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Coal-Based Power Plants of the Future: Electricity and Ammonia Polygeneration Performance Results Report

Submitted To:

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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 Exhibit 1-1:

Exhibit 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
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/hr 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/hr 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 Exhibit 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* 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]. The general and specific siting characteristics are provided in the design basis report.

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.

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[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 Analysis Report*.

[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.

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 Exhibit 2-1 and 2-2 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.

Exhibit 2-1 Polygeneration Plant Block Flow Diagram

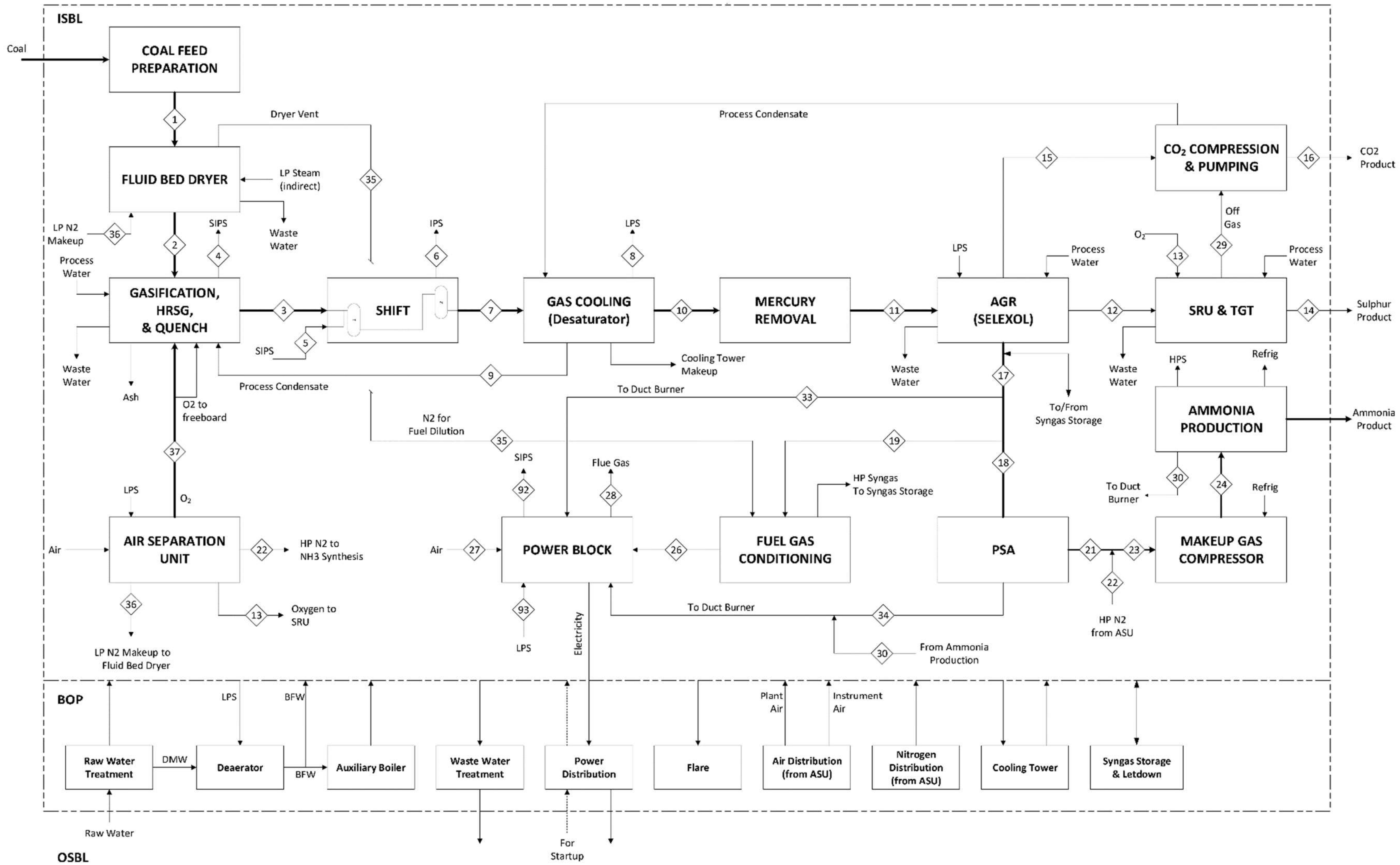

Exhibit 2-2: Balanced Generation, 3 GTs Stream Table/Heat and Mass Balance

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	
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

Exhibit 2-2: 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
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

Exhibit 2-2: 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	
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

Exhibit 2-2: 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
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

Exhibit 2-2: 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	
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

Exhibit 2-2: 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	
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

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- 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. 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[5]

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

[5] 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 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.

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.[6] 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:[7]

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 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.[8]

[6] 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).
[7] This was previously detailed and accepted in the Design Basis Report.
[8] 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

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.

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[9].

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 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

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through pilot plant testing.

[9] 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.

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/hr of nitrogen for system processes and ~50,000 kg/hr 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[10] (Eq. 2-2) reactions operate according to the following equations:

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

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

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 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/hr 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

[10] Note that this refers to steam methane reforming occurring within the gasifier through the aforementioned 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.

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[11]

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 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:

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

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

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/hr comprised primarily of CO_2 (~154,000 kg/hr) and H_2 (~10,000 kg/hr).

[11] 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.

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 H2S 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 H2S 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 H2S absorber, CO_2 absorber, H2S 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/hr 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[12], (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/hr with the balance (11,700 kg/hr) going to the PSA. Of the ~11,700 kg/hr to the PSA, ~4,400 kg/hr 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.

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

[12] The intended use of the storage tanks is to dampen the impacts of lagging system components during the transitions between operating modes. They are able to 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 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.

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.

2.1.10.1 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 Exhibit 2-3.

Exhibit 2-3: 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[13]. 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

[13] 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.

and provide sufficient feed to operate the ammonia trains at full capacity, the gasifier will be 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[14].

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

[14] 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

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 Exhibit 2-4, with a tabular representation in Exhibit 2-5.

Exhibit 2-4: 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

Exhibit 2-5: 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.

Exhibit 2-6: 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)[15]
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-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

[15] 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.

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.

Exhibit 2-7: 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) - Ammonia ramp in 40-50 minutes (0.8-1% per min) - Power Island turndown in ~5-10 minutes - Excess syngas to storage - Utilize aux boiler for feed preheat
Net Zero Power	0 MW	600 MTPD	66% of Capacity	One Turbine @ 67% Capacity	Primary ST @ 40% Load	Both Trains @ 100% Capacity	
Delta	-82 MW	+220 MTPD	34% Turndown	Two @ 100% Turndown, One @ 33% Turndown	Primary ST @ 48% Turndown	Both Trains @ 37% Ramp	

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.

Exhibit 2-8: 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 - Utilize NG or stored syngas for transition - Consider shut down of one ammonia train
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	

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.

Exhibit 2-9: 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 - Utilize aux boiler for feed preheat
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	

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.

Exhibit 2-10: 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

Exhibit 2-11 summarizes the results of the five transient case studies presented.

Exhibit 2-11: 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.[16]

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.

[16] 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.

Exhibit 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
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%[17]	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

[17] 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 Exhibit 3-2.

Exhibit 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 Exhibit 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.[18]

Particulate emissions from normal operation of the LM2500+ turbines has 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%.[19]

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 Exhibit 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[20], the captured CO_2 product, 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).

[18] 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.

[19] 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.)

[20] 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 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.

Exhibit 3-3: Carbon Balance[21]

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)[22]	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 Exhibit 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.

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

Exhibit 3-4: Sulfur Balance[23]

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)

Exhibit 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 Exhibit 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 Exhibit 3-6.

[23] 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 Exhibit 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.

Exhibit 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)
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)**

Exhibit 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**
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 Exhibits 3-7 and 3-8.

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.[24]
- 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 jurisdiction 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, etc.

