points, can be seen in Figure 3-1and Table 3-7.

The following details are meant to provide additional insight regarding these streams and flows:

- The balance of the process condensate from the Gas Cooling section is fed to the Sour Water Stripper (SWS). The SWS bottoms is recycled within the process – to the SRU quench, Desaturator, and AGR makeup. Excess water beyond this is used for Cooling Tower makeup to offset raw water withdrawal.

- There is no waste water stream from the mercury removal section. Wastewater from the AGR and SRU/TGT are directed to waste water treatment.

- After waste water treatment, the fate of the clean water is a return to source. Sludge would be collected and taken away for solid waste disposal.

- The cooling tower makeup requirement is estimated to be 5.2 m^3/min.[25]

- The cooling tower make-up is supplied by raw water and supplemented by recycled water from ASU compression and recycled water from the Desaturator SWS bottoms purge.

- The capacity of the waste water treatment plant is estimated to be 3.3 m^3/min based on the overall water balance.

- Effluent or waste water streams directed to the waste water treatment plant are generally grouped as the following classifications:

    - The first is blowdowns and consist of streams from the cooling tower and steam cycle systems. These effluents contain concentrated salts and minerals that are present in the raw feed water.

    - The second is waste water streams from the process that may contain dissolved solids, trace metals, chloride, fluoride, sulfide and other ionic species.

- The wastewater system is designed to treat the wastewater and reduce / eliminate contaminants to an acceptable level in line with permit and environmental requirements.

- A more detailed analysis of contaminants in each process waste water stream will be performed in the next phase when the specific site location has been identified and thus specific environmental regulations will be known. However, the blowdown and waste water streams are typical of a coal gasification and ammonia production facility and will contain many of the same unit operations including filtration, flocculation, API/CPI, bio.

[25] This cooling tower makeup rate is based on the following assumptions:

- Evaporative losses: 0.0008 * Cooling Tower Temp Range * Water Recirculation Rate
- Drift losses: 0.0002 * Water Recirculation Rate
- Cycles of Concentration: 4
- Blowdown Losses: [Evaporative Losses + Drift Losses – (Cycles of Concentration X Drift Losses)] / (Cycles of Concentration – 1)

Figure 3-1. Polygeneration Plant Water Balance Block Flow Diagram

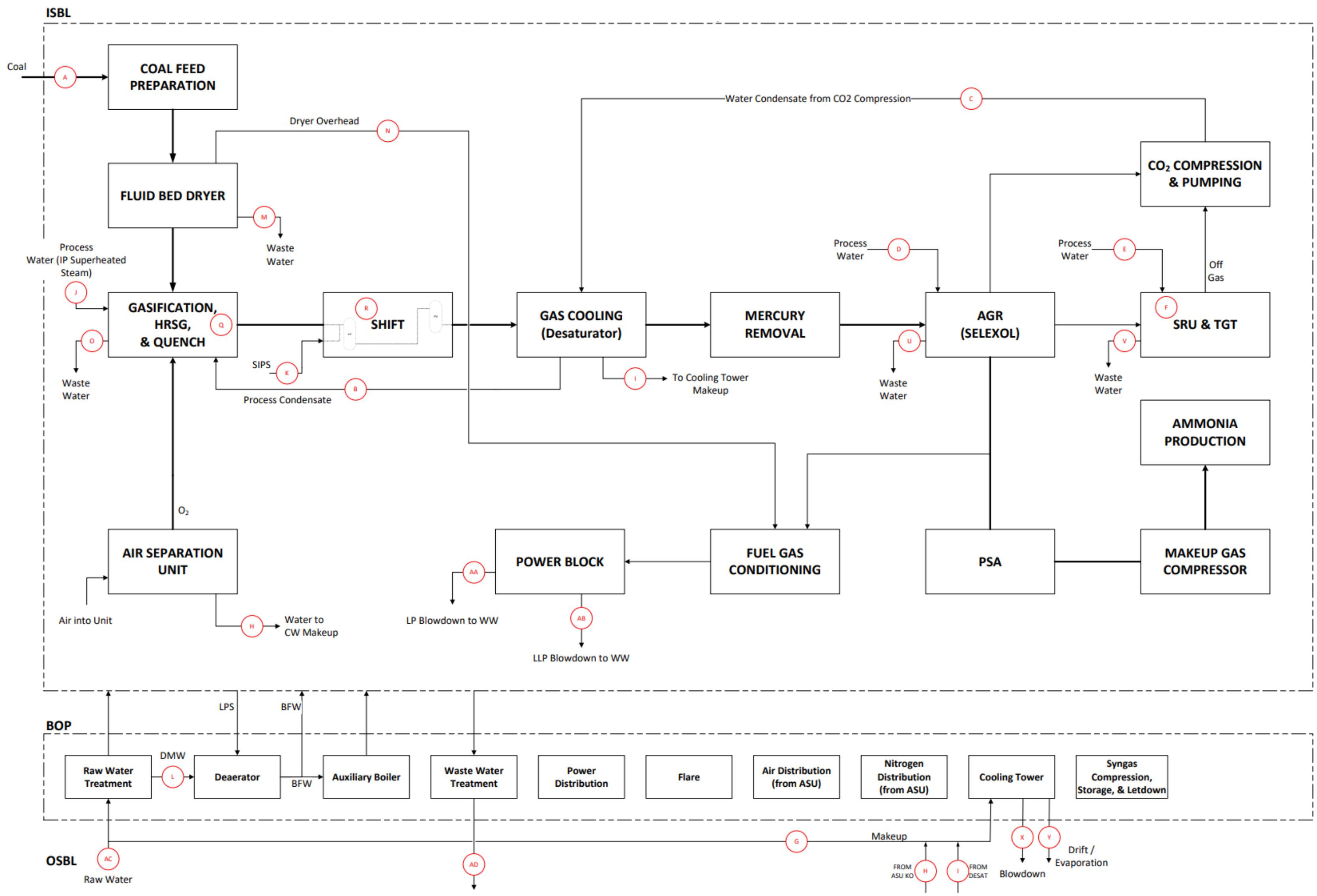


Table 3-7. Polygeneration Plant Water Balance Stream Tables

	Stream Reference and Name	Balanced Production, 3 Turbine	Balanced Production, 2 Turbines	Zero Net Power	High Electricity Production	Maximum Electricity Production
A	Coal Feed (Water supplied with Coal)	0.13 (35)	0.13 (35)	0.13 (35)	0.13 (35)	0.13 (35)
B	Process Condensate Recycle from Gas Cooling (Desaturator) to Gasification	0.33 (88)	0.33 (88)	0.33 (88)	0.33 (88)	0.33 (88)
C	Water Condensate Recycle from CO2 Comp. to Gas Cooling (Desaturator)	0 (1)	0 (1)	0 (1)	0 (1)	0 (1)
D	Process Water to AGR	0.01 (3)	0.01 (3)	0.01 (3)	0.01 (3)	0.01 (3)
E	Process Water to SRU/TGT	0.08 (22)	0.08 (22)	0.08 (22)	0.08 (22)	0.08 (22)
F	Water Generated in SRU/TGT Reaction	0.02 (7)	0.02 (7)	0.02 (4)	0.02 (7)	0.02 (7)
G	Cooling Tower Make-Up (from Raw Water Treatment)	3.43 (907)	3.43 (907)	4.05 (1069)	3.43 (907)	3.43 (907)
H	Cooling Tower Make-Up (Recycle from ASU Knockout)	0.02 (6)	0.02 (6)	0.02 (6)	0.02 (6)	0.02 (6)
I	Cooling Tower Make-Up (Recycle from Desaturator Water (SWS Bottoms)	1.76 (466)	1.76 (466)	1.15 (303)	1.76 (466)	1.76 (466)
J	IP Superheated Steam to Gasification	0.79 (210)	0.79 (210)	0.79 (210)	0.79 (210)	0.79 (210)
K	IP Superheated Steam to Shift	2.11 (557)	2.11 (557)	1.35 (356)	2.11 (556)	2.11 (557)
L	Steam Drum Blowdown Makeup	0.06 (15)	0.06 (15)	0.03 (9)	0.06 (15)	0.06 (15)
M	Fluidized Bed Dryer (Water from KO)	0.07 (19)	0.07 (18)	0.04 (10)	0.08 (20)	0.07 (20)
N	Fluidized Bed Dryer (Water to GT from Dryer Vent)	0.01 (2)	0.01 (2)	0.04 (10)	0 (0)	0 (1)
O	Gasification, HRSG & Quench Waste Water	0.05 (15)	0.05 (15)	0.15 (39)	0.05 (14)	0.05 (15)

	Stream Reference and Name	Balanced Production, 3 Turbine	Balanced Production, 2 Turbines	Zero Net Power	High Electricity Production	Maximum Electricity Production
P	Water Consumed in Gasification Reaction	0.43 (114)	0.43 (114)	0.43 (114)	0.43 (114)	0.43 (114)
Q	Water Consumed in Shift Reaction	0.75 (199)	0.75 (199)	0.47 (124)	0.75 (199)	0.75 (199)
R	AGR Waste Water	0.01 (4)	0.01 (4)	0.01 (4)	0.01 (4)	0.01 (4)
S	SRU/TGT Waste Water	0.12 (32)	0.12 (32)	0.11 (30)	0.12 (32)	0.12 (32)
T	Cooling Tower Blowdown to Waste Water	1.27 (335)	1.27 (335)	1.27 (335)	1.27 (335)	1.27 (335)
U	Cooling Tower Drift / Evaporation	3.95 (1044)	3.95 (1044)	3.95 (1044)	3.95 (1044)	3.95 (1044)
V	IP Steam Blowdown to Waste Water	0.02 (5)	0.02 (5)	0.01 (2)	0.02 (6)	0.02 (5)
W	LP Steam Blowdown to Waste Water	0.04 (9)	0.04 (9)	0.02 (6)	0.04 (9)	0.04 (9)
X	HRSG (LLP Drum) Steam Blowdown to Waste Water	0 (0.3)	0 (0.3)	0 (0.3)	0 (0.3)	0 (0.3)
Y	Raw Water Withdrawal	6.48 (1713)	6.48 (1713)	6.32 (1669)	6.49 (1713)	6.48 (1713)
Z	Treated Waste Water	3.35 (884)	3.35 (884)	2.76 (729)	3.35 (886)	3.35 (885)
AA	Coal Feed (Water supplied with Coal)	0.13 (35)	0.13 (35)	0.13 (35)	0.13 (35)	0.13 (35)
AB	Process Condensate Recycle from Gas Cooling (Desaturator) to Gasification	0.33 (88)	0.33 (88)	0.33 (88)	0.33 (88)	0.33 (88)
AC	Water Condensate Recycle from CO2 Comp to Gas Cooling (Desaturator)	0 (1)	0 (1)	0 (1)	0 (1)	0 (1)
AD	Process Water to AGR	0.01 (3)	0.01 (3)	0.01 (3)	0.01 (3)	0.01 (3)

3.2 Process Flow Diagrams

Process Flow Diagrams can be seen in Figure 3-2 to Figure 3-15.