[24] 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)

Exhibit 3-7 Polygeneration Plant Water Balance Block Flow Diagram

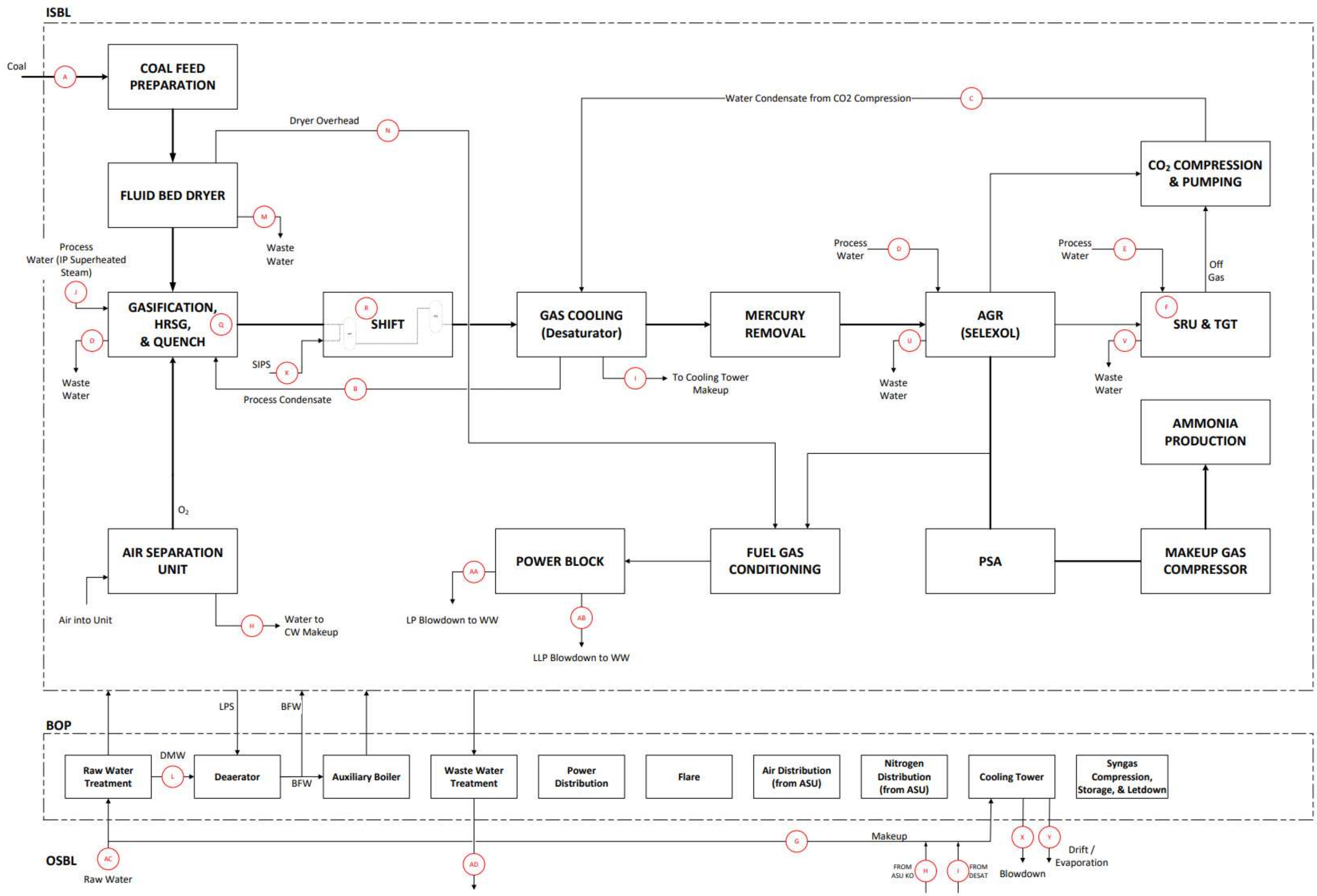


Exhibit 3-8 Polygeneration Plant Water Balance Stream Tables

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)
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 Exhibits 3-9 to 3-22. **Exhibit 3-9: Air Separation Unit Process Flow Diagram**