Figure 3-2. Air Separation Unit Process Flow Diagram

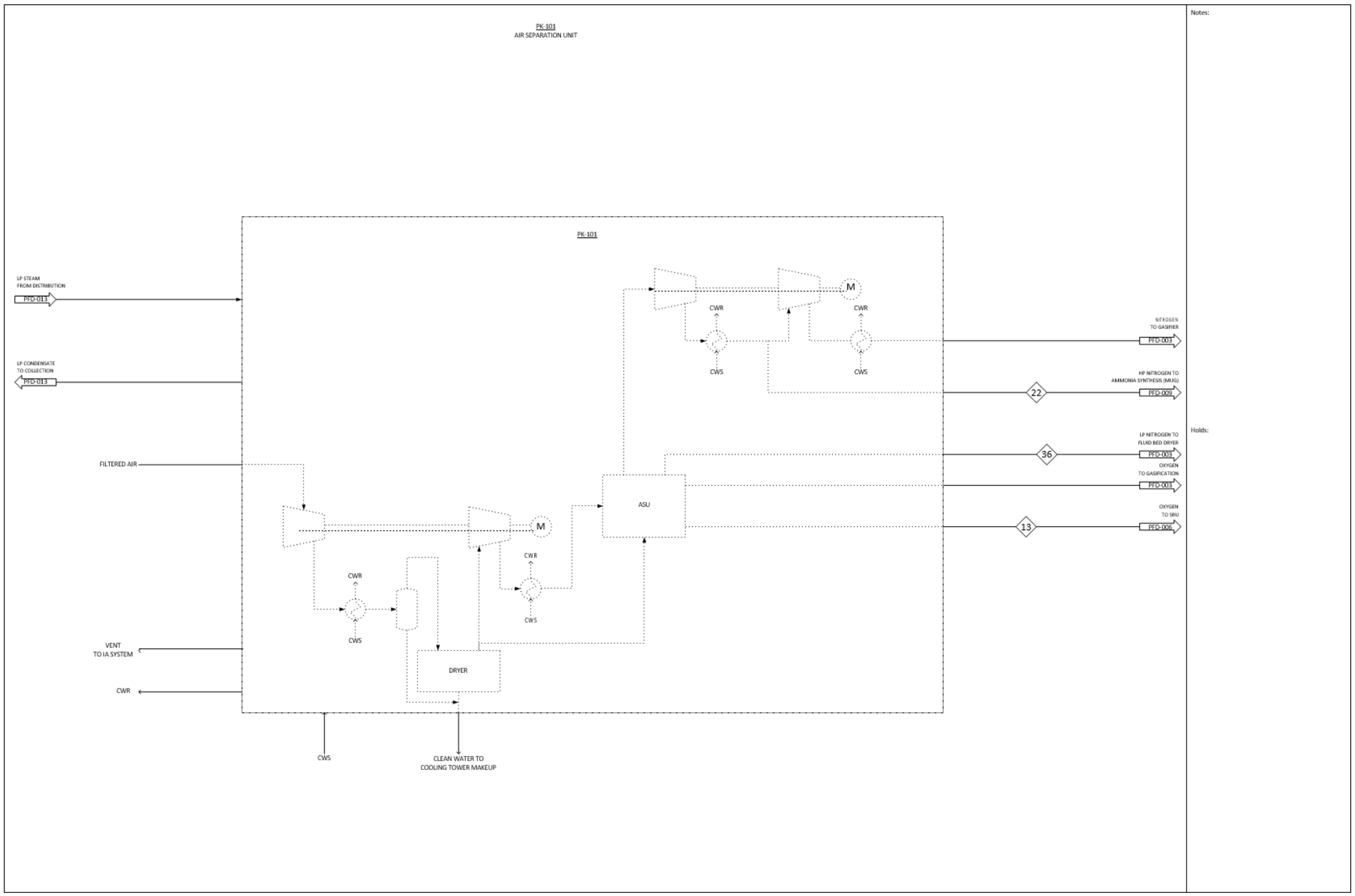


Figure 3-3. Coal Crushing and Handling Process Flow Diagram

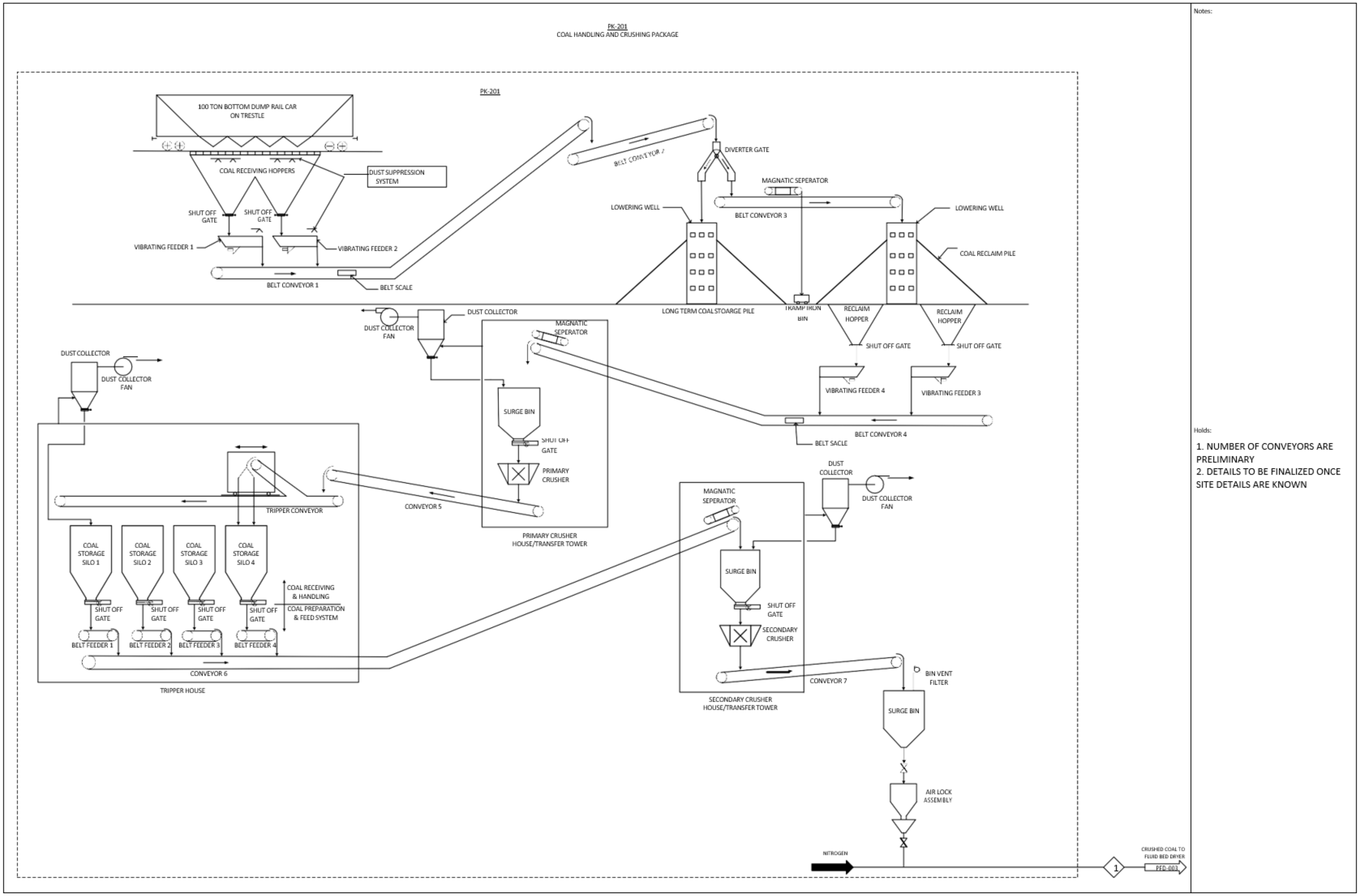


Figure 3-4. Gasifier, HRSG, and Quench Process Flow Diagram

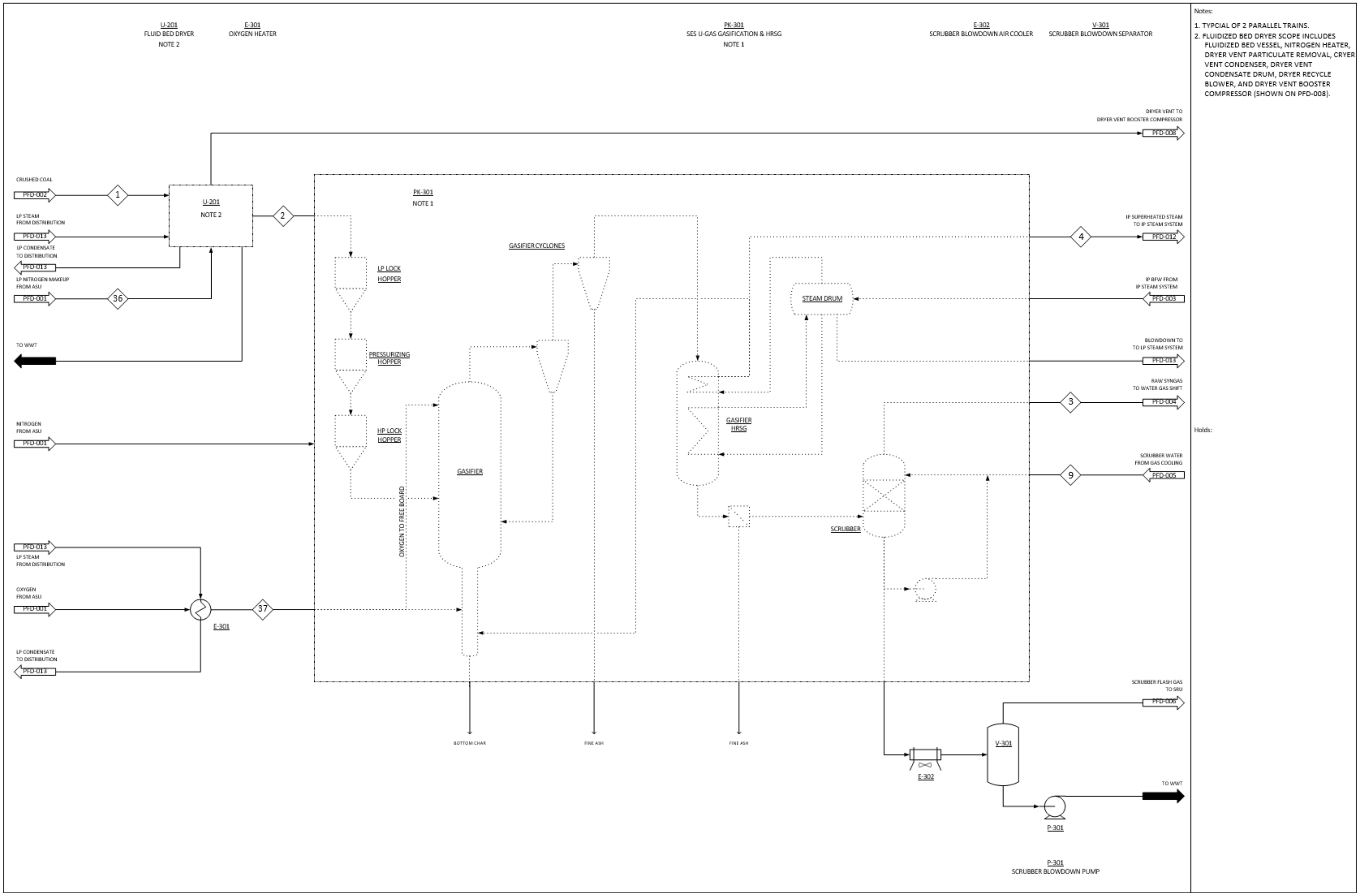


Figure 3-5. Water Gas Shift Process Flow Diagram

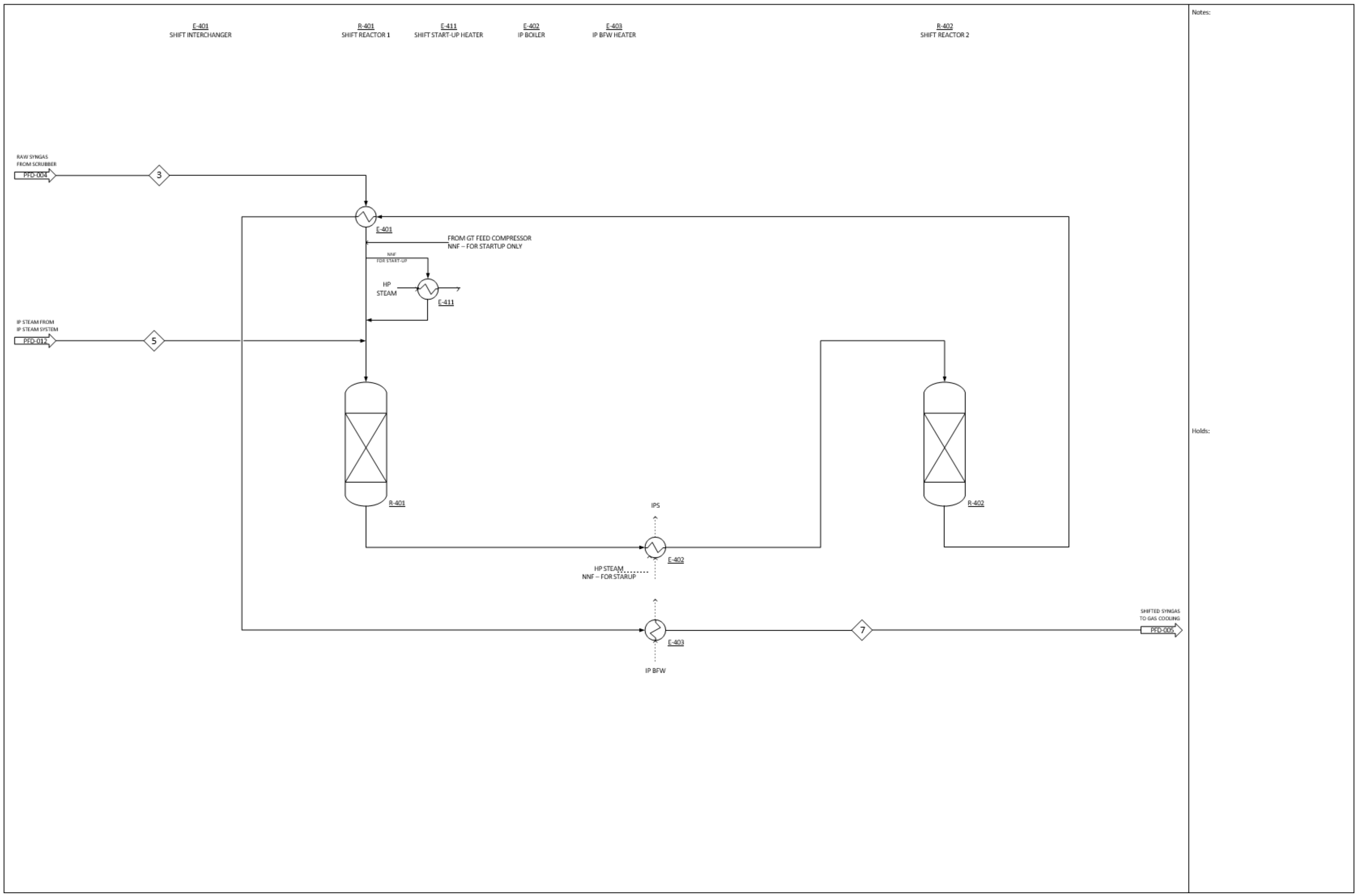


Figure 3-6. Syngas Cooling Process Flow Diagram

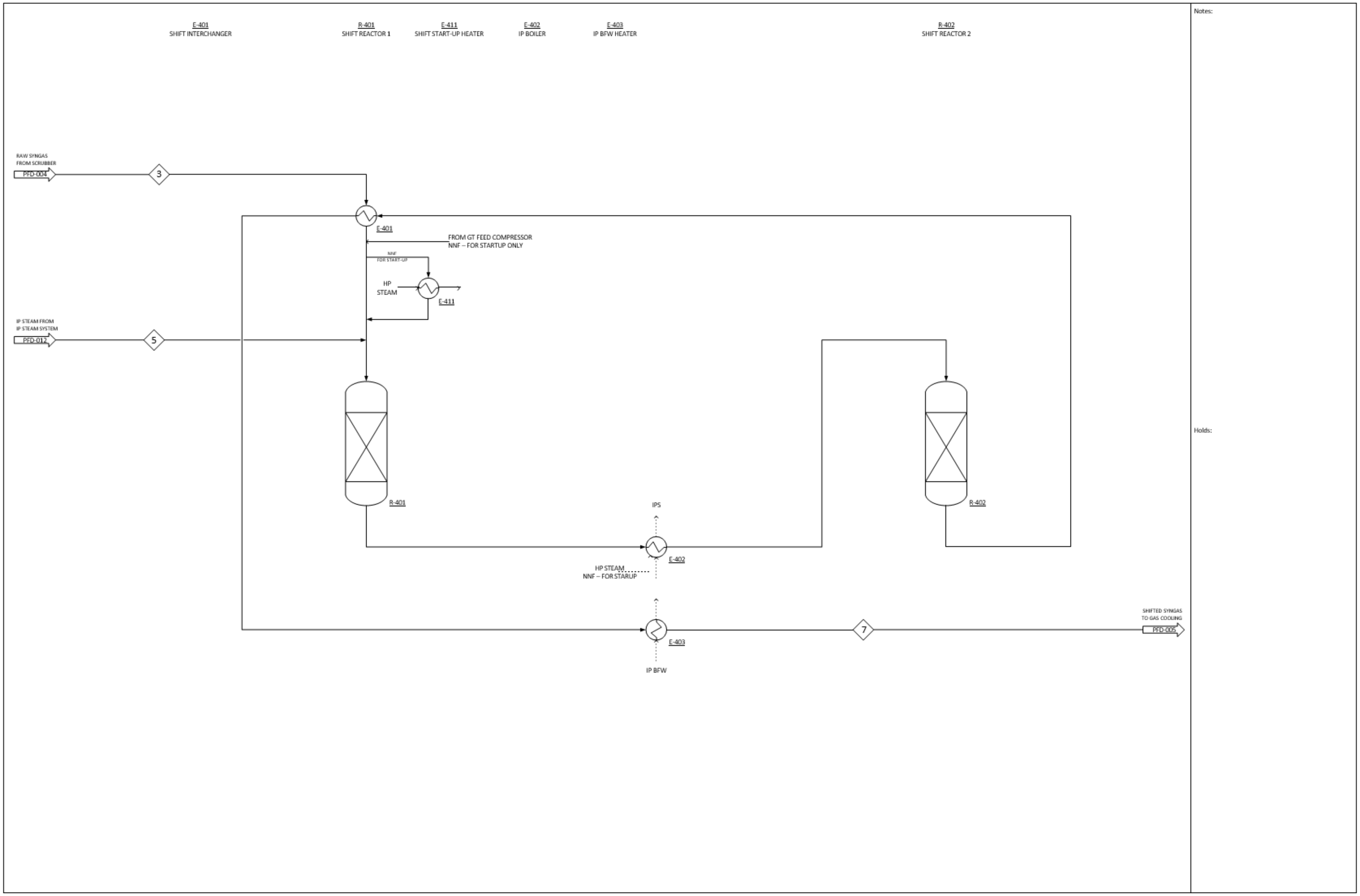


Figure 3-7. Syngas Clean-Up Process Flow Diagram

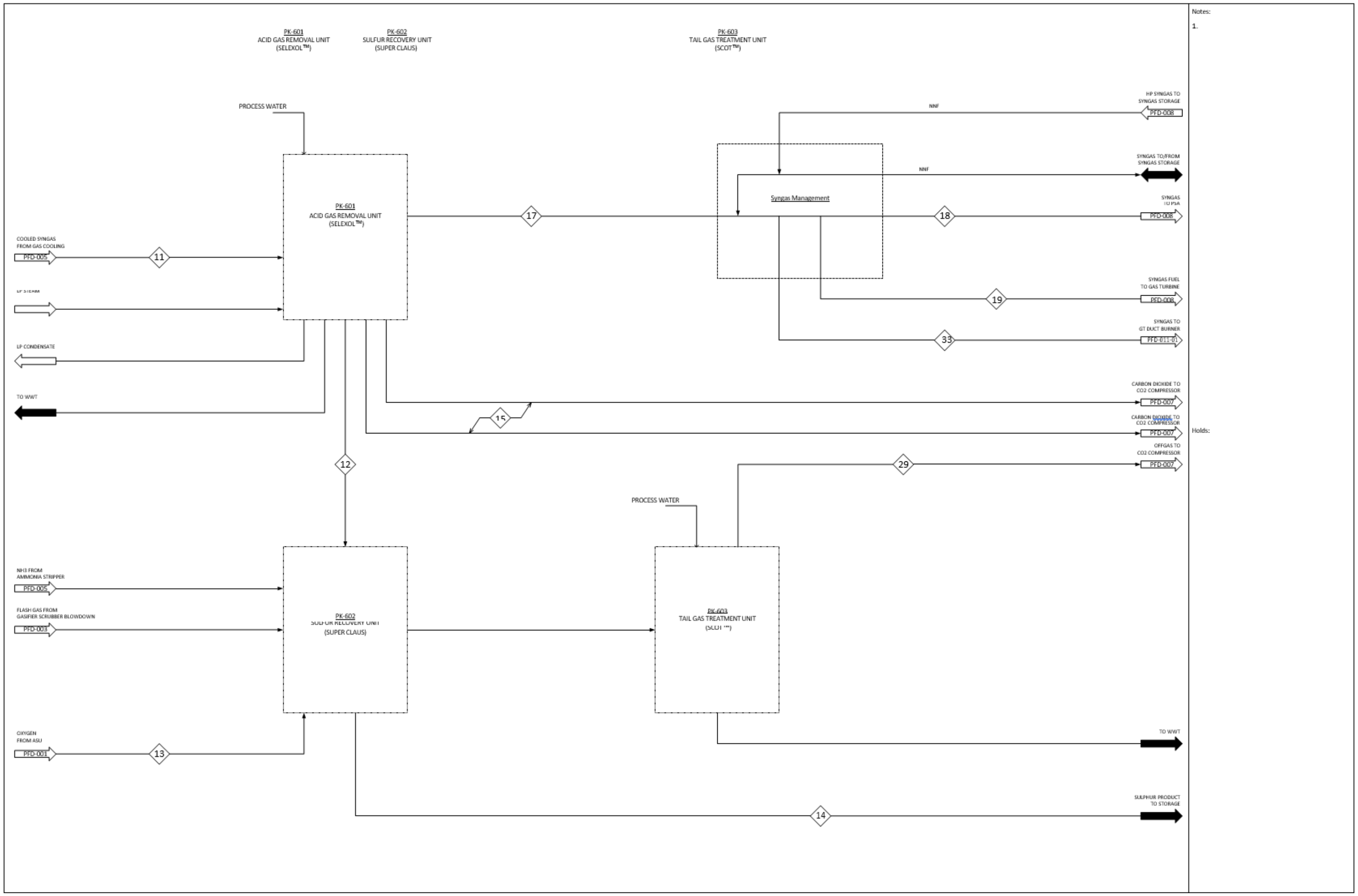


Figure 3-8. CO_2 Compression, Drying, and Pumping Process Flow Diagram

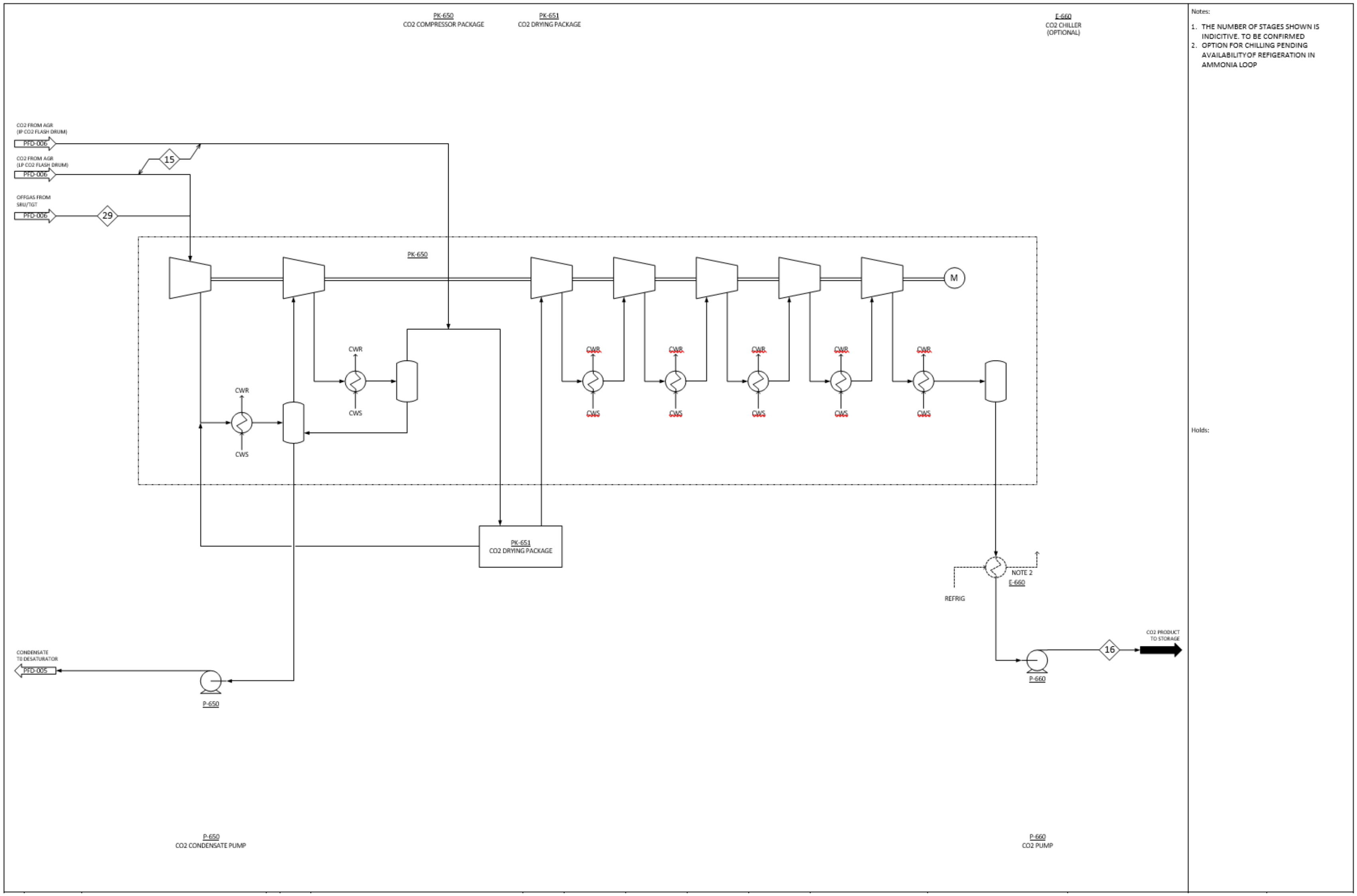


Figure 3-9. Fuel Gas Conditioning Process Flow Diagram

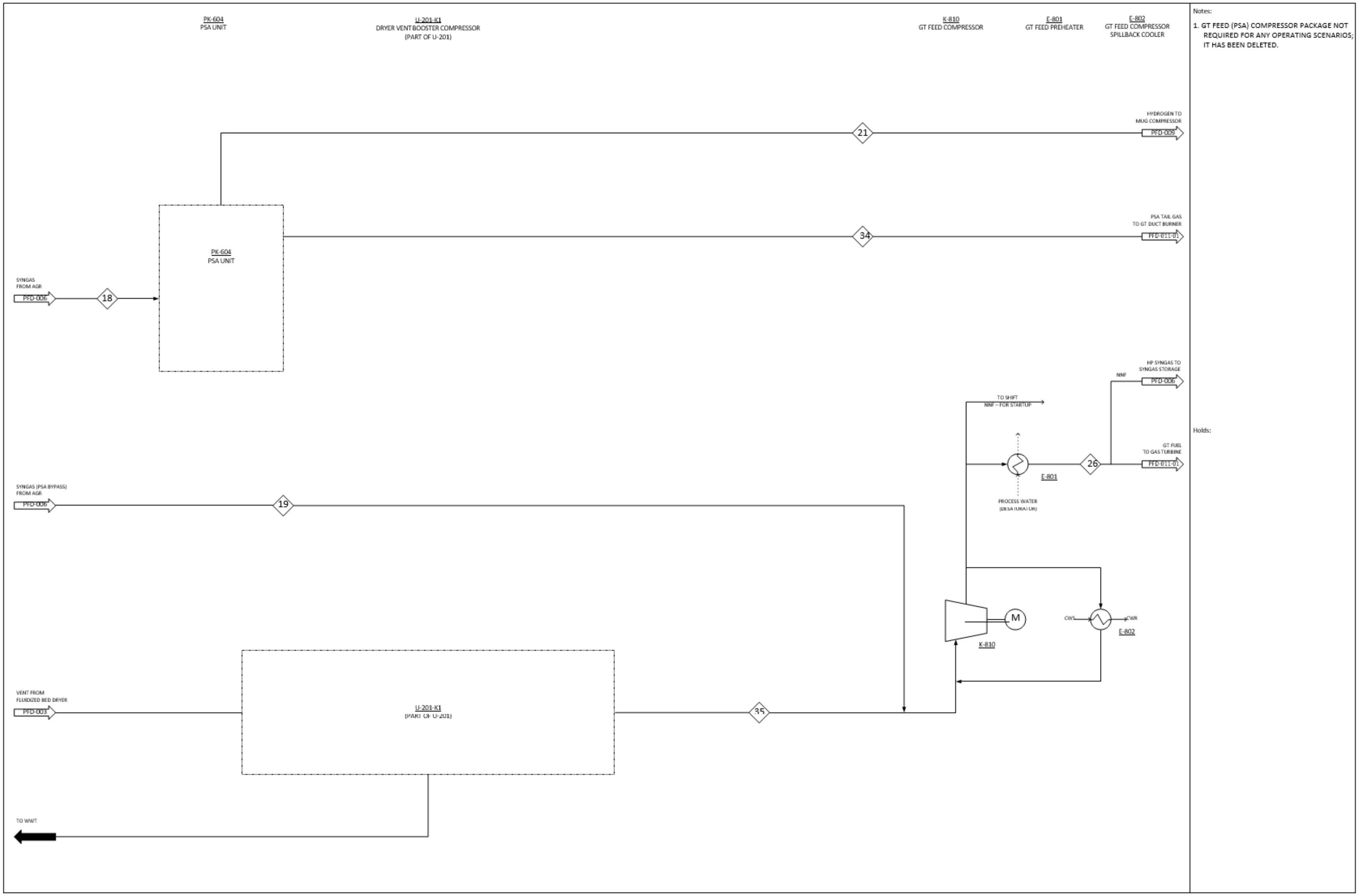


Figure 3-10. Make-up Gas Compressor Process Flow Diagram

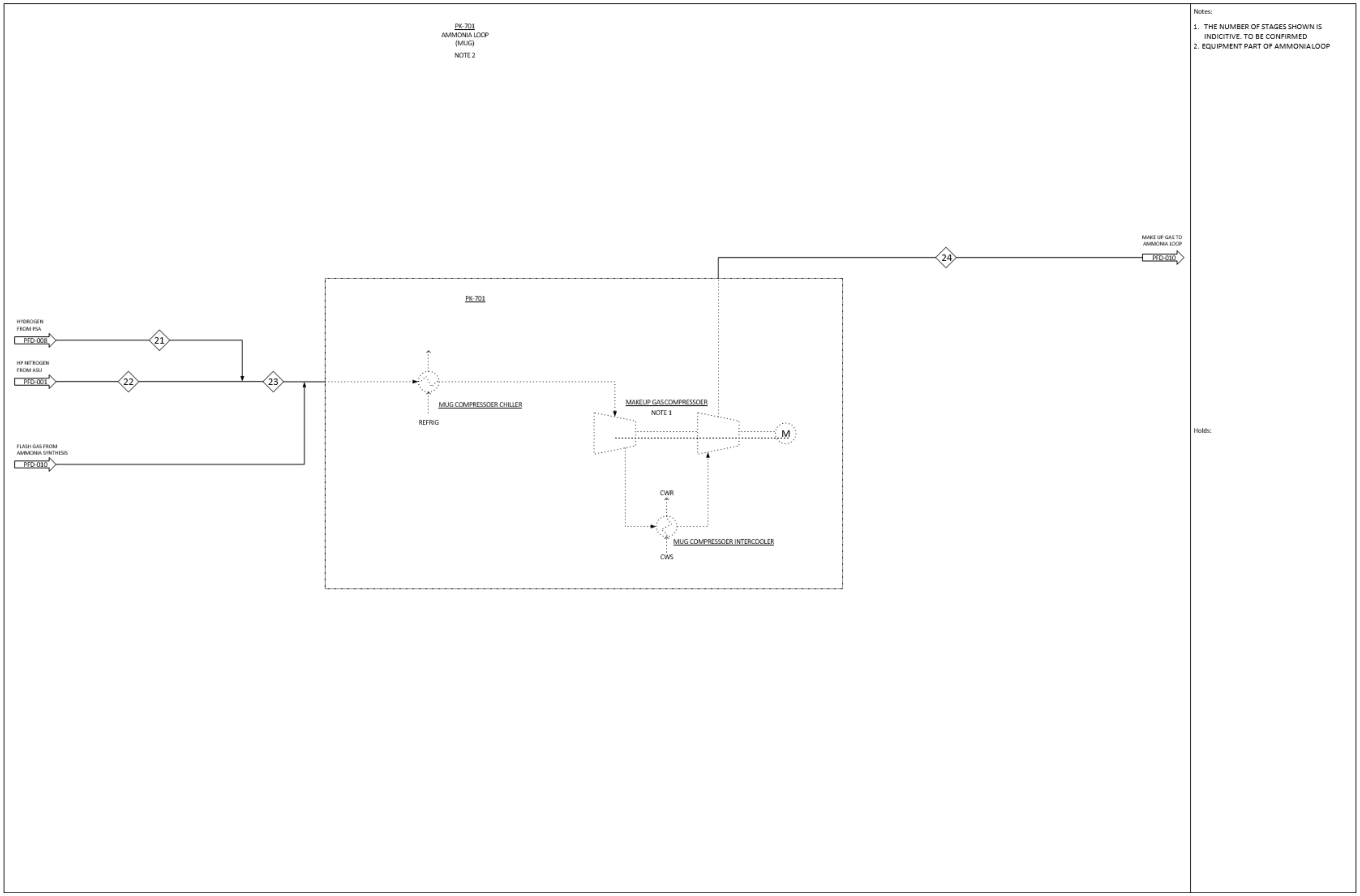


Figure 3-11. Ammonia Loop Process Flow Diagram

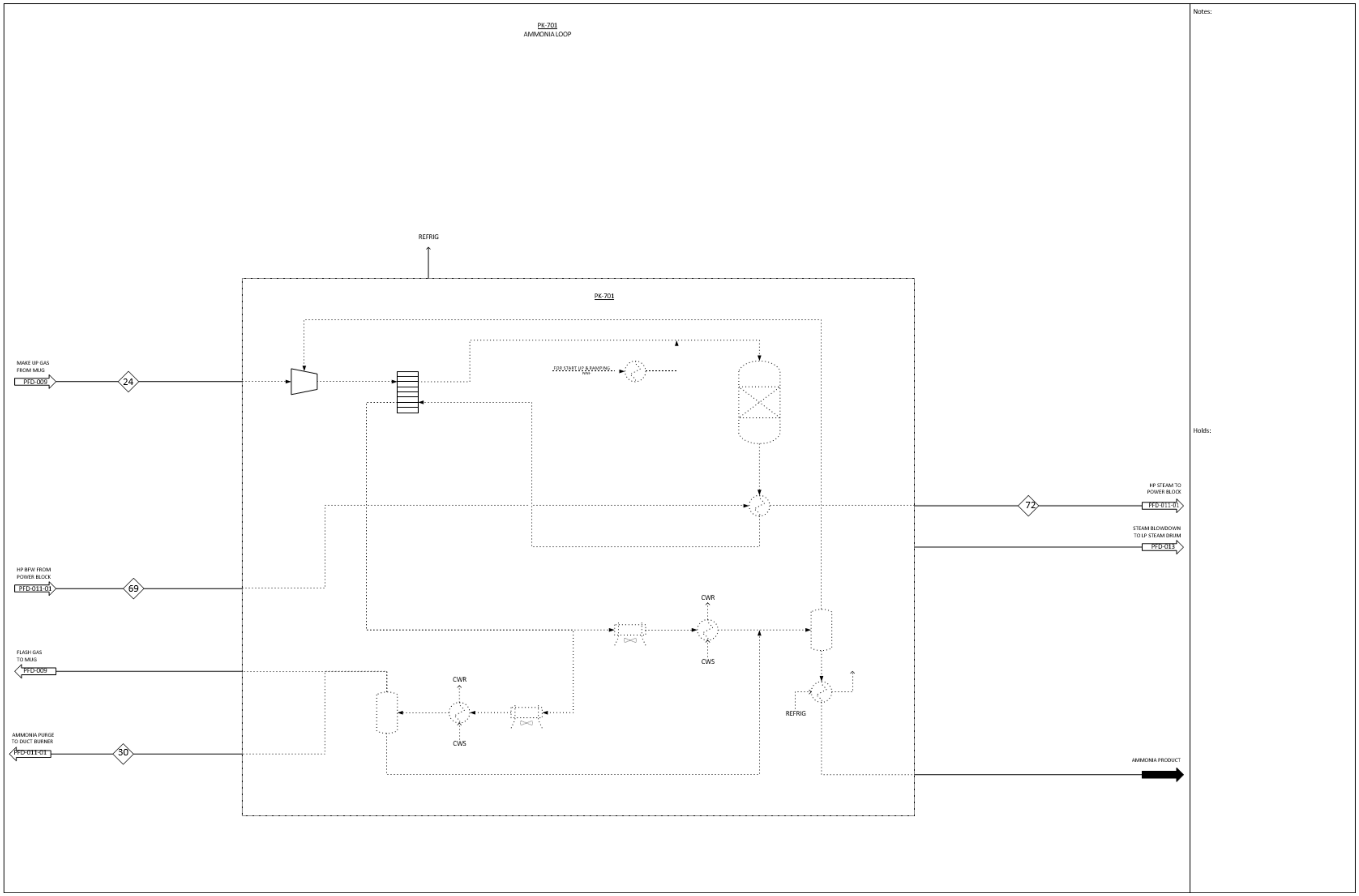


Figure 3-12. Gas Turbine and HRSG Process Flow Diagram

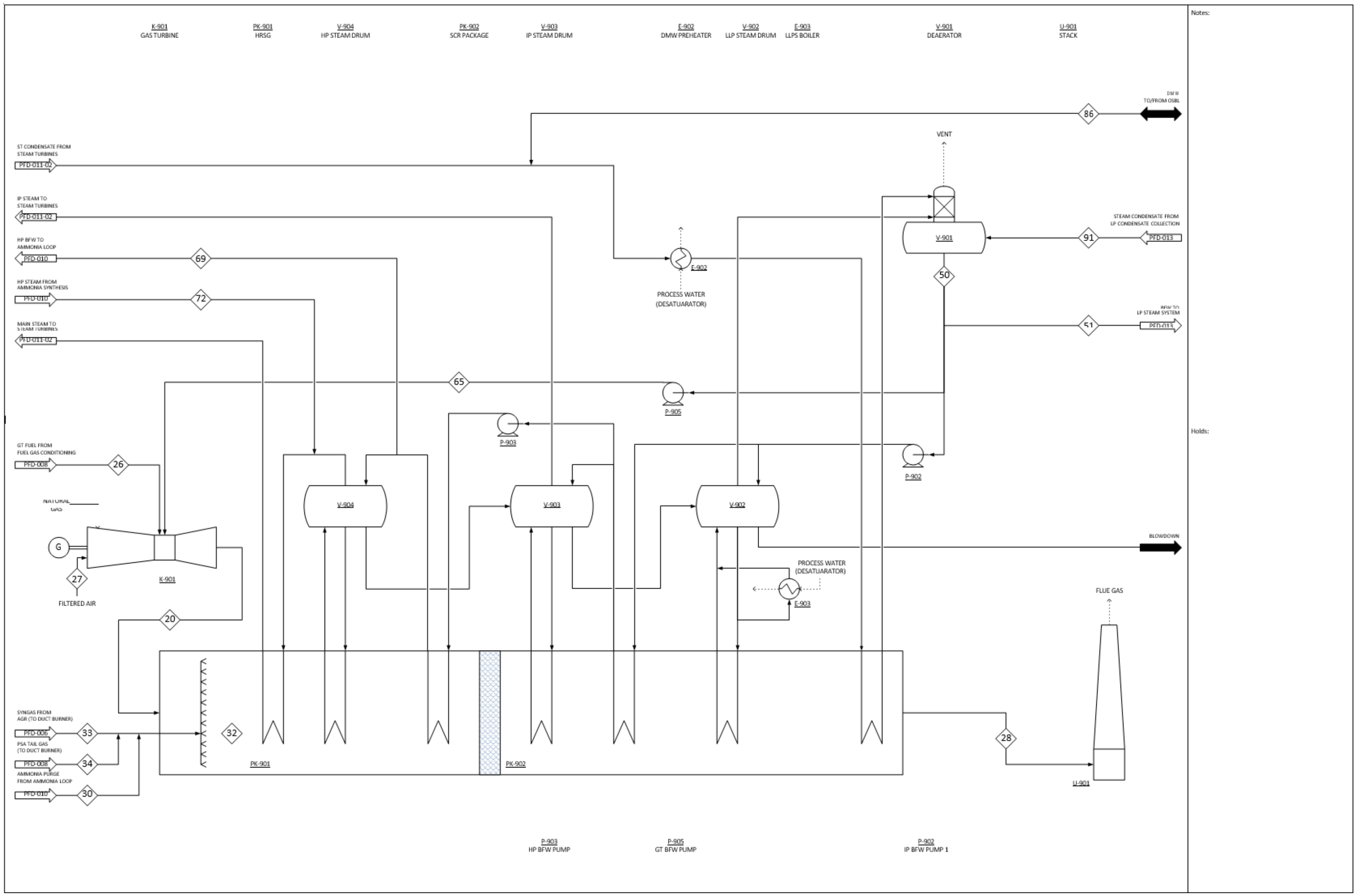


Figure 3-13. Steam Turbine Process Flow Diagram

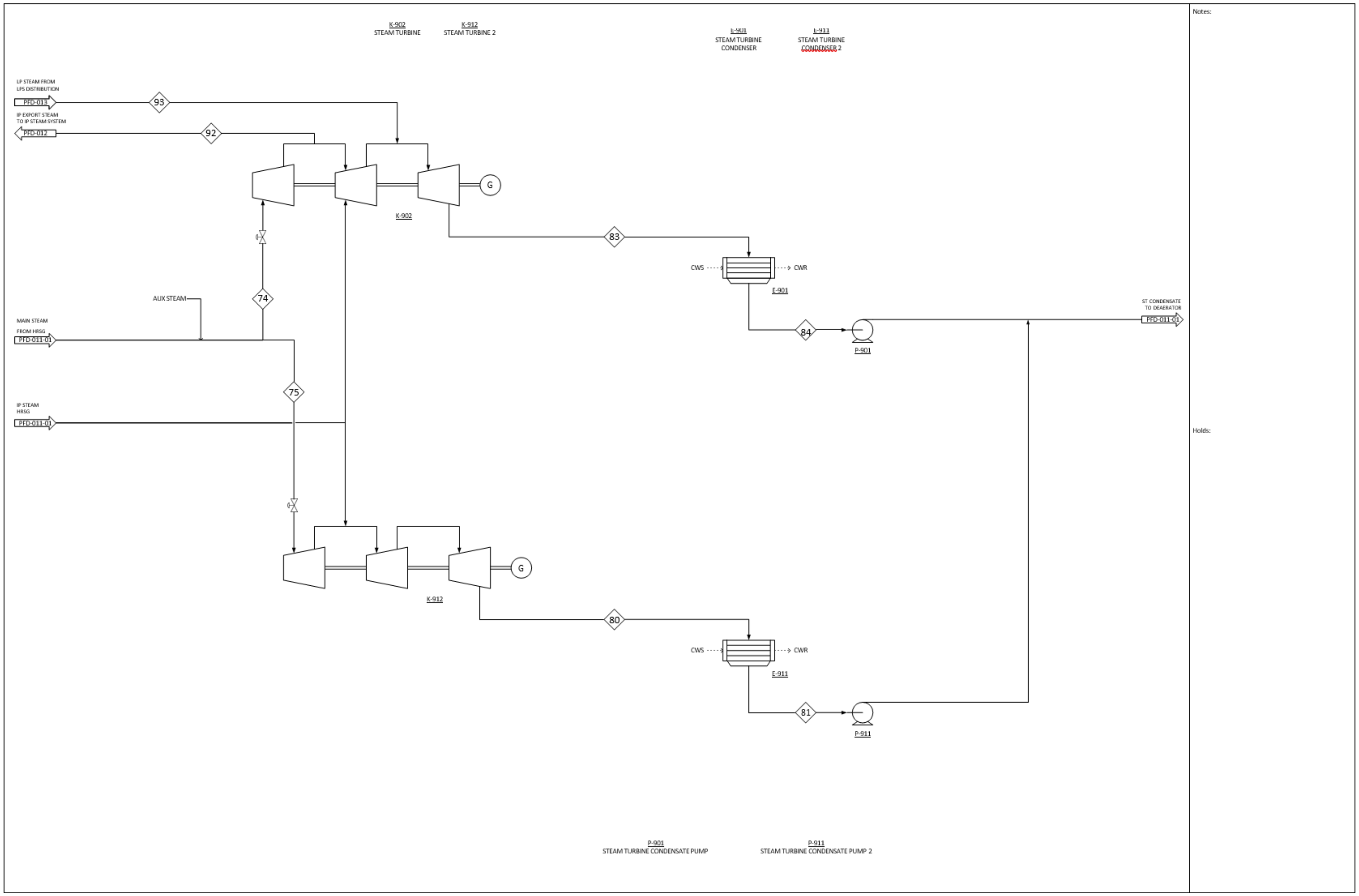


Figure 3-14. IP Steam System Process Flow Diagram

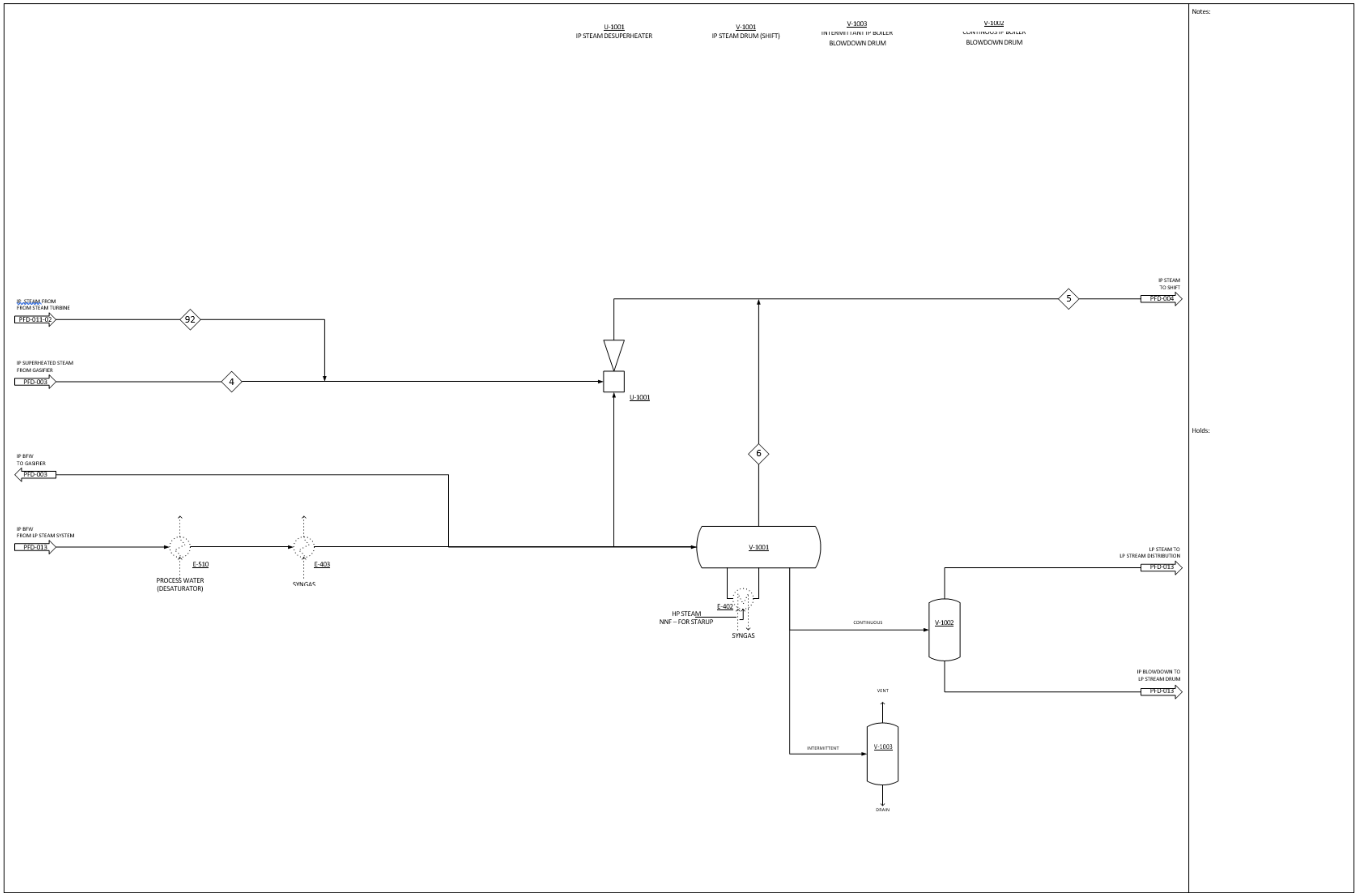
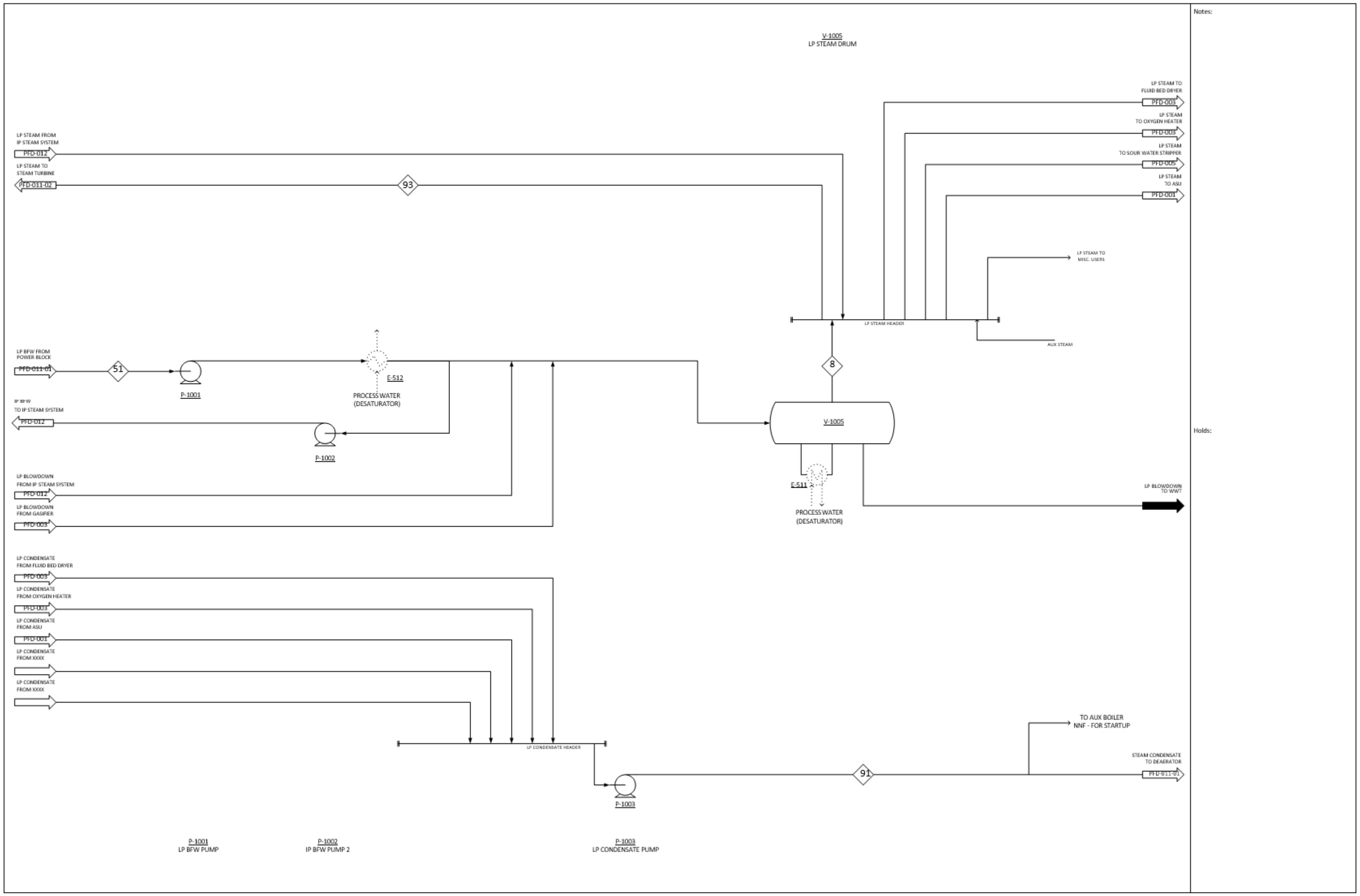


Figure 3-15. LP Steam System Process Flow Diagram



3.3 Major Equipment List

Major equipment items for the polygeneration concept can be found in Appendix C.

3.4 Ability of the Proposed Plant to Meet Coal First Design Criteria

3.4.1 *High Overall Plant Efficiency*

Initiative Objective: *High overall plant efficiency (40%+ HHV or higher at full load, with minimal reductions in efficiency over the required generation range).*

Status: *Preliminarily met - System will have minimal reductions over the operating range and plant can achieve overall HHV efficiency of 40% for non-capture cases.*

The current estimate of net plant efficiency at the *Balanced Production* operating modes is ~38%[26] while achieving 90% carbon capture.

Determining a reasonable HHV efficiency in a non-capture case is difficult as a large number of the design decisions directly support pre-combustion carbon capture (e.g., gasification, characteristics of the water-gas shift, etc.). Because of this, truly optimizing the polygeneration design for a non-capture case would result in a new plant design that it largely dissimilar in operational characteristics to the point that a comparison between the two would be largely meaningless.

However, in the interest of reporting a non-capture case HHV efficiency, one option would be to simply remove the CO_2 compressors and simply vent the CO_2 to atmosphere after it has already been captured. While this is clearly an illogical and inefficient approach to the operation of the polygeneration plant, elimination of the CO_2 compressors would result in a 2.0% gain to HHV efficiency[27] in the *Balanced Generation* cases. This 2% gain in HHV efficiency, combined with the existing HHV efficiencies of 38.3% and 38.8% in the *Balanced Generation, 3 GT* and *Balanced Generation, 2 GT* operating modes, respectively, results in HHV efficiencies in non-capture cases that exceed the 40% target.

The current efficiency is maximized through the combination of electrical generation and chemical storage of energy via ammonia. This is a key component providing a wider band of efficient operation, allowing for greater overall time averaged energy conversion performance than can be achieved by a design focused solely on optimization of "point-in-space" operation.

The 3x2 combined-cycle configuration also supports the goal of efficient operation across a broad range of operating conditions, allowing for improved average efficiencies while effectively following constantly changing load demands. In some respects, the multiple, fast-ramping turbines can be seen as analogous to different gears in an automotive transmission. Essentially, the operator has the choice to meet a given load demand (i.e. a combination of internal, parasitic loads and external grid demand for net export power) by operating fewer turbines at higher individual loads

[26] 38.3% HHV efficiency at the *Balanced Generation, 3 GT* operating point and 38.8% efficiency at the *Balanced Generation, 2 GT* operating point.

[27] CO_2 compressors require 10.7 MW of power relative to 534 MW from the feedstock, equating to 2.00% of overall HHV efficiency.

or operating more turbines at lower individual loads. Much like an automotive transmission selects a given gear to optimize for better fuel efficiency or better transient response, this allows the operator to select the combination (i.e., number of turbines engaged and at what load) to optimize for either efficiency or increased transient response.

By combining multiple systems whose design choices are guided by the desire to establish broader, flatter efficiency curves (e.g., syngas production, syngas combustion turbine for electrical generation, synthesis gas to fuel conversion, and fuel combustion turbine), an overall system with a broadly efficient operating window that is robust to both operational upsets and widely varying load requirements was developed.

The system currently leverages significant heat integration between unit operations to maximize the advantages offered by the various exothermic and endothermic chemical processes as well as the residual heat from the combustion turbine outlet.[28] While the current design basis does not rely on significant technological advances in the near term to improve component system efficiency, later generations of this technology platform should have process intensification options (particularly ammonia synthesis) that will serve to increase overall efficiency.

An additional measure of plant performance and efficiency is the net heat rate. At the *Balanced Generation, 3 GTs* operating point, the polygeneration plant exhibits a net heat rate of 9,384 kJ/kWh (8,895 Btu/kWh).[29][30]

3.4.2 *System Modularity*

Initiative Objective: *Modular (unit sizes of approximately 50 to 350 MW), maximizing the benefits of high-quality, low-cost shop fabrication to minimize field construction costs and project cycle time*

Status: *Met - system capacity chosen such that significant modular construction is anticipated while providing up to ~113 MW of net energy production.*

The designed system is a smaller generation asset capable of serving the spatially diverse requirements for ancillary services (which do not 'travel well' across the grid) and to function competently as a component of a larger distributed system. Due to the modest scale generation systems considered in this concept, the systems may be designed to allow for shop fabrication and use of more standardized components, providing advantages in terms of capital costs, maintenance cost and response, as well as lowered construction times to facilitate limited asset redeployment (i.e. 'semi-mobile'). Specifically, the modularity of the design is based on the selection of component systems and sizes so that all major equipment can be shop fabricated and shipped to the plant site as part of a cohesive unit, ready for integration into the overall plant. Each unit was

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[28] Please refer to Appendices E and F for additional details.

[29] Net heat rates for other defined operating points are as follows:

- *Balanced Generation, 2 GT's:* 9,294 kJ/kWh (8,809 Btu/kWh)
- *Zero Net Power:* 8,211 kJ/kWh (7,782 Btu/kWh)
- *High Electricity:* 10,629 kJ/kWh (10,074 Btu/kWh)
- *Max Electricity:* 15,030 kJ/kWh (14,245 Btu/kWh)

[30] The net heat rate for this plant is calculated as the total input energy of the input coal feed (either in kJ or Btu) relative to the combined kWh of net power for export and the energy chemically stored as NH_3. It should be noted that it is inherently difficult to make a direct and equivalent comparison between the application of this efficiency metric to a polygeneration plant and the application of this metric to a traditional IGCC plant that is only producing electricity.

sized based on the ability to be fabricated off-site and transported to a specific plant site on standard rail and roadway transportation. Additionally, the design including two gasifiers, multiple turbines and two ammonia loops helps enable both the shop fabricability as well as transportation aspects as the capacities and thus sizes of each individual unit are less than had a single unit been chosen. All pressure vessels and pressurized equipment can be transported to site from a remote workshop and many systems are small enough to be modularized as packages complete with piping and instrumentation, FAT complete. The syngas storage sections can be modularized and assembled and tested on site.

The gasifier follows an SES U-Gas design with dimensions limited by the ability to shop fabricate and transport over-land to the site to ensure that modularity is maintained.

Ammonia was chosen as a chemical storage medium as its current state of the art is able to be more efficiently scaled down than methanol synthesis. Additionally, active process intensification research targeting ammonia provides a path for an even more modular system in subsequent generations

3.4.3 *Carbon Capture and Low Emissions*

Initiative Objective: *Near-zero emissions, with options to consider plant designs that inherently emit no or low amounts of carbon dioxide (amounts that are equal to or lower than natural gas technologies) or could be retrofitted with carbon capture without significant plant modifications).*

Status: *Met – The current design achieves 90% carbon capture for multiple modeled operating points*

Team AST's approach makes the ability to implement pre-combustion capture inherent in the polygeneration design through the use of gasification and a water-gas shift reactor. The design leverages an established solvent-based acid gas removal/carbon capture system (i.e. Selexol) as it was determined to have simpler logistics compared to the significant amount of solid material required for a sorbent or Skyonic-like system. Currently, the system adopts and achieves a 90% pre-combustion carbon capture target.

Ammonia, as the chemical storage component, has potential for power generation with limited emissions impact. Specifically, ammonia-based power options have been an area of highly active R&D activities (e.g., fuel cell, internal combustion engines, turbines, and microthrusters) for extracting energy stored in the chemical bonds of ammonia with minimal environmental impact. The proposed approach enables the potential for the specified coal-based generation system to take advantage of complimentary innovations in this space. The current estimate of CO_2 emission is ~20 lb/MMBtu of coal processed in the system for the *Balanced Generation* cases.

3.4.4 *High Ramp Rate Characteristics*

Initiative Objective: *The overall plant must be capable of high ramp rates and achieve minimum loads commensurate with estimates of renewable market penetration by 2050.*

Status: *Met – Projected ramping and turndown characteristics are commensurate with high penetration of renewables.*

The current design combines several systems that provide operational flexibility in order to generate a wide window of operations at reasonable efficiency to facilitate the ability of the plant to absorb grid disturbances and complex market dynamics. Specifically, the syngas production will couple to storage capacity, allowing for adjusted final disposition between the power generation and ammonia production (chemical storage/fuel) options, resulting in the ability to vary the power output without requiring that the entire plant be operated at partial load, effectively reducing the need for the entire plant to operate in a significantly curtailed "turndown" mode in response to a lack of grid demand for export energy. In fact, the "net-zero power" scenario only requires a turndown of the gasifier to ~70% of max load.

The synthesis gas power production will be accomplished by a combined cycle turbine. While a simple cycle turbine generally has a flatter efficiency curve, turndown capabilities, and better response characteristics relative to a combined cycle deployment, it is believed that this specific proposed deployment will mitigate most of the drawbacks related to combined cycle operations through the use of a 3x2 configuration. Specifically, the LM2500+ turbines have an advertised cold start time of ~30 minutes in combined cycle operation with a ramp rate of 30 MW/min in a 1x1 combined cycle configuration.[31]

Additional, surge capacity for electricity production can be achieved through combustion of the syngas in the syngas storage tanks or through the use of natural gas. This can be accomplished either through blending of ammonia in to the feed of the combustion turbine (as needed, on a limited basis) to allow other parts of the system to adjust to demand-load and system upsets or, in specific cases, through deployment of an additional, dedicated ammonia-based power system. The use of ammonia for electrical power generation at small-scale is an active area of research which hopefully can be leveraged in later technology generations.

3.4.5 *Integration of Coal-Based Electricity Generation with Storage*

Initiative Objective: *Integration with thermal or other energy storage to ease intermittency inefficiencies and equipment damage.*

Status: *Met - inherent in the polygeneration approach.*

Polygeneration (co-production with ammonia) was selected so that readily accessible, chemical storage of the energy from coal is inherent in Team AST's design. This choice allows the system to ramp up and down in response to the varying load demands and intermittent power supplied to the grid system without placing unneeded mechanical and/or metallurgical stress on system equipment. The chemical storage options considered in the proposed approach can handle transients in the system.[32] Additionally, the selected option for chemical storage (i.e. ammonia) has multiple disposition options (e.g., combustion for power, readily transported fuel, combined heat and power, vehicle fuel, and/or localized fertilizer production). These multiple dispositions

[31] The advertised 30 MW/min ramp rate is based on a standard 1x1 combined cycle configuration with an advertised net output of 43.0 MW, resulting in a ramp rate of 69.8% per minute *in the advertised configuration.* It is important to note that the polygeneration design employs a different configuration (i.e. a 3x2 combined cycle). However, the ramp rate in the advertised configuration exceeds the minimum program standard ramp rate by such a large amount (i.e., advertised ramp rate of ~70% per minute compared to the required ramp rate of 4% per minute) that it is a virtual certainty that the polygeneration plant will be able to meet the Coal FIRST requirements with respect to ramp rate.

[32] Please refer to *Section 2.4* for detailed discussion of various transient cases.

allow specific project implementations to leverage various potential value streams to facilitate a greater range of economically viable implementations and/or meet mission requirements (e.g., DoD energy and mission resilience options) if the system is deployed in a microgrid or related approach.

The chemical storage medium of ammonia was selected due to it being better aligned with the performance targets of the Coal FIRST initiative. Specifically, overall systems efficiency is enhanced relative to a methanol system due to the higher separation energy (two distillation columns required for a methanol generation system compared to the refrigeration-based system of an ammonia loop) and lower quality heat recovery from a methanol-based system. Current synthesis process technology is known to scale down better for ammonia than methanol. Additionally, developments in the area of renewable energy-derived ammonia are driving process intensification innovations in ammonia synthesis that later generations of this technology platform may leverage. This also indicates that ammonia production is more complimentary to reduced design, construction, and commissioning efforts. Carbon is rejected at a point source in ammonia production allowing more efficient life-cycle carbon dioxide capture (compared to distributed carbon dioxide emissions after methanol end use). Methanol production requires more water than ammonia synthesis. Additionally, ammonia transport costs act as a protective buffer to potential disruptions caused by cheap natural gas-derived mega-plants (cf. methanol), making the ammonia market inherently distributed which is complimentary to a distributed power system.

3.4.6 *Minimized Water Usage*

Initiative Objective: *Minimized water consumption.*

Status: *Met - Significant, sensible water recycle to reduce water consumption*

The design incorporates several water minimizations techniques. These include:

- Recycle of process condensate within the plant
- Reuse of process condensate as CT make-up
- Use of process condensate for process heating duties
- Increase gasifier scrubber temperature

Process condensate is recycled within the plant for use as make-up to the gasifier scrubber, the SRU quench, the AGR and the desaturator reducing fresh water make-up by 46 t/h.

Stripped process condensate is used as CT make-up saving 107 t/h of raw water makeup to the cooling tower. In addition, it is anticipated that this stripped process condensate has a lower TDS and TSS than the fresh water make-up to the cooling tower thus allowing the tower to be operated at higher cycles of concentration than otherwise. This is to be further refined at a later stage of the project once the disposition of the process condensate and the raw water make-up is known.

Hot process condensate is used for heating duties including reboil duty the AGR, GT feed gas preheating and deaerator steam production. These duties would otherwise be done using steam with the attendant consumption of fresh water to make up for system loses.

Process condensate direct from the desaturator bottoms is used for make-up of the gasifier scrubber. Using this hot water increases the temperature of the syngas exiting the scrubber and the water content, thus decreasing the live steam input required for the water gas shift reaction.

Additionally, ammonia was chosen as the chemical energy storage medium partially based on the reduced water and steam requirements relative to methanol synthesis and product recovery.

3.4.7 *Reduced Design, Construction, and Commission Schedules*

Initiative Objective: *Reduced design, construction, and commissioning schedules from conventional norms by leveraging techniques including but not limited to advanced process engineering and parametric design methods.*

Status: *Met - Execution plan provides for completion of plant within CoalFIRST objectives*

The polygeneration design, especially in the selection of components with a high existing Technology Readiness Level (TRL), was selected so that one could rationally select unit operation scales that allow for standardization and parametric design. Additionally, the intention is to leverage advances in process intensification such as those being driven by the American Institute of Chemical Engineers RAPID Manufacturing Institute. Subsequent elements of the pre-FEED study will include a sourcing and manufacturability analysis aimed at establishing the most standardized version of the concept so that it can be replicated with minimum re-engineering and re-specification of equipment. The intent is to have a system that is deployable on timescales similar to those seen by deployment of natural gas combined cycle generation assets rather than the lengthy timelines of baseload coal or nuclear power plants. The proposed Execution Plan provides for the development of a pilot plant and a first-generation plant; in this instance the pilot plant could be complete prior to 2030. Similarly, should a developer choose to begin development with a pioneer plant (bypassing the pilot plant stage), this could also be complete prior to 2030, however would come with somewhat higher risk and thus we would expect the financing terms for this path to be less attractive. Additionally, the execution plan as presented has been developed based on the pilot plant and first-generation plant; it is expected that the design, unit fabrication and construction times for subsequent plants will each benefit from previous experiences and the benefits of modular construction, thus further reducing the development time of subsequent plants.

3.4.8 *Improved Maintainability*

Initiative Objective: *Enhanced maintenance features including technology advances with monitoring and diagnostics to reduce maintenance and minimize forced outages*

Status: *Preliminarily met -*

The approach is designed to respond to curtailed (or even fully reduced) demand for electrical generation capability while remaining on ‘warm stand-by.’ Specifically, the design leverages the intelligent incorporation of storage (synthesis gas and ammonia) capacity in the system. The storage capacity provides the capability to run for a limited time off stored synthesis gas in the event of gasifier curtailment or store produced synthesis gas for future use if the combustion turbine or the ammonia (chemical storage) production train(s) are curtailed. Note that ammonia can be used to augment reduced synthesis gas availability when required to perform both scheduled or unplanned maintenance.

Additionally, multiple trains have been employed, when practical (e.g., gasifier, turbines, ammonia loop, etc.). This allows the ability to respond quickly, minimizes wear and tear on equipment, maximizes utilization of deployed capital, and allows for maintenance on various trains within the system while continuing to provide value. Accomplishing this requires advanced controls and edge computing-enabled asset optimization (such as that deployed in microgrids).

Finally, as the plant overwhelmingly on existing, known, and well-established unit operations, it will benefit from the commensurate wealth of experience and knowledge in the area of maintenance beyond what would normally be expected with a novel unit operation or piece of capital equipment.

3.4.9 *Integration with Other Plant Value Streams*

Initiative Objective: *Integration with coal upgrading, or other plant value streams (e.g., co-production)*

Status: *Met – Inherent in the polygeneration design*

The polygeneration approach inherently links coal-based electricity generation with other value streams (production of ammonia as a chemical fuel or for other beneficial use). These unit operations create multiple options for effective heat integration and dispositions of intermediate streams produced in various operating sections.

3.4.10 *Potential for Natural Gas Integration*

Initiative Objective*: Capable of natural gas co-firing*

Status: *Met*

Natural gas can be incorporated into this approach in a variety of ways to increase reliability, resiliency, and reduce the risks associated with the gasification process. Specifically, the combustion turbines are capable of natural gas co-firing to assist in ramping during transitions between operating modes if sufficient excess syngas is not currently available in the syngas storage tanks.[33] Additionally, natural gas can be fired in the duct burners to increase net power for export during transitions or in periods of high grid demand. Natural gas may also be blended with a portion of the water gas shift reactor effluent directed to the combustion turbine as a means of conditioning the fuel prior to combustion as a control option. Finally, natural gas can also complement the heat requirements of the system as needed

[33] As discussed in *Section 2.4.1, i*t is estimated that a maximum of 80 MMscfd of natural gas would be required to cover both transitions between operating points, as well as supplying additional power to assist in restarting the gasification plant, including the gasifier, shift unit, and utilities. It should be noted that this represent an intermittent and temporary need in transition as opposed to describing a constant consumption of natural gas required for steady state plant operations.

4. Cost Results

4.1 Cost Estimating Methodology[34]

The cost estimates contained in this document are consistent with approved NETL methodologies, as defined in the 2019 revision of the QGESS document *Cost Estimation Methodology for NETL Assessment of Power Plant Performance.* Multiple individual members of the AST Team are well versed in these approaches through their experience serving as Program Managers overseeing past process and cost engineering work for NETL. Additionally, the applied methodology draws on Worley's past experience serving as the EPC supporting NETL strategic analysis functions.