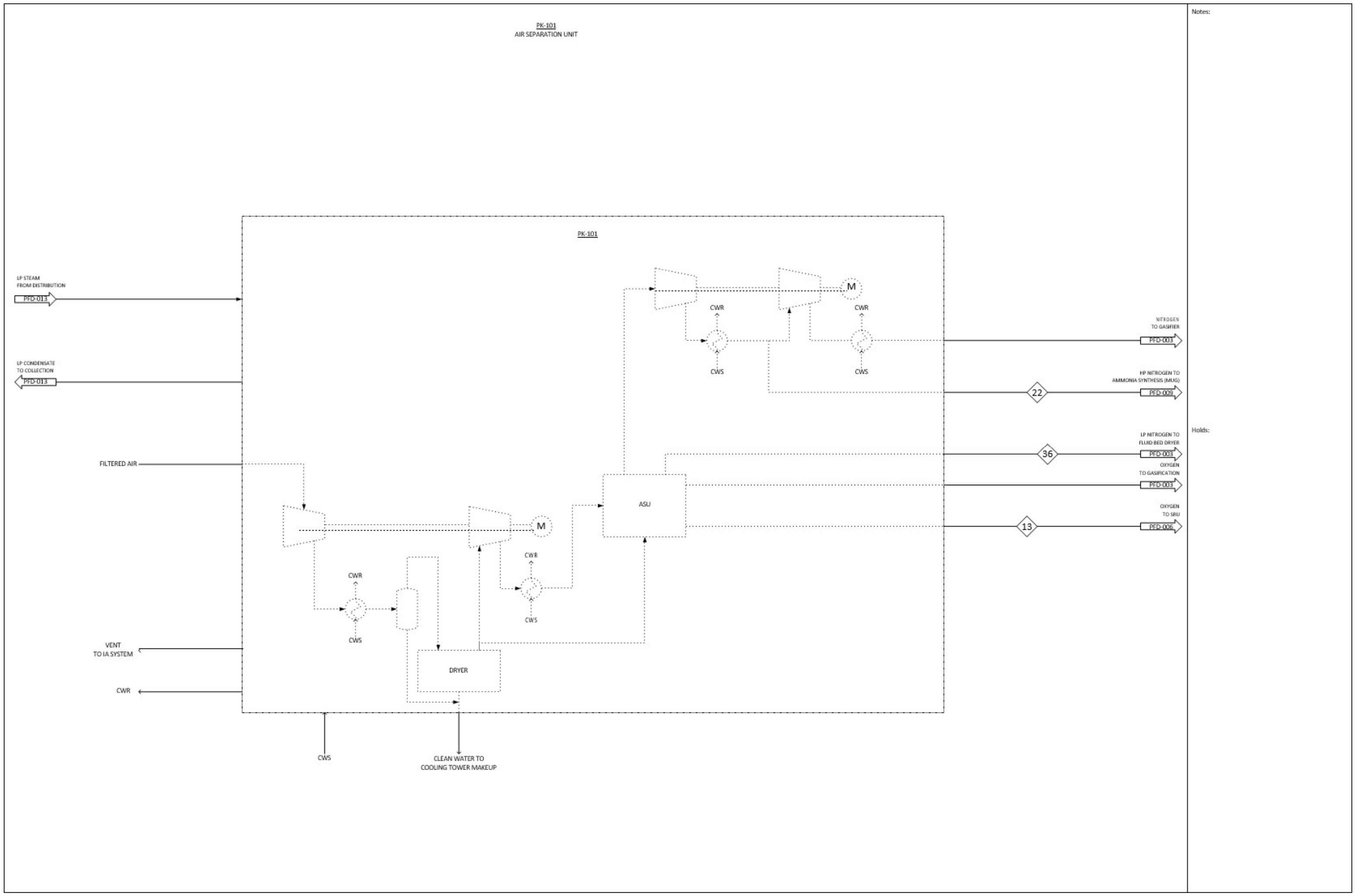


Exhibit 3-10: Coal Crushing and Handling Process Flow Diagram

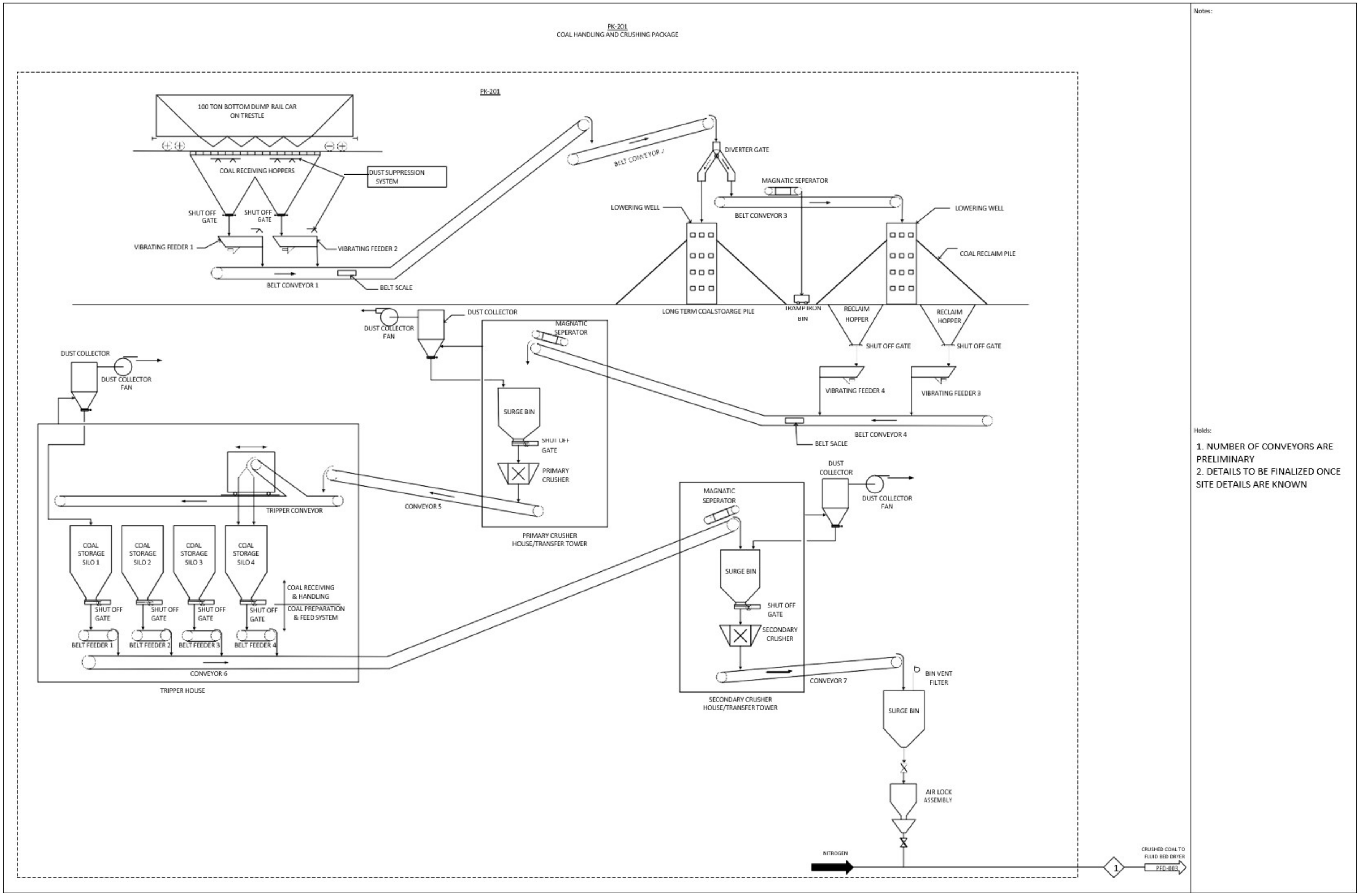


Exhibit 3-11: Gasifier, HRSG, and Quench Process Flow Diagram