Worley has applied their experience, combined with both (1) vendor cost estimates for component technologies and (2) scaling and estimation practices considered to be industry standard to develop and certify a Class 4 capital cost estimate as defined by the Association for the Advancement of Cost Engineering International (AACE).

The individual unit operations and operating sections of the defined polygeneration plant are sufficiently mature to eliminate the need to integrate technologies requiring a high level of new research and development (R&D). However, there is some uncertainty associated with the initial, complex integrations of technologies in a commercial application. It is possible that the integration component may result in higher costs, however this cost variation will be within the cost range as expected for a Class 4 cost estimate.

While these cost estimates represent the best abstract estimate at the current level of engineering, actual reported project costs for the polygeneration design are also expected to deviate from the cost estimates in this report due to differences in real-world implementation (e.g., project- and site-specific considerations) that may impact construction costs. The reported cost uncertainty does not capture changes to site characteristics or added infrastructure costs that would be incurred from changing the design basis of the project.

External supporting innovations (e.g., improvements in ammonia synthesis and pre-combustion capture technology, as mentioned in the Pre-FEED study's technology gap analysis) are expected to result in design and operational improvements for future generations of this polygeneration technology platform (beyond the current scope), resulting in lower costs than those estimated here.

4.2 Capital Costs[35]

Figure 4-1, provides an overview of the five capital cost levels included within this report: BEC, EPCC, TPC, and TOC are "overnight" costs, expressed in December 2018 dollars. TASC is expressed in mixed, current-year dollars over the assumed five-year capital expenditure period for

[34] This section is largely repeated verbatim from the 2019 version of NETL's *Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity."* Changes have been made, were appropriate, to ensure relevance and accuracy with the polygeneration design.

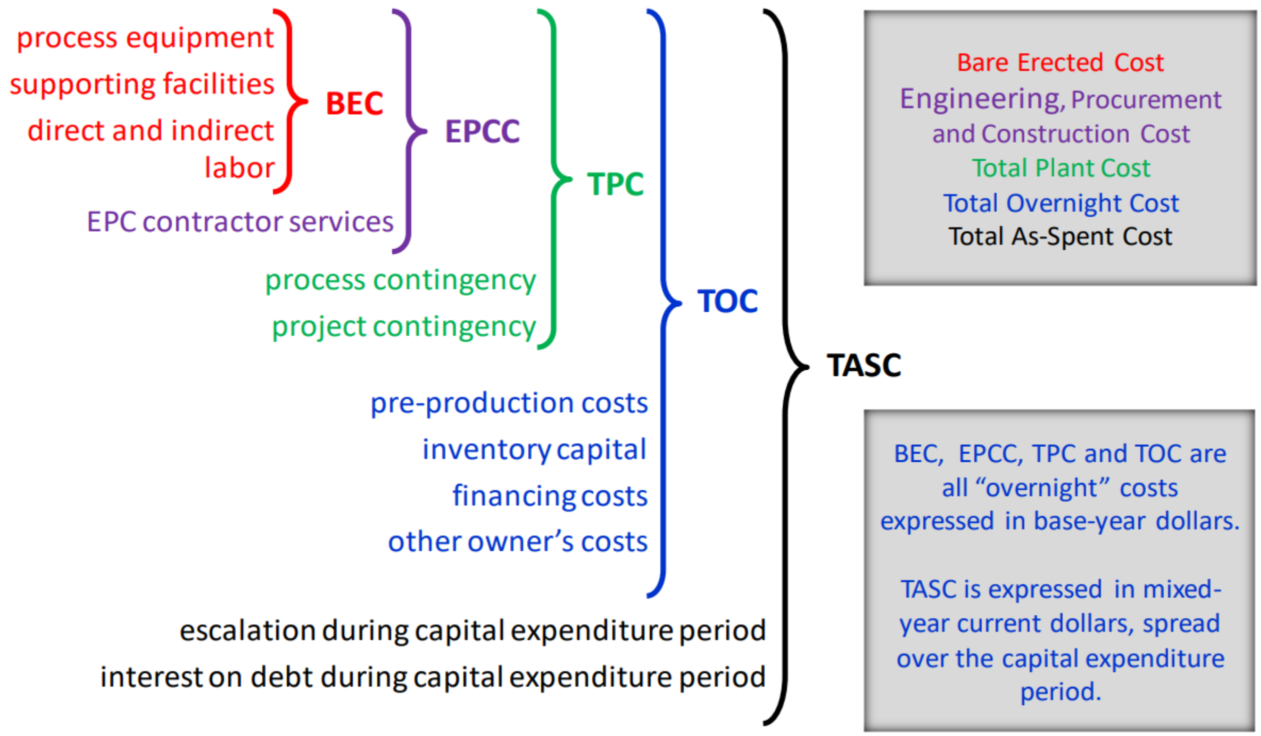
[35] The cost level definitions and graphic appearing in Figure 4-1 are a reproduction of those found in *Section 2.1: Level of Capital Costs* in the June 2019 release of NETL's *Quality Guideline for Energy System Studies: Cost Estimation Methodology for NETL Assessment of Power Plant Performance* (National Energy Technology Laboratory, "Quality Guidelines for Energy System Studies: Cost Estimation Methodology for NETL Assessment of Power Plant Performance," U.S. Department of Energy, Pittsburgh, PA, 2019.)

the polygeneration design. The following definitions have been adopted in accordance with the definitions found in the 2019 version of NETL's *Quality Guideline for Energy System Studies: Cost Estimation Methodology for NETL Assessment of Power Plant Performance:*

> <u>Bare Erected Cost</u> (BEC) comprises the cost of process equipment, on-site facilities and infrastructure that support the plant (e.g., shops, offices, labs, road), and the direct and indirect labor required for its construction and/or installation. Equipment cost estimates are frequently developed for each plant or plant component using in-house database and conceptual estimating models for specific technologies and may differ from values generated by other software packages such as Aspentech's Aspen Economic Analyzer.
>
> <u>Engineering, Procurement, and Construction Cost</u> (EPCC) comprises the BEC plus the cost of services provided by the EPC contractor. The EPC services include detailed design, contractor permitting (i.e., those permits that individual contractors must obtain to perform their scopes of work, as opposed to project permitting, which is not included here), and project/construction management costs.
>
> <u>Total Plant Cost</u> (TPC) comprises the EPCC cost plus project and process contingencies.
>
> <u>Total Overnight Capital</u> (TOC) comprises the TPC plus all other "overnight" costs, including owner's costs. TOC is an overnight cost, expressed in base-year dollars and as such does not include escalation during construction or construction financing costs.

<u>Total As-Spent Capital</u> (TASC) comprises the sum of all capital expenditures as they are incurred during the capital expenditure period for construction including their escalation. TASC also includes interest during construction, comprised of interest on debt and a return on equity (ROE). TASC is expressed in mixed, current-year dollars over the capital expenditure period.

Figure 4-1. Capital Cost Levels and Their Elements[36]



4.2.1 *Cost Estimate Basis and Classification*

Worley used a combination of: (1) in-house database and estimating models, (2) commercial software packages, and (3) scaling based on applying QGESS methodologies to existing NETL reports to develop TPC and operation and maintenance (O&M) costs for the relevant operating modes. Additional discussion and details can be found in *Section 4.4*.

4.2.2 *System Code-of-Accounts*

As with NETL's *Baseline* reports[37], a process/system-oriented code of accounts is used to group relevant costs into logical subaccounts. This approach ensures that all components of a given process or unit operation are logically grouped together.

4.2.3 *Estimate Scope*

The estimates represent the polygeneration plant deployed on a generic site located in the Midwest. The limit of the plant includes the total facility including feedstock receiving and water supply system, ending at the high voltage side of the main power transformers.

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[36] This graphic is a reproduction of one found in existing NETL literature (National Energy Technology Laboratory, "Quality Guidelines for Energy System Studies: Specification for Selected Feedstocks," U.S. Department of Energy, Pittsburgh, PA, 2019) in accordance with fair-use standards.

[37] National Energy Technology Laboratory, "Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity," U.S. Department of Energy, Pittsburgh, PA, 2019.

CO_2 transport and storage (T&S) costs are not considered in the reported capital or O&M costs.

4.2.4 *Capital Cost Assumptions*[38]

Worley developed the capital cost estimates for the polygeneration plant using the company's in-house database, commercial software packages, and relevant QGESS scaling methodologies. This database and approach are maintained by Worley as part of a commercial design base of experience for similar equipment in the company's range of power and chemical process projects. A reference bottom-up estimate for each major component provides the basis for the estimating models.

Other key estimate considerations include the following:

- Labor costs are based on Midwest, Merit Shop. The estimating models are based on a U.S. Gulf Coast location and the labor cost has been factored to a Midwest location. Labor cost data were sourced from recent projects and Worley in-house references/cost databases.

- The estimates are based on a competitive bidding environment, with adequate skilled craft labor available locally.

- Labor is based on a 50-hour work-week (5-10s). No additional incentives such as per-diem allowances or bonuses have been included to attract craft labor.

- While not included at this time, labor incentives may ultimately be required to attract and retain skilled labor depending on the amount of competing work in the region, and the availability of skilled craft in the area at the time the projects proceed to construction.

- The estimates are based on a greenfield site.

- The site is considered to be Seismic Zone 1, relatively level, and free from hazardous materials, archeological artifacts, or excessive rock. Soil conditions are considered adequate for spread footing foundations. The soil bearing capability is assumed adequate such that piling is not needed to support the foundation loads.

- Engineering and Construction Management are estimated based on Worley's historical experience in designing and building power and chemical process projects. The cost of 15% of BEC is representative of Worley's historical engineering/construction management costs for similar plant types. These costs consist of all home office engineering and procurement services as well as field construction management costs. Site staffing generally includes construction manager, resident engineer, scheduler, and personnel for project controls, document control, materials management, site safety, and field inspection.

[38] These are standard assumptions used by Worley for capital cost assumptions. As such they match the assumptions which appear in previous NETL documents on which they have worked, including the previous NETL *Baseline* reports. The text in this section closely mirrors what can be found in Revision 2b of Volume 1b (*National Energy Technology Laboratory, "Cost and Performance Baseline for Fossil Energy Plants Volume 1b: Bituminous Coal (IGCC) to Electricity, Revision 2b – Year Dollar Update" U.S. Department of Energy, Pittsburgh, PA, 2015.*). The only notable exception is the update in the engineering/construction management costs from 8-10% to 15% to reflect prevailing Baseline assumptions.

4.2.5 *Price Fluctuations*

All historic vendor and reference quotes have been updated and adjusted to December 2018 dollars to account for any relevant price fluctuations to equipment and/or materials. Relevant price indices were used as needed for these adjustments.

4.2.6 *Process Contingency*

Notable process contingencies were applied as follows:

- Gasifiers and Syngas Coolers: 15%
- Two-Stage Selexol: 20%
- Mercury Removal: 5%
- CTG: 10%
- Instrumentation and Controls: 5%

4.2.7 *Owner's Costs*

There are three main categories for owner's costs: pre-production costs, inventory capital, and other costs. Pre-production costs are intended to move a given plant through significant completion toward commercial operation.

4.3 Operation and Maintenance Costs[39]

Operating costs and related maintenance expenses (O&M) relate to charges associated with operating and maintaining the plant throughout its expected life, including:

- Operating labor
- Maintenance – material and labor
- Administrative and support labor
- Consumables
- Feedstock
- Waste disposal

[39] 39These are standard assumptions used by Worley for Operation and Maintenance Costs. As such, they match the assumptions which appear in previous NETL documents on which they have worked, including the previous NETL *Baseline* reports. The text in this section very closely mirrors what can be found in Revision 2b of Volume 1b (*National Energy Technology Laboratory, "Cost and Performance Baseline for Fossil Energy Plants Volume 1b: Bituminous Coal (IGCC) to Electricity, Revision 2b – Year Dollar Update" U.S. Department of Energy, Pittsburgh, PA, 2015.*). Notable exceptions include a change in the Operating Labor rate from $39.70/hour to $38.50/hour in *Section 4.3.1* and explicit definition of the waste disposal rates in *Section 4.3.5*.

- Co-product or by-product credit (that is, a negative cost for any by-products sold)

O&M costs can be divided into both "fixed" and "variable" costs.

4.3.1 *Operating Labor*

Operating labor cost was determined based on the number of operators required for the polygeneration plant with an average base labor rate used to determine annual cost is $38.50/hour and an associated labor burden of 30% relative to the base labor rate.

4.3.2 *Maintenance Material and Labor*

Maintenance cost is based on the maintenance costs in relation to the initial capital costs. Due to the aggressive cycling and ramping that this plant is expected to be subjected to, an *additional* 10% maintenance adder has been applied to account for protentional extra wear on the equipment.

4.3.3 *Administrative and Support Labor*

Labor administration and overhead charges are assessed at a rate of 25% of the burdened O&M labor.

4.3.4 *Consumables*

Consumable costs, including plant feedstock, were determined on the using relevant consumption rates, unit costs, and plant capacity factors. Required consumable quantities are based on previously developed energy and mass balances for the polygeneration plant.[40] The quantities for initial fills and daily consumables were calculated on a 100 percent operating capacity basis at the *Balanced Generation, 3 GTs* operating point.[41]

4.3.5 *Waste Disposal*

The approach for estimating waste quantities and disposal costs is similar to consumables, with hazardous waste disposal rates of $80.00/ton and non-hazardous waste disposal rates of $38.00/ton.

4.3.6 *Co-Products and By-Products*

No financial credit was taken to offset costs based on the potential salable value or relevant by-products when calculating system costs. However, as the plant is a polygeneration facility, sensitivity to ammonia prices was examined in *Section 4.4*.

[40] Please refer to the *Performance Results* (Section 3) for the relevant energy and mass balances.

[41] The *Balanced Generation, 3 GTs* operating point represents the maximum values for initial fills and consumables of the five defined operating points.

4.4 Cost Estimates

Applying the previously discussed cost methodologies results in the BEC and TPC seen in Table 4-1. Table 4-2 shows the owner's costs, TOC, and TASC. Table 4-3 through Table 4-9 examine the O&M costs for the polygeneration plant.

It should be noted that no costs in Table 4-1 through Table 4-9 are reported on a per kilowatt (or megawatt) basis due to the inherent design and operating characteristics of a polygeneration plant. There is not a clear kilowatt basis for a system that operates across a broad, adaptive window that includes cogeneration of salable products (e.g., ammonia). Furthermore, the metric has no meaning when there are capital components (such as the ammonia loop) that are not related to electricity generation.

Additional discussion of this decision, as well as the inherent problems of applying a per kilowatt (or per megawatt) metric to a polygeneration plant, is presented following Table 4-11.

The cost estimates for the major sub-systems came from three primary sources:

- Worley in-house data from multiple sources, including:
- *Aspen Capital Cost Estimator* based on the relevant sized equipment list
- Scaling based on the 2019 *Baseline* report, which represents detailed bottoms-up estimates of cost accounts done by qualified firms such as Worley Group and Black and Veatch providing site support services to NETL

In some cases, data points from multiple sources were combined to generate the final reported estimate.

Table 4-1. Polygeneration Capital Plant Cost Details

AST Coal First Polygeneration Plant								Estimate Type:		Class 4
								Cost Base:		Dec 2018
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O & Fee	Contingencies		Total Plant Cost
				Direct	Indirect			Process	Project	$/1000
1.1	Coal Receive & Unload	$492	$0	$237	$0	$730	$109	$0	$168	$1,007
1.2	Coal Stackout & Reclaim	$1,609	$0	$384	$0	$1,994	$299	$0	$459	$2,751
1.3	Coal Conveyors & Yd Crush	$15,351	$0	$3,907	$0	$19,258	$2,889	$0	$4,429	$26,575
1.4	Other Coal Handling	$2,391	$0	$538	$0	$2,929	$439	$0	$674	$4,042
1.5	Sorbent Receive & Unload	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.6	Sorbent Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.7	Sorbent Conveyors	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.8	Other Sorbent Handling	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.9	Coal & Sorbent Hnd. Foundations	$0	$43	$113	$0	$156	$23	$0	$36	$215
	SUBTOTAL 1.	**$19,843**	**$43**	**$5,179**	**$0**	**$25,066**	**$3,760**	**$0**	**$5,765**	**$34,591**
2.1a	Coal Crushing	$376	$23	$54	$0	$453	$68	$0	$104	$625
2.1b	Coal Drying	$9,922	$1,984	$3,382	$0	$15,289	$2,293	$535	$3,623	$21,741
2.2	Prepared Coal Storage & Feed	$2,311	$555	$357	$0	$3,224	$484	$0	$741	$4,448
2.3	Dry Coal Injection System	$2,950	$34	$270	$0	$3,254	$488	$0	$748	$4,491
2.4	Misc. Coal Prep & Feed	$228	$167	$491	$0	$886	$133	$0	$204	$1,223
2.4a	Dryer Vent Booster Compressor & Accessories	$5,511	$473	$1,066	$0	$7,050	$1,057	$0	$1,621	$9,729
2.5	Sorbent Prep Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.6	Sorbent Storage & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.7	Sorbent Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0

AST Coal First Polygeneration Plant								Estimate Type:		Class 4
								Cost Base:		Dec 2018
Item No.	Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Bare Erected Cost	Eng'g CM H.O & Fee	Contingencies Process	Contingencies Project	Total Plant Cost $/1000
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.9	Coal & Sorbent Feed Foundation	$0	$555	$477	$0	$1,032	$155	$0	$237	$1,424
	SUBTOTAL 2.	**$21,299**	**$3,792**	**$6,097**	**$0**	**$31,188**	**$4,678**	**$535**	**$7,280**	**$43,681**
3.1	Feedwater System	$1,994	$3,541	$2,071	$0	$7,606	$1,141	$0	$1,749	$10,496
3.2	Water Makeup & Pretreating	$320	$33	$195	$0	$548	$82	$0	$189	$819
3.3	Other Feedwater Subsystems	$1,549	$541	$656	$0	$2,746	$412	$0	$632	$3,790
3.4	Service Water Systems	$189	$372	$1,429	$0	$1,989	$298	$0	$686	$2,974
3.5	Other Boiler Plant Systems	$3,271	$1,360	$3,201	$0	$7,832	$1,175	$0	$1,801	$10,808
3.6	FO Supply Sys & Nat Gas	$267	$505	$522	$0	$1,295	$194	$0	$298	$1,787
3.7	Waste Treatment Equipment	$416	$0	$298	$0	$713	$107	$0	$246	$1,067
3.8	Misc. Power Plant Equipment	$910	$121	$547	$0	$1,578	$237	$0	$544	$2,359
	SUBTOTAL 3.	**$8,915**	**$6,473**	**$8,919**	**$0**	**$24,306**	**$3,646**	**$0**	**$6,146**	**$34,098**
4.1	Gasifier, Syngas Cooler & Auxiliaries (U-Gas)	$40,045	$23,625	$33,868	$0	$97,538	$14,631	$14,631	$19,020	$145,819
4.2	Syngas Cooling	w/4.1	w/ 4.1	w/ 4.1	$0	$0	$0	$0	$0	$0
4.3	ASU/Oxidant Compression	$23,731	$15,188	$26,199	$0	$65,117	$9,768	$0	$0	$74,885
4.4	LT Heat Recovery & FG Saturation	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.5	Misc. Gasification Equip.	$173	$321	$664	$0	$1,159	$174	$0	$0	$1,333

AST Coal First Polygeneration Plant								Estimate Type:		Class 4
								Cost Base:		Dec 2018
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O & Fee	Contingencies		Total Plant Cost
				Direct	Indirect			Process	Project	$/1000
4.6	Flare Stack System	$343	$193	$108	$0	$643	$96	$0	$148	$887
4.8	Major Component Rigging	w/ 4.1	w/ 4.1	w/ 4.1	$0	$0	$0	$0	$0	$0
4.9	Gasification Foundations	w/ 4.1	w/ 4.1	w/ 4.1	$0	$0	$0	$0	$0	$0
	SUBTOTAL 4.	**$64,292**	**$39,327**	**$60,838**	**$0**	**$164,457**	**$24,669**	**$14,631**	**$19,168**	**$222,924**
5A.1	Double Stage Selexol	$65,792	$0	$27,586	$0	$93,379	$14,007	$18,676	$25,212	$151,274
5A.2	Elemental Sulfur Plant	$23,075	$4,498	$29,566	$0	$57,139	$8,571	$0	$13,142	$78,852
5A.3	Mercury Removal	$317	$0	$240	$0	$557	$84	$28	$134	$802
5A.4	Shift Reactors	$3,741	$2,859	$3,225	$0	$9,824	$1,474	$0	$0	$11,298
5A.5	Particulate Removal	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0
5A.6	Blowback Gas Systems	$343	$193	$108	$0	$643	$96	$0	$0	$739
5A.7	Fuel Gas Piping	$0	$380	$249	$0	$629	$94	$0	$145	$868
5A.8	Gas Cooling	$10,413	$2,355	$4,481	$0	$17,250	$2,587	$0	$3,967	$23,805
5A.9	Sour Water Stripper	$2,394	$1,745	$3,060	$0	$7,199	$1,080	$0	$1,656	$9,935
5A.10	Sulfur Storage	$2,651	$272	$1,238	$0	$4,161	$624	$0	$957	$5,743
5A.11	Syngas Storage	$0	$5,152	$8,872	$0	$14,023	$2,104	$0	$3,225	$19,352
5A.12	Process Interconnects	$0	$10,000	$24,000	$0	$34,000	$5,100	$0	$7,820	$46,920
5A.13	HGCU Foundations	$0	$214	$144	$0	$358	$54	$0	$124	$536
	SUBTOTAL 5A.	**$108,727**	**$27,668**	**$102,769**	**$0**	**$239,164**	**$35,875**	**$18,704**	**$56,382**	**$350,124**
5B.1	CO2 Removal System	$0	$0	$0	$0	$0	$0	$0	$0	$0
5B.2	CO2 Compression & Drying	$13,822	$1,802	$3,468	$0	$19,092	$2,864	$0	$4,391	$26,347
5B.3	CO2 Transport & Storage	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 5B.	**$13,822**	**$1,802**	**$3,468**	**$0**	**$19,092**	**$2,864**	**$0**	**$4,391**	**$26,347**
5C.1	Ammonia Plant	$71,045	$19,563	$13,157	$0	$103,764	$15,565	$0	$23,866	$143,195

AST Coal First Polygeneration Plant										Estimate Type:	Class 4
										Cost Base:	Dec 2018
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O & Fee	Contingencies		Total Plant Cost
				Direct	Indirect			Process	Project	$/1000
5C.2	Ammonia Storage & Loadout	$7,466	$2,146	$12,576	$0	$22,188	$3,328	$0	$5,103	$30,619
	SUBTOTAL 5C.	**$78,510**	**$21,709**	**$25,733**	**$0**	**$125,952**	**$18,893**	**$0**	**$28,969**	**$173,813**
6.1	Combustion Turbine Generator	$33,945	$0	$1,929	$0	$35,874	$5,381	$3,587	$4,484	$49,327
6.2	Combustion Turbine Auxiliaries	$1,796	$429	$813	$0	$3,038	$456	$0	$0	$3,494
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$601	$760	$0	$1,360	$204	$0	$469	$2,034
	SUBTOTAL 6.	**$35,741**	**$1,029**	**$3,502**	**$0**	**$40,273**	**$6,041**	**$3,587**	**$4,954**	**$54,855**
7.1	HRSG	$14,400	$0	$5,623	$0	$20,023	$3,003	$0	$2,303	$25,329
7.2	Selective Catalytic Reduction (SCR) System	w/7.1	w/7.1	w/7.1	$0	$0	$0	$0	$0	$0
7.3	Ductwork	$0	$1,123	$845	$0	$1,967	$295	$0	$453	$2,715
7.4	Stack	$915	$1,304	$3,889	$0	$6,108	$916	$0	$702	$7,727
7.9	HRSG, Duct & Stack Foundations	$0	$324	$356	$0	$680	$102	$0	$234	$1,016
	SUBTOTAL 7.	**$15,315**	**$2,750**	**$10,712**	**$0**	**$28,778**	**$4,317**	**$0**	**$3,692**	**$36,787**
8.1	Steam TG & Accessories	$18,150	$0	$3,101	$0	$21,251	$3,188	$0	$2,444	$26,883
8.2	Turbine Plant Auxiliaries	$262	$0	$665	$0	$928	$139	$0	$107	$1,173
8.3	Condenser & Auxiliaries	$2,016	$1,048	$1,808	$0	$4,872	$731	$0	$560	$6,164
8.4	Steam Piping	$10,354	$0	$4,642	$0	$14,996	$2,249	$0	$4,311	$21,557

AST Coal First Polygeneration Plant								Estimate Type:		Class 4
								Cost Base:		Dec 2018
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O & Fee	Contingencies		Total Plant Cost
				Direct	Indirect			Process	Project	$/1000
8.9	TG Foundations	$0	$189	$365	$0	$554	$83	$0	$191	$828
	SUBTOTAL 8.	**$30,782**	**$1,237**	**$10,582**	**$0**	**$42,601**	**$6,390**	**$0**	**$7,613**	**$56,605**
9.1	Cooling Towers	$2,090	$0	$810	$0	$2,900	$435	$0	$500	$3,835
9.2	Circulating Water Pumps	$803	$0	$44	$0	$848	$127	$0	$146	$1,121
9.3	Circ. Water System Auxiliaries	$87	$0	$14	$0	$101	$15	$0	$17	$133
9.4	Circ. Water Piping	$0	$3,946	$1,042	$0	$4,988	$748	$0	$1,147	$6,884
9.5	Make-up Water System	$215	$0	$325	$0	$540	$81	$0	$124	$745
9.6	Component Cooling Water Sys	$446	$533	$404	$0	$1,383	$207	$0	$318	$1,909
9.9	Circ. Water System Foundations	$0	$1,505	$2,926	$0	$4,431	$665	$0	$1,529	$6,625
	SUBTOTAL 9.	**$3,641**	**$5,985**	**$5,565**	**$0**	**$15,191**	**$2,279**	**$0**	**$3,782**	**$21,251**
10.1	Slag Dewatering & Cooling	$725	$0	$355	$0	$1,080	$162	$0	$124	$1,367
10.2	Gasifier Ash Depressurization	$1,096	$0	$537	$0	$1,633	$245	$0	$282	$2,160
10.3	Cleanup Ash Depressurization	$492	$0	$241	$0	$733	$110	$0	$126	$969
10.4	High Temperature Ash Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0

AST Coal First Polygeneration Plant							Estimate Type:			Class 4
							Cost Base:			Dec 2018
Item No.	Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Bare Erected Cost	Eng'g CM H.O & Fee	Contingencies Process	Contingencies Project	Total Plant Cost $/1000
10.5	Other Ash Rec. Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.6	Ash Storage Silos	$1,104	$0	$1,193	$0	$2,297	$345	$0	$396	$3,038
10.7	Ash Transport/Feed Equip.	$425	$0	$99	$0	$524	$79	$0	$90	$693
10.8	Misc. Ash Handling Equip.	$61	$75	$22	$0	$158	$24	$0	$27	$209
10.9	Ash/Spent Sorbent Foundation	$0	$431	$573	$0	$1,004	$151	$0	$346	$1,501
	SUBTOTAL 10.	**$3,903**	**$506**	**$3,020**	**$0**	**$7,429**	**$1,114**	**$0**	**$1,393**	**$9,936**
11.1	Generator Equipment	$556	$0	$661	$0	$1,217	$183	$0	$140	$1,539
11.2	Station Service Equipment	$3,359	$0	$364	$0	$3,722	$558	$0	$428	$4,709
11.3	Switchgear & Motor Control	$5,986	$0	$1,358	$0	$7,344	$1,102	$0	$1,267	$9,712
11.4	Conduit & Cable Tray	$0	$3,403	$11,439	$0	$14,842	$2,226	$0	$4,267	$21,336
11.5	Wire & Cable	$0	$5,921	$4,353	$0	$10,274	$1,541	$0	$2,954	$14,769
11.6	Protective Equipment	$0	$878	$3,976	$0	$4,854	$728	$0	$837	$6,419
11.7	Standby Equipment	$146	$0	$177	$0	$323	$48	$0	$56	$427
11.8	Main Power Transformers	$9,374	$0	$85	$0	$9,459	$1,419	$0	$1,632	$12,509
11.9	Electrical Foundations	$0	$94	$279	$0	$373	$56	$0	$129	$558
	SUBTOTAL 11.	**$19,421**	**$10,295**	**$22,692**	**$0**	**$52,408**	**$7,861**	**$0**	**$11,709**	**$71,979**
12.1	IGCC Control Equipment	$0	$0	$395	$0	$395	$59	$20	$71	$545
12.2	Combustion Turbine Control	$0	$0	$0	$0	$0	$0	$0	$0	$0
12.3	Steam Turbine Control	$0	$0	$0	$0	$0	$0	$0	$0	$0
12.4	Other Major Component Control	$1,399	$0	$1,163	$0	$2,562	$384	$128	$461	$3,535

AST Coal First Polygeneration Plant									Estimate Type:	Class 4
									Cost Base:	Dec 2018
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O & Fee	Contingencies		Total Plant Cost
				Direct	Indirect			Process	Project	$/1000
12.5	Signal Processing Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0
12.6	Control Boards, Panels & Racks	$322	$0	$257	$0	$578	$87	$29	$139	$833
12.7	Computer & Accessories	$7,462	$0	$297	$0	$7,760	$1,164	$388	$931	$10,243
12.8	Instrument Wiring & Tubing	$0	$2,900	$6,634	$0	$9,535	$1,430	$477	$2,860	$14,302
12.9	Other I & C Equipment	$4,988	$0	$3,015	$0	$8,004	$1,201	$400	$1,441	$11,045
	SUBTOTAL 12.	**$14,171**	**$2,900**	**$11,762**	**$0**	**$28,834**	**$4,325**	**$1,442**	**$5,903**	**$40,504**
13.1	Site Preparation	$0	$141	$3,630	$0	$3,771	$566	$0	$1,301	$5,638
13.2	Site Improvements	$0	$2,513	$4,014	$0	$6,527	$979	$0	$2,252	$9,759
13.3	Site Facilities	$4,504	$0	$5,712	$0	$10,216	$1,532	$0	$3,524	$15,273
	SUBTOTAL 13.	**$4,504**	**$2,655**	**$13,356**	**$0**	**$20,514**	**$3,077**	**$0**	**$7,077**	**$30,669**
14.1	Combustion Turbine Area	$0	$202	$123	$0	$326	$49	$0	$75	$449
14.2	Steam Turbine Building	$0	$1,678	$2,579	$0	$4,257	$639	$0	$734	$5,630
14.3	Administration Building	$0	$1,190	$931	$0	$2,120	$318	$0	$366	$2,804
14.4	Circulation Water Pumphouse	$0	$149	$85	$0	$234	$35	$0	$40	$309
14.5	Water Treatment Buildings	$0	$271	$285	$0	$556	$83	$0	$96	$735
14.6	Machine Shop	$0	$629	$464	$0	$1,093	$164	$0	$189	$1,446
14.7	Warehouse	$0	$1,016	$707	$0	$1,723	$258	$0	$297	$2,279
14.8	Other Buildings & Structures	$0	$608	$511	$0	$1,119	$168	$0	$257	$1,545
14.9	Waste Treating Building & Str.	$0	$1,228	$2,532	$0	$3,760	$564	$0	$865	$5,189

AST Coal First Polygeneration Plant								Estimate Type:		Class 4
								Cost Base:		Dec 2018
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O & Fee	Contingencies		Total Plant Cost
				Direct	Indirect			Process	Project	$/1000
	SUBTOTAL 14.	$0	$6,972	$8,216	$0	$15,188	$2,278	$0	$2,919	$20,386
	TOTAL COST	$442,887	$135,144	$302,411	$0	$880,441	$132,066	$38,899	$177,144	$1,228,550

Estimates related to syngas storage capacity used a syngas storage capacity of 1,000 m^3. The design basis for the storage capacity was motivated by the desire to ease transitions between plant operating points, as well as assisting in handling process upsets (i.e. syngas to be diverted to storage while the gasifier is backdown in event of an issue with the PSA or ammonia train). These transition needs set the capacity requirement, primarily by evaluating the lag in the transition time of the ammonia loop relative to the gasifier trains and the power island. The capacity selected will provide 40 minutes of storage which is sufficient to handle the most drastic operating point transition, and this storage time can be extended to 60 – 80 minutes by performing other operational adjustments during the transition period.

Table 4-2 reports the TOC and TASC using the pre-production and inventory capital requirements required to operate across the whole operating window as strategically desired. As previously noted, the Owner's Costs are based on assumptions found in NETL's *Quality Guideline for Energy System Studies: Cost Estimation Methodology for NETL Assessment of Power Plant Performance*[42].

[42] The cost estimation contained in this report assumes:

- A Debt/Equity split of 55%/45%
- Real current dollar cost of debt of 2.94%
- Real current dollar cost of equity of 7.84%
- A total weighted average cost of capital of 5.14%
- A 5-year capital expenditure period, with a distribution of total overnight capital over the capital expenditure period (before escalation) of: 10%, 30%, 25%, 20%, 15%

Please refer to Exhibits 2-4, 3-1, 3-2, and 3-11 in the referenced QGESS document for additional details and assumptions (National Energy Technology Laboratory, "Quality Guidelines for Energy System Studies: Cost Estimation Methodology for NETL Assessment of Power Plant Performance," U.S. Department of Energy, Pittsburgh, PA, 2019.).

Table 4-2. Polygeneration Owner's Costs

Description	Balanced, 3 GTs
	$/1,000
6 Months All Labor	$12,090
1 Month Maintenance Materials	$1,492
1 Month Non-Feedstock Consumables	$132
1 Month Waste Disposal	$197
25% of 1 Month's Feedstock at 100% CF	$740
2% of TPC	$24,571
Total Pre-production	**$39,222**
60 Day Supply Feedstock & Consumables at 100% CF	$6,095
0.5% of TPC (spare parts)	$6,143
Total Inventory Capital	**$12,238**
Initial Cost for Catalysts & Chemicals	$10,456
Land	$900
Financing Costs	$33,171
Other Owner's Costs	$184,282
Total Other Costs	**$228,809**
Total Overnight Cost (TOC)	**$1,508,818**
TASC Multiplier (IOU, 35 year)	1.154
Total As-Spent Cost (TASC)	**$1,741,177**

Table 4-3 represents the fixed *annual* operating and maintenance costs. These are operating and maintenance costs which are independent of operational choices (i.e. the distribution of time spent in various portions of the operating window defined by the five operating points).

Table 4-3. Polygeneration Fixed O&M Costs

Operating Point: All Cases	**AST Coal First Polygeneration Plant**			**Cost Base:**	**Dec 2018**
				Capacity Factor	**100**
Operating Labor				**Operating Labor Requirements per Shift**	
Operating Labor Rate (base):		38.50	$/hour	Skilled Operator:	4
Operating Labor Burden		30.00	% of base	Operator:	11
Labor O-H Charge Rate		25.00	% of Labor	Foreman:	2
				Lab Techs, etc.:	3
				Total:	20
					Annual Cost
					($)
Annual Operating Labor Cost					$8,768,760
Maintenance Labor Cost					$10,575,577
Administrative & Support Labor					$4,836,084
Property Taxes and Insurance					$24,570,995
TOTAL FIXED OPERATING COSTS					**$48,751,416**

Table 4-4 presents a summary of the Fixed O&M costs on an *hourly* basis, the Variable O&M costs (defined again on an hourly basis) and hourly Feedstock costs for each of the five defined operating points.[43]

One key takeaway from this summary table is that the total O&M costs are primarily driven by the operating capacity of the gasifier. As long as it is operating at 100% capacity, the total O&M, including Feedstock, costs will be ~$12,200 per hour. If the gasifier is turned down (e.g., the *Zero Net Power* case operates the gasifier at 66% capacity), then one starts to see meaningful reduction in the total O&M cost.

[43] Representing Hourly Costs ($/hr) is a deviation from the Annual Costs approach commonly seen in the *Baseline Reports.* This decision is meant to more accurately reflect the expected real-world operating conditions of this polygeneration plant. While the *Baseline Reports'* approach of selecting a single operating point (e.g., max net export power generation) and assuming the plant operates at that point for the entire year at a set capacity factor (e.g., 80% in the 2019 revision of *Volume 1*) is sensible for evaluating a traditional PC, NGCC, or IGCC power plant, it is a poor metric for a polygeneration design that is specifically designed to frequently and rapidly vary the amount of net power and cogeneration product (i.e. ammonia) produced in order to meet market demand.

Table 4-4. Summary of Hourly Polygeneration O&M and Feedstock Costs

Cost Component	Balanced, 3 GTs	Balanced, 2 GTs	Zero Net Power	High Elec Prod	Max Elec Prod
Fixed O&M ($/hr)	**$5,565**	**$5,565**	**$5,565**	**$5,565**	**$5,565**
Variable O&M ($/hr)	**$2,545**	**$2,556**	**$2,409**	**$2,552**	**$2,543**
Maintenance Material Cost ($/hr)	$2,044	$2,044	$2,044	$2,044	$2,044
Water ($/hr)	$50	$63	$38	$64	$74
Chemicals ($/hr)	$180	$179	$148	$173	$154
Waste Disposal ($/hr)	$270	$270	$179	$270	$270
By-Products and Emissions ($/hr)	$0	$0	$0	$0	$0
Feedstock ($/hr)	**$4,059**	**$4,059**	**$2,680**	**$4,059**	**$4,059**
Total:	**$12,169**	**$12,180**	**$10,654**	**$12,176**	**$12,167**

Table 4-5 through Table 4-9 present the detailed breakdown of the Variable O&M and Feedstock costs that are summarized in Table 4-4. The analysis focuses on the December 2018 Dollars per hour since the hours spent in various portions of the operating window are not known a priori. These per hour cost vectors are a key input to AST's investment analysis which uses a reduced form model for the evaluating the profit potential for this polygeneration platform at the five defined operating points.