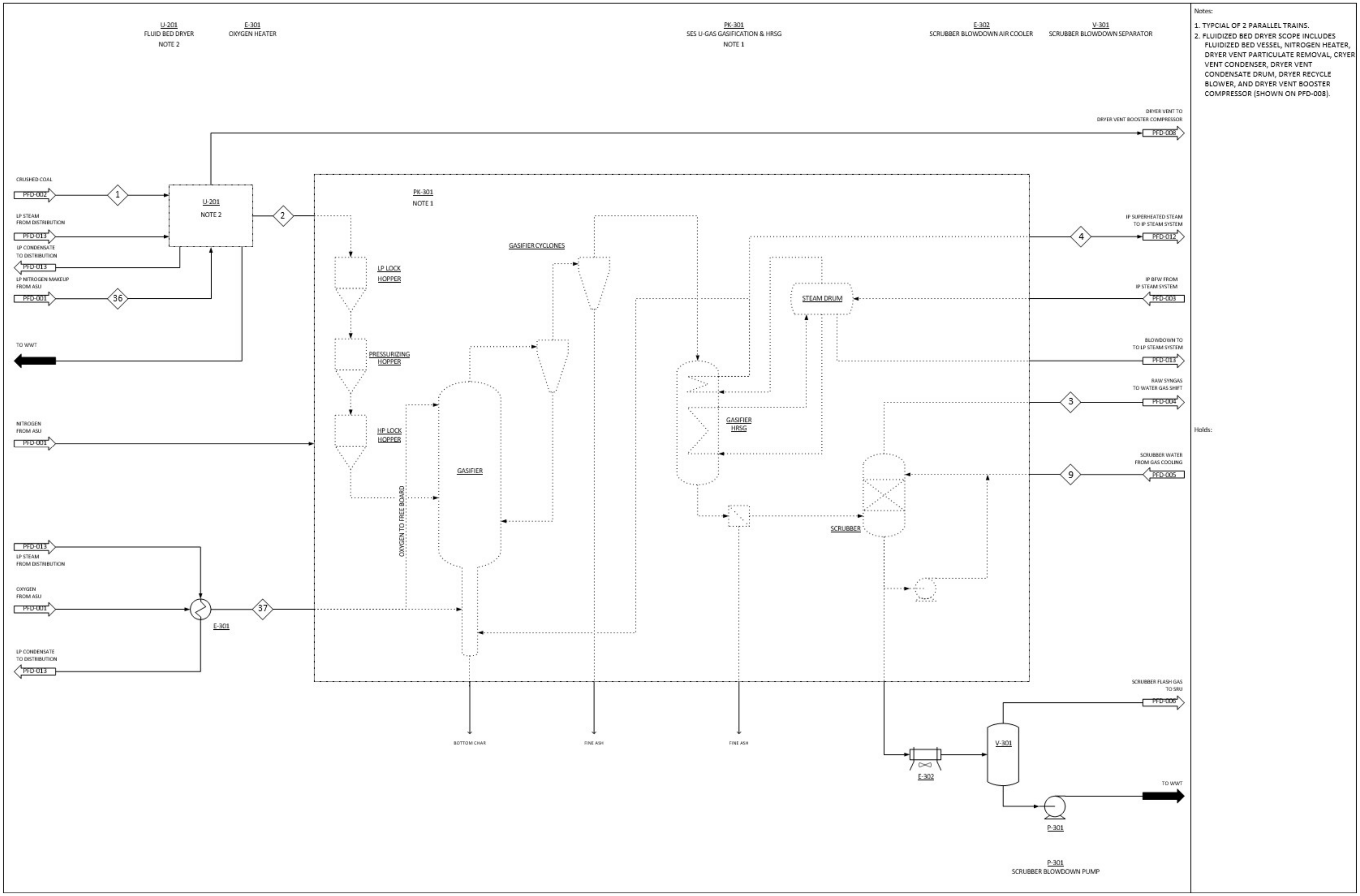


Exhibit 3-12: Water Gas Shift Process Flow Diagram

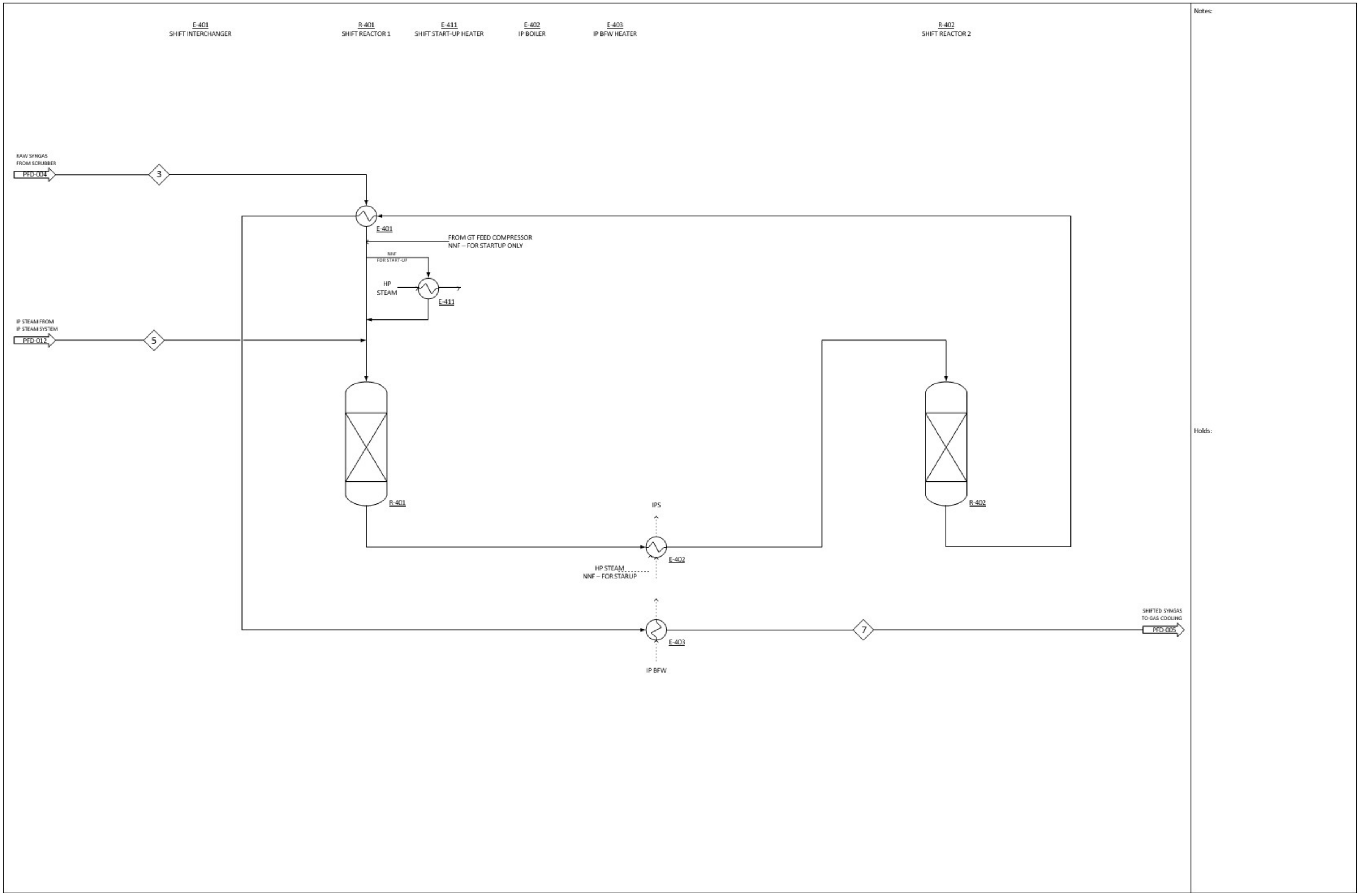


Exhibit 3-13: Syngas Cooling Process Flow Diagram

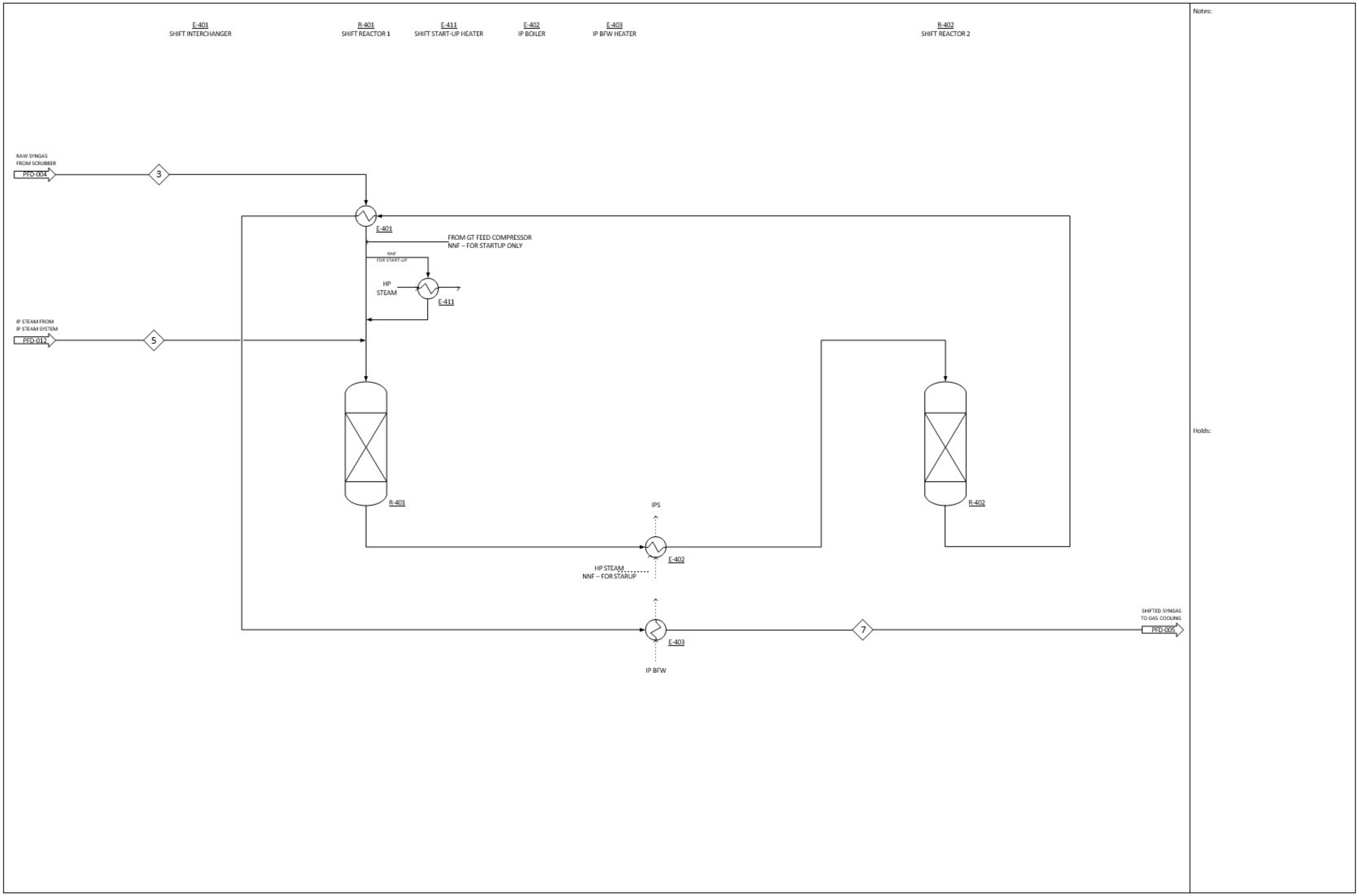


Exhibit 3-14: Syngas Clean-Up Process Flow Diagram

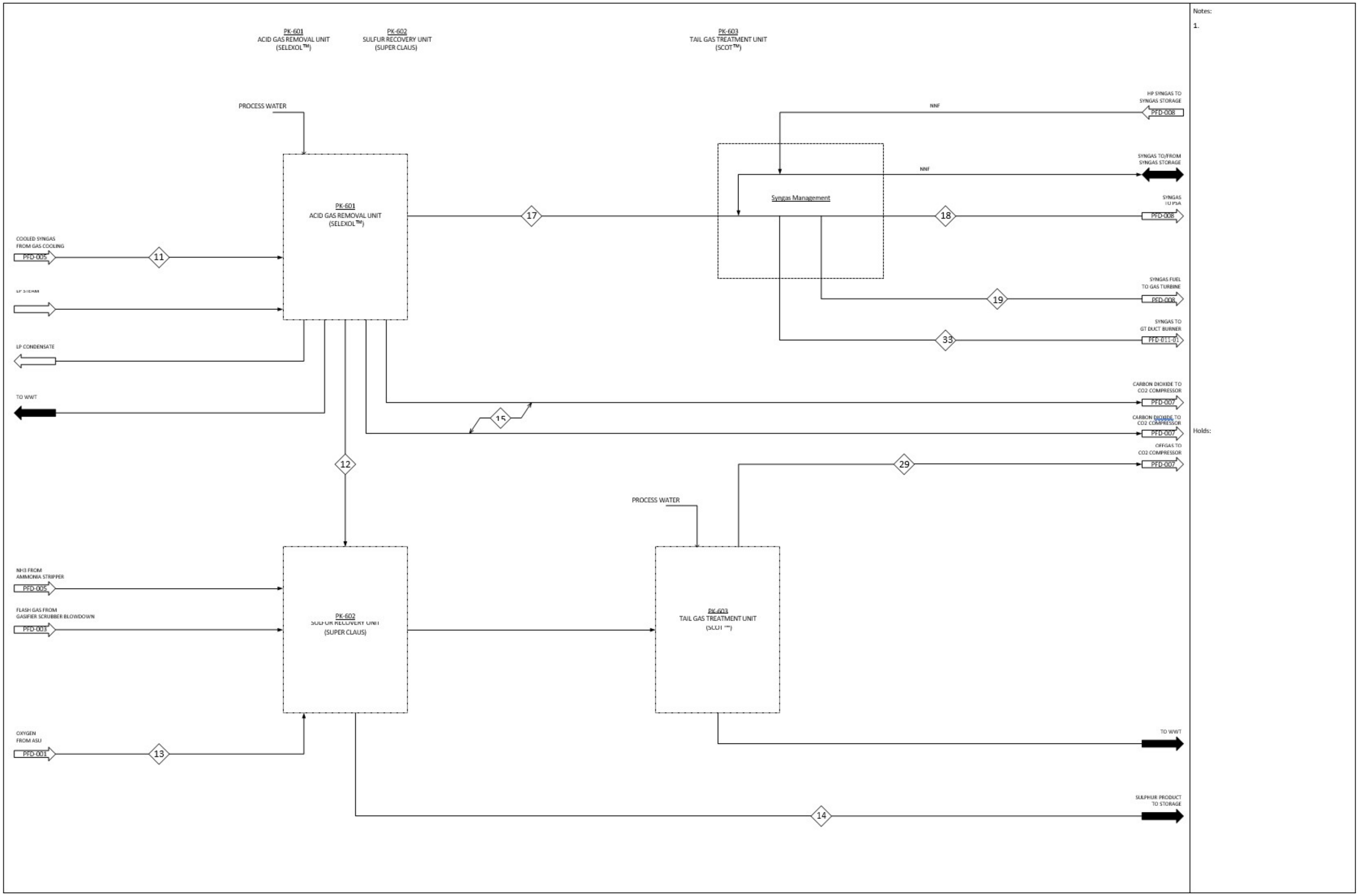


Exhibit 3-15: CO_2 Compression, Drying, and Pumping Process Flow Diagram