Table 4-5. Variable Polygeneration O&M and Feedstock Costs for Balanced, 3 GTs Operating Point

Operating Point	**Bal, 3 GT**	**AST Coal First**		**Cost Base:**	**Dec 2018**
Electrical Generation (MW, net)	**48**	**Polygeneration Plant**			
					($)/hr
Maintenance Material:					**$2,044.46**
Consumables					
	Initial Fill	Per Hour	Per Unit	Initial Fill Cost	
Water (gal/1000)	-	26.3910	$1.90	**$0**	**$50.14**
MU & WT Chem. (ton)	-	0.0197	$550.00	$0	$10.81
Carbon (Mercury Removal) (ton)	73	0.0042	$12,000.00	$873,031	$50.27
Water Gas Shift Catalyst (ft^3)	3,320	0.0636	$480.00	$1,593,398	$30.51
Selexol Solution (gal)	118,613	0.6868	$38.00	$4,507,304	$26.10
SCR Catalyst (ft^3)	w/ equip	0.0886	$48.00	$0	$4.25
Ammonia (19% NH_3) (ton)	95	0.0562	$300.00	$28,440	$16.87
Ammonia Synthesis Catalyst (ft^3)	1,766	0.0202	$1,956.00	$3,453,774	$39.44
Claus Catalyst (ft^3)	w/ equip	0.0403	$48.00	$0	$1.93
Subtotal:			$550.00	**$10,455,946**	**$180.18**
Waste Disposal					
Spent Mercury Catalyst (ton)		0.0042	$80.00	$0	$0.34
Water Gas Shift Catalyst (ft^3)		0.0636	$2.50	$0	$0.16
Selexol Solution		0.6868	$0.35	$0	$0.24
SCR Catalyst (ft^3)		0.0886	$3.10	$0	$0.27
Ammonia Synthesis Catalyst (ft^3)		0.0202	$16.00	$0	$0.32
Claus Catalyst (ft^3)		0.0011	$2.00	$0	$0.00
Slag (ton)		7.0713	$38.00	$0	$268.71
Subtotal:		0.0042	$80.00	**$0**	**$270.04**
By-Products Disposal					
Sulfur (ton)		1.9842	$0.00	$0	$0
Ammonia (ton)		27.5000	$0.00	$0	$0
Subtotal:				**$0**	**$0**
Variable Operating Costs Total:				**$10,455,946**	**$2,544.83**
Feedstock Cost					
Illinois #6 (ton)		78.1142	$51.96	**$0**	**$4,058.82**

Table 4-6. Variable Polygeneration O&M and Feedstock Costs for Balanced, 2 GTs Operating Point

Operating Point	Bal, 2 GT	AST Coal First		Cost Base:	Dec 2018
Electrical Generation (MW, net)	51	Polygeneration Plant			
					($)/hr
Maintenance Material:					**$2,044.46**
		Consumables			
	Initial Fill	Per Hour	Per Unit	Initial Fill Cost	
Water (gal/1000)	-	33.1202	$1.90	**$0**	**$62.93**
MU & WT Chem. (ton)	-	0.0247	$550.00	$0	$13.57
Carbon (Mercury Removal) (ton)	73	0.0042	$12,000.00	$873,031	$50.27
Water Gas Shift Catalyst (ft^3)	3,320	0.0636	$480.00	$1,593,398	$30.51
Selexol Solution (gal)	118,613	0.6868	$38.00	$4,507,304	$26.10
SCR Catalyst (ft^3)	w/ equip	0.0812	$48.00	$0	$3.90
Ammonia (19% NH_3) (ton)	95	0.0430	$300.00	$28,440	$12.90
Ammonia Synthesis Catalyst (ft^3)	1,766	0.0202	$1,956.00	$3,453,774	$39.44
Claus Catalyst (ft^3)	w/ equip	0.0403	$48.00	$0	$1.93
Subtotal:				**$10,455,946**	**$178.61**
		Waste Disposal			
Spent Mercury Catalyst (ton)		0.0042	$80.00	$0	$0.34
Water Gas Shift Catalyst (ft^3)		0.0636	$2.50	$0	$0.16
Selexol Solution		0.6868	$0.35	$0	$0.24
SCR Catalyst (ft^3)		0.0812	$3.10	$0	$0.25
Ammonia Synthesis Catalyst (ft^3)		0.0202	$16.00	$0	$0.32
Claus Catalyst (ft^3)		0.0011	$2.00	$0	$0.00
Slag (ton)		7.0713	$38.00	$0	$268.71
Subtotal:				**$0**	**$270.02**
		By-Products Disposal			
Sulfur (ton)		1.9842	$0.00	$0	$0
Ammonia (ton)		27.5000	$0.00	$0	$0
Subtotal:				**$0**	**$0**
Variable Operating Costs Total:				**$10,455,946**	**$2,556.02**
		Feedstock Cost			
Illinois #6 (ton)		78.1142	$51.96	**$0**	**$4,058.82**

Table 4-7. Variable Polygeneration O&M and Feedstock Costs for Zero Net Power Operating Point

Operating Point	**Zero Net Power**	**AST Coal First Polygeneration Plant**		**Cost Base:**	**Dec 2018**
Electrical Generation (MW, net)	**0**				
					($)/hr
Maintenance Material:					**$2,044.46**
Consumables					
	Initial Fill	Per Hour	Per Unit	Initial Fill Cost	
Water (gal/1000)	-	19.9075	$1.90	**$0**	**$37.82**
MU & WT Chem. (ton)	-	0.0148	$550.00	$0	$8.15
Carbon (Mercury Removal) (ton)	73	0.0030	$12,000.00	$873,031	$35.71
Water Gas Shift Catalyst (ft^3)	3,320	0.0636	$480.00	$1,593,398	$30.51
Selexol Solution (gal)	118,613	0.6868	$38.00	$4,507,304	$26.10
SCR Catalyst (ft^3)	w/ equip	0.0290	$48.00	$0	$1.39
Ammonia (19% NH_3) (ton)	95	0.0187	$300.00	$28,440	$5.62
Ammonia Synthesis Catalyst (ft^3)	1,766	0.0202	$1,956.00	$3,453,774	$39.44
Claus Catalyst (ft^3)	w/ equip	0.0268	$48.00	$0	$1.29
Subtotal:				**$10,455,946**	**$148.22**
Waste Disposal					
Spent Mercury Catalyst (ton)		0.0030	$80.00	$0	$0.24
Water Gas Shift Catalyst (ft^3)		0.0636	$2.50	$0	$0.16
Selexol Solution		0.6868	$0.35	$0	$0.24
SCR Catalyst (ft^3)		0.0290	$3.10	$0	$0.09
Ammonia Synthesis Catalyst (ft^3)		0.0202	$16.00	$0	$0.32
Claus Catalyst (ft^3)		0.0008	$2.00	$0	$0.00
Slag (ton)		4.6727	$38.00	$0	$177.56
Subtotal:				**$0**	**$178.61**
By-Products Disposal					
Sulfur (ton)		1.3228	$0.00	$0	$0
Ammonia (ton)		27.5000	$0.00	$0	$0
Subtotal:				**$0**	**$0**
Variable Operating Costs Total:				**$10,455,946**	**$2,409.12**
Feedstock Cost					
Illinois #6 (ton)		51.5815	$51.96	**$0**	**$2,680.17**

Table 4-8. Variable Polygeneration O&M and Feedstock Costs for High Electricity Production Operating Point

Operating Point	High Elec Prod	AST Coal First Polygeneration Plant		Cost Base:	Dec 2018
Electrical Generation (MW, net)	**82**				
					($)/hr
Maintenance Material:					**$2,044.46**
Consumables					
	Initial Fill	Per Hour	Per Unit	Initial Fill Cost	
Water (gal/1000)	-	33.8956	$1.90	**$0**	**$64.40**
MU & WT Chem. (ton)	-	0.0252	$550.00	$0	$13.88
Carbon (Mercury Removal) (ton)	73	0.0042	$12,000.00	$873,031	$50.27
Water Gas Shift Catalyst (ft^3)	3,320	0.0636	$480.00	$1,593,398	$30.51
Selexol Solution (gal)	118,613	0.6868	$38.00	$4,507,304	$26.10
SCR Catalyst (ft^3)	w/ equip	0.1218	$48.00	$0	$5.85
Ammonia (19% NH_3) (ton)	95	0.0661	$300.00	$28,440	$19.84
Ammonia Synthesis Catalyst (ft^3)	1,766	0.0128	$1,956.00	$3,453,774	$25.01
Claus Catalyst (ft^3)	w/ equip	0.0403	$48.00	$0	$1.93
Subtotal:				**$10,455,946**	**$173.39**
Waste Disposal					
Spent Mercury Catalyst (ton)		0.0042	$80.00	$0	$0.34
Water Gas Shift Catalyst (ft^3)		0.0636	$2.50	$0	$0.16
Selexol Solution		0.6868	$0.35	$0	$0.24
SCR Catalyst (ft^3)		0.1218	$3.10	$0	$0.38
Ammonia Synthesis Catalyst (ft^3)		0.0128	$16.00	$0	$0.20
Claus Catalyst (ft^3)		0.0011	$2.00	$0	$0.00
Slag (ton)		7.0713	$38.00	$0	$268.71
Subtotal:				**$0**	**$270.03**
By-Products Disposal					
Sulfur (ton)		1.9842	$0.00	$0	$0
Ammonia (ton)		17.4350	$0.00	$0	$0
Subtotal:				**$0**	**$0**
Variable Operating Costs Total:				**$10,455,946**	**$2,552.28**
Feedstock Cost					
Illinois #6 (ton)		78.1142	$51.96	**$0**	**$4,058.82**

Table 4-9. Variable Polygeneration O&M and Feedstock Costs for Max Electricity Production Operating Point

Operating Point	**Max Elec Prod**	**AST Coal First Polygeneration Plant**		**Cost Base:**	**Dec 2018**
Electrical Generation (MW, net)	**112**				
Variable Operating Costs					
					($)/hr
Maintenance Material:					**$2,044.46**
Consumables					
	Initial Fill	Per Hour	Per Unit	Initial Fill Cost	
Water (gal/1000)	-	39.1546	$1.90	**$0**	**$74.39**
MU & WT Chem. (ton)	-	0.0292	$550.00	$0	$16.04
Carbon (Mercury Removal) (ton)	73	0.0042	$12,000.00	$873,031	$50.27
Water Gas Shift Catalyst (ft^3)	3,320	0.0636	$480.00	$1,593,398	$30.51
Selexol Solution (gal)	118,613	0.6868	$38.00	$4,507,304	$26.10
SCR Catalyst (ft^3)	w/ equip	0.1218	$48.00	$0	$5.85
Ammonia (19% NH_3) (ton)	95	0.0650	$300.00	$28,440	$19.51
Ammonia Synthesis Catalyst (ft^3)	1,766	0.0020	$1,956.00	$3,453,774	$3.87
Claus Catalyst (ft^3)	w/ equip	0.0403	$48.00	$0	$1.93
Subtotal:				**$10,455,946**	**$154.08**
Waste Disposal					
Spent Mercury Catalyst (ton)		0.0042	$80.00	$0	$0.34
Water Gas Shift Catalyst (ft^3)		0.0636	$2.50	$0	$0.16
Selexol Solution		0.6868	$0.35	$0	$0.24
SCR Catalyst (ft^3)		0.1218	$3.10	$0	$0.38
Ammonia Synthesis Catalyst (ft^3)		0.0020	$16.00	$0	$0.03
Claus Catalyst (ft^3)		0.0011	$2.00	$0	$0.00
Slag (ton)		7.0713	$38.00	$0	$268.71
Subtotal:				**$0**	**$269.86**
By-Products and Emissions					
Sulfur (ton)		1.9842	$0.00	$0	$0
Ammonia (ton)		2.71167	$0.00	$0	$0
Subtotal:				**$0**	**$0**
Variable Operating Costs Total:				**$10,455,946**	**$2,542.78**
Feedstock Cost					
Illinois #6 (ton)		78.1142	$51.96	**$0**	**$4,058.82**

Table 4-10 represents the calculated required first-year cost of electricity (COE) in dollars per MWh required at the five representative operating points based on the previously discussed financial assumptions and cost estimates and accounting for revenue obtained through the sale of ammonia produced by the polygeneration plant.[44] It should be noted that Table 4-10 below does not reflect additional revenue for any potential CO_2 sales prices and emissions penalties, profit from sale of sulfur, etc.

The first year COE estimate was evaluated over the range from the current ammonia retail cost ($551/ton, representing the "high end" estimate) and the current United States Gulf Coast (USGC) ammonia contract price ($195/ton, representing the "low end" estimate). The retail price represents a reasonable upper bound estimate on potential ammonia revenue (i.e., full capture of the distributed ammonia production advantage), whereas the USGC contract price represents a current reasonable lower bound estimate, for this stage of evaluation, on the potential ammonia revenue (i.e., no capture of the distributed ammonia production advantage). An intermediate choice, such as 75% of the retail price, is a more plausible basis for evaluation. In practice, the plant would most likely capture a different level of locational advantage based on the geographical distribution of customers relative to the specific plant siting.

Table 4-10. First Year COE ($/MWh-net) at Five Defined Operating Points for Various Ammonia Price Sensitivities

Balanced Gen, 3 GTs	Balanced Gen, 2 GTs	Zero Net Power	High Electricity Production	Max Electricity Production	NH_3 Revenue ($/ton)	
$245	$234	N/A	$213	$227	$551	Full Retail Ammonia Price w/o Distribution Costs
$323	$308	N/A	$243	$231	$473	75% of Retail
$401	$383	N/A	$272	$234	$276	50% of Retail
$448	$427	N/A	$290	$236	$195	Current U.S. Gulf Coast Delivery; No locational Advantage

A breakdown of these first year COE's across previously discussed cost components is presented in Table 4-11 using the "75% of Retail" price point for ammonia.

[44] It is appropriate to present the required first-year COE in Table 4-10 through Table 4-12 in terms of $/MWh-net (in contrast to the approach adopted for Table 4-5 through Table 4-9) as the results presented in these tables take into account the financial value that can be provided by the ammonia production aspect of the polygeneration plant. Additional discussion on this point can be seen following Table 4-12.

Table 4-11. First Year COE ($/MWh-net) Breakdown with Ammonia Price Set at 75% of Retail

First Year COE Component	Balanced 3 GTs	Balanced 2 GTs	Zero Net Power	High Elec Prod	Max Elec Prod	Percent[45]
Capital	$302	$288	N/A	$179	$130	54%
Fixed O&M	$120	$114	N/A	$71	$52	22%
Variable O&M	$53	$50	N/A	$31	$23	9%
Feedstock	$84	$80	N/A	$50	$36	15%
Total (Excluding Ammonia Revenue)	**$559**	**$532**	**N/A**	**$331**	**$241**	N/A
Ammonia Revenue	$235	$224	N/A	$88	$10	N/A
Total (Including Ammonia Revenue)	**$323**	**$308**	**N/A**	**$243**	**$231**	N/A

Table 4-12 provides an alternative representation on the information contained in Table 4-11. Rather than including "Ammonia Revenue" as a separate line item, it has been pro-rated and included as a credit in each of the other cost components. For example, 54% of the $235 "Ammonia Revenue" in the *Balanced, 3 GTs* case was applied as a credit to reduce the First Year COE of the "Capital" cost component, 22% of the $235 "Ammonia Revenue" was applied as a credit to the "Fixed O&M" cost component, etc.

Table 4-12. First Year COE ($/MWh-net) Breakdown with Pro-Rated Ammonia Revenue

First Year COE Component	Balanced 3 GTs	Balanced 2 GTs	Zero Net Power	High Elec Prod	Max Elec Prod	Percentage
Capital	$175	$167	N/A	$132	$125	54%
Fixed O&M	$69	$66	N/A	$52	$49	22%
Variable O&M	$30	$29	N/A	$23	$22	9%
Feedstock	$49	$46	N/A	$36	$35	15%
Total	**$323**	**$308**	**N/A**	**$243**	**$231**	N/A

It is important to note the impact of including the pro-rated ammonia revenues to the various cost components. For example, the *Capital* cost component at the *Balanced, 3 GTs* operating point is $302/MWh-net without ammonia revenue considered (Table 4-11). However, when the pro-rated ammonia revenue is included (Table 4-12), the *Capital* cost component is reduced to $175/MWh-net.

This large change in the apparent *Capital* cost component (as well as similar comparisons between Table 4-11 and Table 4-12) should provide justification for the previous decision to forego inclusion of metric based on a net export power (e.g., $/MWh-net) in Table 4-4 to Table 4-9.

The COE metric is inherently challenging for use in comparing a polygeneration plant to other power producing facilities. For example, a cursory glance may make it appear that the *Max Electricity Production* representative operating point is superior to the other four at ammonia

[45] This represents the percent of each cost component relative to the *Total (Excluding Ammonia Revenue)*. While the percentage was not exact across all five operating points, the variance fell within the bounds of round-off error (e.g., Capital Cost percentages ranged from 54.12% to 54.15%).

prices at "75% of Retail" and below, but realistically this is simply a construct of the calculation. In this operating mode, while the gasifier and power island are being fully utilized to capacity, there is capital cost for the ammonia loop which is not being fully utilized. Similarly, in the *Balanced* representative operating modes the power island capital equipment is not being run to capacity (hence higher COE, as capital costs are being spread among fewer MWh), but the ammonia loop is being run to capacity (hence no idle capacity or capital costs in the ammonia loop). Since COE fails to adequately capture or evaluate the value of the multiple product, multiple operating point polygeneration facility, a multivariate financial analysis is necessary to support the technology platform and future project decisions.

5. Technology Gap Analysis & Commercial Pathway

5.1 Performance Gap Assessment

Section 3, Performance Results provides a detailed assessment of the polygeneration system's performance over a broad range of operating points that define the overall plant operating window. Additionally, *Section 3.4* of *Performance Results* above provides an explicit evaluation on how the polygeneration system meets the objectives of the Coal FIRST program. Of these, the only objective that is not unambiguously satisfied is system efficiency. The efficiency of the system varies with operating mode and an aggregate efficiency is difficult to define for a polygeneration system without more detailed analysis and significant forecasting of assumptions on what percentage of time the plant will spend at each operational point.[46] As such, any estimate of overall HHV efficiency of the plant incorporates a level of subjectivity in choosing a representative reference operating point. However, since the system reaches high efficiency numbers across a wide operational range, it is believed that no significant performance gap will exist, especially when considering a non-capture case. As such, no further research and development efforts are *required* for the polygeneration system to meet the objectives of the Coal FIRST program.

Additionally, as described below, all of the equipment in the designed polygeneration system can be commercially procured at this time. The specifications of equipment sent for bid may be adjusted through process validation and piloting to complement the current systems analysis, but there is no anticipated novel function or equipment that will need to be created. As such, no further research and development is *required* to deploy the equipment currently detailed on the polygeneration flowsheet.

While no inventive or inherently risky development is required to deploy the described system, the system can still benefit from additional process development and engineering activities aimed at lowering the technical risk to full-scale deployment through validation of modeled and simulated system performance and operating characteristics. Given the mature nature of the core unit operations and the manageable level of technical risks associated with system integration, a higher risk approach of omitting the pilot phase could be considered to accelerate deployment. In this instance, the first commercial application would be a pioneer plant with the understanding that evaluation of the pilot plant objectives would come during initial pioneer plant operations leading to an improvement-based turnaround that implements the learning of the pilot stage investigations undertaken during the early operations of the pioneer plant. However, this approach is inherently risky, and, as such, the current recommendation incorporates separate piloting and commercial phases, particularly as successful pilot operations will prove out a reduced project risk level and lead to better financing terms. Additionally, while not required for deployment, the performance of polygeneration technology platform may be improved by supporting innovation outside of the current work and commercially available offerings.

[46] A financial and investment model has been developed for inclusion in the final report that is capable of forecasting time spent at each defined operating point under a given set of economic conditions. However, as the results of this model are highly dependent on the specifics of the user-defined scenario and forecasting market evolution over a 10+ year time frame is inherently difficult, it is not reasonable to present a definitive statement regarding projected time spent at each operating point at this time.

A pilot plant program serves both of these goals by providing an opportunity to validate modeled and simulated results in a real-world setting, as well as providing an option to explore integration of novel innovations into the system. These benefits are captured in the pilot plant objectives section below.

5.2 Commercialization Assessment[47]

The section will provide an assessment of the ability to move forward and implement the polygeneration system described herein. This section would typically contain discussion of further research and development required to move forward and implement the designed system, had there been any required. Discussion on additional systems integration, validation, and opportunity for incorporating further innovation during pilot plant operations is covered in *Section 5.3: Pilot Plant Operations*. The later sub-section of the *Commercialization Assessment* is to document the basis and resources that Team AST (Allegheny Science and Technology, Worley, and Catalyte) used in undertaking that assessment.

5.2.1 *High-Level Commercial Readiness*

Table 5-1 provides a high-level assessment of the commercial readiness of each major unit operation in the design basis, as well as high-level notes on application of generally available commercial components to the specific polygeneration plant design. The technology readiness levels (TRLs) are based on the Department of Energy definitions. By design, the polygeneration system integrates mature, stable, and fully commercialized subsystems (TRLs of 9).

With this being said, there are two subsystems with specific, minor demonstration needs related directly to the designed use that may warrant a slightly lower TRL designation:

- While fluidized bed drying of coal is an established and demonstrated process, it has not been demonstrated for this specific coal basis.
- The operational design of the gasifier includes partial oxidation in the freeboard to reduce the methane content of the produced syngas. While this combines two well-established commercial operations and the selected vendor has specific experience with such application, the operation requires some specific operational validation under the polygeneration design conditions.

The TRL designations of 9 on all other subsystems are justified both by extensive relevant commercial operations and commercial availability of all relevant subsystem equipment.

[47] To assure the technology gap analysis assessment are interpreted properly, it is important to distinguish that the presented technology readiness assessment is focused on the technical maturity and commercial availability of the individual unit operations and equipment. The desire to include commercially available equipment and unit operations with long operating histories was inherent in the design approach. However, it is understood that the specific proposed system requires the process development actions articulated below related primarily to integration in order to be applied commercially. Specifically, the TRL of the overall system is not an 'average' of the individual components but of the state of integration.

Table 5-1. Commercial Readiness of Plant Operating Sections

Operating Section	Component Availability	Pathway Forward
Coal Receiving and Handling	Commercially Available	Mature, stable, and established technology. Technology Readiness Level: 9
Coal Preparation and Feed Systems	Commercially Available	Mature, stable, and established technology. Technology Readiness Level: 9
Coal Drying System	Commercially Available	Bubbling bed drying and desorption is an established and demonstrated process. Disciplined detailed engineering and scale-up of this bubbling bed is required. This process can be fabricated by standard qualified, coded vessel fabrication shops based on a design provided during a future detailed design phase or by competitive solicitations based on a duty specification Technology Readiness Level: 9/7; same scale, but different coal feed basis, have operated commercially. Minor demonstration for this coal in the context of this system will be helpful to fully mitigate technical risk.
Air Separation Unit	Commercially Available	Mature, stable, and established technology. If future generations of this technology platform change scale then pressure swing adsorption options should be reconsidered; however, this is also an established commercial technology. Technology Readiness Level: 9

Operating Section	Component Availability	Pathway Forward
Gasifier	Commercially Available	SES U-Gas design is a mature, stable, and established commercial design. This gasifier has successful been deployed at this scale on utilizing Illinois #6 coal. Partial oxidation is commercially known and deployed technology. This reduces the need for significant supporting experimentation (cold flow), modeling (CFD), or analysis based on fluid bed design and scale-up methodologies as the vendor has already completed this process. Minor demonstration of the partial oxidation in the freeboard would be helpful to fully mitigate technical risk. SES is in the middle of corporate restructure and ownership changes, but based on previous conversations with representatives, there is a high degree of confidence that some entity will retain and support the licensing of the SES U-gas design. Technology Readiness Level: 9/8.
Water Gas Shift	Commercially Available	Mature, stable, and established technology. Technology Readiness Level: 9
Syngas Cooling	Commercially Available	Mature, stable, and established technology. Technology Readiness Level: 9
Syngas Clean Up	Commercially Available	Mature, stable, and established technology options. Honeywell UOP Selexol technology forms the basis of the carbon capture component. Future generations of this platform can look to integrate improvements in pre-combustion capture when their technical maturity is sufficient to warrant the risk. Pre-combustion capture has the potential to provide a step change improvement in financial performance via reduced capital expenditures, reduced parasitic load, and an easier to handle stream of CO_2. Technology Readiness Level: 9
Syngas Management	Commercially Available	Mature, stable, and established technology. Technology Readiness Level: 9
Ammonia Generation	Commercially Available; Current R&D offers significant	Mature, stable, and established technology options at scale relevant to this project. Active R&D in areas such as catalysis and process intensification offer potential innovation opportunities for future generations of this

Operating Section	Component Availability	Pathway Forward
	opportunities for future designs	technology platform (see description in the Pilot Plant Objectives section below). Multiple vendors (KBR, Proton Ventures, Linde, ThyssenKrupp, Casale, JGC, and Haldor Topsøe) have offerings at scales at or greater than the scale of interest, albeit their "standard" packages would need to be adapted to this application. The cycling of the NH_3 train will complicate vendor negotiations with respect to warranty and performance assurance, regardless the current system interacts with these trains within known dynamic performance. Technology Readiness Level: 9
Power Block	Commercially Available	Mature, stable, and established technology. Selected turbine (LM2500+) has displayed capability to operate with high H_2 content fuels. General Electric has been leveraged by Worley experts in modeling their performance for this application. The more detailed heat integration and recovery during this pre-FEED study has led to more optimized operation and integration of steam turbine generation based on well-established techniques. Please refer to the *Performance Results* (Section 3) for a more detailed discussion of the heat integration strategy. Technology Readiness Level: 9

A detailed *Major Equipment List* associated with these operating sections is provided in Appendix C. None of the detailed equipment requires significant research and development activities in order to be procured and delivered. The process description, modeling, and specifications (functional duty specification) are substantially developed to the point where they could be used to structure competitive bids on the equipment in the time frames outlined in *Appendix I, Execution Plan*, when financing for a pilot or pioneer plant supports such activity.

There are systems integration and validation piloting activities that will lower the risk associated with developing the functional specifications of this equipment solely based on process modeling and simulation. These development activities that focus on systems integration and validation are described below in *Section 5.3: Pilot Plant Objectives*. There are no equipment development or issues in the performance of the equipment to preclude implementing the system within the timelines targeted by the Coal FIRST program. The project implementation and execution timelines have been developed and presented in *Appendix I, Execution Plan*.

5.2.2 *Summary of Team Experience with Relevant OEMs*

Team AST member Worley has extensive experience with relevant OEM and project databases to translate the commercially ready options listed above to this project. Additionally, Team AST member Catalyte has extensive experience in ammonia projects, including interfacing and interacting with ammonia licensors and vendors to complement Worley's expertise.

Worley Group Inc. (formally WorleyParsons Group, Inc.) provides the Architecture and Engineering (A&E) firm component of Team AST. The A&E functions serves to assure designs that reflect state of the art commercial practice, leverages relevant vendor relationships, and can draw on learnings from an extensive and relevant set of projects. From Worley's experience working on a range of similar study type projects and commercial power generation projects, Worley has developed a range of contacts with Original Equipment Manufacturers (OEMs). The following provides an overview of Worley and Team AST's experience with OEMs for the critical equipment of this polygeneration system. Also, since the conceptual design phase and the pre-FEED study phase, the merger of Worley and Jacobs Engineering's Energy, Chemicals, and Resources group. This merger doubled the size of Worley and has resulted in a substantial expansion of the resources, network of OEM contracts, past reference projects, databases, and experience that can be provided relative to the resources available during the Conceptual Design Phase.

5.2.2.1 Gasifier

Worley has worked extensively with the licensor and fabricators to develop the capital costs for this unit. The Worley lead has been in contact with SES for guidance on application of the U-gas design to support modeling efforts.

5.2.2.2 Gas Turbine & Steam Turbine

Worley has successfully built many power projects that utilizes gas turbines from various OEMs including that of General Electric, Siemens, Mitsubishi, and Alstom. From small aero-derivative gas turbine to the largest advanced class H, J and JAC class gas turbines, Worley had been involved with major OEMs and projects spanning throughout the world. These relationships have been leveraged in assessing turbine choices from the conceptual design process through the current more formed description and model of the polygeneration system. Worley conducts annual technology meetings with major OEMs, during which each OEM will showcase their latest advancements in their gas turbine products and lessons learnt from their projects worldwide. Worley tracks current advancements in the Gas Turbine Technologies. For this project, Worley received input on the suitability and performance from General Electric on a range of aero derivative gas turbines before selecting the LM2500+ GT for this project, specifically in the area of ensuring compatibility with high hydrogen content fuel. The design of the power block and selection of the steam turbines was completed using software cross referenced with the Worley Group Project Library database.

5.2.2.3 Air Separation Unit (ASU)

Worley has worked as an EPC as well as at the FEED and Pre-FEED study level on many ASU projects, working with all the major vendors including Air Liquide, Air Products and Linde.

5.2.2.4 *Syngas Cooler*

Worley has worked with major Syngas cooler equipment suppliers on various study projects as well as on some of the combined cycle gas turbine power projects.

5.2.2.5 *Water Gas Shift*

Worley has worked with all of the major water gas shift vendors including Johnson Matthey, Haldor Topsøe and Clariant on a variety of projects including power generation with carbon capture and coal to ammonia projects.

5.2.2.6 *Gas Cooling and Desaturator*

The desaturator, which is a packed column used as a direct contact heat exchanger, is commonly used in our chemical plant designs. Worley has worked closely on many projects with vendors for the column as well as the internal distributors and packing to optimize the design.

5.2.2.7 *Acid Gas Removal*

Worley has evaluated all of the major selective acid gas removal technologies for multiple clients, and for both IGCC and coal to ammonia projects, the SelexolTM technology for UOP is the most cost-effective solution. These projects have involved multiple sets of performance data from UOP (in one case over 20) as the design is optimized.

5.2.2.8 *Sulfur Recovery Unit*

Worley has supplied and / or licensed over 60% of the sulfur recovery units in the world.

5.2.2.9 *Pressure Swing Adsorption (PSA)*

Worley has worked with all of the major PSA vendors including Air Products, Air Liquide, Linde, UOP and others

5.2.2.10 *CO_2 Compressors*

Worley has interfaced with the compressor manufacturers like Kobelco, Atlas Copco, MAN Turbo, Ingersoll Rand etc. on our current projects involving gas compression duty for various gases including natural gas, CO2, other product gases.

5.2.3 *Equipment Information Resources*

Worley has provided Team AST resources for equipment information including Vendor Data & Interfaces, the Worley Project Library, budgetary quotes, and past reference projects.

5.2.3.1 *Vendor Data & Interface*

Worley has direct key vendor contacts for major critical equipment in the gasification process.

Worley interfaces with the OEMs directly on a regular basis. Some of the OEMs have given access to Worley Engineers to be able to run the OEM's performance estimation software on OEM's computer portals. Worley relationships with OEMs is also leveraged to have the OEMs provide the emission and performance estimates for given ambient conditions, fuel types, various load points and different cooling system configurations. This is useful completing vessel components such as the fluid bed dryer.

5.2.3.2 Past Project References

In addition to the above sources, Worley also has access to generic published data from previously completed studies performed by Worley on various gasification study projects.

Additionally, Worley and AST's limited experience with commercial ammonia process and catalyst equipment and technology licensors is complemented by the subject matter experts at Catalyte who are actively engaged in this area. Catalyte's contacts and experience was leveraged to verify offerings exist near our intended train size and that provided confidence that the vendors list above would be amenable to adapting their standard offerings to bid on a functional duty specification when it comes time to procure equipment.

5.3 Pilot Plant Objectives

The foundation Team AST's proposed polygeneration concept is the paradigm that meeting CFI's objectives is best accomplished by combining intelligent systems analysis, engineering process development, and novel applications of existing, proven technology platforms. This approach provides greater confidence in modelling and analysis since it does not rely on attempting multiple, significant, high-risk inventive steps based on emerging or nascent technology. The end result is that the polygeneration concept is more based on sound development than research.

Nonetheless, even the integration of established components benefits from pilot operation activities focused on validating systems modeling and analysis, gaining operational efficiency, and mitigating the risks of systems integration. Besides supporting the detailed engineering and equipment refinement inherent in proceeding with a complex engineering project, pilot plant operations provide fertile ground for additional supporting innovations (generally low to medium risk innovations) that improve operational understanding, inspire modifications to the flowsheet for further costs savings, increase operational flexibility, and improve performance. The pilot scale recommended at this time to be 1/10th of the design basis, which would essentially entail a single gasifier and single ammonia loop (as opposed to two, as contemplated in the design basis) at 1/5th capacity. This scale retains the essential system characteristics to properly assess dynamics, response, controls, and performance while not committing to a more expensive, higher-risk full-scale pioneer application. The activities identified to advance these objectives are below in *Section 5.3.1 Risk Mitigation and Validation of the Current Design Basis*.

Additionally, pilot operations provide the opportunity for the technology platform to capture external innovations for potential implementation in future generations. While such activities are not necessary for executing the first generation of the polygeneration concept described in the current design basis and process description, potential pilot plant activities aimed at improving future iterations of the polygeneration concept are captured below in *Section 5.3.2 Incorporation of Higher-risk Supporting Innovation into Future Generations of the Technology Platform*. This section is provided to show the potential arc of improvement for the polygeneration platform beyond the initial deployment envisioned as part of the Coal FIRST program.

5.3.1 *Risk Mitigation and Validation of the Current Design Basis (First Generation)*

While extensive modeling, analysis, and vendor interactions have occurred to optimize the integration of mature components into a polygeneration system response to the Coal FIRST program, validation and system integration is greatly enhanced through targeted pilot plant activities. Additionally, operational understanding and full appreciation of system dynamics are a secondary desired product of the pilot plant operations.

5.3.1.1 Process Controls Development

A core outcome of pilot plant operations is the development of the process control strategy and corresponding validation in real-world operation. Additionally, the controls strategy for a system designed for system flexibility and frequent, rapid transitions between states is an inherently different challenge than targeting steady-state name plate capacity for a “monogeneration” process (i.e., IGCC or ammonia-only production as opposed to polygeneration). One of the challenges in developing the controls for this polygeneration plant is the limited ability to fully capture the complex dynamic performance of multiple connected systems. The various state transitions and high turn down anticipated in the ammonia train are expected to require close attention, particularly with respect to the recovering the heat from the process.

The discipline of completing the hazard-operability reviews required for a pilot plant operation has the added benefit of forcing the detailed assessment of how the system truly operates and how the transitions occur, providing validation of the transitions described in Sections 2.3 and 2.4. This ancillary benefit of proper process safety is invaluable in developing a complete and full set of control loops to allow the system to perform as intended, as well as identify needed automated alarms, shutdowns, and emergency response.

Operational experience of this initial set of control loops allow validation and then improvement of controller strategy, design and tuning. The strategy and design of controls entails a detailed understanding of the system dynamics and understanding not only how a specific action creates reactions throughout the system, but also a conscious decision of how one intends to make individual control actions to initiate (i.e., selection of the manipulated variable-control variable pairings and the associated controls methodology) intended system reactions. Additionally, pilot demonstration allows unit operators and engineers the ability to start to innovate alternate control methodology which can be considered and, eventually, lead to evolution of the overall process control scheme.

While none of the activities described in developing the control scheme require innovative research, they are fundamental, required process development steps that are critical to safely and effectively making the transitions that this polygeneration system was designed to do.

5.3.1.2 Operations and Transitions Validation

Related to the operations that allow control system development are operations to validate that the system operates at steady-state as intended. Additionally, the transition between steady-states (e.g. the operating modes described in Performance Results) both in terms of timing and avoidance of problems also must be validated. Transitions between operating points are best validated due to difficulties in reliably and (more often) comprehensively modeling dynamic system operations (permutations grow geometrically with the number of unit operations, systems, and potential operator choices).

Operational data allows analysis of unit operations to determine short-comings and determine if the specifications determined through modeling require additional functionality to operate as intended. Often the issues are that the specifications derived from modeling are incomplete and there are other operational aspects (e.g. minimal velocity to prevent entrained solids from 'salting out' at an inconvenient location) that need to be added to the specification. While the Pre-FEED study included detailed heat integration analysis and careful consideration of the implementation and frequency of anticipated transitions, actual operating experience has historically proven to be the only true manner to assess and optimize a complex engineered systems' response to transitions and disturbances. This operational guidance allows an understanding of the process beyond the five modeled operational modes. Additionally, this allows for definition of operational points and transitions with enough data to do true root case analyses (fishbone, 5 whys, Pareto analysis, Kepner-Tregoe analysis, etc.) to truly understand operations.

Again, such activities do not require innovative research but are fundamental activities of process development driving a design basis from a paper exercise to a commercially complete design basis with minimal technical risk.

5.3.1.3 Validate Fluid Bed Dryer Operations

The primary purpose of the fluid bed dryer is to (1) facilitate drying of the as-received feedstock to meet the requirements of the SES U-Gas gasifier and to (2) release any hydrocarbons that are adsorbed in the pores of the crushed coal.
The fluid bed dryer meets these objectives by:

- Reducing the moisture content of the coal prior to delivery to the gasifier
- Reducing the amount of light hydrocarbons adsorbed in the pores of the coal.[48]

Through these functions, the fluid bed dryer assures a more consistent feedstock for the gasifier. Specifically, the wet coal (11.12% moisture content by weight) is dried within the fluid bed dryer to a ~5% moisture content by weight through indirect heating supplied by excess low-pressure steam that is generated in other plant processes. ASU-supplied nitrogen will be introduced as a stripping gas into the fluid bed dryer to aid in stripping of the removed moisture and absorbed light hydrocarbons from the system. In addition to serving as the sweep gas, this nitrogen forms the

[48] The coal selected for this study, as defined by DOE, is assumed to be "adsorbed hydrocarbon free." However, it is believed that the potential exists for trace amounts of adsorbed hydrocarbons in real-world feedstocks.

bulk of the diluent that will be required to ensure that the syngas composition meets the requirements of the turbine that has been selected.

The resulting overhead stream from this drying and desorption process contains the stripping gas, the moisture driven off of the as-received coal, and any desorbed hydrocarbons. The resulting overhead stream from this drying and desorption process contains the stripping gas, the moisture driven off of the as-received coal, and any desorbed hydrocarbons.[49] Water is knocked-out from the overhead stream by condensation through a transfer line exchanger prior to re-integration of the overhead stream with the post-water gas shifted (WGS) syngas stream. This re-integration occurs after the acid gas removal (AGR) system and before fuel gas conditioning.[50]

The core product of the fluid bed dryer is the sufficiently dried coal stream. The final disposition of this solid stream is delivered to the gasifier for conversion to syngas.

The details of the fluid bed dryer need to be validated during pilot plant operations. Specifically, we need to assure that the size of this overhead stream is very small, predominately comprised of nitrogen and water, and devoid of operationally difficult tars, ash, particulates and entrained atomized hydrocarbons. The default specification of Illinois #6 coal for NETL systems studies does not facilitate modeling of adsorbed hydrocarbons. The minute production of tars and atomized hydrocarbons are not well captured in process simulation analysis but do create significant operational issues when accumulated.

It is currently believed that the fluid bed dryer will act as nothing more than a robust and resilient coal dryer. Thus, based on the known physical chemistry of the feedstock and at the operating temperatures there will not be cracking of the hydrocarbons contained in the coal. The designed operating temperature is well below that at which thermal cracking can occur, thus regardless of the presence of light hydrocarbons adsorbed in the pores of the coal. The dryer will be below the temperatures at which coal devolatilizes and releases tars, oils and other components. Even should these undesirable species release from the coal, the dryer will be operating below temperatures where these thermally crack. Provided operations stay within the parameters mentioned, the overhead stream should only contain the relatively small amounts of hydrocarbons that were previously adsorbed in the pores of the coal. Significant cracking of hydrocarbons in the coal will impact the amount of hydrocarbon feed to the gasifier, based on the current routing of the overhead stream.

The piloting plant will sample this overhead stream to evaluate the amount of hydrocarbons present. If the hydrocarbon content exceeds an acceptable limit, the operation of the fluid bed dryer will be evaluated to see if the drying of the coal can still be accomplished without the production of increased hydrocarbons in the overhead.[51] If that proves unsuccessful, alternative methods will

[49] It is the intention and belief that the overhead stream will only contain minimal amounts of desorbed hydrocarbons with pilot plant testing to quantify and characterize hydrocarbons that wind up in the fluid bed dryer overhead stream (most likely desorbed hydrocarbons from the pore volume of the coal, but possibly generated but unintended chemical transformation of the coal in "hots spots" or other poor operation transients).

[50] If piloting reveals significant hydrocarbons, heteroatoms compounds, etc., then this approach may need to be re-visited. Possible inclusion of a baghouse prior to fuel gas conditioning can be used to protect down-stream equipment.

[51] The most likely possible sources of unintended hydrocarbon cracking include unintended chemical transformation of the coal in "hots spots" within the fluid bed or other poor operation transients.

be evaluated, including use of a different drying method or changes to the reintegration strategy for the overhead stream. There also will be active attempts to track tars and atomized hydrocarbons—this is often difficult and often requires either significant operating hours or deliberate routing to filters and other items that would not be part of a commercial flow sheet.

Additionally, pilot operations will monitor for potential unintended release of mercury for the coal. The mercury content of the overhead stream will be monitored and evaluated during the piloting process. As the current design basis includes re-integration of the overhead stream after the mercury removal bed[52], any mercury contained in the overhead stream will not be captured and will eventually be exhausted as part of the power island's flue gas. As high mercury emission levels are undesirable, it is important to measure the mercury content of the overhead stream in the pilot plant to validate that it is low enough to be considered "acceptable" for plant operation. If high mercury content is found, fluid bed dryer operation will be examined to attempt to eliminate the creation or driving off of mercury into the overhead stream. If this proves unsuccessful, the feasibility and impacts of reintegrating the overhead stream upstream of the mercury removal bed will be examined to ensure that the exhaust gas from the power island exhibits a low mercury content.

5.3.1.4 Methane Control in the SES U-Gas Gasifier

Gasification will be completed through the use of an SES U-Gas style gasifier. This style of gasifier has been selected as it is vendor supported as well as the fact that there are a number of existing and recent commercial operations, helping to ensure a flow of active and fresh operating knowledge. Additionally, both the vendor and selected gasifier design have demonstrated experience operating with the selected Illinois #6 feedstock. These factors combine to lower the technological risk associated with piloting and commercialization of the overall plant design.

One area of concern with the selection of this gasifier design is the fact that standard operating conditions result in syngas with a methane content of ~7%. As the selected carbon mitigation approach (i.e., a standard, dual-stage Selexol system) will not capture any of the carbon contained in the methane, this level of methane content in the syngas will make it very difficult to economically achieve the 90% carbon capture goal of the design basis.

In order to address this concern, oxygen injection into the freeboard of the gasifier can result in a reduction of the methane content to ~1% through partial oxidation of the methane. While SES has indicated that they have utilized a similar approach before (in fact, it was SES who originally recommended this potential solution) and while the partial oxidation of methane is well known, validation of this approach during piloting ensures that this approach is an effective means to control methane content and does not have any spurious operational issues (such as afterburning due to the lack of thermal mass in the freeboard) and to identify the required operating characteristics, both at steady state and during system transients.

[52] This point has been selected for re-integration because it allows for use of an existing compressor, reducing the need for additional capital expenditures.

5.3.2 *Incorporation of Higher-risk Supporting Innovation into Future Generations of the Technology Platform*

This section documents potential improvements to the polygeneration system that could be developed in pilot plant operations. These potential improvements are too risky to deploy in the first generation of the polygeneration platform and not necessary (based on our Performance Results) to meet the objectives of the Coal FIRST program. Potential improvements are documented below so the future arc of platform performance can be understood and that any pilot plant design considers the capability to support investigation of these options.

5.3.2.1 Fluid Bed Dryer's Impact on Coal Feed Flexibility

One initial justification for using a bubbling fluid bed dryer was the potential for leveraging existing, deployed capital equipment to further increase operational flexibility and value opportunities throughout the plant's lifecycle. Specifically, the fluid bed vessel provides an opportunity to handle high-sulfur content coals with minimal modifications and capital outlay through the use of limestone injection. Sulfur removal via limestone injection is a known method of desulfurizing coal feeds, that typically requires more intense (higher temperature) conditions than used in the current drying process. The design basis, equipment specifications and plant cost estimate include a dryer vessel with sufficient size, material and pressure specifications to handle the necessary temperature and pressure to investigate sulfur removal via limestone injection. If successful, this additional sulfur mitigation opportunity can enable the use of high sulfur coal sources at some point in the plant's lifecycle without the need to expand the fixed capacity of the acid gas removal system beyond the size of the originally installed system. This capability is analogous to the ability of refineries to accept various qualities of crude oil feedstocks, this unit operation increases overall plant flexibility and supports potential future arbitrage opportunities among different available coal feedstocks.