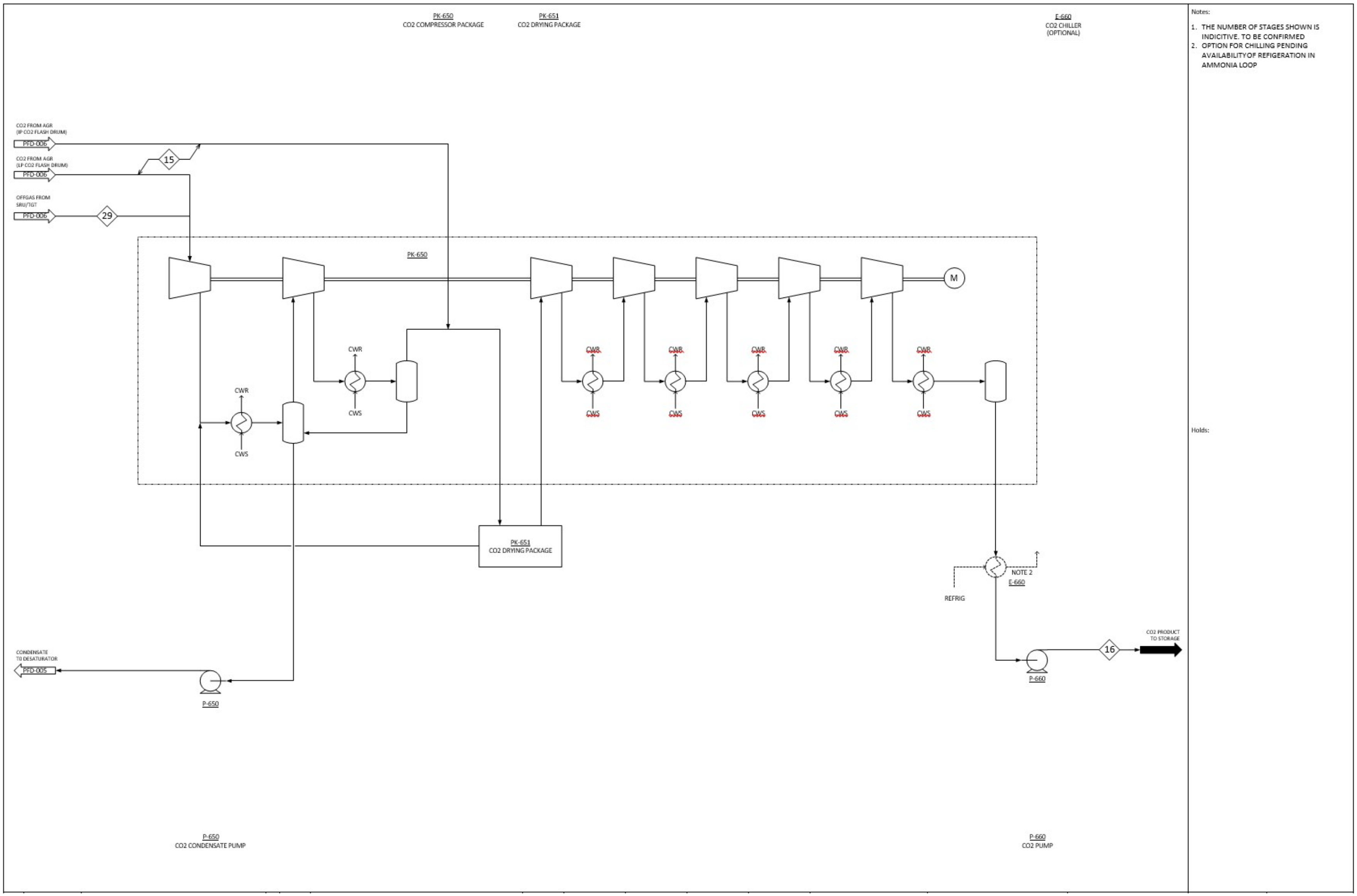


Exhibit 3-16: Fuel Gas Conditioning Process Flow Diagram

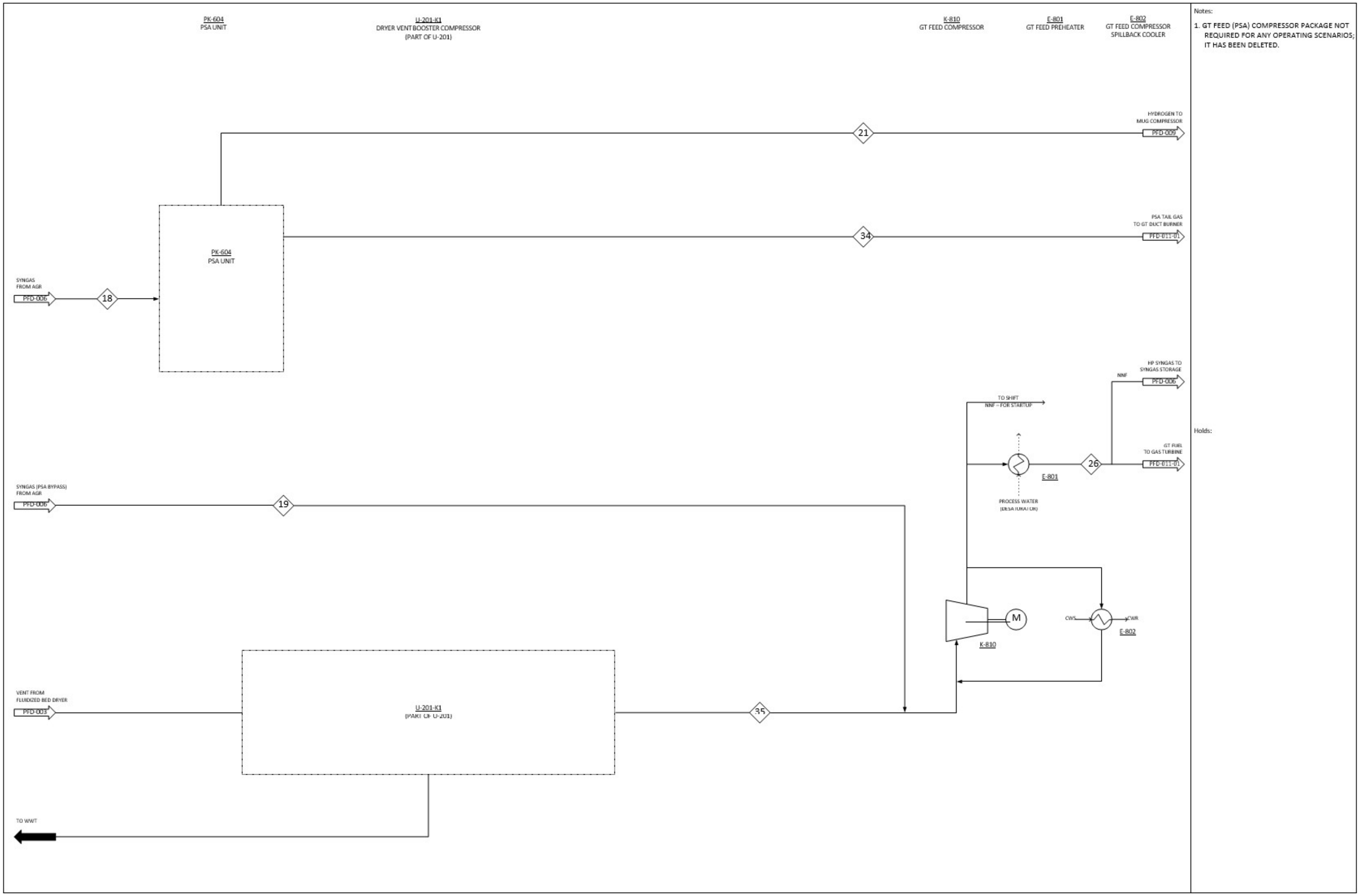


Exhibit 3-17: Make-up Gas Compressor Process Flow Diagram

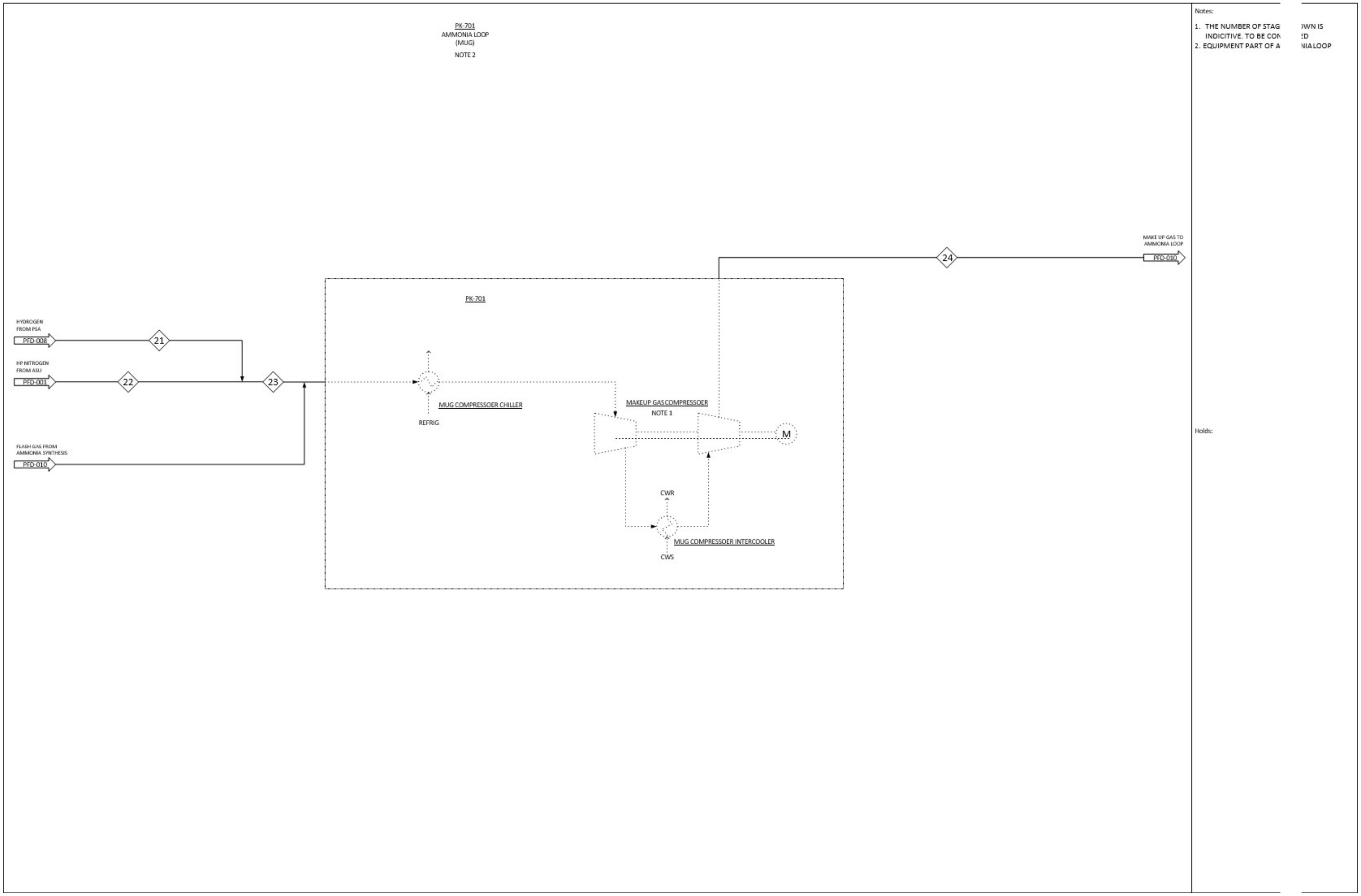


Exhibit 3-18: Ammonia Loop Process Flow Diagram

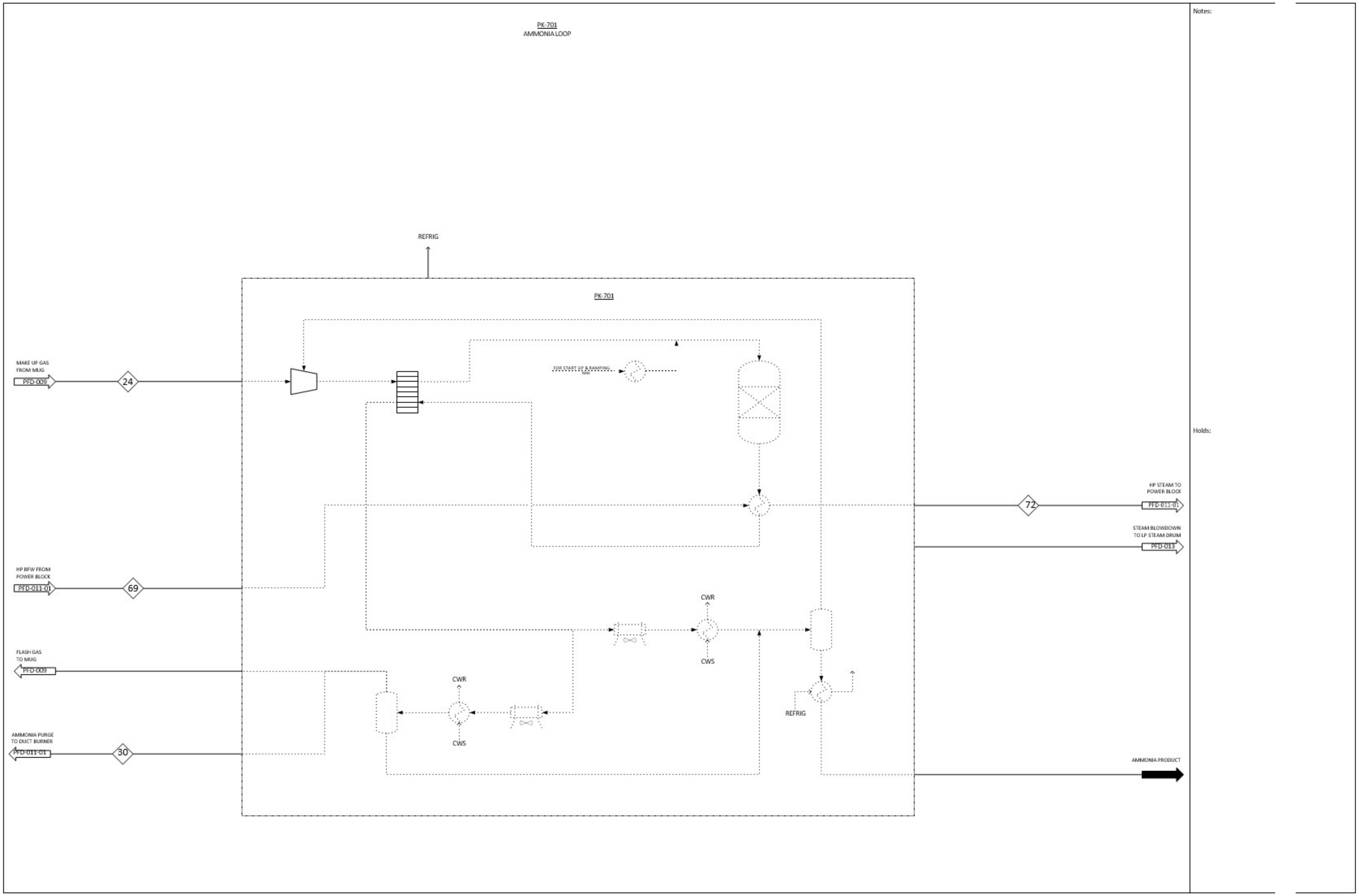


Exhibit 3-19: Gas Turbine and HRSG Process Flow Diagram

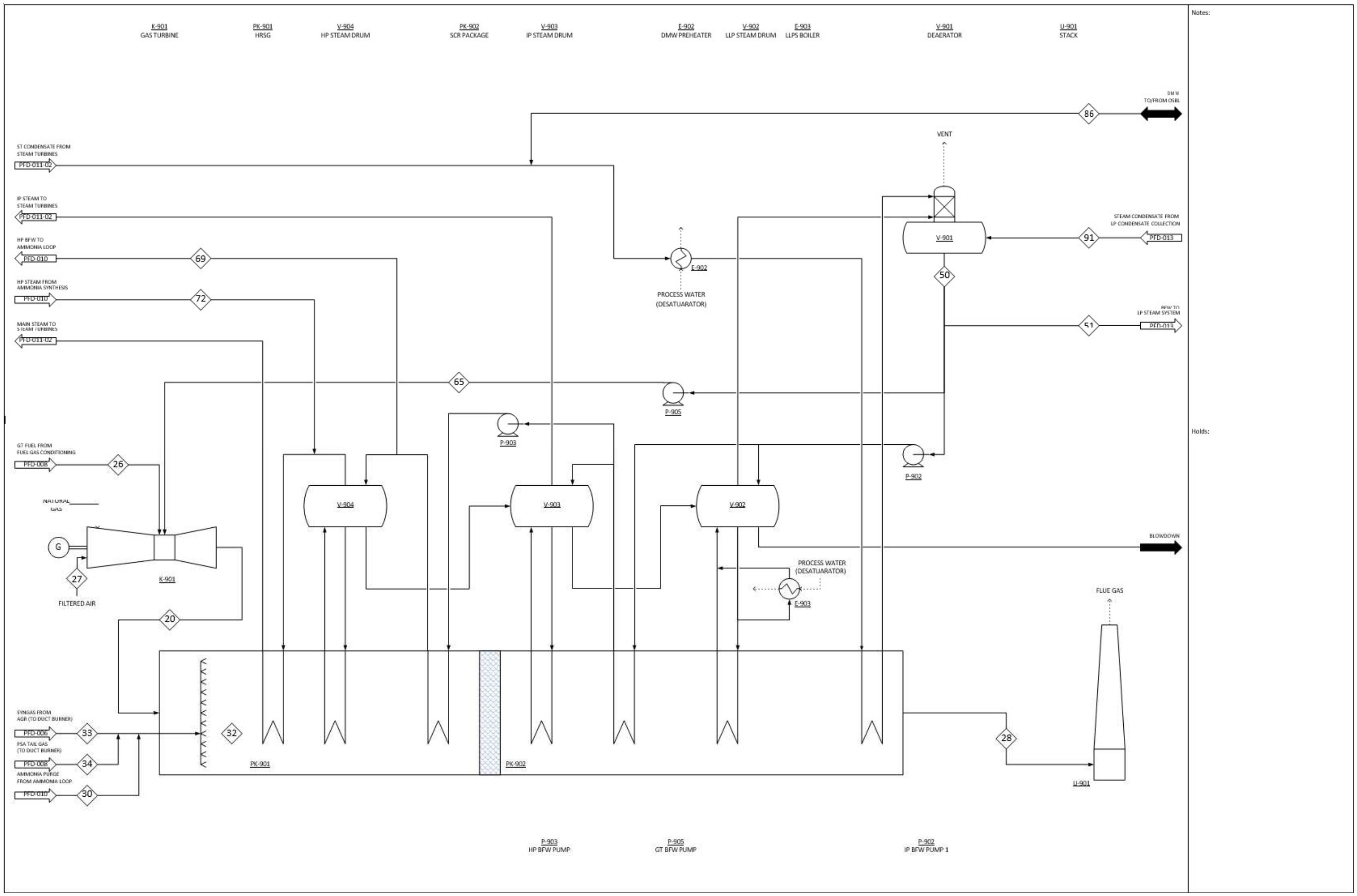