While current efforts have focused on the use of Illinois #6 as the primary fuel feedstock, initial analysis in the Conceptual Design phase suggests that this approach could support the use of additional coal feedstocks, including waste coal streams. However, as this significantly changes the operating characteristics of the fluid bed dryer and expands the operational goals of the unit process, it is important to perform adequate piloting efforts to verify that it will still operate successfully under these new conditions.

The first piloting goal in this area will be to confirm that limestone injection will be effective in removing sufficient sulfur to the point that expansion of the existing AGR system is not needed to handle such coal feedstocks. While limestone injection is a fairly standard process with generally well understood chemical interactions, it is important to verify the operating details in the specific system in order to properly control and operate all downstream systems, as well as to adequately size limestone feeds and waste disposal systems. Additionally, sulfur removal in this manner occurs at more severe processing conditions, the fluid bed dry begins to approach 'devolatilization' operations which requires extensive re-analysis of the 'dryer' overhead and understanding of the coal outlet to determine if devolatilization is occurring. If devolatilization occurs during this sulfur removal step, the system needs to re-evaluate where to reintroduce the devolatilized hydrocarbons into the system, quantify the impact on gasifier performance, and track cascading effects. A very detailed hazard and operability analysis should occur before undertaking this pilot plant activity. However, material specifications of the unit operations and extra flanges and nozzles on the dryer, gasifier, and other equipment should be considered in the pilot application to create options for conduction such work.

Piloting different coal feedstocks must involve a rigorous management of change process that assures the safety, health, and environmental impacts of feed switching are addressed and subsequently validated in pilot plant operations. For instances, if the feedstock change under consideration was from Illinois #6 to Powder River Basin coal, the level of drying would need to be changed to avoid the potential of creating an explosive dust. Additionally, the impact of the resultant change in the coal moisture level on gasifier operations would need to be assessed and validate in pilot plant operations.

5.3.2.2 Potential for Biomass-Coal Co-feed (Additional Feed Flexibility)

An additional opportunity for added flexibility to be examined during the pilot plant phase is the ability to utilize biomass as an alternative or supplemental feedstock for the gasifier. SES U-Gas advertises the ability to utilize biomass and, in fact, there have been multiple demonstrations of the U-Gas design fueled by biomass.

Aside from the logistical issues of finding a steady and suitable biomass feedstock in large enough quantities to significantly augment the coal-based feedstock, a number of process and operational elements will need to be evaluated, including:

- Moisture content of the biomass feed and the potential need for additional drying capabilities
- Evaluation of co-feed vs. separate feed trains
- Evaluation of need for partial oxidation to control methane generation

While this is an interesting opportunity, more pre-work will be required to sufficiently adapt the flowsheet and validate the operational costs and benefits through modeling and simulation. If these efforts suggest the potential for a net benefit to the system, even if only in niche deployments, development of a rational piloting plan may be justified.

5.3.2.3 Potential for Urea Production

Urea production was not explicitly considered as part of this Coal FIRST pre-FEED study because committing to urea production diverges from the concept of NH_3 as chemical energy storage mechanism and commits the facility to commodity chemical production. Nonetheless, while expansion of this project's scope requires significantly more capital and thus increasing the venture's risk profile, the outputs and byproducts of the plant, as designed, lend themselves to integration with a urea production facility. Combining NH_3 and CO_2 creates ammonium carbamate, which is then dehydrated to create urea. This integration would help provide a natural and stable disposition of the captured CO_2 (other than venting, expensive on-site storage, or requiring mature CO_2 transport infrastructure and markets). Such expansion of the facility would require significant adjustment to the business models and design basis. For example, the plant developers would need to decide if it makes more financial sense to invest in the urea production plant capital expenditures or if identifying an off-taker for the inputs to the urea process is most financially advantageous. If the decision is made to invest in the capital equipment, additional design and heat integration with the existing plant and facilities operations would be required.

5.3.2.4 Potential for Improved Pre-Combustion Capture

Commercially available pre-combustion capture systems are not optimized for the temperatures, pressures, and scales required this polygeneration system. As such, potential exists to develop pre-combustion capture systems targeted to this application that have efficiency and cost improvements relative to adapted conventional pre-combustion capture systems. Incorporation of such systems could improve financial performance via reduced capital expenditures, reduced parasitic load, and simplifying the handling of captured CO_2. The incorporation of such systems into the flowsheet, including an additional heat integration, and their impact on system dynamics (particularly during transitions) and control schemes would greatly benefit from pilot plant studies and validation activities.

5.4 Innovation Opportunities for Ammonia Generation

5.4.1 *Overview*

There is currently a large amount of innovation activity in the area of ammonia generation that align with the goals of the polygeneration concept (e.g., fast ramping, ammonia trains operating at 300 MTPD or less, etc.). This is predominately focused on using renewable energy sources to create hydrogen for ammonia synthesis but includes other potential innovations as well.

Catalyte LLC was chosen for participation in Team AST due to their unique insights from the international and domestic ammonia technology and market landscapes. A major driving force for the intensification of ammonia innovation are the European Union and Japanese climate mitigation efforts, with targets ranging from an 80% to 95% reduction in greenhouse gas (GHG) emission by 2050 when compared to 1990 or 2010 levels.[53] These policy goals have spurred international development for dynamic, low-CO_2, renewable NH_3 production which should provide complementary innovation and technology developments that offer potential advantages to the polygeneration design.

The leading United States-based technology licensor is KBR, while several other players are involved at some level, including: RTI (ammonia partner Casale SA, Switzerland), University of Minnesota, Air Products (US and UK), Praxair (with Linde in Germany), Catalyte, Colorado School of Mines, and Kansas State. Additional entities funded by the ARPA-e REFUEL program, include: West Virginia University, University of Delaware, Giner, Rensselaer Polytechnic Institute, and Wichita State. The forefront of non-USA ammonia engineering, procurement, and construction (EPC) and research and development companies are primarily located in Germany, Netherlands, Denmark and Japan. The European Commission and Japanese government seek to

[53] Jensterle, Miha; Jana Narita, Raffaele Piria, Sascha Samadi, Magdolna Prantner, Kilian Crone, Stefan Siegemund, Sichao Kan, Tomoko Matsumoto, Yoshiaki Shibata and Jill Thesen 2019: The role of clean hydrogen in the future energy systems of Japan and Germany. Berlin: adelphi.

use ammonia as a hydrogen carrier,[54] which drives focus on ammonia generation as an important part of a future Hydrogen Economy. To this extent, the European Commission has funded Horizon 2020 grants on the future of ammonia, [55] as well as other and major players in the ammonia space, such as the Yara plan's solar-and-wind-driven NH_3 facilities.

Various new renewable ammonia technologies are funded by the Japanese National Institute of Advanced Industrial Science and Technology's (AIST) Cross Ministerial Strategic Innovation Promotion Program (SIP) whose developments may translate to more traditional ammonia processes such as those used in this polygeneration platform. AIST's research and development includes a demonstration plant by JGC at 20 kg/day starting in 2019,[56] using new, low-temperature (< 400 ^oC) and low-pressure enabling catalysts. AIST and JGC C&C developed a new Ru/CeO_2 catalyst[57] as a first step, however the demonstration of this catalyst was in a low pressure (5 MPa) synthesis loop.

The burgeoning international research and development landscape in ammonia synthesis provides opportunities for some innovations could be incorporated in future generations of the polygeneration concept. Some new technology developments in the NH_3 generation space are so dramatic that one may no longer be able to refer to ammonia synthesis as strictly a Haber-Bosch process. The high degree of integration between the ammonia train and the other elements of the polygeneration system means that there will be multiple effects to the flowsheet when incorporating any advancements in ammonia synthesis. While modeling and analysis will provide the foundational assessment of such engineering opportunities, pilot plant operations would need to validate the altered operations and transitions.

[54] Path to Hydrogen Competitiveness, A Cost Perspective, 20 January 2020, Hydrogen Council

[55] https://ec.europa.eu/programmes/horizon2020/en/news/commission-invest-€11-billion-new-solutions-societal-challenges-and-drive-innovation-led

[56] https://www.jgc.com/en/news/assets/pdf/20181019e.pdf

[57] https://www.ammoniaenergy.org/wp-content/uploads/2019/08/20191112.0826-AIChE2019_NH3_EnergyJGC_Final.pdf Fujimura, Kai, Fujimoto, Atsui, Nishi, Mochizuki and Nanba.

Table 5-2 below identifies some ammonia synthesis process improvements that are being tracked for future (not the current design basis) manifestations of this polygeneration technology platform. Pertinent details are provided in order to demonstrate the flourishing ammonia innovation landscape. Additional details are also provided below regarding the catalyst (*Section 5.4.2)* and ammonia separation improvements (*Section 5.4.3)* that are being tracked for future inclusion in the technology platform.

Table 5-2. Potential Ammonia Synthesis Improvements

Process Component	Readiness of New Alternative	Area to Improve	Proposed Alternative
Catalyst	<TRL 3 [58] < TRL5 [5] TRL 6 -7 [13]	-High Pressure/temperature	-Low temp Ru-Ba-Cs MOF -Ru/Ba-Ca$(NH_2)_2$ -Ru/CeO_2 -Ru (10%)-Cs/MgO -Chemical looping
NH_3 separation & Absorbent-enhances Haber-Bosch	TRL 5 - 8	-Recycle Heat-up (lowers parasitic load)	-Absorbents [Alkaline Earth Chlorides, such as, $MgCl_2$] -600 kg NH_3/m^3 target -Efficiency > flash drums.
Make up gas compressor	Commercial	-Energy consumption -Spare on-site	-Centrifugal -Lower reactor pressure (catalyst material improvements)
Synthesis Loop Design	< TRL 8	-Massive -180 in 600 MTPD turndown	-Lower pressure and temperature (increase catalyst activity) -At temperature NH_3 absorption
NH_3 Liquid storage	TRL 6 to 8	-Potential fugitive gas	-Borohydrides and Metal halides

5.4.2 *Potential Catalyst Improvements*

Motivating much of the research and development in ammonia synthesis is the notion that, while the thermodynamic equilibrium partial pressure for producing NH_3 is improved at low temperature, the N_2 triple bond is difficult to kinetically cleave at low temperature. Researchers have focused on catalysts that increase NH_3 yield at lower temperature and pressure, with the hope of reducing compression burden.

Catalyst research has shown large increases in rate of reaction with experimental materials, as compared with commercial magnetite.[59] However, rates of reaction must be compared at similar TRL's, since highly active catalysts may quickly deactivate or have significant performance reduction once commercially formulated or stabilized. Ruthenium catalysts have been in use for some time, most notably by KBR, who completed development of a combined magnetite and graphite-supported, Ru/Rb low-pressure ammonia process around 1990.[60] The graphitized-carbon-supported Ru catalyst enables 16% per pass conversion at 90 bar and is 10 to 20 times more active than magnetite. Since 2000, KBR has licensed 35 new grass root ammonia plants,[61] but applying

[58] Catal. Sci. Technol., 2020, 10, 105-112, Ignacio Luz, Sameer Parvathikar, Timothy Bellamy, Kelly Amato, J Carpenter and Marty Lail, MOF-derived nanostructured catalysts for low-temperature NH3 synthesis

[59] 7.5 mmol/h•g (catalyst) at 260 °C and 9 bar in laboratory experimentation compared to commercial values of 2.2 mmol/h•g (catalyst) at 450 °C and 200 bar

[60] CEP, Sept. 2016, J. Richardson and V. Pattabathula

[61] Proceedings form Rotterdam NH3 2019, Summit D. Morris, G. Patel

these systems (or similar systems) across the broad operating window of the polygeneration process requires significant process development work.

Since the ammonia synthesis reaction is exothermic, development of a catalyst that could be used at even lower temperature could provide further benefits beyond the aforementioned process. In 2001, PDIL in India was able to achieve excellent catalyst activity at 100 ^{o}C using a cobalt/ruthenium catalyst, but recent efforts have intensified to balance all the competing parameters of temperature, stability, pressure, manufacturability and capital expenditures. Recently, the Tokyo Institute of Technology studied, published, and patent-applied Ru/Ba-Ca $(NH_2)_2$ and found it is one hundred times more active than that of conventional ruthenium catalysts at < 300 °C and 9 bar,[62,63] and is considered to be TRL 3. Several experimental catalysts are currently being tested, including Ru (10%)-Cs/MgO,[64] which is considered TRL 6-7.[65] These catalyst improvements can lead to significant advantageous changes in the ammonia synthesis train, although (again) careful engineering and targeted pilot plan studies would be required before commercial integration with the polygeneration platform.

5.4.3 *Potential Ammonia Separation Improvements*

Absorbents may be used to capture ammonia after it is produced at elevated temperatures. This is in contrast to the current industry-standard refrigeration methods of separation, resulting in lower parasitic loads. This technique serves important purposes, including improving reactor performance as shown by Cussler[66], since most ammonia catalysts are kinetically NH_3 inhibited and the reaction equilibrium is inherently NH_3 inhibited.

Relevantabsorbents[14] include: magnesium chloride ($MgCl_2$), calcium chloride ($CaCl_2$), strontium chloride ($SrCl_2$), zinc chloride ($ZnCl_2$), and zinc nitrate ($Zn(NO_3)_2$).[67] US20170152149 indicates an improvement in performance by strategic ammonia removal near autogenic (700 K reactor outlet, with 460 K $MgCl_2$ absorber) conditions. These absorbents typically range between TRL 3-5.

Applying the absorbents to ammonia production is anticipated to provide process advantages that may be applied to this polygeneration platform to lower pressure (10 – 30 bar), potentially lower capital expenses, improve operational safety, and replace refrigeration, resulting in lower costs and parasitic loads.[14,68,69] Incorporating these features in the polygeneration platform requires detailed techno-economic analysis, engineering analysis, and pilot plant validation of the altered system dynamics.

[62] Angewandte Chemie International Edition (2018)
[63] EP 2650047A1 (2011), Tokyo Institute of Technology
[64] Angew Chem Int Ed, 57 (10) (2018), pp. 2648-2652
[65] Energy Reviews, Vol 114, October 2019, Kevin H.R.Rouwenhorst, Guido Mul, Sascha R.A. Kersten
[66] Converting Wind Energy to Ammonia at Lower Pressure, Mahdi Malmali, Alon V. McCormick, and E. L. Cussler
[67] US 20170152149 $MgCl_2$ Absorption System
[68] Palys M, McCormick A, Daoutidis P. Design optimization of a distributed Ammonia generation system. NH3 fuel conference. 2017. Minneapolis (MN). Retrieved from. https://nh3fuelassociation.org/2017/10/01/design-optimization-of-a-distributedammonia-generation-system/.
[69] Malmali M, McCormick A, Cussler EL, Prince J, Reese M. Lower pressure ammonia synthesis. NH3 fuel conference. 2017. Minneapolis (MN). Retrieved from. https://nh3fuelassociation.org/2017/10/01/lower-pressure-ammonia-synthesis/.

This short survey of innovation activity in the area of ammonia production establishes the potential technology lift to the polygeneration process that could be provided by external research and development. As such, these developments are being tracked for possible future incorporation to further improve the polygeneration platform.

6. Business Case

The general business philosophy of the polygeneration conceptual design centers on offering multiple potential revenue streams, including (1) commercial electricity available for sale to the grid, (2) salable ancillary services (e.g., capacity markets, frequency stability, voltage regulation, etc.), (3) and NH_3 for commercial delivery. Other potential revenue streams include sale or credits related to captured CO_2 and elemental sulfur by-products. By combining these various revenue streams with the polygeneration platform's emphasis on overall plant flexibility, it is possible to modulate plant operations on a very short time scale to meet emerging market signals and opportunities. This ability to correctly match production to market demand will allow for optimization of plant profitability.

The discussion below is organized into two parts. The first section of the business case focuses on the general viability evaluation of the polygeneration platform. This section documents and establishes general assumptions and parameters used in our performance and costs analysis. Additionally, the section covers, predominately by reference, the general evaluation of the economic viability of the polygeneration technology platform. To complete the basis for future consideration of advancing the polygeneration technology platform, the second section of the business case focuses on the technology's market positions, advantages, and our general competitive strategy. This section covers project specific site characteristics, concerns, and supporting analysis that must be considered for a specific application of the technology.

6.1 General Business Case Analysis

To maximize cross-comparison against existing studies, and to maintain full compliance with the terms of the awarded contract, site characteristics and ambient conditions are defined are presented in Appendix B.

Additional market scenario assumptions that define the business case can be seen Table 6-1.

Table 6-1. Market Scenario Assumptions

Description	Values
Coal Type	Illinois #6
Coal Price, Current $'s per short ton	51.96
Natural Gas Price, Current $'s per million BTU's	4.42
Estimated Renewables Penetration, %	25
CO_2 Constraint and/or Price	90% Pre-Combustion Capture
Ammonia Contract Price, Gulf Coast, Current $'s per ton	$195
Ammonia Retail Price, Current $'s per ton	$551

The Class 4 estimates of capital and operating expenditures (Cost Results Section), performance data (i.e. production and required inputs from the Performance Results Section) were combined with market data to understand the economic viability and competitiveness of the technology platform[70]. First, these inputs were used in conjunction with NETL's Power Systems Financial Model to calculate the first year COE which is presented in further detail, along with discussion of the difficulty of this metric for polygeneration system, in the Cost Results section above.[71]

These first year COE metrics were consistently higher than the $139-167/MWh[72] LCOE for IGCC cases in the 2019 NETL Baseline studies.[73] This indicates that specific applications of this platform will need to consider capacity market (i.e. firm dispatchable support for renewables), ancillary services revenue (not included), sales of elemental sulfur or CO_2, or expanding the facility to produce urea. However, this also indicates the need for electricity markets with more frequent and higher magnitude variations in locational marginal electricity prices (such as those that could occur in a power system with high renewable penetration, lower load inertia, and limited reliable energy storage options) in order to create the peaks in electricity price that will generate capital recovery to pay for the equipment providing the flexible operations capability of the polygeneration platform.

As discussed in the Cost Results section, COE metrics are limited to providing high-level insight for polygeneration concepts. Consequently, in order to support continued sound decisions making as the technology platform progresses in development, demonstration, and deployment AST has the Class 4 estimates of capital and operating expenditures (Cost Results Section), performance data (i.e. production and required inputs from the Performance Results section) and market data as inputs for creating an investment analysis model. These modeling methods are discussed in Appendix H. This infrastructure enables more robust exploration of sensitivities, business models, and will aid in evaluation of specific sites and site-specific concerns. Modeling efforts indicate that the anticipated price volatility in electricity markets with higher renewable penetration are most likely required to improve the plant's financial performance[74] (i.e. this technology fulfills an anticipated 2050 market need versus a 2020 market need). Additionally, current electricity and ammonia market demand lead to less movement through the operating window than expected, which may lead to site-by-site consideration of 'how much flexibility' to invest capital in (i.e. 2 combustion turbines instead of 3).

[70] Please see Appendix H for further discussion of the Investment Model

[71] Please refer to the Cost Results (Section 4) section of this report for further information on assumptions and methodology related to calculating the COE metric

[72] This COE estimate represents the price required to achieve a zero net present value (NPV) based on the capital and operating expenditures, revenue, and financing assumptions just described at the required internal rate of return on equity of 10% (i.e. this is not a "profitless" or "breakeven price" but a price that yields a 10% return). These results are on a 2018-dollar basis assuming a 30-year operational life. A 2018-dollar basis was chosen to facilitate crosschecking and benchmarking with relevant NETL systems analysis studies.

[73] While it is understood that "first-year COE" and "LCOE" are not the same metric, this comparison has been presented only to provide generally illustrate that the required cost of electricity of the polygeneration concept is higher than what is required in the IGCC cases presented in the Baseline report under current market conditions.

[74] This is based on the general results for a variety of scenarios based on current electricity market dynamics and preliminary attempts to evaluate higher market volatility (high frequency and higher magnitude of price swings) due to renewable electricity penetration. However, these initial explorations are only preliminary given the difficulty in accurately representing future electricity markets—but were sufficient to form this hypothesis.

6.2 Market Positioning, Advantages and Competitive Strategy

As discussed in the Concept Background, the philosophy behind the polygeneration plant is based on the premise that by being able to flexibly shift between production of two relatively unrelated product streams with a single feedstock stream, the plant will have the advantage of being able to both plan production based on the most economically beneficial product (such as seasonal demand increases for ammonia) and quickly respond to market changes (such as weather events that disrupt renewable generation in the market). From a business model and market participation perspective, the project concept is well insulated from external forces, such as supply disruption, buyer power, competition and new entrants.

Supply: The plant is designed to operate on Illinois #6 coal feedstock as per NETL guidelines for project development, however in execution could operate on similar coal feedstocks with additional design and engineering which is not anticipated to add significant capital cost relative to the total plant cost. Thus, while the plant is in theory vulnerable to shocks in the coal supply chain, this risk is no greater than any production facility that relies on a commodity coal feedstock. Additionally, should the developer of the facility choose, the risk of being beholden to a specific coal feedstock can be mitigated through additional design and engineering, but it would be up to each developer’s risk appetite as to if this cost / benefit tradeoff would fit within their risk profile. The plant will also have the capability to run the power island on natural gas, however it is not currently anticipated to operate in this manner and thus the price fluctuations of natural gas would have minimal impact on the plant’s financial performance.

Buyers: The facility’s outputs, electricity and ammonia, are both essentially commodity products that are sold into existing wholesale and retail markets. Typically, this means that a power plant or an ammonia plant would be a price-taker from the market, assuming their supply is not sufficient to impact the supply / demand price dynamics of a market. However, in this case our facility has more options than to simply be a price taker, as the facility has the ability to shift to higher electricity output when electricity prices are elevated, can shift to ammonia production when ammonia prices are higher, or, should both markets be unattractive, can produce ammonia for storage and sale at a later date. This enables the facility to operate at a relatively steady state thus behaving more like a process plant or a baseload power plant and thus taking advantage of operating at the most efficient operating points. Additionally, this flexibility removes some of the typical buyer power that is held over standalone power or ammonia plants. In the future, use of ammonia as a fuel is another energy market that may change current ammonia market dynamics, increasing the demand for ammonia; this future market shift was part of the motivation of the project team to investigate ammonia as the chemical storage medium for this polygeneration concept.

Competition: While the polygeneration nature of the plant by default causes it to compete with twice the normal number of other firms (i.e., firms in both the power generation industry and ammonia production industry), the ability to flexibly navigate between these two markets enables the plant to essentially choose its competition at any given time. Additionally, the fact that both electricity and ammonia are essentially commodity markets means that the competition from substitute products is generally low. While energy storage could eventually compete, or ammonia-use as a fuel competes as against a wide range of fuels], substitution for these products in the market is generally difficult since the current bulk power system has achieved dominant design status and there is no direct, trust substitute for ammonia disposition as a chemical. Additionally, while competition could come from increasing renewables penetration, it is also expected that price volatility that could occur based on the variability of renewable generation will produce the swings in locational marginal pricing of electricity that the plant is well suited to capitalize on based on its flexible production nature. In fact, it is this high renewable penetration that motivated the exploration of this polygeneration platform and the high capital investments required to achieve flexible operations.

New Entrants: The main barrier to entry for new entrants will predominantly be the scale of capital investment required to develop a competing facility. Similarly, once a plant is developed and operating, it is unlikely that a second plant would be developed in the same close geographical area, as the competing facilities would be selling commodity products into the same commodity market. Should this happen, the local supply of each commodity would increase, thus driving the unit price down given a stable level of demand, thus reducing the profitability of both plants. In the development stage, the second facility would model these lower commodity prices and determine their facility to provide a low internal rate of return on investment. In some industries this strategy would still be engaged, in order to drive the first facility out of business – similar to how large businesses have subsidized their footprint expansion into unprofitable areas in order to reduce competition, only to later increase prices. In this case the investment - $1.2 billion – would serve as a barrier to that strategy.

Our investment model (Cf. Appendix H) was developed to enable intelligent evaluation of locations where a local market can be developed and dominated. The main facility risk is potential disruption by ammonia production from natural gas-fed mega-scale ammonia plants, should infrastructure be developed lowering their cost to serve the markets that we identify as location advantaged (where our plant has lower distribution and sales costs due to closer proximity to the end user). This disruption would lower the margins achievable in ammonia sales, while participation in the electricity market would remain under the same circumstances. The general mechanics of adding thermal generation or transmission and distribution infrastructure to the bulk electric power system is anticipated to protect the market characteristics of the electricity market where the plant is built. The addition of local renewables near our facility while on average, most likely, reducing the demand for our electricity production may also lead to the spikes in locational marginal pricing volatility and firm dispatchable capacity payments that make this polygeneration platform valuable.

Markets: Domestic agricultural-driven markets for ammonia and ammonia derivatives are established and are well-behaved and predictable. These characteristics are beneficial for developers of projects that require large capital investments such as this project. Additionally, international markets with less abundant supplies of natural gas still grow their commodity chemical base from coal feedstocks (cf. China) such that our technology platform integrates well into the growing agricultural and electricity needs of the developing world.

Our current economic assessments suggest that electricity markets with high frequency price spikes are needed for the flexibility capability of this platform to be warranted—higher variation than is currently seen in the US market is more beneficial. This variation is anticipated to occur as renewable penetration increases or in international applications where their power grid and markets are less reliable and evolving.

The addressable market of the polygeneration platform increases if efforts to utilize ammonia as a fuel (including as a hydrogen storage medium) cross the tipping point to broader application either domestically or internationally. Specifically, ammonia fuel options open up a wider range of electricity generation options to support emerging markets, or other areas such as Department of Defense installation and forward area energy resilience goals. These evolutions are very hard to project given the nascent state of ammonia as a fuel options and markets. Additionally, such potential markets are competing with other options such as the evolving methanol economy.

In our modeling (please see Appendix H for additional details on the investment model) we accounted for a number of key factors influencing the commercial and economic viability of a coal-based power plant with co-generation of ammonia product for export, including:

1. Electricity need and demand
2. Access to reliable coal supply
3. Ammonia need and demand
4. Ability to export ammonia product to relevant customer segments (e.g., adjacency to ammonia end-users, ammonia production sites, or ammonia pipelines)

While determining an exact location that offers the best combination of these primary specific siting factors is a complex activity beyond the scope of this report, it is important to provide high-level thoughts and analysis related to these various factors. This is a similar thought process that a project developer will go through, albeit in greater detail, to identify a specific parcel of land for project development. Additionally, the business model is neutral to the price of CO_2; our investment model has taken into account both CO_2 credit and CO_2 sales, but does not rely on these streams of income as the CO_2 markets are relatively difficult to predict. For metrics on COE and financial viability please refer to Section 4, Cost Results, and to Appendix H.

6.2.1 *Electricity Need and Demand*

To determine the most advantages conditions for electricity need/demand, information has been pulled from PJM's publicly available data repositories regarding hourly prices for electricity and ancillary services over the past year, both as an average across the RTO as well as at individual nodes. Of particular interest is identification of nodes that have a high average price premium relative to the RTO average, which can indicate an unmet demand.

Within this subgroup of nodes with a high relative price premium, it is desirable to identify nodes that occupy either the low or high ends of price premium variability, as they offer two potential advantages to the polygeneration concept. A high average price premium with variability should indicate a node that has a consistent need for cheaper generation alternatives. This type of location should allow the polygeneration plant to consistently operate at high net electricity generation and sell power to the grid at consistently high prices.

While these characteristics would often been seen as ideal for a more traditional, base-load power plant, there is concern that it might not be ideally suited to maximize the benefits of the current design definition. Specifically, there are a number of capital-intensive design characteristics incorporated into the defined design concept to enable significant flexibility in plant operational flexibility. By deploying to a location with a relatively steady, high price premium for electricity, the benefits and competitive advantages of the defined polygeneration plant may not be fully captured.

To serve as a point of comparison to this scenario, it is also important to consider locations with a high price premium and high premium variability. It is expected that these characteristics would indicate a location that has a high willingness to pay a high price premium on average as well as a need for a flexible plant that can respond efficiently to highly variable price premiums. As superior performance in this type of scenario (e.g., need for increased responsiveness, operational flexibility, and broad, efficient operating window to meet high variability in grid demand) was one of the key targets that the plant was designed to address, it is logical to assume that the plant will see significant competitive advantages in this type of deployment.

6.2.2 *Access to a Reliable Coal Supply*

As the production and delivery of coal within the Midwestern US is fairly mature, it is believed that this should not be an overly limiting factor in site selection, assuming that the general characteristics detailed in Appendix B are met (e.g., sufficient available land, sufficient rail access, etc.)

6.2.3 *Ammonia Need and Demand*

One inherent advantage with virtually any site selected in the Midwestern US is the fact that inland production of NH_3 can sell at a price premium over ammonia delivered to ports locations, typically situated on the US Gulf Coast (USGC), as it avoids many of the expensive costs associated with long overland transportation. It is this reality that allows relatively smaller scale ammonia cogeneration plants to compete with much larger operations while still maintaining the ability to charge prices that are closer to retail values.

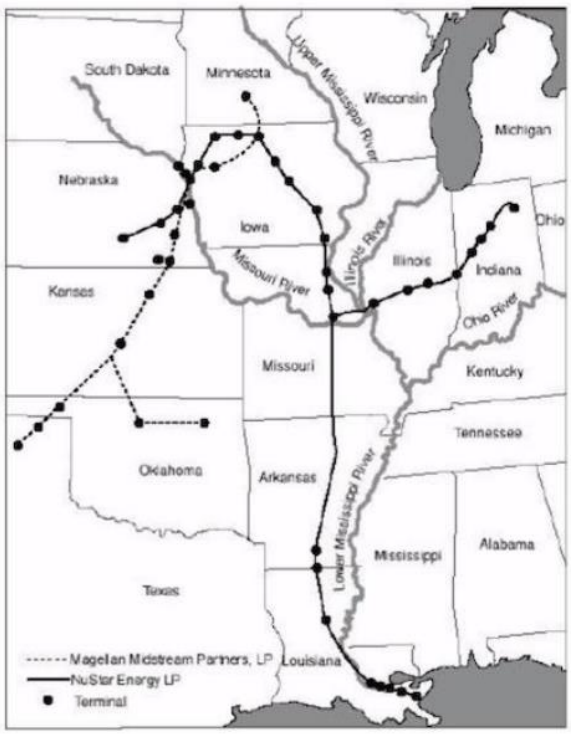
An additional inland market of particular relevance to this study is the US Corn Belt as this market offers opportunities to sell ammonia at significant price premium over USGC sales. While this price premium typically is ~$150/MT, it can go as high as ~$300/MT.[75] The ability to sell ammonia at this Corn Belt price premium is an effective tool to help maximize plant revenue and profit even when operating at modes that exhibit a reduction in net power export.

[75] The November market forecast from Farm Futures (https://www.farmprogress.com/story-weekly-fertilizer-review-0-30765) includes the following passage: *"Terminal prices edged higher following a $5 boost in settlements for November that took the Gulf price to $236. That followed a $27 increase for October, based on ideas farmers will plant more corn in 2020 – a notion USDA's first baseline forecast supported by forecasting 94.5 million acres. While that wasn't far off the 94.1 million our first survey of planting intentions found, we talked to farmers in late July and early August, when the ration of new crop soybean to corn futures favored corn. That benchmark has since turned in favor of beans. Our average retail cost for ammonia was unchanged last week at $472, only $15 off the forecast based on wholesale prices, though offers vary widely, running anywhere from $415 or less on the southern Plains to $555 or more in parts of the Corn Belt."*

6.2.4 *Ability to Export Ammonia Product to Relevant Customer Segments*

While not as extensive as the coal delivery and transport resources (essentially just available rail line), the existence of existing transportation assets such as the Magellan and Kaneb pipelines help to facilitate movement from the Midwestern generation site of the proposed polygeneration plant to other Midwestern locations and, more importantly for reasons detailed above, relevant Corn Belt locations. A map of these pipelines can be seen below Figure 6-1 with high-level details of relevant existing ammonia transportation and storage assets presented in Table 6-2.

Figure 6-1. Relevant Pipeline Asset Map[76]



[76] Graphic is from the following Department of Energy report https://www.energy.gov/sites/prod/files/2015/01/f19/fcto_nh3_h2_storage_white_paper_2006.pdf Accessed on April 17, 2020

Table 6-2 Existing Ammonia Transportation and Storage Asset Summary

Existing Ammonia Assets	Regional Coverage
Magellan Ammonia Pipeline	• 1,100 miles • 20 Terminals • 528,000 tons Storage of ammonia • Services Texas to Minnesota • Delivery Capacity: 900,000 MT/year
Kaneb Pipeline	• 2,000 miles • 24 Terminals • 1 Million tons Storage • Services Louisiana to Nebraska & Indiana • Delivery Capacity: 2 Million
Pipeline Terminals	• 44 Terminals • 2.9 Million Tons of Capacity
River Storage Terminals	• 30 Terminals • 780,000 Tons • Services Mississippi, Illinois & Ohio Rivers
USA Production Points	• 23 Plants • 767,000 Tons of Available Storage
Storage Terminals (>1000 Tons)	• 1,500,000 Tons of Available Storage
Total Storage	• 4,575,000 Tons of Available Storage

Finally, there is a growing market for both ammonia as a fuel and for “GHG Free” or “Green” ammonia. Ammonia is becoming more widely viewed as a transport mechanism for energy, as it is stable to transport via pipeline, rail or truck, and can be converted to H_2 for use as a direct fuel source. Additionally, by using electricity from renewable generators to create ammonia, a GHG

Free ammonia market is growing internationally, expected to be approximately 325 million mt/year by 2030 by the International Renewable Energy Agency[77]. While it is unclear whether the current project's 90% carbon capture would qualify as "GHG Free," what is clear is there is a growing worldwide market for ammonia as a commodity, for expanding and new use cases.

[77] Hydrogen A Renewable Energy Perspective, IRENA, Report prepared for the 2nd Hydrogen Energy Ministerial Meeting in Tokyo, Japan, September 2019

Appendix A. Coal Feed Design Characteristics

The characteristics of the Illinois #6 design coal are as follows:

Exhibit A-1 Design Coal - Bituminous (Illinois No. 6, Herrin)

Rank	**Bituminous**	
Seam	**Illinois No. 6 (Herrin)**	
Source	**Old Ben Mine**	
Proximate Analysis (weight %)[A]		
	As Received	**Dry**
Moisture	11.12	0.00
Ash	9.70	10.91
Volatile Matter	34.99	39.37
Fixed Carbon	44.19	49.72
Total	100.00	100.00
Sulfur	2.51	2.82
HHV, kJ/kg (Btu/lb)	27,113 (11,666)	30,506 (13,126)
LHV, kJ/kg (Btu/lb)	26,151 (11,252)	29,544 (12,712)
Ultimate Analysis (weight %)		
	As Received	**Dry**
Moisture	11.12	0.00
Carbon	63.75	71.72
Hydrogen	4.50	5.06
Nitrogen	1.25	1.41
Chlorine	0.29	0.33
Sulfur	2.51	2.82
Ash	9.70	10.91
Oxygen[B]	6.88	7.75
Total	100.00	100.00

Appendix B. Site Design Characteristics

B-1 General Site Characteristics

To maximize cross-comparison against existing studies, and to maintain full compliance with the terms of the awarded contract, site characteristics and ambient conditions are defined as follows:

Exhibit B-1.1 Site Characteristics

Parameter	Value
Location	Greenfield, Midwestern USA
Topography	Level
Size, Acres	300
Transportation	Rail or Highway
Ash Disposal	Off Site
Water	Municipal (50%) / Groundwater (50%)

Exhibit B-1.2 Site Ambient Conditions

Parameter	Values
Elevation, m, (ft)	0, (0)
Barometric Pressure, MPa, (psia)	0.101 (14.696)
Design Ambient Temperature, Dry Bulb, °C, (°F)	15 (59)
Design Ambient Temperature, Wet Bulb, °C, (°F)	10.8 (51.5)
Design Ambient Relative Humidity, %	60
Cooling Water Temperature, °C, (°F)^	15.6 (60)
Air composition based on published psychrometric data, mass %	
N_2	72.429
O_2	25.352
Ar	1.761
H_2O	0.382
CO_2	0.076
Total	100.00

^The cooling water temperature is the cooling tower water exit temperature.
This is set to 8.5°F above ambient wet bulb conditions in ISO cases.

As assumed for gasification-based cases in the NETL baseline studies, the required land area is estimated as 30 acres for the plant proper with the balance providing a buffer of approximately 0.25 miles between the plant and the fence line. While this land area estimation is generous for this distributed small-scale concept, the 'extra land' provides for a potential rail loop, product storage and distribution, and a greenspace barrier between the facility and the surrounding community.

In all cases, it was assumed that the steam turbine is enclosed in a turbine building. The gasifiers, reformers, ammonia synthesis reactors, and the combustion turbines are not enclosed.

Allowances for normal conditions and construction are included in the cost estimates. The following design parameters are considered site-specific and are not quantified for this study. Costs associated with the site-specific parameters can have significant impact on capital cost estimates.

- Flood plain considerations
- Existing soil/site conditions
- Water discharges and reuse
- Rainfall/snowfall criteria
- Seismic design
- Buildings/enclosures
- Local code height requirements
- Noise regulations – Impact on site and surrounding area

Appendix C. Major Equipment List

Exhibit C-1: Equipment Schedule

		PLANT AREAS						REMARKS
01		Air Separation Unit						
02		Coal Handling and Crushing						
03		Gasifier, HRSG & Quench						
04		Water Gas Shift						
05		Syngas Cooling						
06		Syngas Clean Up						
07		Ammonia Production						
08		Fuel Gas Conditioning						
09		Power Generation						
10		Utilities						

Exhibit C-2: Compressors[78]

ITEM NO.	No OFF	DESCRIPTION	TYPE	FLOW Am³/h	FLUID & Density kg/m³	PRESSURE SUCTION BAR(G)	PRESSURE DISCHARGE BAR(G)	ABSORBED POWER kW	MATERIAL C = Casing	MATERIAL I = Internals	Remarks
K-810	1	GT Feed Compressor	C	3688.0	Syngas 11.12	31.7	44.0	1380			
PK-650	1	CO2 Compressor Package (Booster)	C	25186.0	CO2 2.22	0.2	5.0	10630 (total)			Total Stages: 7 total (est) Includes Intercoolers (40°C) / Aftercooler (35°C)
		CO2 Compressor Package (Main Compressor)	C	16685.0	CO2 8.77	4.8	89.9				

[78] Code: A – Axial; C – Centrifugal; M – Metering; R – Reciprocating; S – Screw
All drives are electric motors unless specified otherwise.

Exhibit C-3: Heat Exchangers

ITEM NO.	No OFF	DESCRIPTION	TYPE	DUTY (kW)	DIMENSIONS DIAMETER m	DIMENSIONS LENGTH m	SURFACE AREA m^2	FLUID T = Tube S = Shell		MATERIAL	DESIGN CONDITIONS PRES BAR(G)	DESIGN CONDITIONS TEMP °C	Remarks
E-301	1	Oxygen Heater	S&T	1554	0.51	6.1	128	T	LP Steam	304 SS	6	182	BEU assumed
								S	Oxygen	304 SS	50	178	
E-302	1	Scrubber Blowdown Air Cooler	Air Cooler	449	N/A	N/A	26 (Bare)	T	Process Water	CS	41	209	
E-401	2	Shift Interchanger	S&T	9474 total	1.50 per shell	6.10 per shell	2779 total	T	Syngas	304 SS	42	332	AES assumed. HOLD may consider plate and frame
								S	Syngas	304 SS	42	318	
E-402	1	IP Boiler	S&T	23744	0.64	6.1	216	T	Syngas	304 SS	51	460	BEU assumed
								S	IP BFW / Steam	CS/3mm	51	287	
E-403	1	IP BFW Heater	S&T	8194	1.13	6.1	728	T	IP BFW	CS	68	258	AES assumed
								S	Syngas	304 SS	39	275	
E-411	1	Shift Start-Up Heater	S&T	663	0.16	6.1	10	T	HP Steam	CS/3mm	76	314	BEU assumed
								S	Syngas	304 SS	41	338	
E-501	1	Desaturator Air Cooler	Air Cooler	297	N/A	N/A	36 (Bare)	T	Process Water	304 SS	48	113	
E-502	1	Desaturator Water Cooler	S&T	6389	0.84	6.1	383	T	Cooling Water	CS	6	65	BEU assumed
								S	Process Water	304 SS	48	93	
E-510	1	IP BFW Heater	S&T	8194	0.88	6.1	424	T	Process Water	304 SS	50	222	BEU assumed
								S	IP BFW	CS	69	211	
E-511	1	LP Boiler	S&T	36471	1.55	6.1	1498	T	Process Water	304 SS	49	218	BEU assumed
								S	LP BFW / Steam	CS/3mm	10	181	
E-512	1	LP BFW Heater	S&T	10730	1.37	6.1	1114	T	LP BFW	CS	13	178	AES assumed
								S	Process Water	304 SS	49	191	
E-660	0	CO2 Chiller (Optional)						T	S. Critical CO2	SS			Not to be costed
								S	Refrigerant	SS			

Exhibit C-4: Heat Exchangers (continued)

ITEM NO.	No OFF	DESCRIPTION	TYPE	DUTY (kW)	DIMENSIONS DIAMETER m	LENGTH m	SURFACE AREA m^2	FLUID T = Tube S = Shell		MATERIAL	DESIGN CONDITIONS PRES BAR(G)	TEMP °C	Remarks
E-801	2	GT Feed Preheater	S&T	1179	0.38 per shell	6.1 per shell	68 per shell	T	Process Water	304 SS	50	191	BEU assumed
								S	Fuel Gas	CS/3mm	50	149	
E-802	1	GT Feed Compressor Spillback Cooler	S&T	898	0.37	6.1	62	T	Cooling Water	CS	6	65	BEU assumed
								S	Fuel Gas	CS/3mm	50	103	
E-901	1	Steam Turbine Condenser	S&T	94622		9.40	3600	T	Cooling Water	304 SS			Part of HRSG
								S	Steam / Cond	CS/3mm			
E-902	2	DMW Preheater	S&T	20467	0.83 per shell	6.1 per shell	379 per shell	T	Process Water	304 SS	50.0	149.0	BEU assumed
								S	DMW / Condensate	CS/3mm	10.0	112.0	
E-903	1	LLPS Boiler	S&T	2959	0.40	6.1	74	T	Process Water	304 SS	50.0	177.0	BEU assumed
								S	LLP BFW / Steam	CS/3mm	10.0	131.0	
E-911	1	Steam Turbine Condenser 2	S&T	48193		9.40	1800	T	Cooling Water	304 SS			Part of HRSG
								S	Steam / Cond	CS/3mm			
								T					
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Exhibit C-5: Pumps[79]

ITEM NO.	No OFF	DESCRIPTION	TYPE	FLOW m^3/h	FLUID & Density (kg/m3)	PRESSURE SUCTION BAR(G)	PRESSURE DISCHARGE BAR(G)	ABSORBED POWER kW	MATERIAL C = Casing I = Internals		Remarks
P-301	2	Scrubber Blowdown Pump	C	10.2	Water 955.9	0.6	7.2	4.1	C I	CS/3mm 12% Cr	
P-501	2	Desaturator Circulation Pump	C	1561.0	Water 809.8	33.4	46.4	726.3	C I	304 SS 12% Cr	
P-650	2	CO2 Condensate Pump	C	0.3	Water 997.0	4.0	37.2	0.37	C I	304 SS 12% Cr	
P-660	2	CO2 Pump	C	303.2	Supercrit CO2 530.6	89.6	148.2	683.2	C I	304 SS 12% Cr	
P-901	2	Steam Turbine Condensate Pump	C	314.4	Water 993.0	-0.6	4.0	55.7	C I	CS/3mm	Part of HRSG
P-902	9	IP BFW Pump 1	C						C I		Part of HRSG 2 operating / 1 stand-by per HRSG
P-903	9	HP BFW Pump	C						C I		Part of HRSG 2 operating / 1 stand-by per HRSG
P-905	6	GT BFW Pump	C						C I		Part of HRSG 1 operating / 1 stand-by per HRSG
P-911	2	Steam Turbine Condensate Pump 2	C	160.1	Water 993.0	-0.6	4.0	29.3	C I	CS/3mm	Part of HRSG
P-1001	2	LP BFW Pump	C	235.9	Condensate 954.6	0.2	10.5	94.8	C I	CS/3mm	
P-1002	2	IP BFW Pump 2	C	169.1	Condensate 920.2	8.0	63.4	374.0	C I	CS/3mm	
P-1003	2	LP Condensate Pump	C	5.4	Condensate 916.0	1.0	7.8	3.07	C I	CS/3mm	

[79] Code: C – Centrifugal; D – Diaphragm; M - Metering

Exhibit C-6: Pressure Vessels

ITEM NO.	No OFF	DESCRIPTION	ORI.	SECTION	DIMENSIONS DIAMETER m	DIMENSIONS LENGTH m	DESIGN CONDITION PRES BAR(G)	DESIGN CONDITION TEMP °C	MATERIAL SHELL	MATERIAL INTERNALS	Remarks
R-401	1	Shift Reactor 1	V		3.4	7.30	40.0	475	CS SS Lined		Catalyst Bed Volume = 46 m^3
R-402	1	Shift Reactor 2	V		3.5	7.30	40.0	330	CS SS Lined		Catalyst Bed Volume = 48 m^3
T-501	1	Desaturator	V	TOP BOTTOM	3.14 4.69	13.1 15.2	45.0 47.0	200 215	CS 3mm SS Lined	SS Packing	Stainless Steel Packing Total T/T = 31.5 m including transition height
V-301	1	Scrubber Blowdown Separator	V		1.10	3.00	3.5	180	CS 3 mm CA		
V-501	1	Mercury Guard Bed	V		By Vendor		45.0	200	CS 3mm SS Lined	SS	Flow: 6713 m^3/h, 171,800 kg/h; Inlet Hg : 55 ppbm; Mercury Removal: 95%
V-901	1	Deaerator	H		By Vendor						Flow: 467,100 kg/h
V-902	1	LLP Steam Drum	H								Included in HRSG Package
V-903	1	IP Steam Drum	H								Included in HRSG Package
V-904	1	HP Steam Drum	H								Included in HRSG Package
V-1001	1	IP Steam Drum (Shift)	H		2.1	8.40	51.0	287	CS 3 mm CA		
V-1002	1	Continuous IP Boiler Blowdown Drum	V		1.00	2.60	7.5	180	CS 3 mm CA		
V-1003	1	Intermittent IP Boiler Blowdown Drum	V		1.00	2.60	7.5	180	CS 3 mm CA		
V-1005	1	LP Steam Drum	H		2.0	6.00	6.0	182	CS 3 mm CA		

Exhibit C-7: Packaged Equipment

ITEM NO.	No OFF	DESCRIPTION	DUTY SPECIFICATION	REMARKS
PK-101	1	Air Separation Unit	Air Flow: 210,000 kg/h @ 0 barg; O_2 Flow: 50,000 kg/h; O_2 Press: 44 barg; O_2 Purity 99.5%; Nitirogen Product: 44 barg; Power Cons (est): 20.62 MW	Includes N_2 (NH_3 / Diluent) Compessor
PK-201	1	Coal Handling and Crushing Package	Delivery 100 x 100 ton Trains, 8 cm x 0; Initial Crushing to Storage Silo: (3 cm x 0); Final Crushing 70,900 kg/h, Moisture 11.12 wt%, Size 1/8" x 0	Based on Illinois #6 coal Refer to Process Description for scope.
PK-301	2	SES U-Gas Gasification & HRSG	Raw Syngas Output (each identical train): 85,900 kg/h ea., 4525 m³/h ea.; Outlet Pressure: 35 barg;	SES U-Gas Process Includes Lock Hoppers, HRSG, & Scubber
PK-501	1	Sour Water Stripper Package	Total Feed Flow: 106,400 kg/h; MOC: 304 SS, Pump Impeller 12% Cr	
PK-601	1	Acid Gas Removal Unit (SELEXOL)	Feed Flow Rate: 171,500 kg/h; Volumetric Flow (actual): 6713 m³/h; Inlet Pressure: 33 barg; Inlet Temperature; 40 °C	Selexol Process
PK-602	1	Sulfur Recovery Unit (Super Claus)	Acid Gas Flow: 9,600 kg/h; Recoverd Sulfur: 1,800 kg/h	Super Claus Process
PK-603	1	Tail Gas Treatment Unit (SCOT)	Feed Flow Rate: 7,800 kg/h;	
PK-604	1	PSA Unit	Feed Flow Rate: 11,700 kg/h; Volumetric Flow (actual): 2213 m³/h; Inlet Pressure: 32 barg; Hydgrogen Flow: 4400 kg/h (86% Recovery)	
PK-651	1	CO_2 Drying Package	Total Gas Flowrate (wet): 146,500 kg/h; Dried Gas Flowrate: 146,200 kg/h; Pressure: 4 barg; Temperature: 21.6 °C	
PK-701	2	Ammonia Loop & MUG	300 MTPD Ammonia Production/train; Power Cons (est): 4.07 MW/train	Includes Makeup Gas Compressor
PK-901	3	HRSG	HRSG Duty: 46,440 kW; Duct Burner Duty HHV (norm/max): 14,000/41,500 kW; Pressure Levels: 3; HP Steam Flow: 79,000 kg/h @ 65 barg & 500°C	Duty Specifications are per HRSG
PK-902	3	SCR Package	Feed Rate: 330,000 kg/h ea; SOx Inlet 2 ppmvd; NOx Inlet: 25 ppmvd; NOx Outlet: 5 ppmvd	Integrated with HRSGs

Exhibit C-8: Miscellaneous Equipment

ITEM NO.	No OFF	DESCRIPTION	DUTY SPECIFICATION	REMARKS
U-201	1	Fluid Bed Dryer Package	Flowrate: 62,984 kg/h (dry); Size 1/8" x 0; Moisture in/out: 11.12 wt%/ 5 wt%; Nitrogen Flow (est): 30,000 kg/h; N2 Temperature 140 °C; MOC: 316H; Dryer Size: 2.5 m ID x 9.0 m T/T (Preliminary); DP: 50 barg; DT: 700 °C Internal Tube Bundle; Dryer Duty (est): 3.2 MW	Includes associated auxiliary equipment, Nitrogen Heater, vent condenser, water and particulate removal, recycle blower, booster compressor (suplemental info below)
U-201-K1	1	Drier Vent Booster Compressor	Centrif. Compressor - Flowrate: 22600 am³/h; Suction Pressure: 0.1 barg; Discharge Pressure: 32 barg; Absorbed Power: 3.8 MW; Stages: 4; Intercooling and aftercooling Duty (Total) 3800 kW	Part of U-201
U-201-E1	1	Drier Vent Condenser	Type: S&T HX; Duty: 2900 kW; Area: 193 m²; Tube Side: DP: 6 barg, DT: 65°C, MOC: SS; Shell Side: DP: 6 barg, DT: 150 °C, MOC: SS	Part of U-201
U-201-E2	1	Nitrogen Heater	Type: S&T HX; Duty: 800 kW; Area: 44 m²; Tube Side: DP: 6 barg, DT: 155°C, MOC: SS; Shell Side: DP: 6 barg, DT: 170 °C, MOC: SS	Part of U-201
U-201-V1	1	Drier Vent Condensate Drum	Type: Vertical Pressure Vessel; MOC: SS; Diameter: 2.4 m; T/T: 2.4 m; Design Pressure 6 bar g; Design Temperature: 150 °C	Part of U-201
U-201-K2	1	Drier Recycle Blower	Centrif. Compressor - Flowrate: 19800 am³/h; Suction Pressure: 0.1 barg; Discharge Pressure: 1.4 barg; Absorbed Power: 0.7 MW; Stages: 1;	Part of U-201
U-201-U1	1	Dryer Vent Particulate Removal	Flow (est): 35,000 kg/h; Temperature 75 °C; MOC: 316H; Pressure 0.4 barg	Part of U-201
U-901	3	Stack	Flowrate: 330,000 kg/h ea; 105 m³/s; Temperature: 105 °C; MW: 27; Height: 22.9 m; Diameter: 2.7 m;	
U-1001	1	IP Steam Desuperheater	Steam Inlet: 72.5 Mt/h @ 413°C & 39.26 barg; BFW: 60 barg & 228 °C Steam Outlet: 79 Mt/h @ 336°C & 39 barg	
K-901	3	Gas Turbine	Output: 30.18 MW ea / 90.53 MW Total	Model: GE
K-902	1	Steam Turbine	Output: 47 MW; HP Steam Flow: 155,500 kg/h @ 63 barg / 500 °C	
K-912	1	Steam Turbine 2	Output: 25 MW; HP Steam Flow: 79,200 kg/h @ 63 barg / 500 °C	

Appendix D. Stream Tables for Alternative Operating Points

D-1 Balanced Generation, 2 GTs

Exhibit D-1: Balanced Generation, 2 GTs Stream Table/Heat and Mass Balance

STREAM NUMBER		1		2		3		4		5		6	
STREAM NAME		AR coal feed		Dried Coal Feed		Scrubbed Syngas		Net Steam from Gasifier		Steam to Shift 1		Steam Raised in Shift	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	0.00	0.00	2552.07	40.23	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	28.013	0.00	0.00	0.00	0.00	24.50	0.39	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Monoxide	28.010	0.00	0.00	0.00	0.00	2596.96	40.94	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	44.010	0.00	0.00	0.00	0.00	998.24	15.74	0.00	0.00	0.00	0.00	0.00	0.00
Methane	16.042	0.00	0.00	0.00	0.00	91.95	1.45	0.00	0.00	0.00	0.00	0.00	0.00
Argon	39.948	0.00	0.00	0.00	0.00	7.67	0.12	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Sulfide	34.082	0.00	0.00	0.00	0.00	55.57	0.88	0.00	0.00	0.00	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	16.46	0.26	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h	62984		62984									
HHV / LHV (MW)		533.72	514.77	533.72	515.10	429.50	396.06	0.00	0.00	0.00	0.00	0.00	0.00
Temperature	°C	15.00		75.00		178.32		398.89		300.00		258.79	
Pressure	bara	1.01		1.01		36.35		41.00		41.00		46.00	

STREAM NUMBER		1		2		3		4		5		6	
STREAM NAME		AR coal feed		Dried Coal Feed		Scrubbed Syngas		Net Steam from Gasifier		Steam to Shift 1		Steam Raised in Shift	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Total Dry Molar Flow (kg.mol/h)			0.00		0.00	6343.43	100.00	0.00	0.00	0.00	0.00	0.00	0.00
Water	kg.mol/h	437.42		184.01		2511.80		2074.21		7003.26		2615.78	
Total Wet (kg.mol/h)		437.42		184.01		8855.23		2074.21		7003.26		2615.78	
Total Mass Flow (kg/h)			70,900		66,300		171,700		37,400		126,200		47,100
Molecular Weight							19.39		18.02		18.02		18.02

Exhibit D-1: Balanced Generation, 2 GTs Stream Table/Heat and Mass Balance

STREAM NUMBER		7		8		9		10		11		12	
STREAM NAME		Hot Syngas		LPS from Cooling Train		Process Cond rec'le to sc'ber		Cold Syngas		Syngas (Hg free)		Sour Gas to SRU	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	5050.44	57.12	0.00	0.00	0.11	3.07	5049.65	57.28	5049.65	57.28	16.11	6.68
Nitrogen	28.013	24.50	0.28	0.00	0.00	0.00	0.02	24.49	0.28	24.49	0.28	0.05	0.02
Carbon Monoxide	28.010	98.60	1.12	0.00	0.00	0.00	0.05	98.58	1.12	98.58	1.12	0.52	0.22
Carbon Dioxide	44.010	3496.48	39.55	0.00	0.00	1.04	29.72	3488.85	39.57	3488.85	39.57	169.21	70.12
Methane	16.042	91.95	1.04	0.00	0.00	0.01	0.23	91.89	1.04	91.89	1.04	0.75	0.31
Argon	39.948	7.67	0.09	0.00	0.00	0.00	0.02	7.67	0.09	7.67	0.09	0.03	0.01
Hydrogen Sulfide	34.082	55.45	0.63	0.00	0.00	0.11	3.20	54.63	0.62	54.63	0.62	54.48	22.58
Carbonyl Sulfide	60.076	0.12	0.00	0.00	0.00	0.00	0.00	0.12	0.00	0.12	0.00	0.04	0.02
Ammonia	17.031	16.46	0.19	0.00	0.00	2.23	63.68	0.12	0.00	0.12	0.00	0.12	0.05
Oxygen	31.999	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		431.44	367.47	0.00	0.00	0.01	0.01	431.36	367.40	431.36	367.40	1.51	1.29

STREAM NUMBER		7		8		9		10		11		12	
STREAM NAME		Hot Syngas		LPS from Cooling Train		Process Cond rec'le to sc'ber		Cold Syngas		Syngas (Hg free)		Sour Gas to SRU	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Temperature	°C	303.80		153.02		192.19		39.30		39.30		39.30	
Pressure	bara	34.95		5.16		44.35		34.05		34.05		34.05	
Total Dry Molar Flow (kg.mol/h)		8841.66	100.00	0.00	0.00	3.50	100.00	8816.00	100.00	8816.00	100.00	241.32	100.00
Water	kg.mol/h	7016.82		3218.70		1105.29		17.35		17.35		5.40	
Total Wet (kg.mol/h)		15858.48		3218.70		1108.78		8833.35		8833.35		246.71	
Total Mass Flow (kg/h)		297,900		58,000		20,000		171,100		171,100		9,500	
Molecular Weight		18.78		18.02		18.04		19.37		19.37		38.37	

Exhibit D-1: Balanced Generation, 2 GTs Stream Table/Heat and Mass Balance

STREAM NUMBER		13		14		15		16		17		18	
STREAM NAME		O_2 to SRU		Sulfur Product		Feed to CO_2 Comp		CO_2 Product		Total Sweet Syngas		Syngas to PSA	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	0.00	0.00	15.75	0.50	26.42	0.79	5017.78	92.71	2560.36	92.71
Nitrogen	28.013	0.00	0.00	0.00	0.00	0.07	0.00	7.20	0.21	24.38	0.45	12.44	0.45
Carbon Monoxide	28.010	0.00	0.00	0.00	0.00	0.98	0.03	1.33	0.04	97.08	1.79	49.54	1.79
Carbon Dioxide	44.010	0.00	0.00	0.00	0.00	3143.70	99.41	3320.00	98.87	175.94	3.25	89.77	3.25
Methane	16.042	0.00	0.00	0.00	0.00	1.75	0.06	2.26	0.07	89.39	1.65	45.61	1.65
Argon	39.948	0.21	0.50	0.00	0.00	0.08	0.00	0.33	0.01	7.55	0.14	3.85	0.14
Hydrogen Sulfide	34.082	0.00	0.00	0.00	0.00	0.10	0.00	0.16	0.00	0.05	0.00	0.02	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.08	0.00	0.08	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	0.00	0.00	0.02	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	41.95	99.50	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	55.16	100.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		0.00	0.00	0.00	0.00	1.76	1.53	2.76	2.38	428.10	364.59	218.44	186.03
Temperature	°C	20.00		135.00		38.61		49.90		38.61		38.64	

STREAM NUMBER		13		14		15		16		17		18	
STREAM NAME		O_2 to SRU		Sulfur Product		Feed to CO_2 Comp		CO_2 Product		Total Sweet Syngas		Syngas to PSA	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara	3.00		1.01		34.05		145.00		34.05		33.05	
Total Dry Molar Flow (kg.mol/h)		42.16	100.00	55.16	100.00	3162.51	100.00	3357.82	100.00	5412.17	100.00	2761.59	100.00
Water	kg.mol/h	0.00		0.00		5.49		0.00		0.00		0.00	
Total Wet (kg.mol/h)		42.16		55.16		3168.00		3357.82		5412.17		2761.59	
Total Mass Flow (kg/h)			1,400		1,800		138,600		146,500		23,000		11,700
Molecular Weight			32.04		32.07		43.73		43.62		4.25		4.25

Exhibit D-1: Balanced Generation, 2 GTs Stream Table/Heat and Mass Balance

STREAM NUMBER		19		20		21		22		23		24	
STREAM NAME		Syngas to GT		Total Exhaust from GTs (x3)		PSA H_2 to NH_3 loop		N_2 to NH_3 loop		Feed to MUG Comp		Feed to NH_3 loop	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	2262.79	92.71	0.00	0.00	2201.91	100.00	0.00	0.00	2201.91	75.00	2201.91	75.00
Nitrogen	28.013	10.99	0.45	16375.29	80.21	0.00	0.00	733.97	100.00	733.97	25.00	733.97	25.00
Carbon Monoxide	28.010	43.78	1.79	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	44.010	79.34	3.25	173.91	0.85	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methane	16.042	40.31	1.65	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Argon	39.948	3.40	0.14	272.21	1.33	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Sulfide	34.082	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	0.00	0.00	3595.20	17.61	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		193.05	164.41	0.00	0.00	174.81	147.89	0.00	0.00	174.81	147.89	174.81	147.89
Temperature	°C	38.64		439.70		38.64		40.00		37.93		123.30	

STREAM NUMBER		19		20		21		22		23		24	
STREAM NAME		Syngas to GT		Total Exhaust from GTs (x3)		PSA H_2 to NH_3 loop		N_2 to NH_3 loop		Feed to MUG Comp		Feed to NH_3 loop	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara	32.75		1.05		33.05		33.30		33.05		142.00	
Total Dry Molar Flow (kg.mol/h)		2440.64	100.00	20416.63	100.00	2201.91	100.00	733.97	100.00	2935.88	100.00	2935.88	100.00
Water	kg.mol/h	0.00		3713.70		0.00		0.00		0.00		0.00	
Total Wet (kg.mol/h)		2440.64		24130.32		2201.91		733.97		2935.88		2935.88	
Total Mass Flow (kg/h)			10,371		659,200		4,400		20,600		25,000		25,000
Molecular Weight			4.25		27.32		2.02		28.01		8.52		8.52

Exhibit D-1: Balanced Generation, 2 GTs Stream Table/Heat and Mass Balance

STREAM NUMBER		25		26		27		28		29		30	
STREAM NAME		PSA Tail Gas to Recompression		Diluted Fuel to GT (x1)		Air to GGT (x1)		Flue Gas (total)		SRU Off gas to CO2 Compressor		Ammonia Purge to Duct Burner	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016			1131.40	74.31	0.00	0.00	0.00	0.00	10.67	5.46	4.40	50.03
Nitrogen	28.013			307.75	20.21	7879.90	75.52	16391.61	81.03	7.13	3.65	1.47	16.69
Carbon Monoxide	28.010			21.89	1.44	0.00	0.00	0.00	0.00	0.35	0.18	0.00	0.00
Carbon Dioxide	44.010			39.67	2.61	5.24	0.05	372.89	1.84	176.31	90.27	0.00	0.00
Methane	16.042			20.16	1.32	0.00	0.00	0.00	0.00	0.51	0.26	0.00	0.00
Argon	39.948			1.70	0.11	134.40	1.29	276.36	1.37	0.25	0.13	0.00	0.00
Hydrogen Sulfide	34.082			0.01	0.00	0.00	0.00	0.00	0.00	0.06	0.03	0.00	0.00
Carbonyl Sulfide	60.076			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031			0.00	0.00	0.00	0.00	0.00	0.00	0.02	0.01	2.93	33.28
Oxygen	31.999			0.00	0.00	2414.57	23.14	3189.41	15.77	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065			0.00	0.00	0.00	0.00	0.05	0.00	0.01	0.00	0.00	0.00
Sulphur	32.070			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)				96.53	82.21	0.00	0.00	0.00	0.00	1.00	0.86	0.35	0.30
Temperature	°C			121.00		15.00		101.70		39.79		6.00	

STREAM NUMBER		25		26		27		28		29		30	
STREAM NAME		PSA Tail Gas to Recompression		Diluted Fuel to GT (x1)		Air to GGT (x1)		Flue Gas (total)		SRU Off gas to CO2 Compressor		Ammonia Purge to Duct Burner	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara			45.00		1.01		1.01		1.20		20.00	
Total Dry Molar Flow (kg.mol/h)				1522.57	100.00	10434.11	100.00	20230.31	100.00	195.31	100.00	8.80	100.00
Water	kg.mol/h			0.64		41.69		4373.76		9.62		0.00	
Total Wet (kg.mol/h)				1523.21		10475.80		24604.07		204.93		8.80	
Total Mass Flow (kg/h)					13,700		304,400		667,500		8,185		100
Molecular Weight					8.97		29.05		27.13		39.94		11.35

Exhibit D-1: Balanced Generation, 2 GTs Stream Table/Heat and Mass Balance

STREAM NUMBER		32		33		34		35		36		37	
STREAM NAME		Duct Burner Exhaust		Syngas to Duct Burner		PSA Tail to Duct Burner		HP N_2 Diluent to GT Feed		Sweep N_2 to Dryer		Total Oxygen Feed to Gasifier	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	194.63	92.71	358.45	64.04	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	28.013	16391.61	75.52	0.95	0.45	12.44	2.22	604.50	100.00	604.50	100.00	0.00	0.00
Carbon Monoxide	28.010	0.00	0.00	3.77	1.79	49.54	8.85	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	44.010	372.89	0.05	6.82	3.25	89.77	16.04	0.00	0.00	0.00	0.00	0.00	0.00
Methane	16.042	0.00	0.00	3.47	1.65	45.61	8.15	0.00	0.00	0.00	0.00	0.00	0.00
Argon	39.948	276.36	1.29	0.29	0.14	3.85	0.69	0.00	0.00	0.00	0.00	7.67	0.50
Hydrogen Sulfide	34.082	0.00	0.00	0.00	0.00	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	3189.41	23.14	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1526.47	99.50
Sulphur Dioxide	64.065	0.05	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		0.00	0.00	16.61	14.14	43.63	38.14	0.00	0.00	0.00	0.00	0.00	0.00
Temperature	°C	659.70		38.64		40.00		40.00		40.00		150.00	

STREAM NUMBER		32		33		34		35		36		37	
STREAM NAME		Duct Burner Exhaust		Syngas to Duct Burner		PSA Tail to Duct Burner		HP N_2 Diluent to GT Feed		Sweep N_2 to Dryer		Total Oxygen Feed to Gasifier	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara	1.04		33.05		1.30		32.90		2.30		45.00	
Total Dry Molar Flow (kg.mol/h)		20230.31	100.00	209.93	100.00	559.69	100.00	604.50	100.00	604.50	100.00	1534.14	100.00
Water	kg.mol/h	4373.76		0.00		0.00		1.28		0.00		0.00	
Total Wet (kg.mol/h)		24604.07		209.93		559.69		605.78		604.50		1534.14	
Total Mass Flow (kg/h)			667,500		900		7,300		17,000		16,900		49,200
Molecular Weight			27.13		4.25		13.04		27.99		28.01		32.04

D-2 Net Zero Power Operating Point

Exhibit D-2: Net Zero Power Stream Table/Heat and Mass Balance

STREAM NUMBER		1		2		3		4		5		6	
STREAM NAME		AR coal feed		Dried Coal Feed		Scrubbed Syngas		Net Steam from Gasifier		Steam to Shift 1		Steam Raised in Shift	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	0.00	0.00	1844.27	42.39	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	28.013	0.00	0.00	0.00	0.00	16.19	0.37	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Monoxide	28.010	0.00	0.00	0.00	0.00	1618.07	37.19	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	44.010	0.00	0.00	0.00	0.00	772.66	17.76	0.00	0.00	0.00	0.00	0.00	0.00
Methane	16.042	0.00	0.00	0.00	0.00	45.92	1.06	0.00	0.00	0.00	0.00	0.00	0.00
Argon	39.948	0.00	0.00	0.00	0.00	5.07	0.12	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Sulfide	34.082	0.00	0.00	0.00	0.00	36.74	0.84	0.00	0.00	0.00	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	11.40	0.26	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h	41620		41620									
HHV / LHV (MW)		352.68	340.16	352.68	340.38	284.97	261.31	0.00	0.00	0.00	0.00	0.00	0.00
Temperature	°C	15.00		75.00		183.51		398.89		300.00		258.79	
Pressure	bara	1.01		1.01		36.35		41.00		41.00		46.00	

STREAM NUMBER		1		2		3		4		5		6	
STREAM NAME		AR coal feed		Dried Coal Feed		Scrubbed Syngas		Net Steam from Gasifier		Steam to Shift 1		Steam Raised in Shift	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Total Dry Molar Flow (kg.mol/h)			0.00		0.00	4350.33	100.00	0.00	0.00	0.00	0.00	0.00	0.00
Water	kg.mol/h	289.05		121.59		2045.80		720.88		4479.56		1579.72	
Total Wet (kg.mol/h)		289.05		121.59		6396.14		720.88		4479.56		1579.72	
Total Mass Flow (kg/h)			46,800		43,800		122,700		13,000		80,700		28,500
Molecular Weight							19.19		18.02		18.02		18.02

Exhibit D-2: Net Zero Power Stream Table/Heat and Mass Balance

STREAM NUMBER		7		8		9		10		11		12	
STREAM NAME		Hot Syngas		LPS from Cooling Train		Process Cond rec'le to sc'ber		Cold Syngas		Syngas (Hg free)		Sour Gas to SRU	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	3401.53	57.58	0.00	0.00	0.09	2.70	3401.01	57.75	3401.01	57.75	10.85	6.77
Nitrogen	28.013	16.19	0.27	0.00	0.00	0.00	0.01	16.19	0.27	16.19	0.27	0.03	0.02
Carbon Monoxide	28.010	60.82	1.03	0.00	0.00	0.00	0.04	60.81	1.03	60.81	1.03	0.32	0.20
Carbon Dioxide	44.010	2329.84	39.44	0.00	0.00	1.07	30.68	2323.99	39.46	2323.99	39.46	112.71	70.34
Methane	16.042	45.92	0.78	0.00	0.00	0.01	0.16	45.89	0.78	45.89	0.78	0.37	0.23
Argon	39.948	5.07	0.09	0.00	0.00	0.00	0.02	5.06	0.09	5.06	0.09	0.02	0.01
Hydrogen Sulfide	34.082	36.67	0.62	0.00	0.00	0.13	3.69	35.97	0.61	35.97	0.61	35.87	22.38
Carbonyl Sulfide	60.076	0.07	0.00	0.00	0.00	0.00	0.00	0.07	0.00	0.07	0.00	0.02	0.01
Ammonia	17.031	11.40	0.19	0.00	0.00	2.18	62.69	0.04	0.00	0.04	0.00	0.04	0.02
Oxygen	31.999	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		286.18	243.49	0.00	0.00	0.01	0.01	286.13	243.45	286.13	243.45	0.98	0.84

STREAM NUMBER		7		8		9		10		11		12	
STREAM NAME		Hot Syngas		LPS from Cooling Train		Process Cond rec'le to sc'ber		Cold Syngas		Syngas (Hg free)		Sour Gas to SRU	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Temperature	°C	301.52		153.02		176.27		38.38		38.38		38.38	
Pressure	bara	34.95		5.16		44.35		34.05		34.05		34.05	
Total Dry Molar Flow (kg.mol/h)		5907.51	100.00	0.00	0.00	3.47	100.00	5889.03	100.00	5889.03	100.00	160.25	100.00
Water	kg.mol/h	4968.19		1412.69		1105.24		11.00		11.00		3.42	
Total Wet (kg.mol/h)		10875.70		1412.69		1108.72		5900.03		5900.03		163.67	
Total Mass Flow (kg/h)		203,400		25,400		20,000		113,700		113,700		6,300	
Molecular Weight		18.71		18.02		18.04		19.26		19.26		38.40	

Exhibit D-2: Net Zero Power Stream Table/Heat and Mass Balance

STREAM NUMBER		13		14		15		16		17		18	
STREAM NAME		O_2 to SRU		Sulfur Product		Feed to CO_2 Comp		CO_2 Product		Total Sweet Syngas		Syngas to PSA	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	0.00	0.00	10.61	0.50	17.77	0.79	3379.55	93.30	2560.36	93.30
Nitrogen	28.013	0.00	0.00	0.00	0.00	0.04	0.00	4.97	0.22	16.11	0.44	12.20	0.44
Carbon Monoxide	28.010	0.00	0.00	0.00	0.00	0.60	0.03	0.82	0.04	59.89	1.65	45.37	1.65
Carbon Dioxide	44.010	0.00	0.00	0.00	0.00	2094.08	99.42	2211.95	98.88	117.20	3.24	88.79	3.24
Methane	16.042	0.00	0.00	0.00	0.00	0.87	0.04	1.13	0.05	44.65	1.23	33.82	1.23
Argon	39.948	0.14	0.50	0.00	0.00	0.05	0.00	0.22	0.01	4.99	0.14	3.78	0.14
Hydrogen Sulfide	34.082	0.00	0.00	0.00	0.00	0.07	0.00	0.11	0.00	0.03	0.00	0.02	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.05	0.00	0.05	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	28.00	99.50	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	36.44	100.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		0.00	0.00	0.00	0.00	1.11	0.96	1.75	1.51	284.05	241.65	215.20	183.08
Temperature	°C	20.00		135.00		37.69		49.91		37.69		37.72	

STREAM NUMBER		13		14		15		16		17		18	
STREAM NAME		O_2 to SRU		Sulfur Product		Feed to CO_2 Comp		CO_2 Product		Total Sweet Syngas		Syngas to PSA	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara	3.00		1.01		34.05		145.00		34.05		33.05	
Total Dry Molar Flow (kg.mol/h)		28.14	100.00	36.44	100.00	2106.38	100.00	2237.03	100.00	3622.41	100.00	2744.35	100.00
Water	kg.mol/h	0.00		0.00		5.05		0.00		0.00		0.00	
Total Wet (kg.mol/h)		28.14		36.44		2111.43		2237.03		3622.41		2744.35	
Total Mass Flow (kg/h)			900		1,200		92,300		97,600		15,000		11,400
Molecular Weight			32.04		32.07		43.72		43.62		4.15		4.15

Exhibit D-2: Net Zero Power Stream Table/Heat and Mass Balance

STREAM NUMBER		19		20		21		22		23		24	
STREAM NAME		Syngas to GT		Total Exhaust from GTs (x3)		PSA H_2 to NH_3 loop		N_2 to NH_3 loop		Feed to MUG Comp		Feed to NH_3 loop	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	819.16	93.30	0.00	0.00	2201.91	100.00	0.00	0.00	2201.91	75.00	2201.91	75.00
Nitrogen	28.013	3.90	0.44	6876.86	79.51	0.00	0.00	733.97	100.00	733.97	25.00	733.97	25.00
Carbon Monoxide	28.010	14.52	1.65	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	44.010	28.41	3.24	58.16	0.67	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methane	16.042	10.82	1.23	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Argon	39.948	1.21	0.14	114.63	1.33	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Sulfide	34.082	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	0.00	0.00	1599.08	18.49	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		68.85	58.57	0.00	0.00	174.81	147.89	0.00	0.00	174.81	147.89	174.81	147.89
Temperature	°C	37.72		422.50		37.72		40.00		37.25		123.30	

STREAM NUMBER		19		20		21		22		23		24	
STREAM NAME		Syngas to GT		Total Exhaust from GTs (x3)		PSA H_2 to NH_3 loop		N_2 to NH_3 loop		Feed to MUG Comp		Feed to NH_3 loop	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara	32.75		1.05		33.05		33.30		33.05		142.00	
Total Dry Molar Flow (kg.mol/h)		878.02	100.00	8648.73	100.00	2201.91	100.00	733.97	100.00	2935.88	100.00	2935.88	100.00
Water	kg.mol/h	0.00		1213.29		0.00		0.00		0.00		0.00	
Total Wet (kg.mol/h)		878.02		9862.03		2201.91		733.97		2935.88		2935.88	
Total Mass Flow (kg/h)		3,640		272,800		4,400		20,600		25,000		25,000	
Molecular Weight		4.15		27.66		2.02		28.01		8.52		8.52	

Exhibit D-2: Net Zero Power Stream Table/Heat and Mass Balance

STREAM NUMBER		25		26		27		28		29		30	
STREAM NAME		PSA Tail Gas to Recompression		Diluted Fuel to GT (x1)		Air to GGT (x1)		Flue Gas (total)		SRU Off gas to CO2 Compressor		Ammonia Purge to Duct Burner	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016			819.16	74.37	0.00	0.00	0.00	0.00	7.16	5.48	4.40	50.04
Nitrogen	28.013			227.30	20.64	6649.56	75.52	6891.99	80.50	4.93	3.77	1.47	16.67
Carbon Monoxide	28.010			14.52	1.32	0.00	0.00	0.00	0.00	0.22	0.17	0.00	0.00
Carbon Dioxide	44.010			28.41	2.58	4.42	0.05	226.15	2.64	117.88	90.22	0.00	0.00
Methane	16.042			10.82	0.98	0.00	0.00	0.00	0.00	0.25	0.19	0.00	0.00
Argon	39.948			1.21	0.11	113.42	1.29	118.41	1.38	0.17	0.13	0.00	0.00
Hydrogen Sulfide	34.082			0.01	0.00	0.00	0.00	0.00	0.00	0.04	0.03	0.00	0.00
Carbonyl Sulfide	60.076			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	2.93	33.29
Oxygen	31.999			0.00	0.00	2037.57	23.14	1325.07	15.48	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065			0.00	0.00	0.00	0.00	0.03	0.00	0.01	0.00	0.00	0.00
Sulphur	32.070			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)				68.85	58.57	0.00	0.00	0.00	0.00	0.65	0.55	0.35	0.30
Temperature	°C			121.00		15.00		102.90		37.48		6.00	

STREAM NUMBER		25		26		27		28		29		30	
STREAM NAME		PSA Tail Gas to Recompression		Diluted Fuel to GT (x1)		Air to GGT (x1)		Flue Gas (total)		SRU Off gas to CO2 Compressor		Ammonia Purge to Duct Burner	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara			45.00		1.01		1.01		1.20		20.00	
Total Dry Molar Flow (kg.mol/h)				1101.41	100.00	8804.97	100.00	8561.65	100.00	130.65	100.00	8.80	100.00
Water	kg.mol/h			0.47		35.18		1648.25		5.60		0.00	
Total Wet (kg.mol/h)				1101.88		8840.15		10209.89		136.26		8.80	
Total Mass Flow (kg/h)					9,900		256,800		279,800		5,460		100
Molecular Weight					8.99		29.05		27.41		40.07		11.35

Exhibit D-2: Net Zero Power Stream Table/Heat and Mass Balance

STREAM NUMBER		32		33		34		35		36		37	
STREAM NAME		Duct Burner Exhaust		Syngas to Duct Burner		PSA Tail to Duct Burner		HP N_2 Diluent to GT Feed		Sweep N_2 to Dryer		Total Oxygen Feed to Gasifier	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	0.03	93.30	358.45	66.08	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	28.013	6891.99	75.52	0.00	0.44	12.20	2.25	223.39	100.00	223.39	100.00	0.00	0.00
Carbon Monoxide	28.010	0.00	0.00	0.00	1.65	45.37	8.36	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	44.010	226.15	0.05	0.00	3.24	88.79	16.37	0.00	0.00	0.00	0.00	0.00	0.00
Methane	16.042	0.00	0.00	0.00	1.23	33.82	6.24	0.00	0.00	0.00	0.00	0.00	0.00
Argon	39.948	118.41	1.29	0.00	0.14	3.78	0.70	0.00	0.00	0.00	0.00	5.07	0.50
Hydrogen Sulfide	34.082	0.00	0.00	0.00	0.00	0.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	1325.07	23.14	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1008.33	99.50
Sulphur Dioxide	64.065	0.03	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		0.00	0.00	0.00	0.00	40.39	35.18	0.00	0.00	0.00	0.00	0.00	0.00
Temperature	°C	781.20		37.72		40.00		40.00		40.00		150.00	

STREAM NUMBER		32		33		34		35		36		37	
STREAM NAME		Duct Burner Exhaust		Syngas to Duct Burner		PSA Tail to Duct Burner		HP N_2 Diluent to GT Feed		Sweep N_2 to Dryer		Total Oxygen Feed to Gasifier	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara	1.04		33.05		1.30		32.90		2.30		45.00	
Total Dry Molar Flow (kg.mol/h)		8561.65	100.00	0.04	100.00	542.44	100.00	223.39	100.00	223.39	100.00	1013.40	100.00
Water	kg.mol/h	1648.25		0.00		0.00		0.47		0.00		0.00	
Total Wet (kg.mol/h)		10209.89		0.04		542.44		223.86		223.39		1013.40	
Total Mass Flow (kg/h)		279,800		0		6,900		6,300		6,300		32,500	
Molecular Weight		27.41		4.15		12.79		27.99		28.01		32.04	

D-3 High Electricity Generation Operating Point

Exhibit D-3: High Electricity Generation Stream Table/Heat and Mass Balance

STREAM NUMBER		1		2		3		4		5		6	
STREAM NAME		AR coal feed		Dried Coal Feed		Scrubbed Syngas		Net Steam from Gasifier		Steam to Shift 1		Steam Raised in Shift	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	0.00	0.00	2552.08	40.23	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	28.013	0.00	0.00	0.00	0.00	24.50	0.39	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Monoxide	28.010	0.00	0.00	0.00	0.00	2596.96	40.94	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	44.010	0.00	0.00	0.00	0.00	998.23	15.74	0.00	0.00	0.00	0.00	0.00	0.00
Methane	16.042	0.00	0.00	0.00	0.00	91.95	1.45	0.00	0.00	0.00	0.00	0.00	0.00
Argon	39.948	0.00	0.00	0.00	0.00	7.67	0.12	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Sulfide	34.082	0.00	0.00	0.00	0.00	55.57	0.88	0.00	0.00	0.00	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	16.46	0.26	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h	62984		62984									
HHV / LHV (MW)		533.72	514.77	533.72	515.10	429.50	396.06	0.00	0.00	0.00	0.00	0.00	0.00
Temperature	°C	15.00		75.00		178.36		398.89		300.00		258.79	
Pressure	bara	1.01		1.01		36.35		41.00		41.00		46.00	

STREAM NUMBER		1		2		3		4		5		6	
STREAM NAME		AR coal feed		Dried Coal Feed		Scrubbed Syngas		Net Steam from Gasifier		Steam to Shift 1		Steam Raised in Shift	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Total Dry Molar Flow (kg.mol/h)			0.00		0.00	6343.42	100.00	0.00	0.00	0.00	0.00	0.00	0.00
Water	kg.mol/h	437.42		184.01		2514.70		2074.31		7000.28		2618.86	
Total Wet (kg.mol/h)		437.42		184.01		8858.12		2074.31		7000.28		2618.86	
Total Mass Flow (kg/h)			70,900		66,300		171,800		37,400		126,100		47,200
Molecular Weight							19.39		18.02		18.02		18.02

Exhibit D-3: High Electricity Generation Stream Table/Heat and Mass Balance

STREAM NUMBER		7		8		9		10		11		12	
STREAM NAME		Hot Syngas		LPS from Cooling Train		Process Cond rec'le to sc'ber		Cold Syngas		Syngas (Hg free)		Sour Gas to SRU	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	5050.41	57.12	0.00	0.00	0.11	3.11	5049.61	57.28	5049.61	57.28	16.11	6.67
Nitrogen	28.013	24.50	0.28	0.00	0.00	0.00	0.02	24.49	0.28	24.49	0.28	0.05	0.02
Carbon Monoxide	28.010	98.63	1.12	0.00	0.00	0.00	0.05	98.61	1.12	98.61	1.12	0.52	0.22
Carbon Dioxide	44.010	3496.44	39.55	0.00	0.00	1.03	29.60	3488.86	39.57	3488.86	39.57	169.21	70.11
Methane	16.042	91.95	1.04	0.00	0.00	0.01	0.23	91.89	1.04	91.89	1.04	0.75	0.31
Argon	39.948	7.67	0.09	0.00	0.00	0.00	0.02	7.67	0.09	7.67	0.09	0.03	0.01
Hydrogen Sulfide	34.082	55.45	0.63	0.00	0.00	0.11	3.15	54.64	0.62	54.64	0.62	54.50	22.58
Carbonyl Sulfide	60.076	0.12	0.00	0.00	0.00	0.00	0.00	0.12	0.00	0.12	0.00	0.04	0.02
Ammonia	17.031	16.46	0.19	0.00	0.00	2.23	63.82	0.13	0.00	0.13	0.00	0.13	0.05
Oxygen	31.999	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		431.44	367.47	0.00	0.00	0.01	0.01	431.36	367.40	431.36	367.40	1.51	1.29
Temperature	°C	303.83		153.02		193.51		39.97		39.97		39.97	

STREAM NUMBER		7		8		9		10		11		12	
STREAM NAME		Hot Syngas		LPS from Cooling Train		Process Cond rec'le to sc'ber		Cold Syngas		Syngas (Hg free)		Sour Gas to SRU	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara	34.95		5.16		44.35		34.05		34.05		34.05	
Total Dry Molar Flow (kg.mol/h)		8841.62	100.00	0.00	0.00	3.49	100.00	8816.03	100.00	8816.03	100.00	241.34	100.00
Water	kg.mol/h	7016.77		3432.72		1105.31		18.01		18.01		5.60	
Total Wet (kg.mol/h)		15858.40		3432.72		1108.80		8834.04		8834.04		246.94	
Total Mass Flow (kg/h)			297,900		61,800		20,000		171,100		171,100		9,500
Molecular Weight			18.78		18.02		18.04		19.37		19.37		38.36

Exhibit D-3: High Electricity Generation Stream Table/Heat and Mass Balance

STREAM NUMBER		13		14		15		16		17		18	
STREAM NAME		O_2 to SRU		Sulfur Product		Feed to CO_2 Comp		CO_2 Product		Total Sweet Syngas		Syngas to PSA	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	0.00	0.00	15.75	0.50	26.42	0.79	5017.75	92.71	1623.15	92.71
Nitrogen	28.013	0.00	0.00	0.00	0.00	0.07	0.00	7.19	0.21	24.38	0.45	7.89	0.45
Carbon Monoxide	28.010	0.00	0.00	0.00	0.00	0.98	0.03	1.33	0.04	97.11	1.79	31.41	1.79
Carbon Dioxide	44.010	0.00	0.00	0.00	0.00	3143.71	99.41	3319.97	98.87	175.94	3.25	56.91	3.25
Methane	16.042	0.00	0.00	0.00	0.00	1.75	0.06	2.26	0.07	89.39	1.65	28.92	1.65
Argon	39.948	0.21	0.50	0.00	0.00	0.08	0.00	0.33	0.01	7.55	0.14	2.44	0.14
Hydrogen Sulfide	34.082	0.00	0.00	0.00	0.00	0.10	0.00	0.16	0.00	0.05	0.00	0.01	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.08	0.00	0.08	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	0.00	0.00	0.02	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	41.95	99.50	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	55.16	100.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		0.00	0.00	0.00	0.00	1.76	1.53	2.76	2.38	428.10	364.59	138.48	117.94
Temperature	°C	20.00		135.00		39.25		49.90		39.25		39.28	

STREAM NUMBER		13		14		15		16		17		18	
STREAM NAME		O_2 to SRU		Sulfur Product		Feed to CO_2 Comp		CO_2 Product		Total Sweet Syngas		Syngas to PSA	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara	3.00		1.01		34.05		145.00		34.05		33.05	
Total Dry Molar Flow (kg.mol/h)		42.16	100.00	55.16	100.00	3162.52	100.00	3357.79	100.00	5412.17	100.00	1750.74	100.00
Water	kg.mol/h	0.00		0.00		5.53		0.00		0.00		0.00	
Total Wet (kg.mol/h)		42.16		55.16		3168.06		3357.79		5412.17		1750.74	
Total Mass Flow (kg/h)			1,400		1,800		138,600		146,500		23,000		7,400
Molecular Weight			32.04		32.07		43.73		43.62		4.25		4.25

Exhibit D-3: High Electricity Generation Stream Table/Heat and Mass Balance

STREAM NUMBER		19		20		21		22		23		24	
STREAM NAME		Syngas to GT		Total Exhaust from GTs (x3)		PSA H_2 to NH_3 loop		N_2 to NH_3 loop		Feed to MUG Comp		Feed to NH_3 loop	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	3394.59	92.71	0.00	0.00	1395.91	100.00	0.00	0.00	1395.91	75.00	1395.91	75.00
Nitrogen	28.013	16.49	0.45	24562.6 8	80.21	0.00	0.00	465.30	100.00	465.30	25.00	465.30	25.00
Carbon Monoxide	28.010	65.70	1.79	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	44.010	119.03	3.25	260.89	0.85	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methane	16.042	60.47	1.65	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Argon	39.948	5.11	0.14	408.32	1.33	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Sulfide	34.082	0.03	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	0.00	0.00	5392.75	17.61	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.03	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		289.61	246.65	0.00	0.00	110.82	93.76	0.00	0.00	110.82	93.76	110.82	93.76

STREAM NUMBER		19		20		21		22		23		24	
STREAM NAME		Syngas to GT		Total Exhaust from GTs (x3)		PSA H_2 to NH_3 loop		N_2 to NH_3 loop		Feed to MUG Comp		Feed to NH_3 loop	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Temperature	°C	39.29		439.70		39.28		40.00		38.41		123.30	
Pressure	bara	32.75		1.05		33.05		33.30		33.05		142.00	
Total Dry Molar Flow (kg.mol/h)		3661.42	100.00	30624.67	100.00	1395.91	100.00	465.30	100.00	1861.21	100.00	1861.21	100.00
Water	kg.mol/h	0.00		5570.61		0.00		0.00		0.00		0.00	
Total Wet (kg.mol/h)		3661.42		36195.27		1395.91		465.30		1861.21		1861.21	
Total Mass Flow (kg/h)			15,559		988,800		2,800		13,000		15,800		15,800
Molecular Weight			4.25		27.32		2.02		28.01		8.52		8.52

Exhibit D-3: High Electricity Generation Stream Table/Heat and Mass Balance

STREAM NUMBER		25		26		27		28		29		30	
STREAM NAME		PSA Tail Gas to Recompression		Diluted Fuel to GT (x1)		Air to GGT (x1)		Flue Gas (total)		SRU Off gas to CO2 Compressor		Ammonia Purge to Duct Burner	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016			1131.42	74.31	0.00	0.00	0.00	0.00	10.67	5.46	2.79	50.04
Nitrogen	28.013			307.66	20.21	7879.90	75.52	24572.42	80.40	7.13	3.65	0.93	16.67
Carbon Monoxide	28.010			21.90	1.44	0.00	0.00	0.00	0.00	0.35	0.18	0.00	0.00
Carbon Dioxide	44.010			39.67	2.61	5.24	0.05	378.13	1.24	176.26	90.27	0.00	0.00
Methane	16.042			20.16	1.32	0.00	0.00	0.00	0.00	0.51	0.26	0.00	0.00
Argon	39.948			1.70	0.11	134.40	1.29	410.76	1.34	0.25	0.13	0.00	0.00
Hydrogen Sulfide	34.082			0.01	0.00	0.00	0.00	0.00	0.00	0.06	0.03	0.00	0.00
Carbonyl Sulfide	60.076			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031			0.00	0.00	0.00	0.00	0.00	0.00	0.02	0.01	1.86	33.29
Oxygen	31.999			0.00	0.00	2414.57	23.14	5202.78	17.02	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065			0.00	0.00	0.00	0.00	0.05	0.00	0.01	0.00	0.00	0.00
Sulphur	32.070			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)				96.53	82.21	0.00	0.00	0.00	0.00	1.00	0.86	0.22	0.19
Temperature	°C			121.00		15.00		105.00		39.79		6.00	

STREAM NUMBER		25		26		27		28		29		30	
STREAM NAME		PSA Tail Gas to Recompression		Diluted Fuel to GT (x1)		Air to GGT (x1)		Flue Gas (total)		SRU Off gas to CO2 Compressor		Ammonia Purge to Duct Burner	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara			45.00		1.01		1.01		1.20		20.00	
Total Dry Molar Flow (kg.mol/h)				1522.51	100.00	10434.11	100.00	30564.14	100.00	195.26	100.00	5.58	100.00
Water	kg.mol/h			0.64		41.69		5861.28		9.62		0.00	
Total Wet (kg.mol/h)				1523.15		10475.80		36425.42		204.88		5.58	
Total Mass Flow (kg/h)					13,700		304,400		993,500		8,183		63
Molecular Weight					8.97		29.05		27.27		39.94		11.35

Exhibit D-3: High Electricity Generation Stream Table/Heat and Mass Balance

STREAM NUMBER		32		33		34		35		36		37	
STREAM NAME		Duct Burner Exhaust		Syngas to Duct Burner		PSA Tail to Duct Burner		HP N_2 Diluent to GT Feed		Sweep N_2 to Dryer		Total Oxygen Feed to Gasifier	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	0.01	92.71	227.24	64.04	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	28.013	24572.42	75.52	0.00	0.45	7.89	2.22	906.58	100.00	906.58	100.00	0.00	0.00
Carbon Monoxide	28.010	0.00	0.00	0.00	1.79	31.41	8.85	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	44.010	378.13	0.05	0.00	3.25	56.91	16.04	0.00	0.00	0.00	0.00	0.00	0.00
Methane	16.042	0.00	0.00	0.00	1.65	28.92	8.15	0.00	0.00	0.00	0.00	0.00	0.00
Argon	39.948	410.76	1.29	0.00	0.14	2.44	0.69	0.00	0.00	0.00	0.00	7.67	0.50
Hydrogen Sulfide	34.082	0.00	0.00	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	5202.78	23.14	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1526.47	99.50
Sulphur Dioxide	64.065	0.05	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		0.00	0.00	0.00	0.00	27.66	24.18	0.00	0.00	0.00	0.00	0.00	0.00
Temperature	°C	509.00		39.28		40.00		40.00		40.00		150.00	

STREAM NUMBER		32		33		34		35		36		37	
STREAM NAME		Duct Burner Exhaust		Syngas to Duct Burner		PSA Tail to Duct Burner		HP N_2 Diluent to GT Feed		Sweep N_2 to Dryer		Total Oxygen Feed to Gasifier	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara	1.04		33.05		1.30		32.90		2.30		45.00	
Total Dry Molar Flow (kg.mol/h)		30564.14	100.00	0.01	100.00	354.83	100.00	906.58	100.00	906.58	100.00	1534.14	100.00
Water	kg.mol/h	5861.28		0.00		0.00		1.92		0.00		0.00	
Total Wet (kg.mol/h)		36425.42		0.01		354.83		908.49		906.58		1534.14	
Total Mass Flow (kg/h)		993,500		0		4,600		25,400		25,400		49,200	
Molecular Weight		27.27		4.25		13.04		27.99		28.01		32.04	

D-4 Max Electricity Generation Operating Point

Exhibit D-4: Max Electricity Generation Stream Table/Heat and Mass Balance

STREAM NUMBER		1		2		3		4		5		6	
STREAM NAME		AR coal feed		Dried Coal Feed		Scrubbed Syngas		Net Steam from Gasifier		Steam to Shift 1		Steam Raised in Shift	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	0.00	0.00	2552.07	40.23	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	28.013	0.00	0.00	0.00	0.00	24.50	0.39	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Monoxide	28.010	0.00	0.00	0.00	0.00	2596.96	40.94	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	44.010	0.00	0.00	0.00	0.00	998.24	15.74	0.00	0.00	0.00	0.00	0.00	0.00
Methane	16.042	0.00	0.00	0.00	0.00	91.95	1.45	0.00	0.00	0.00	0.00	0.00	0.00
Argon	39.948	0.00	0.00	0.00	0.00	7.67	0.12	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Sulfide	34.082	0.00	0.00	0.00	0.00	55.57	0.88	0.00	0.00	0.00	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	16.46	0.26	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h	62984		62984									
HHV / LHV (MW)		533.72	514.77	533.72	515.10	429.50	396.06	0.00	0.00	0.00	0.00	0.00	0.00
Temperature	°C	15.00		75.00		178.32		398.89		300.00		258.79	
Pressure	bara	1.01		1.01		36.35		41.00		41.00		46.00	

STREAM NUMBER		1		2		3		4		5		6	
STREAM NAME		AR coal feed		Dried Coal Feed		Scrubbed Syngas		Net Steam from Gasifier		Steam to Shift 1		Steam Raised in Shift	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Total Dry Molar Flow (kg.mol/h)			0.00		0.00	6343.43	100.00	0.00	0.00	0.00	0.00	0.00	0.00
Water	kg.mol/h	437.42		184.01		2511.80		2074.21		7003.26		2615.78	
Total Wet (kg.mol/h)		437.42		184.01		8855.23		2074.21		7003.26		2615.78	
Total Mass Flow (kg/h)			70,900		66,300		171,700		37,400		126,200		47,100
Molecular Weight							19.39		18.02		18.02		18.02

Exhibit D-4: Max Electricity Generation Stream Table/Heat and Mass Balance

STREAM NUMBER		7		8		9		10		11		12	
STREAM NAME		Hot Syngas		LPS from Cooling Train		Process Cond rec'le to sc'ber		Cold Syngas		Syngas (Hg free)		Sour Gas to SRU	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	5050.44	57.12	0.00	0.00	0.11	3.07	5049.65	57.28	5049.65	57.28	16.11	6.68
Nitrogen	28.013	24.50	0.28	0.00	0.00	0.00	0.02	24.49	0.28	24.49	0.28	0.05	0.02
Carbon Monoxide	28.010	98.60	1.12	0.00	0.00	0.00	0.05	98.58	1.12	98.58	1.12	0.52	0.22
Carbon Dioxide	44.010	3496.48	39.55	0.00	0.00	1.04	29.72	3488.85	39.57	3488.85	39.57	169.21	70.12
Methane	16.042	91.95	1.04	0.00	0.00	0.01	0.23	91.89	1.04	91.89	1.04	0.75	0.31
Argon	39.948	7.67	0.09	0.00	0.00	0.00	0.02	7.67	0.09	7.67	0.09	0.03	0.01
Hydrogen Sulfide	34.082	55.45	0.63	0.00	0.00	0.11	3.20	54.63	0.62	54.63	0.62	54.48	22.58
Carbonyl Sulfide	60.076	0.12	0.00	0.00	0.00	0.00	0.00	0.12	0.00	0.12	0.00	0.04	0.02
Ammonia	17.031	16.46	0.19	0.00	0.00	2.23	63.68	0.12	0.00	0.12	0.00	0.12	0.05
Oxygen	31.999	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		431.44	367.47	0.00	0.00	0.01	0.01	431.36	367.40	431.36	367.40	1.51	1.29
Temperature	°C	303.80		153.02		192.19		39.30		39.30		39.30	

STREAM NUMBER		7		8		9		10		11		12	
STREAM NAME		Hot Syngas		LPS from Cooling Train		Process Cond rec'le to sc'ber		Cold Syngas		Syngas (Hg free)		Sour Gas to SRU	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara	34.95		5.16		44.35		34.05		34.05		34.05	
Total Dry Molar Flow (kg.mol/h)		8841.66	100.00	0.00	0.00	3.50	100.00	8816.00	100.00	8816.00	100.00	241.32	100.00
Water	kg.mol/h	7016.82		3218.70		1105.29		17.35		17.35		5.40	
Total Wet (kg.mol/h)		15858.48		3218.70		1108.78		8833.35		8833.35		246.71	
Total Mass Flow (kg/h)			297,900		58,000		20,000		171,100		171,100		9,500
Molecular Weight			18.78		18.02		18.04		19.37		19.37		38.37

Exhibit D-4: Max Electricity Generation Stream Table/Heat and Mass Balance

STREAM NUMBER		13		14		15		16		17		18	
STREAM NAME		O_2 to SRU		Sulfur Product		Feed to CO_2 Comp		CO_2 Product		Total Sweet Syngas		Syngas to PSA	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	0.00	0.00	15.75	0.50	26.42	0.79	5017.78	92.71	252.15	92.71
Nitrogen	28.013	0.00	0.00	0.00	0.00	0.07	0.00	7.20	0.21	24.38	0.45	1.23	0.45
Carbon Monoxide	28.010	0.00	0.00	0.00	0.00	0.98	0.03	1.33	0.04	97.08	1.79	4.88	1.79
Carbon Dioxide	44.010	0.00	0.00	0.00	0.00	3143.70	99.41	3320.00	98.87	175.94	3.25	8.84	3.25
Methane	16.042	0.00	0.00	0.00	0.00	1.75	0.06	2.26	0.07	89.39	1.65	4.49	1.65
Argon	39.948	0.21	0.50	0.00	0.00	0.08	0.00	0.33	0.01	7.55	0.14	0.38	0.14
Hydrogen Sulfide	34.082	0.00	0.00	0.00	0.00	0.10	0.00	0.16	0.00	0.05	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.08	0.00	0.08	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	0.00	0.00	0.02	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	41.95	99.50	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.00	0.00	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	55.16	100.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		0.00	0.00	0.00	0.00	1.76	1.53	2.76	2.38	428.10	364.59	21.51	18.32
Temperature	°C	20.00		135.00		38.61		49.90		38.61		38.64	

STREAM NUMBER		13		14		15		16		17		18	
STREAM NAME		O_2 to SRU		Sulfur Product		Feed to CO_2 Comp		CO_2 Product		Total Sweet Syngas		Syngas to PSA	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara	3.00		1.01		34.05		145.00		34.05		33.05	
Total Dry Molar Flow (kg.mol/h)		42.16	100.00	55.16	100.00	3162.51	100.00	3357.82	100.00	5412.17	100.00	271.97	100.00
Water	kg.mol/h	0.00		0.00		5.49		0.00		0.00		0.00	
Total Wet (kg.mol/h)		42.16		55.16		3168.00		3357.82		5412.17		271.97	
Total Mass Flow (kg/h)			1,400		1,800		138,600		146,500		23,000		1,200
Molecular Weight			32.04		32.07		43.73		43.62		4.25		4.25

Exhibit D-4: Max Electricity Generation Stream Table/Heat and Mass Balance

STREAM NUMBER		19		20		21		22		23		24	
STREAM NAME		Syngas to GT		Total Exhaust from GTs (x3)		PSA H_2 to NH_3 loop		N_2 to NH_3 loop		Feed to MUG Comp		Feed to NH_3 loop	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	3394.73	92.71	0.00	0.00	216.85	100.00	0.00	0.00	216.85	75.00	216.85	75.00
Nitrogen	28.013	16.49	0.45	24563.01	80.21	0.00	0.00	72.28	100.00	72.28	25.00	72.28	25.00
Carbon Monoxide	28.010	65.68	1.79	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	44.010	119.03	3.25	260.88	0.85	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Methane	16.042	60.48	1.65	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Argon	39.948	5.11	0.14	408.32	1.33	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Sulfide	34.082	0.03	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	0.00	0.00	5392.68	17.61	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065	0.00	0.00	0.03	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		289.63	246.66	0.00	0.00	17.22	14.57	0.00	0.00	17.22	14.57	17.22	14.57

STREAM NUMBER		19		20		21		22		23		24	
STREAM NAME		Syngas to GT		Total Exhaust from GTs (x3)		PSA H_2 to NH_3 loop		N_2 to NH_3 loop		Feed to MUG Comp		Feed to NH_3 loop	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Temperature	°C	38.64		439.70		38.64		40.00		37.93		123.30	
Pressure	bara	32.75		1.05		33.05		33.30		33.05		142.00	
Total Dry Molar Flow (kg.mol/h)		3661.55	100.00	30624.92	100.00	216.85	100.00	72.28	100.00	289.14	100.00	289.14	100.00
Water	kg.mol/h	0.00		5570.76		0.00		0.00		0.00		0.00	
Total Wet (kg.mol/h)		3661.55		36195.67		216.85		72.28		289.14		289.14	
Total Mass Flow (kg/h)			15,559		988,800		400		2,000		2,500		2,500
Molecular Weight			4.25		27.32		2.02		28.01		8.52		8.52

Exhibit D-4: Max Electricity Generation Stream Table/Heat and Mass Balance

STREAM NUMBER		25		26		27		28		29		30	
STREAM NAME		PSA Tail Gas to Recompression		Diluted Fuel to GT (x1)		Air to GGT (x1)		Flue Gas (total)		SRU Off gas to CO2 Compressor		Ammonia Purge to Duct Burner	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016			1131.47	74.31	0.00	0.00	0.00	0.00	10.67	5.46	0.43	50.03
Nitrogen	28.013			307.77	20.21	7879.90	75.52	24571.18	81.97	7.13	3.65	0.14	16.68
Carbon Monoxide	28.010			21.89	1.44	0.00	0.00	0.00	0.00	0.35	0.18	0.00	0.00
Carbon Dioxide	44.010			39.67	2.61	5.24	0.05	378.10	1.26	176.31	90.27	0.00	0.00
Methane	16.042			20.16	1.32	0.00	0.00	0.00	0.00	0.51	0.26	0.00	0.00
Argon	39.948			1.70	0.11	134.40	1.29	410.76	1.37	0.25	0.13	0.00	0.00
Hydrogen Sulfide	34.082			0.01	0.00	0.00	0.00	0.00	0.00	0.06	0.03	0.00	0.00
Carbonyl Sulfide	60.076			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031			0.00	0.00	0.00	0.00	0.00	0.00	0.02	0.01	0.29	33.29
Oxygen	31.999			0.00	0.00	2414.57	23.14	4615.60	15.40	0.00	0.00	0.00	0.00
Sulphur Dioxide	64.065			0.00	0.00	0.00	0.00	0.05	0.00	0.01	0.00	0.00	0.00
Sulphur	32.070			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461			0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)				96.53	82.21	0.00	0.00	0.00	0.00	1.00	0.86	0.03	0.03
Temperature	°C			121.00		15.00		105.00		39.79		6.00	

STREAM NUMBER		25		26		27		28		29		30	
STREAM NAME		PSA Tail Gas to Recompression		Diluted Fuel to GT (x1)		Air to GGT (x1)		Flue Gas (total)		SRU Off gas to CO2 Compressor		Ammonia Purge to Duct Burner	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Pressure	bara			45.00		1.01		1.01		1.20		20.00	
Total Dry Molar Flow (kg.mol/h)				1522.67	100.00	10434.11	100.00	29975.69	100.00	195.31	100.00	0.87	100.00
Water	kg.mol/h			0.64		41.69		7035.66		9.62		0.00	
Total Wet (kg.mol/h)				1523.31		10475.80		37011.36		204.93		0.87	
Total Mass Flow (kg/h)					13,700		304,400		995,800		8,185		10
Molecular Weight					8.97		29.05		26.91		39.94		11.35

Exhibit D-4: Max Electricity Generation Stream Table/Heat and Mass Balance

STREAM NUMBER		32		33		34		35		36		37	
STREAM NAME		Duct Burner Exhaust		Syngas to Duct Burner		PSA Tail to Duct Burner		HP N_2 Diluent to GT Feed		Sweep N_2 to Dryer		Total Oxygen Feed to Gasifier	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Hydrogen	2.016	0.00	0.00	1370.90	92.71	35.30	64.04	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	28.013	24571.18	75.52	6.66	0.45	1.23	2.22	906.91	100.00	906.91	100.00	0.00	0.00
Carbon Monoxide	28.010	0.00	0.00	26.52	1.79	4.88	8.85	0.00	0.00	0.00	0.00	0.00	0.00
Carbon Dioxide	44.010	378.10	0.05	48.07	3.25	8.84	16.04	0.00	0.00	0.00	0.00	0.00	0.00
Methane	16.042	0.00	0.00	24.42	1.65	4.49	8.15	0.00	0.00	0.00	0.00	0.00	0.00
Argon	39.948	410.76	1.29	2.06	0.14	0.38	0.69	0.00	0.00	0.00	0.00	7.67	0.50
Hydrogen Sulfide	34.082	0.00	0.00	0.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbonyl Sulfide	60.076	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Ammonia	17.031	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Oxygen	31.999	4615.60	23.14	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1526.47	99.50
Sulphur Dioxide	64.065	0.05	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Sulphur	32.070	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen Chloride	36.461	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
coal feed (dry)	kg/h												
HHV / LHV (MW)		0.00	0.00	116.96	99.61	4.30	3.76	0.00	0.00	0.00	0.00	0.00	0.00

STREAM NUMBER		32		33		34		35		36		37	
STREAM NAME		Duct Burner Exhaust		Syngas to Duct Burner		PSA Tail to Duct Burner		HP N_2 Diluent to GT Feed		Sweep N_2 to Dryer		Total Oxygen Feed to Gasifier	
Component	Molecular Weight	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)	kg.mol/h	mol% (dry)
Temperature	°C	725.80		38.64		40.00		40.00		40.00		150.00	
Pressure	bara	1.04		33.05		1.30		32.90		2.30		45.00	
Total Dry Molar Flow (kg.mol/h)		29975.69	100.00	1478.64	100.00	55.12	100.00	906.91	100.00	906.91	100.00	1534.14	100.00
Water	kg.mol/h	7035.66		0.00		0.00		1.92		0.00		0.00	
Total Wet (kg.mol/h)		37011.36		1478.64		55.12		908.83		906.91		1534.14	
Total Mass Flow (kg/h)		995,800		6,300		700		25,400		25,400		49,200	
Molecular Weight		26.91		4.25		13.04		27.99		28.01		32.04	

Appendix E. Process Integration

E-1 Air Integration

Some IGCC plants integrate the GT air compressor with the ASU to save part (or all) of the capital cost of the ASU main air compressor there are also some overall plant efficiency gains as long as most or all of the nitrogen from the ASU is fed to the GT for fuel dilution. For an IGCC burning a high hydrogen fuel in a large frame Gas Turbine, the fuel is typically diluted to about 40% H_2, most of not all of the N_2 from the ASU is used for this service.

The subject plant design uses aero-derivative gas turbines which require diluting the hydrogen in the fuel to 75%, therefore only a small portion of the available nitrogen is consumed by the power island. The ammonia loop also consumes nitrogen, but together the two consumptions represent about 25% of the total nitrogen available from the ASU.

E-2 Dryer Nitrogen Integration

The coal dryer uses nitrogen for fluidization and drying of the coal. This nitrogen is cooled, and the water condensed, with some nitrogen recycled to be re-used in the dryer. The balance (effectively the vent from the dryer) is compressed and fed to the GTs to dilute the high hydrogen fuel to meet the speciation from GE.

E-3 SRU Tail Gas Integration

The tail gas from the SRU, or Tail Gas Treatment, is typically combusted and vented to atmosphere.

In the subject plant design the SRU tail gas is fed to the suction of the CO_2 compressor to capture all of the CO_2 fed to the SRU. This is because the syngas feed to the AGR contains a large quantity of CO_2 which results in a significant quantity of CO_2 presenting in the feed gas to the SRU. This allows the overall carbon capture target to be met with a smaller shift unit and smaller AGR decreasing capital costs and reducing steam consumption which in turn increases power production.

Appendix F. Heat Integration

Heat is liberated in several units within the plant and is consumed in others. The AST team has designed the plant to efficiently utilize the available heat without compromising plant functionality. Some IGCC plants have been highly integrated hampering start-up and operation resulting in significant reduction in availability.

F-1 Coal Drying

The coal is dried to meet the gasifier feedstock speciation using low pressure steam. The low pressure steam is generated in the gas cooling section of the plant.

F-2 Gasifier

The gasifier consumes energy in the form of IP superheated steam and most of the sensible heat in the syngas is recovered in the form of IP superheated steam. Most of this steam is fed back to the gasifier to provide the fluidization of the coal and as a reagent. The balance is mixed with IP steam from the shift unit boiler and IP steam extracted from the HRSG and BFW for superheating before being fed as process steam to the inlet of the shift reactor. The gasifier also uses LP steam from the gas cooling unit to preheat the oxygen feed to the gasifier.

Part of the sensible heat in the syngas exiting the gasifier is used to generate steam in the syngas scrubber. This steam is intimately mixed with the syngas as is fed to the shift unit reducing the demand for IP process steam in the shift unit thereby increasing efficiency and reducing water consumption, in the form of make-up to the demineralized water system.

The temperature of the scrubber is increased by using hot water makeup recycled from the bottom of the desaturator further increasing steam production.

F-3 Sour Shift Unit

The water gas shift reaction is exothermic, and the heat of the reaction is recovered as IP steam which is combined with IP steam from the HRSG and the gasifier and fed back to the process stream of the shift unit.

F-4 Syngas Cooling

The syngas cooling unit recovers the sensible and latent heat from the syngas exiting the shift unit. The first step is to preheat BFW feeding the heat recovery boilers in the gasifier and the shift units. The syngas is then fed to the desaturator, which allows the sensible and latent heat in the syngas to be recovered and reused in a cost effective manner.

Steam is taken from the hot water outlet of the desaturator and used as make-up for the scrubber in the gasifier increasing the temperature of the scrubber and the amount of water vaporized by it. The circulating hot water from the desaturator is used to:

- Preheat BFW and generate LP steam for use in the coal dryer, gasifier oxygen preheating and the utilities.
- Provide the reboiler duty in the AGR
- Preheat the GT fuel gas steam

- Preheat DMW and provide LP steam for the deaerator in the power block.

F-5 AGR

The AGR uses low grade heat in the regenerator reboilers to strip the solvent of sulfur compounds. There is a potential to integrate the AGR refrigeration unit with the ammonia synthesis refrigeration unit.

F-6 Ammonia Synthesis

The ammonia synthesis reaction is exothermic. The heat of reaction is recovered generating HP steam. This steam is superheated in the HRSG and integrated into the steam power cycle. The refrigeration unit in the ammonia synthesis unit is integrated with the makeup gas compressor to reduce the number of stages required and the parasitic power load by chilling the feed to the first stage.

Integration of the ammonia syntheses refrigeration unit with the product CO_2 compressor was considered but ruled out as not cost effective.

There is a potential to integrate the ammonia synthesis refrigeration AGR refrigeration unit in further generations of the facility.

F-7 Power Block

The power bock is integrated with almost all the unit operations in the plant.

It provides:

- Deaerated BFW to the gas cooling unit to produce IP steam in the gasifier and the shift unit and LP steam in the gas cooling unit.
- IP steam to the shift unit
- Preheated BFW to the ammonia synthesis loop to raise HP steam

It receives:

- Hot condensate from the coal dryer and oxygen preheaters
- Energy from the gas cooling unit to preheat the DMW feeding the deaerator
- Energy from the gas cooling unit to preheat the gas turbine fuel
- LP steam from the gas cooling unit for use in the deaerator
- HP saturated steam from the ammonia synthesis loop for use in the steam cycle

The energy integration of the plant is intended to improve the overall efficiency of the plant. Care is taken in the design so that it is flexible, allows the plant to start up smoothly and to be able to move between operating points with ease.

Appendix G. Carbon and Sulfur Balance

Appendix G presents Carbon Balance (Exhibits G-1 to G-4) and Sulfur Balance (Exhibit G-5) tables for operating points beyond what was presented previously in *Section 3.1*.

Exhibit G-1: *Balanced Generation, 2 GTs* Carbon Balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	45,172 (99,588)	Emitted to Atmosphere	4,461 (9,835)
Air (CO_2)	112 (246)	CO_2 Product	39,920 (88,008)
		Gasifier Waste	903 (1,991)
Total	45,284 (99,834)	***Total***	45,284 (99,834)

$$\left(1-\left(\frac{Carbon\ to\ Atmosphere}{(Total\ Carbon\ In)-(Carbon\ in\ Gasifer\ Waste)}\right)\right)*100 \qquad \textit{Eq. 3-1}$$

$$\left(1-\left(\frac{4{,}461}{(45{,}284-903)}\right)\right)*100=90\%$$

Exhibit G-2: *Zero Net Power* Carbon Balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	29,850 (65,808)	Emitted to Atmosphere	2,717 (5,989)
Air (CO_2)	54 (119)	CO_2 Product	26,591 (58,623)
		Gasifier Waste	596 (1,314)
Total	29,904 (65,927)	***Total***	29,904 (65,927)

$$\left(1-\left(\frac{Carbon\ to\ Atmosphere}{(Total\ Carbon\ In)-(Carbon\ in\ Gasifer\ Waste)}\right)\right)*100 \qquad \textit{Eq. 3-1}$$

$$\left(1-\left(\frac{2{,}717}{(29{,}904-596)}\right)\right)*100=91\%$$

Exhibit G-3: *High Electricity* Carbon Balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	45,172 (99,588)	Emitted to Atmosphere	4,503 (9,928)
Air (CO_2)	153 (338)	CO_2 Product	39,919 (88,007)
		Gasifier Waste	903 (1,991)
Total	45,326 (99,926)	***Total***	45,326 (99,926)

$$\left(1-\left(\frac{Carbon\ to\ Atmosphere}{(Total\ Carbon\ In)-(Carbon\ in\ Gasifer\ Waste)}\right)\right)*100 \qquad \textit{Eq. 3-1}$$

$$\left(1-\left(\frac{4{,}503}{(45{,}326-903)}\right)\right)*100=90\%$$

Exhibit G-4: *Max Electricity* Carbon Balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	45,172 (99,588)	Emitted to Atmosphere	4,503 (9,928)
Air (CO_2)	153 (338)	CO_2 Product	39,920 (88,008)
		Gasifier Waste	903 (1,991)
Total	45,326 (99,926)	***Total***	45,326 (99,926)

$$\left(1-\left(\frac{Carbon\ to\ Atmosphere}{(Total\ Carbon\ In)-(Carbon\ in\ Gasifer\ Waste)}\right)\right)*100 \qquad \textit{Eq. 3-1}$$

$$\left(1-\left(\frac{4{,}503}{(45{,}326-903)}\right)\right)*100=90\%$$

Exhibit G-5: *Zero Net Power* Sulfur Balance

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Coal	1,174 (2,588)	Emitted to Atmosphere	-
		CO_2 Product	5 (11)
		Elemental Sulfur	1,169 (2,576)
Total	1,174 (2,588)	***Total***	1,174 (2,588)

Appendix H. Execution Plan

The Execution Plan has been developed to present a plan to implement the polygeneration plant subsequent to the completion of this report. While the plan presents no lapse in time between the completion of this report and the beginning of subsequent activities, this is presented in this manner for illustrative purposes only, and as such the first activity in the Execution Plan effectively begins with line item #2. Additionally, the Execution Plan has been developed in advance of specific details such as the identification and contract negotiation with a pilot plant host facility or the identification of a specific plant location. As such, the plan necessarily includes assumptions, such as the specifics of integrating with the host site, as well as identifying a project sponsor with the financial capability to provide pre-development funds and attract construction financing. Similarly, as there are no specific historical examples of developing and constructing this particular polygeneration plant, the Execution Plan schedule was developed in consultation with experts in project development, design, EPC, permitting and other relevant disciplines, and is developed based on average timelines for activities for plants of similar size and complexity. The authors recognize that each individual instance of development of these plants will have specific nuances that could result in longer or shorter timeframes for specific items.

Allegheny Science & Technology

Coal FIRST: Pre-FEED Study
Coal-Based Power Plants of the Future: Electricity and Ammonia Polygeneration
Execution Plan

April 14, 2020

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Agenda

- Current Project Description
- Execution Phases and Categorization
- Pilot Plant Phase
- Execution Plan Timeline – Pilot Plant Phase
- Execution Plan Timeline – Commercial Plant Phase
- Execution Plan Description
- Site Selection Details

Current Project Description

Coal-based polygeneration system to meet the needs of the evolving bulk power system

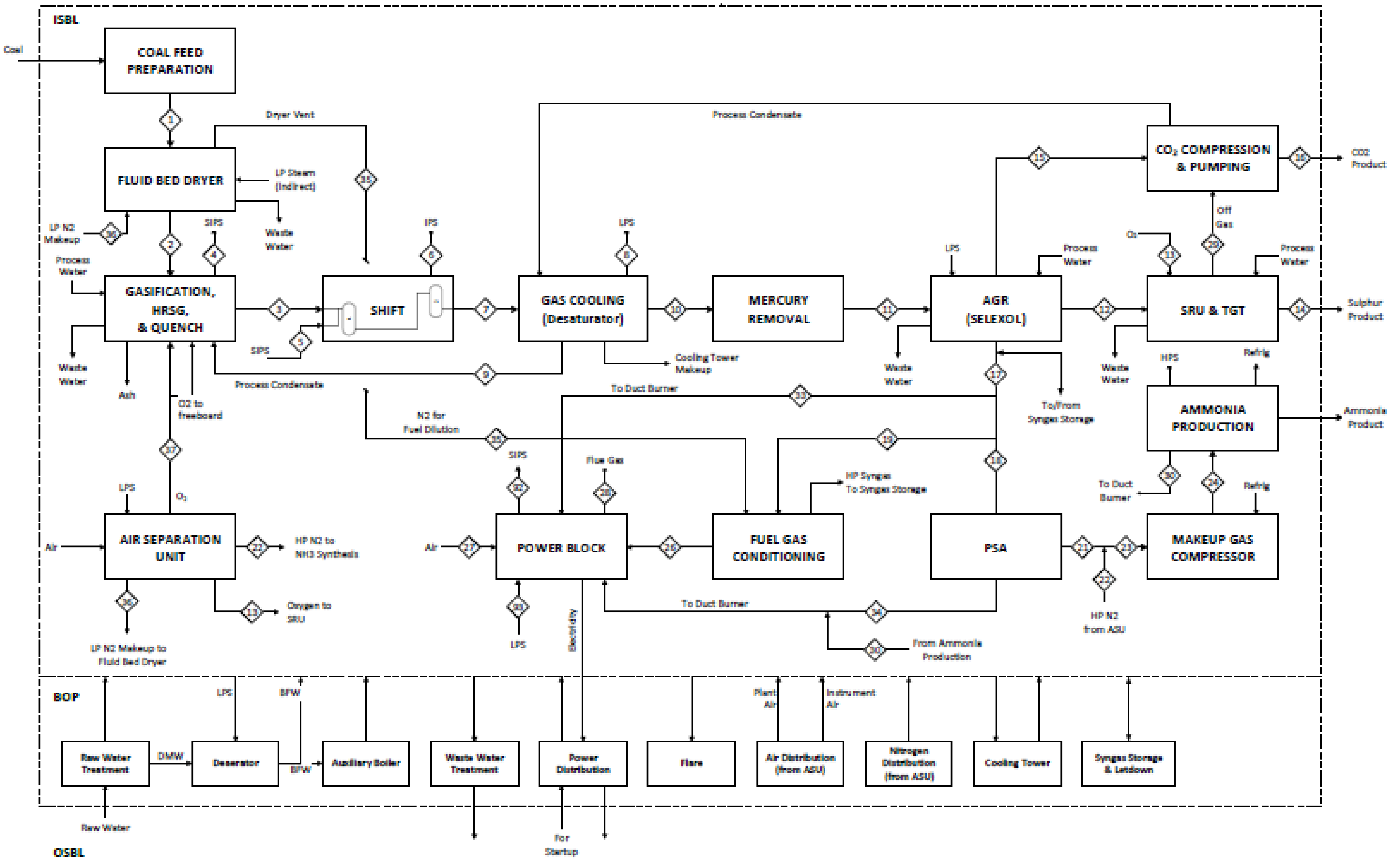
- High operational flexibility to respond rapidly to market conditions and signals, offer the ability to correctly match production to market demand
- Provide high operational efficiency while incorporating carbon capture rates of 90%

Business philosophy centers on offering multiple potential revenue streams

- Commercial electricity available for sale to the grid
- NH_3 for commercial delivery at or near retail (as opposed to wholesale) prices
- Saleable ancillary services

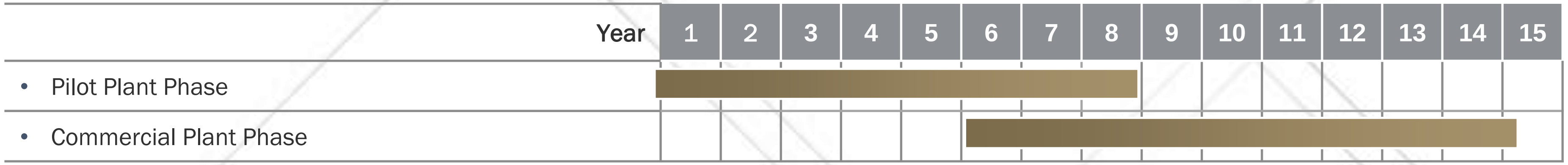
Operating Point	Net Export Power	Ammonia Production	Gasifier Operation	GT Operation	ST Operation	Ammonia Loop Operation
Balanced Generation, 3 GT's	48 MW	600 MTPD	100% of Capacity	Three Turbines @ 67% Capacity	Primary ST @ 86% load	Both Trains @ 100% Capacity
Balanced Generation, 2 GT's	51 MW	600 MTPD	100% of Capacity	Two Turbines @ 100% Capacity	Primary ST @ 91% Load	Both Trains @ 100% Capacity
Net Zero Power	0 MW	600 MTPD	66% of Capacity	One Turbine at 67% Capacity	Primary ST @ 40% Load	Both Trains @ 100% Capacity
High Electricity Production	82 MW	380 MTPD	100% of Capacity	Three Turbines @ 100% Capacity	Primary ST @ 88% Load	Both Trains @ 63% Capacity
Max Electricity Production	112 MW	59 MTPD	100% of Capacity	Three Turbines @ 100% Capacity	Primary ST @ 100% Load, Secondary ST @ 85% Load	Both Trains @ 10% Capacity

Current Project Description - Block Flow Diagram



Execution Plan Phases and Categorization

Execution Plan Timeframe



Execution Plan Step Categorization

Key	Element	Execution Plan Steps
	Non-Commercial Component Development	18-19
	Project Financing	2, 15, 20, 31, 43, 45, 49
	Site Selection	4, 26-30, 44
	Partner with Technology	8-12, 32-35
	Permitting	36-38, 50-51, 56
	Detailed Design	1, 3, 5-7, 14, 21-25, 39-42, 46-48
	Construction	13, 16-17, 52-55

Execution Plan Comments

- Given the mature nature of our core unit operations and the manageable level of technical risks associated with system integration, a higher risk approach of omitting the pilot phase could be considered to accelerate deployment
- In this instance, the first commercial application would be a pioneer plant with the understanding that evaluation of the pilot plant objectives would come during initial pioneer plant operations leading to an improvement based turnaround that implements the learning of the pilot stage investigations undertaken during the early operations of the pioneer plant
- We recommend separate piloting and commercial phases, particularly as successful pilot operations will prove out a reduced project risk level and lead to better financing terms

Pilot Plant Phase

Allow for rational development of controls, validate transitions and performance, and spur additional innovation

- As the plant is designed to be subject to frequent and rapid transitions, it is critical to validate transient characteristics agree with projected performance
- Implementation, observation and refinement of plant controls in the pilot plant will serve to mitigate risks of poor plant controls while verifying real-world operational performance, especially in regards to transient performance

Verify the ability to use partial oxidation in the freeboard oxygen-injection to limit methane content in the syngas to ~1%

Ensure the fluid bed dryer operates as intended, particularly with respect to the content of the overhead stream

- Verify that the conditions required to adequately dry the coal from ~11% moisture content to ~5% moisture content does not result in significant concentrations of hydrocarbons in the dryer overhead stream
- Verify mercury content of the overhead stream. If significant mercury exists, reassess overhead stream re-integration strategy

Key Pilot Plant Assumptions

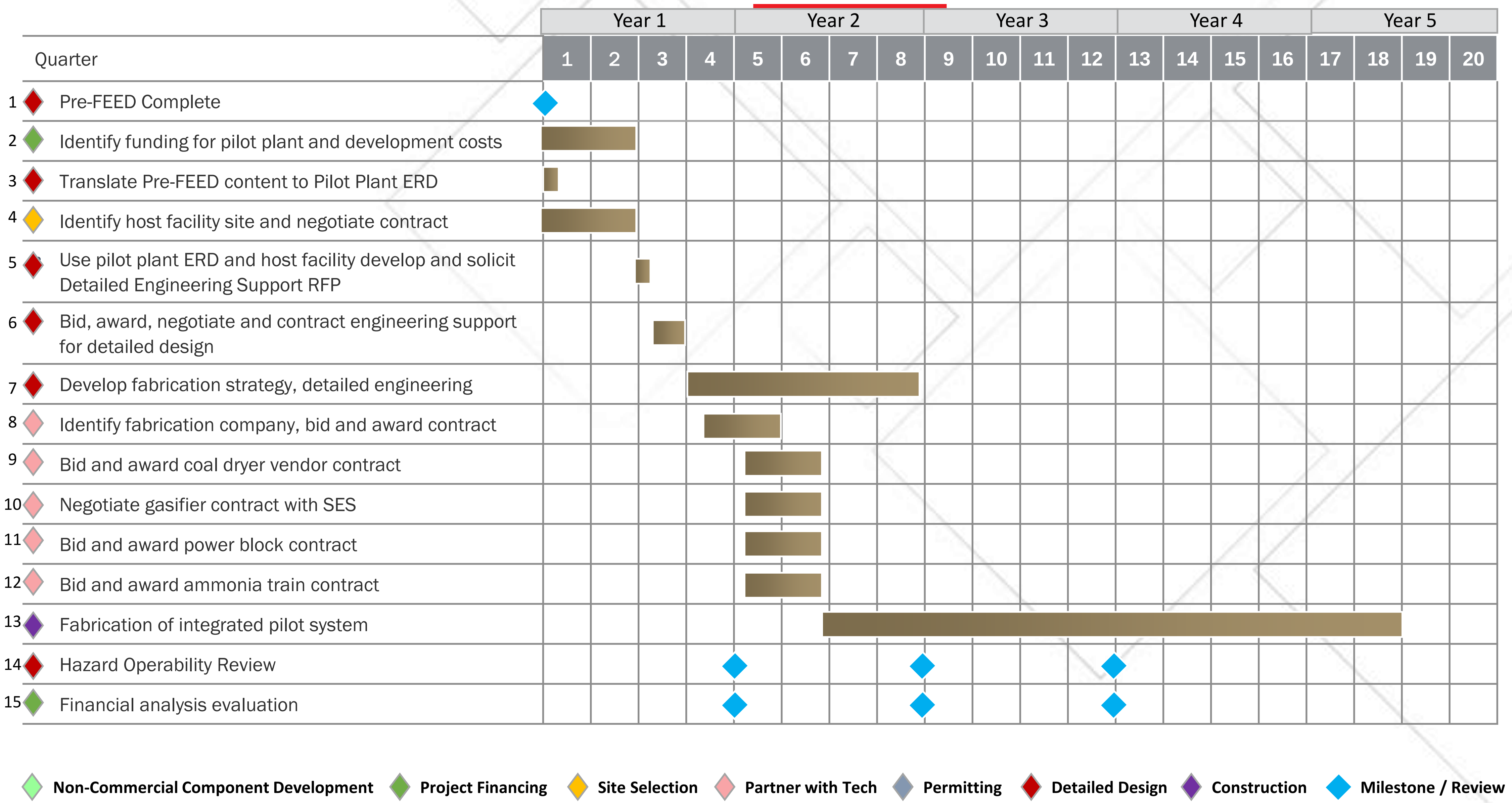
Available host site that can be used for pilot plant operations

All required infrastructure provided by/already in place at the host site

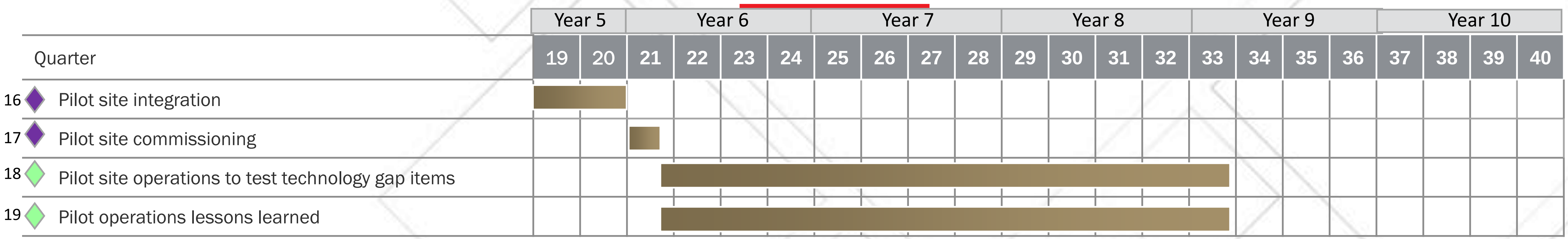
- Transportation infrastructure (roads, rail spurs, etc.)
- Complete complement of offsites, utilities, and electrical support systems (i.e. balance of plant and OSBL needs are met)

Permitting part of the general host facility permit (take advantage of permit by rule, facility level permitting constructs)

Execution Plan Timeline – Pilot Plant

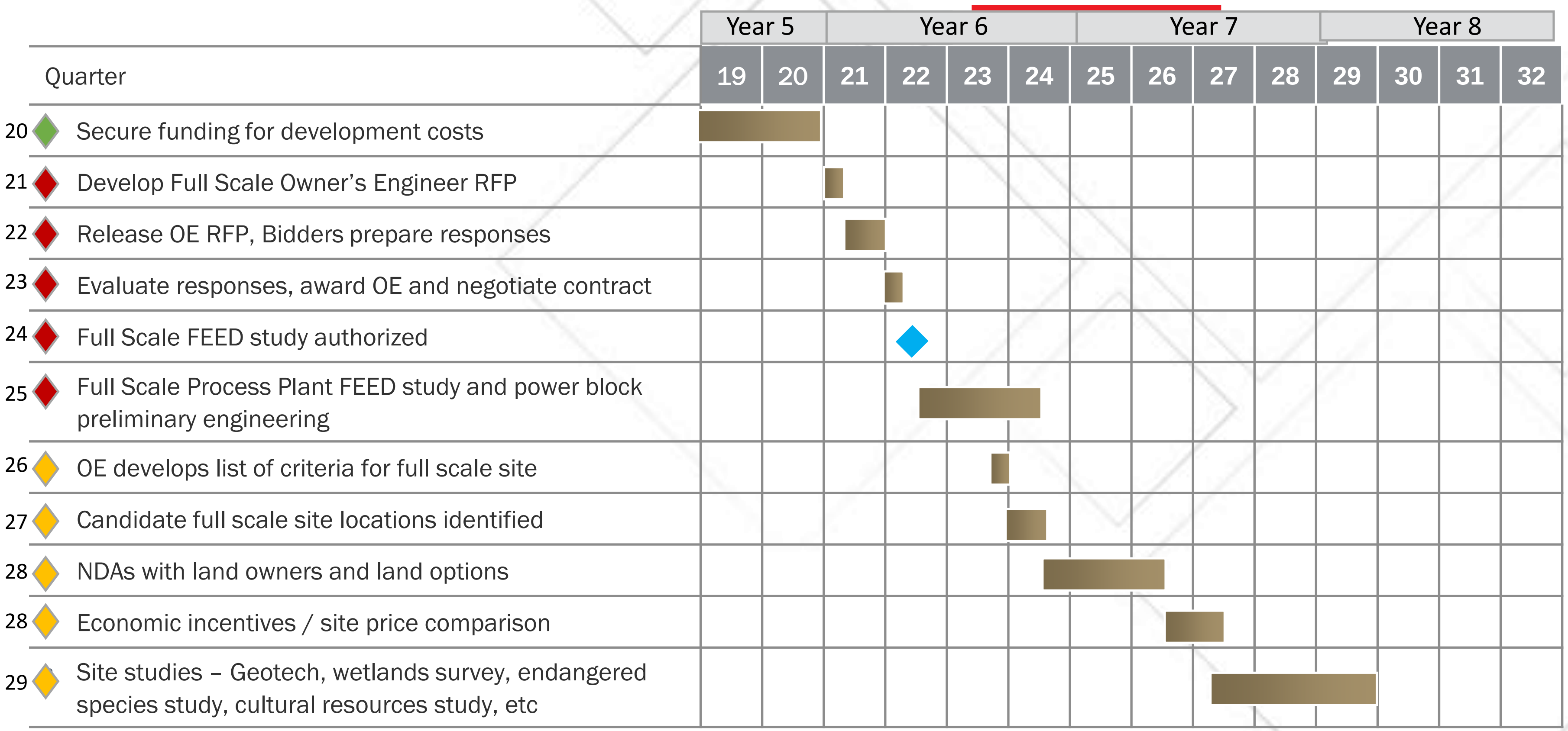


Execution Plan Timeline – Pilot Plant



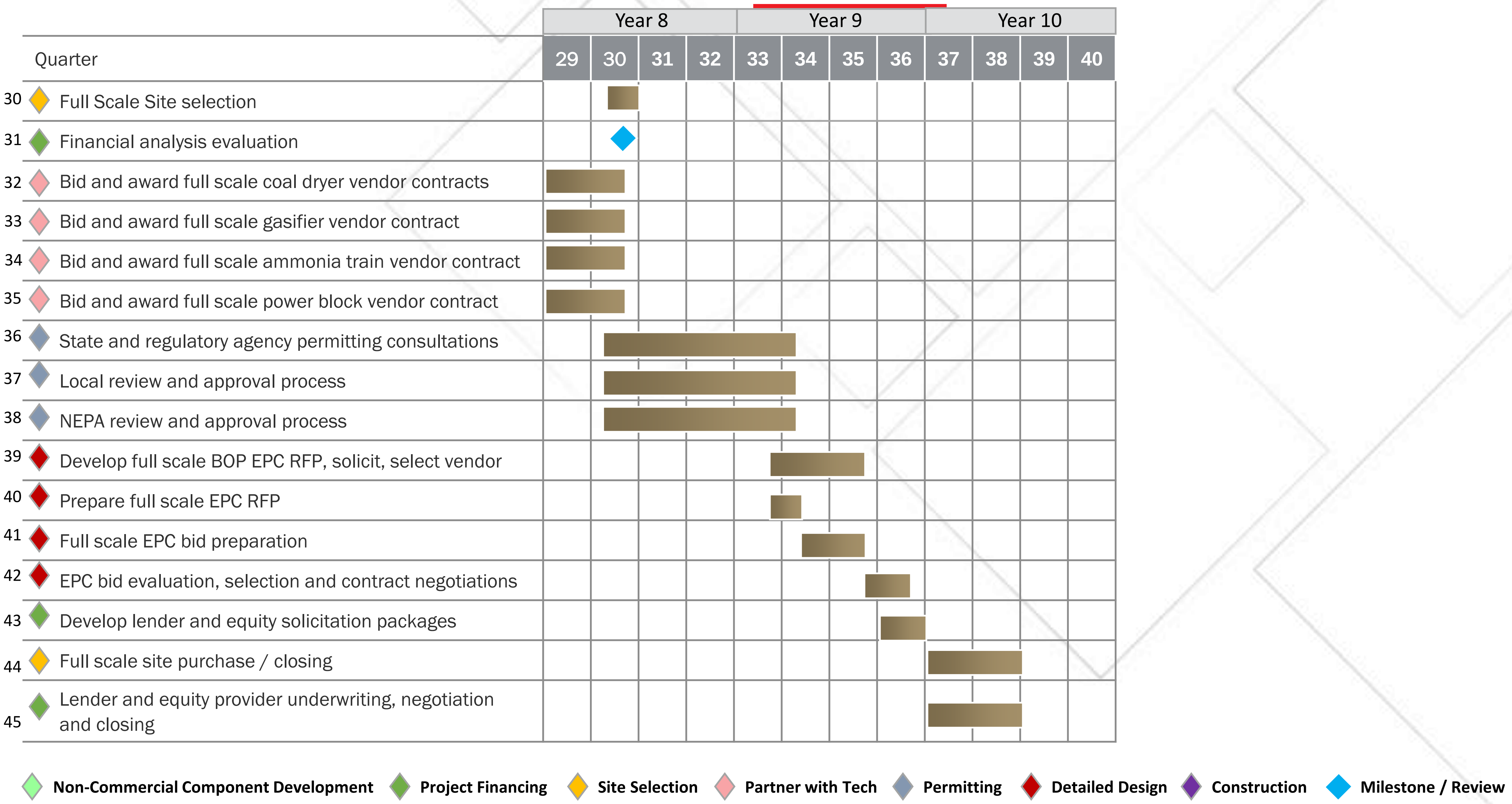
Non-Commercial Component Development · Project Financing · Site Selection · Partner with Tech · Permitting · Detailed Design · Construction · Milestone / Review

Execution Plan Timeline – Commercial Plant

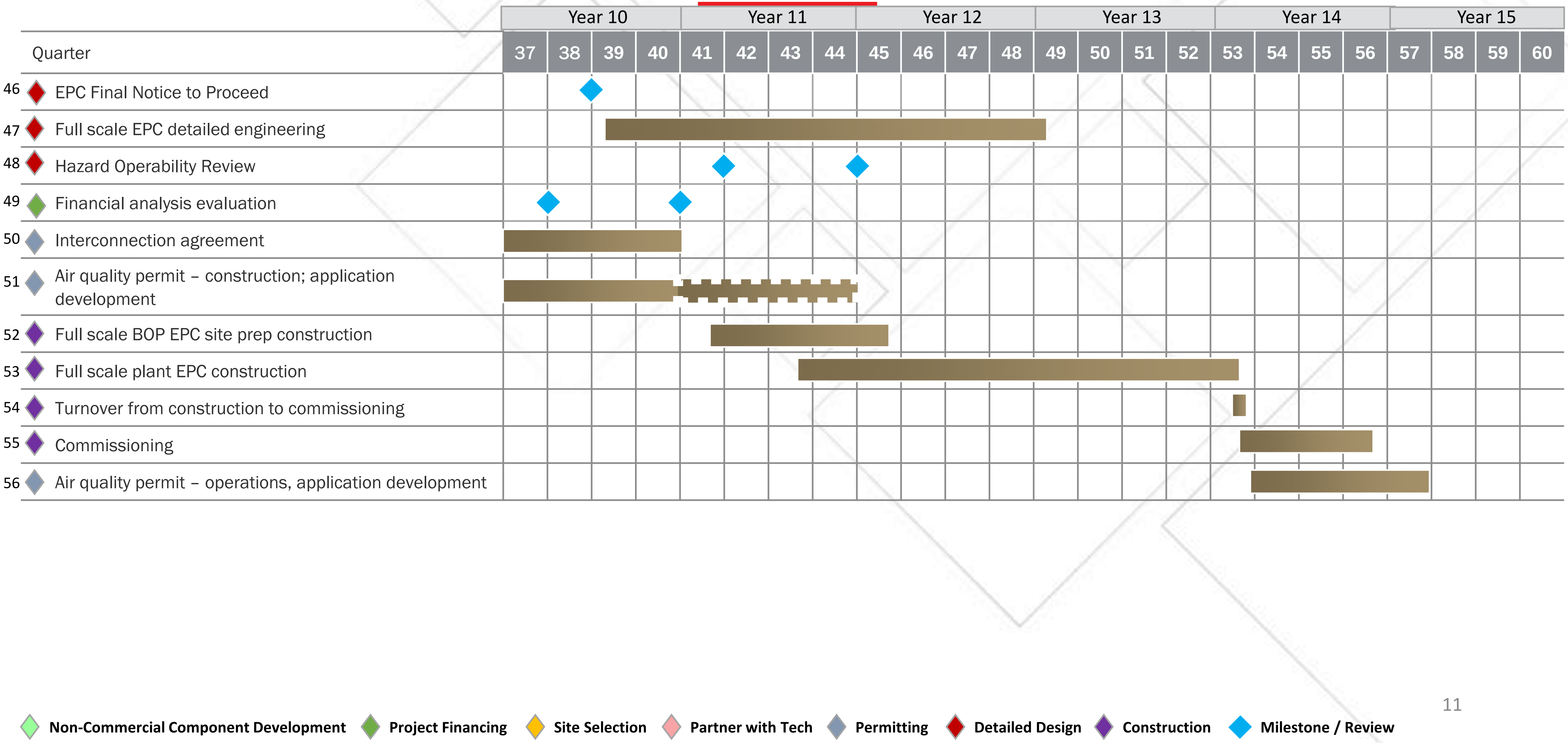


Non-Commercial Component Development · Project Financing · Site Selection · Partner with Tech · Permitting · Detailed Design · Construction · Milestone / Review

Execution Plan Timeline – Commercial Plant



Execution Plan Timeline – Commercial Plant



Execution Plan Description

#	Line Item	Description	Outcome
1	Pre-FEED Complete	Conclusion of CoalFIRST phase II	Completed Pre-FEED study
2	Identify funding for pilot plant and development costs	As the objectives of the pilot plant are not based on achieving economic profit, funding for the pilot plant and development costs will by necessity be equity funds from an investor, rather than debt financing which would be expected for later phases of the project. It is possible for funding activities to take up to 24 months, however in these cases development funding is typically provided by the developer to begin early activities such as RFP development, host site identification, etc	Pilot plant funding
3	Translate Pre-FEED content to Pilot Plant Engineering Requirements Document (ERD)	Pre-FEED study output will provide the detailed specifications for an engineering services firm to support project and work with pilot facility and fabricator, and work with fabricator on detailed engineering of the combined facility	Detailed specifications and requirements for engineering services support contract
4	Identify host site and negotiate contract	Piloting timeline assumes the ability to find an amenable functioning pilot plant hosting facility (cf. National Carbon Capture Center, GTI, U-ND EERC, UPARC, or a partner site), which can provide existing infrastructure including offsites, utilities, electrical, civil works, and safety can be leveraged. Additionally host site is assumed to have overall blanket permits and site work complete, such that site due diligence activities are not necessary at this stage and air permits are covered by the host site's existing permits ("permit by rule," inclusion in the facility permit, etc.). This enables the pilot plant phase to focus on capturing core technical aspects and facility integration experience to lower the risk of commercial development, rather than spending time during the piloting phase focused on the tactical issues of permits and optimizing development activities	Host site identified and contracted, understand what elements of pilot development are included in host site and what will need to be provided (blanket permits, civil infrastructure, etc.)
5	Use pilot plant ERD and host facility develop and solicit detailed engineering support RFP	Time allotted for pre-bid meeting, defined period for questions, answers prepared and presented by development team. During response time developer should begin to prepare for bid evaluation, including preparing evaluation materials, meetings, and ensuring development team is fully staffed	Proposals for vendors to perform engineering support tasks (mix of OE and EPC-like tasks) for the core development team

Execution Plan Description

#	Line Item	Description	Outcome
6	Bid, award, negotiate and contract engineering support for detailed design	Evaluation based on predetermined scoring criteria, as well as contract terms and conditions redlines and cost proposal; engineering support will likely come from an EPC firm, however at this stage neither the procurement nor construction is anticipated to be in the detailed engineering support scope of work, as those elements will be driven by the project developer, host site and fabrication company for the pilot plant	Executed engineering support contract
7	Develop fabrication strategy, detailed engineering	Engineering support contractor will develop detailed engineering for unit integration as well as strategy to have units fabricated and constructed on a skid to be delivered to host site; specifics of fabrication strategy are directly tied to host site selection regarding scheduling, what infrastructure exists, etc	Fabrication strategy and detailed engineering
8	Identify fabrication company, bid and award contract	Fabrication strategy work will inform the requirements for a fabrication company to assemble the unit operations on a skid and deliver to the host site; as seen on the timeline graphics identifying the fabrication company happens in parallel with the development of the fabrication strategy, this is to ensure the fabrication company is able to provide their input on strategy development and best practices	Fabrication company contract
9	Bid and award coal dryer vendor contract	In conjunction with fabrication strategy and fabrication vendor, engineering support company will develop duty specs based on the Pre-FEED work and detailed design to develop coal dryer RFP, solicit and evaluate bids, award and negotiate a purchase contract	Coal dryer vendor contract
10	Negotiate gasifier contract with SES	In conjunction with fabrication strategy and fabrication vendor, engineering support company will use duty specs from Pre-FEED work and detailed design and work with SES to negotiate a contract for the gasifier unit	Gasifier contract
11	Bid and award power block vendor contract	In conjunction with fabrication strategy and fabrication vendor, engineering support company will use duty specs from Pre-FEED work and detailed design to develop power block RFP, solicit and evaluate bids, award and negotiate a purchase contract	Power block vendor contract

Execution Plan Description

#	Line Item	Description	Outcome
12	Bid and award ammonia train vendor contract	In conjunction with fabrication strategy and fabrication company, engineering support company will use duty specs from Pre-FEED work and detailed design to develop ammonia train RFP, solicit and evaluate bids, award and negotiate a purchase contract	Ammonia train vendor contract
13	Fabrication of integrated pilot system	Fabrication company will assemble the unit operations into a single system and deliver to the site	Skid fabricated unit
14	Hazard Operability (HAZOP) Review	Safety and operability review of design to ensure intended operations do not pose a safety hazard and equipment will function as intended; especially relevant to assess additional hazards of rapid, repeated ramping and turndown; develops list of action items and required design modifications to assure safety and operability	Hazard Operability Review report and associated action items tracked to closure
15	Financial analysis evaluation	Periodic exercising and updates of the financial model as project specific parameters and costs continue to be informed and refined, market conditions for commodity inputs / outputs change, and overall economic landscape evolves. Informs detailed design choices, vendor/site selections, and negotiations. Helps obtain funding (debt and equity) and serves as "gate review" on continuing the technology deployment.	Updated financial model, improved guidance for commercial deployment and negotiation
16	Pilot site integration	After delivery of fabricated skid to host site, engineering support firm will work with host site to integrate the skid into the existing infrastructure of the host site, and construct / procure / install any integration components necessary	Integrated pilot plant
17	Pilot site commissioning	Preparation for commissioning can start when pilot site integration is ~60% complete, acceptance follows performance testing and punch list completion	Equipment capable of running safely and robustly
18	Pilot site operations to test technology gap items	Operating pilot plant to ensure smooth performance as expected, as well as testing operational procedures as outlined in technology gap report, ensuring controls perform appropriately, determine appropriate methods for transitioning between operating points, etc	Mitigation of technical risks, improved control strategy, more complete operating procedures/manuals
19	Pilot operations lessons learned	Document all lessons learned and results of pilot operations to ensure full scale plant design incorporates needed experience	Translate experience into operating manual and final detail design choices

Execution Plan Description

#	Line Item	Description	Outcome
20	Secure funding for development costs	Identify and allocate funding to fund development costs until financing for plant can be secured, including Owners Engineer contract, potentially some portion of EPC contract, and site options and/or purchase; not all funds may be delivered at this stage but firm commitments must be obtained	Funding for development costs
21	Develop Owner's Engineer RFP	OE scope includes FEED study completion, preliminary engineering for power block, turbine selection RFP, EPC bid RFP, BOP EPC RFP, permitting, site selection, review of EPC detailed design, construction surveillance, on site / off site QA/QC support, commissioning assistance. During this time developer will also prepare the evaluation criteria, both the public facing elements to be presented in the RFP but also the forms and format of bidder response evaluation. Additionally, RFP should include a draft contract Terms and Conditions for bidders to review	Specifications for Owners Engineer work, including RFP documents and evaluation forms
22	Release OE RFP, Bidders prepare responses	Time allotted for pre-bid meeting, defined period for questions, answers prepared and presented by development team. During response time developer should begin to prepare for bid evaluation, including preparing evaluation materials, meetings, and ensuring development team is fully staffed	Proposals from companies to perform Owners Engineering tasks
23	Evaluate responses, award OE and negotiate contract	Evaluation based on predetermined scoring criteria, as well as contract terms and conditions redlines and cost proposal	Finalized Owners Engineer contract
24	FEED study authorized	Coincides with final contract signed, budget for OE work must be secured	OE may begin FEED study

Execution Plan Description

#	Line Item	Description	Outcome
25	Full scale process plant FEED study, power block engineering, BOP engineering, system integration engineering	This plan anticipates the OE completing the FEED study for the process plant and integration, and engineering for coal dryer, gasifier, ammonia train and power block to get to an RFP for fixed price quotes from an EPC/Vendor. Work includes conceptual and detailed engineering for modularization, EPC bid RFP, refined cost estimate, air permit support, construction planning study, noise engineering, general arrangement drawings, construction planning study, civil works study, foundation designs, grading plans, underground piping & electrical engineering. Additional detailed engineering for bubbling bed coal drying at scale and with proposed coal feed will be required; demonstration of partial oxidation in the gasifier freeboard; process controls development; operations and transition detailed dynamic modeling; note these elements will also be addressed as part of pilot plant operations and lessons learned	FEED study outputs, such as functional specs for unit operations RFPs, process engineering, general arrangement drawings, cost estimate, construction planning, foundation designs, grading plans, civil works study, underground piping and electrical engineering, support for permits and site studies
26	OE develops list of criteria for site	Acreage, access to feedstocks, general topography, greenfield/brownfield requirements, distance to / access to rail (expected Class 1 service needed), electrical infrastructure, ammonia infrastructure if desired, CO2 pipeline, water, natural gas; zoning needs	Prioritized list of characteristics needed for project site
27	Candidate site locations identified	Pilot scale site location driven by most amenable and available host facility location. Many firms specialize in site identification, including EPC firms. Companies may have databases of potentially available land, and / or contacts with a variety of economic development agencies across jurisdictions to both help identify potential sites and begin to develop economic incentives provided by municipalities or states for the job creation and increased tax basis provided via the development	Pilot stage – this step provides final site location. Full scale stage produces short list (<~10) of potential sites identified

Execution Plan Description

#	Line Item	Description	Outcome
28	NDAs with land owners and land options	Multiple sites will be put under contract, typically via options to purchase the parcels which allow for the developer to decline to purchase the property as well as perform a variety of environmental surveys and site investigation studies. NDAs are particularly important, as large tracts of land are often owned by multiple different owners, such that the selection team will need to negotiate purchase options with several owners simultaneously, and need to avoid existing land owners communicating and potentially colluding to increase the land offer price.	Nondisclosure agreements with land owners and purchase options executed; ~3 – 5 sites
28	Economic incentives / site price comparison	Negotiations with municipalities and states can produce economic incentives to develop the project in a specific area; these incentives are viewed in conjunction with the land purchase option prices in order to fully understand the costs of specific parcels	Detailed price for each site, including purchase price and local economic incentives
29	Site studies – Geotech, wetlands survey, endangered species study, cultural resources study, background air quality measurements	Site purchase option should include ability to perform tests and surveys in order to determine financial suitability of site; all point source air emissions should be known at this point such that background air quality measurements and computational fluid dynamic modeling can be done to see how the new emissions sources will affect the ambient conditions. Unfavorable results likely rule out proposed parcels, while successful modeling enables preparation of the air permit application. Other studies typically must consider: feedstock supply and product delivery availability review, raw water sources and characteristics, water supply due diligence, waste water discharge provisions/agreements, air permit and other permit preliminary review to determine ability of site to receive permits, surveys/topography of site, initial threatened/endangered species consultation, wetland delineation and stream investigation, cultural resources investigation, development of preliminary environmental impacts, hydrographic surveys, phase I / II ESA	Site due diligence studies

Execution Plan Description

#	Line Item	Description	Outcome
30	Full scale site selection	Final site selection will be done in consideration of financial modeling of the specific impacts a site has on the project's financial return, including local economic incentives, costs of civil improvements, access (or cost to access) to feedstock and product delivery points, as well as the results of site studies. Firms exist to help with the site selection process (many EPC firms also provide this service), these companies can work through local economic development agencies, or may have their own databases, to identify available land, surrounding infrastructure, etc; some of these firms will also do wetlands surveying, endangered species studies, cultural resources studies, and other of the site due diligence studies	Final site selection
31	Financial analysis evaluation	Periodic exercising and updates of the financial model as project specific parameters and costs continue to be informed and refined, market conditions for commodity inputs / outputs change, and overall economic landscape evolves. Informs detailed design choices, vendor/site selections, and negotiations. Helps obtain funding (debt and equity) and serves as "gate review" on continuing the technology deployment.	Updated financial model, improved guidance for commercial deployment and negotiation
32	Bid and award coal dryer vendor contract	Work facilitated by OE in advance of EPC contract bids in order to reduce EPC firm's risk and thus improve EPC fixed price contract pricing; RFP will indicate intent to have unit shop fabricated versus assembled on site	Final price and contract for coal dryer
33	Bid and award gasifier vendor contract	Work facilitated by OE in advance of EPC contract bids in order to reduce EPC firm's risk and thus improve EPC fixed price contract pricing; RFP will indicate intent to have unit shop fabricated versus assembled on site	Final price and contract for gasifier
34	Bid and award ammonia train vendor contract	Work facilitated by OE in advance of EPC contract bids in order to reduce EPC firm's risk and thus improve EPC fixed price contract pricing; RFP will indicate intent to have unit shop fabricated versus assembled on site	Final price and contact for ammonia train
35	Bid and award power block vendor contracts	Work facilitated by OE in advance of EPC contract bids in order to reduce EPC firm's risk and thus improve EPC fixed price contract pricing; RFP will indicate intent to have unit shop fabricated versus assembled on site	Final price and contract for power block

Execution Plan Description

#	Line Item	Description	Outcome
36	State and Federal regulatory agency permitting consultations	Note that different states have different permit application processes; a multi-technology application will allow the project to provide details for multiple generators in the permit application, and then revise the permit when a specific vendor has been selected. Other states will revert the application to the beginning of the process once the specific vendor is selected, thus impacting the timing of permit issuance. Based on the state's application process, permitting may need to begin after technology vendor is selected. Proposed power block is fossil fuel-fired combustion devices used to generate electricity for sale and serve as a generator over 25 MWe. Therefore, plant will meet the definition of an affected Phase II "utility unit" under the Acid Rain Program pursuant to the Clean Air Act, and require a phase II acid rain permit.	Detailed list of necessary permits, application processes and application timelines
37	Local review and approval process	Specific city or county jurisdictional agencies may require planning and land use approvals that require independent review and approval processes from the State and Federal environmental entitlements. Land approvals may include zoning change, conditional use permits, or various planning approvals specific to setbacks, siting and other general plan exemptions. Additionally, local agencies will be involved in transportation and traffic permits; building and engineering reviews and permits; grading and drainage plan approvals; stormwater pollution protection plan approval; hazardous waste materials generation, collection, handling or transport; onsite waste water treatment facilities; infrastructure; and wastewater discharge activities	Local approvals
38	NEPA review and approval process	Environmental documentation for either Categorical Exclusion (CA), Environmental Assessment (EA) or Environmental Impact Statement (EIS), based on project's potential to have significant environmental impacts and the involvement of Federal funds or permitting. Many of the studies completed in the Site Studies step will be applicable to this process	NEPA categorization, application and approval
39	Develop full scale BOP EPC RFP, solicit, select vendor	RFP for site construction activities outside of main plant components and integration, such as site preparation, road / parking infrastructure, administrative buildings, rail infrastructure	BOP EPC firm contract finalized

Execution Plan Description

#	Line Item	Description	Outcome
40	Prepare full scale EPC RFP	Request for fixed price EPC bid in order to transfer construction risk to EPC firm, however to do this all detailed design work and site selection activities must be complete and permitting largely de-risked. During this time developer will also prepare the evaluation criteria, both the public facing elements to be presented in the RFP but also the forms and format of bidder response evaluation. Additionally, RFP should include a draft contract Terms and Conditions for bidders to review	Specifications for EPC work, including RFP documents and evaluation forms
41	Full scale EPC bid preparation	Time allotted for pre-bid meeting, site visit for bidders to walk the identified parcel, defined period for questions, answers prepared and presented by development team. During response time developer should begin to prepare for bid evaluation, including preparing evaluation materials and clear evaluation methodology	Final proposals from EPC firms
42	EPC bid evaluation, selection and contract negotiations	Evaluation based on predetermined scoring criteria, as well as contract terms and conditions redlines and cost proposal	Negotiated final contract for EPC vendor
43	Develop lender and equity solicitation packages	Economic and business model details, financial analysis of project, sensitivities to key risks and mitigation plans, expected returns for investors, required amount from equity and debt, preferred types of equity (i.e., preferred equity, non-voting, etc, depending on legal entity), detailed budget including fixed price EPC contract, progress towards permitting, studies from site selection activities. Financial projections are built from the periodic financial analysis evaluations (lines 15, 31 and 49). Packages will likely be delivered to several lenders and potential equity partners	Detailed business plan, financial model and project description
44	Full scale site purchase / closing	Land purchase price is likely to be a component of project developer's upfront equity contribution to the project, and thus option execution may occur prior to loan closing; purchase of land may also happen concurrent with loan closing but no later than loan closing as lender will place a lien on the property	Final site purchase
45	Lender and equity provider underwriting, negotiation and closing	Bank finance is likely to be a syndicate of lenders. While negotiation will occur with only the lead lender, the syndicate must come to consensus on loan terms and conditions, which adds time to the underwriting and negotiation phase.	Financing commitments

Execution Plan Description

#	Line Item	Description	Outcome
46	EPC Final Notice to Proceed	Final notice to the EPC firm to begin work and begin accumulating charges; note this notice will happen after a financial analysis evaluation and at or just after closing of the financing instrument	EPC start
47	EPC Engineering	Final detailed integration engineering by EPC firm	Detailed engineering
48	Hazard Operability Review	Safety and operability review of design to ensure intended operations do not pose a safety hazard; especially relevant to assess additional hazards of rapid, repeated ramping and turndown; Develops list of action items and required design modifications to assure safety and operability	Hazard Operability Review report and associated action items tracked to closure
49	Financial analysis evaluation	Periodic exercising and updates of the financial model as project specific parameters and costs continue to be informed and refined, market conditions for commodity inputs / outputs change, and overall economic landscape evolves. Informs detailed design choices, vendor/site selections, and negotiations. Helps obtain funding (debt and equity) and serves as "gate review" on continuing the technology deployment.	Updated financial model, improved guidance for commercial deployment and negotiation
50	Interconnection agreement	Prepare interconnection request with preliminary site documentation, expected in-service date and deposit; perform interconnection studies, negotiate schedule for constructing interconnecting facilities and network upgrades, finalize interconnection agreement	Interconnection agreement
51	Air quality permit – construction; application development	Depending on emissions expectations of FEED design, several requirements may be needed to be met for construction permit (from air quality perspective): best available control technology analysis, air quality analysis, and additional impact analysis. Application will include process description, process flow diagrams, plot identification, identification of applicable federal and state air regulations and emissions limitations, emissions quantification and application forms. Facility will also be required to apply for a phase II acid rain permit, install CEMS to demonstrate compliance with the ARP provisions meeting the requirements specified in 40 CFR 75, and hold allowance equivalent to annual NOx and SO2 emissions	Air permits
52	Full scale BOP EPC site prep construction	Mobilization, construction management trailers, site clearing, civil infrastructure, roads, grading, retaining walls, foundations, underground piping and electrical	Balance of plant construction

Execution Plan Description

#	Line Item	Description	Outcome
53	Full scale plant EPC construction	Integration of shop-fabricated coal drying, gasifier, power block, process plant, storage units, integration components, development of control room, underground utilities, electrical equipment installation, interconnection infrastructure, etc. A conservative timeframe based on standard construction timeframes for a plant of this scale is used to account for this being the first application, however for subsequent plants construction time would be reduced as shop fabricated unit construction and integration times will reduce	Mechanically complete plant
54	Turnover from construction to commissioning	Hand over of all relevant project documentation, completion of final punch list items and deconstruction of construction staging equipment	Start of commissioning work
55	Commissioning	Preparation for commissioning can start when plant is ~60% complete, acceptance follows performance testing and punch list completion	Equipment capable of running safely and robustly
56	Air quality permit – operations, application development	Anticipated that an operating permit under the Clean Air Act Permit Program (CAAPP) or Federally Enforceable State Operating Permit (PESOP) will be required. Title V Part 70 identifies the standard permit requirements that each permit shall include, including all monitoring and analysis procedures or test methods required; potential exclusions of testing or monitoring and compliance certification. CAAPP is generally required for “major source” emissions as defined by criteria pollutants (NOx, SOx, CO, Lead, Ozone, VOC, PM 2.5, PM10), limits based on Hazardous Air Pollutants (HAPs) in the Clean Air Act; FESOP is applicable when a facility can voluntarily limit emissions by accepting limits on operations.	Air permits for ongoing operations

Site Selection Details

Project Stage	*Comments*
Pilot Site Selection	• Pilot site selection includes an inherent assumption that the project sponsor can identify potential host sites, and as such site selection process does not start from scratch • Piloting timeline assumes the ability to find an amenable functioning pilot plant hosting facility (cf. National Carbon Capture Center, GTI, U-ND EERC, UPARC, or a partner site), which can provide existing infrastructure including offsites, utilities, electrical, civil works, and safety can be leveraged • Host site is assumed to have overall blanket permits and site work complete, such that site due diligence activities are not necessary at this stage and air permits are covered by the host site's existing permits ("permit by rule," inclusion in the facility permit, etc.)
Full Scale Site Selection	• Identifying a site for the full scale plant will be more complicated than the pilot plant site, as confidentiality during the search process is essential to ensure land prices are not inflated – One challenge in acquiring a large tract of land is the high likelihood that the project team will need to secure multiple smaller parcels of land to aggregate into a sufficient size. In this scenario, it is crucial to avoid landowners holding out on selling or signing purchase options with the hopes that they will be the last holdout – and thus inflate the cost of their land. For this reason nondisclosure agreements are key – Similarly, executing property purchase options (as opposed to immediately purchasing land) is typically done in order to have several sites moving down the due diligence and procurement path simultaneously, in the event there is a problem with a due diligence study or if a landowner reneges on the purchase option for any reason • There are firms that specialize in executing site selection, including some EPC firms. These companies may have databases of potentially available land, and additionally are typically informed as to the potential economic incentives that different municipalities/counties/states might be willing to provide • Typically the OE will provide the site selection firm with a list of criteria, such as access to feedstock / product offtake sources, distance to transmission lines, parcel size, preferred topology, etc • In most cases the project developer will purchase the land using equity funds, and as such the land closing would occur prior to the closing of a loan facility. By doing it in this order the lender is able to place a lien on the property as part of the loan collateral. It is possible to close the land purchase and loan closing simultaneously, but this does open the project developer to additional transactional risk should one of the two transactions need to be delayed