Exhibit 3-20: Steam Turbine Process Flow Diagram

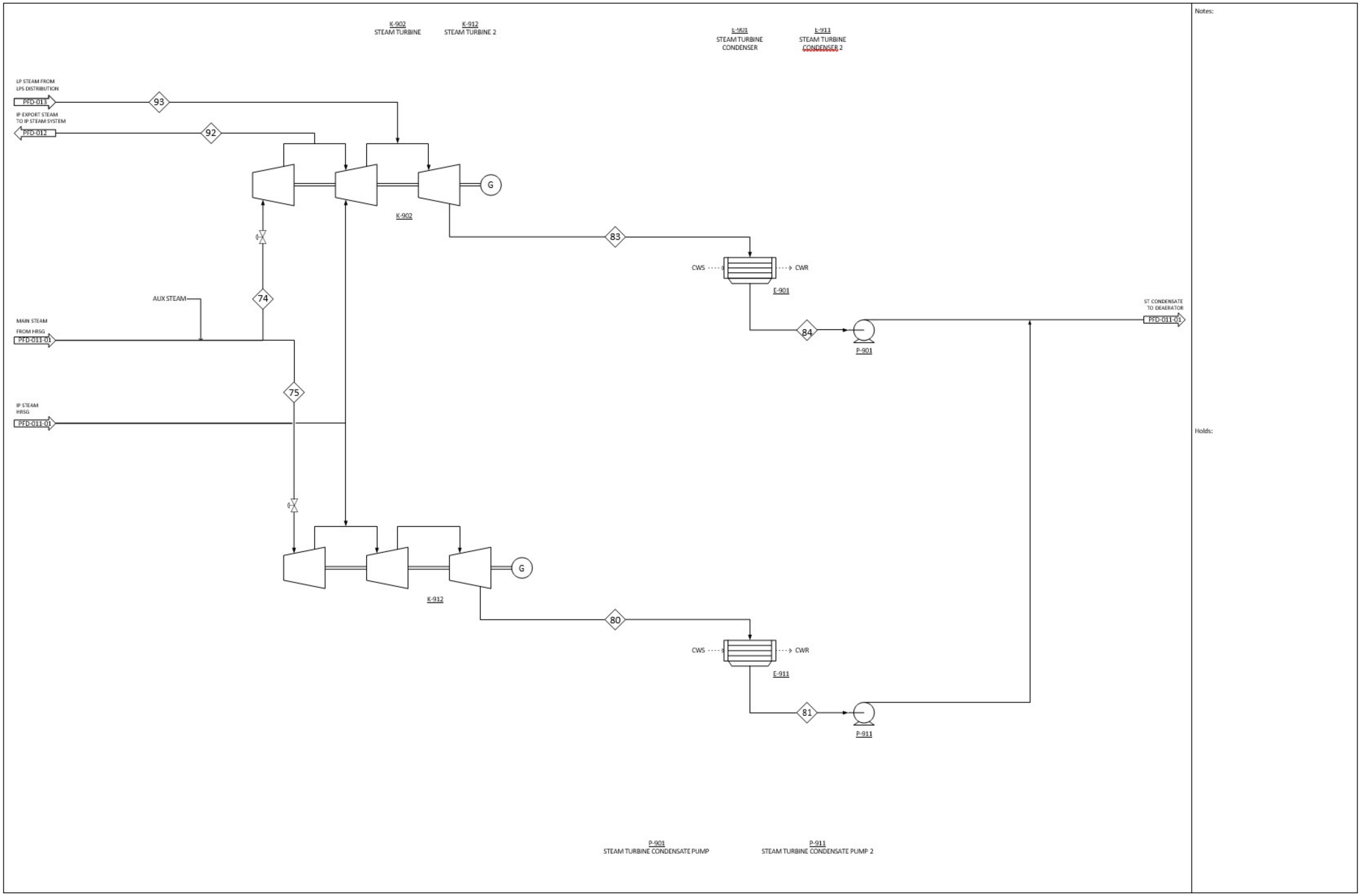


Exhibit 3-21: IP Steam System Process Flow Diagram

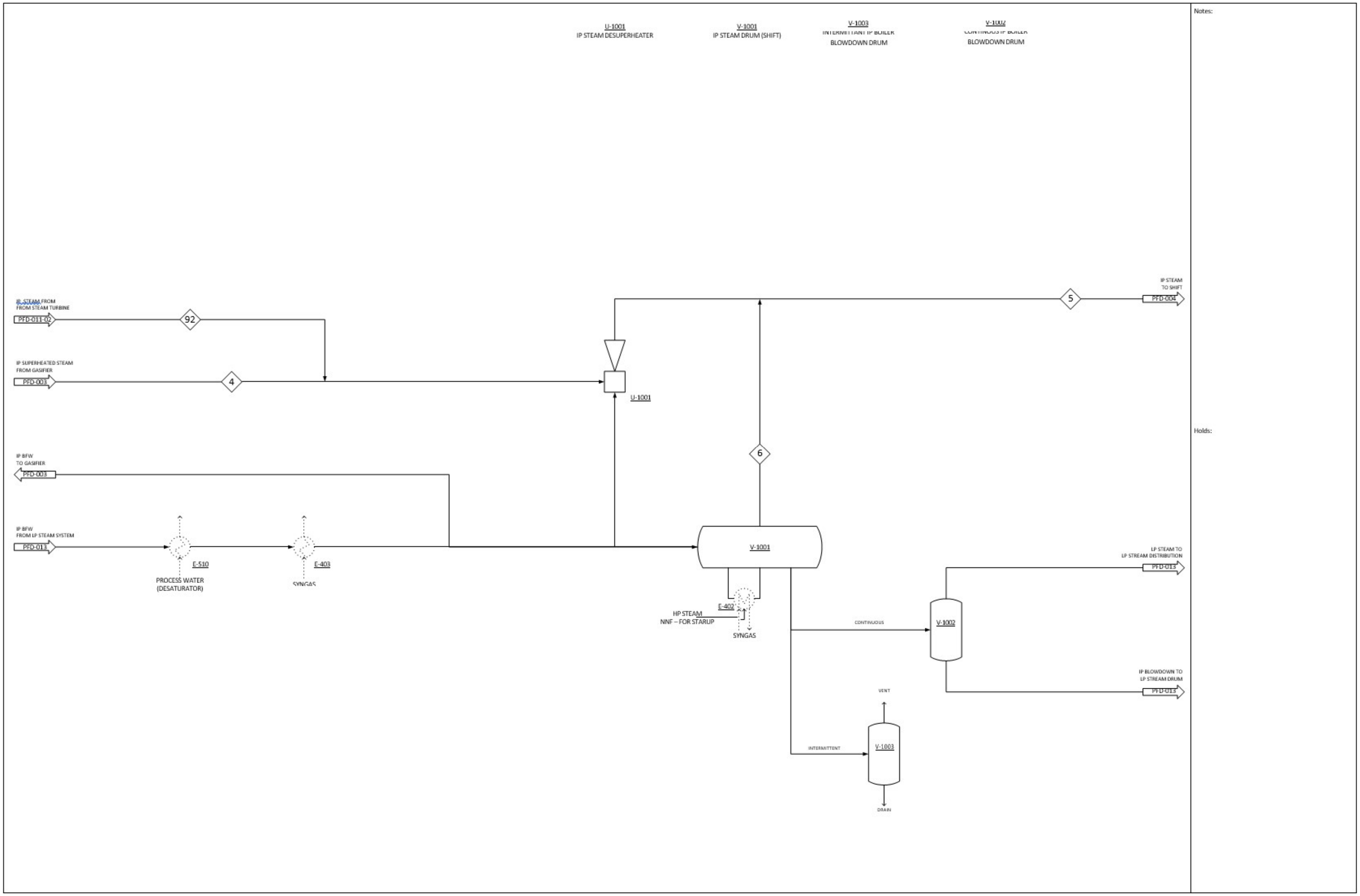
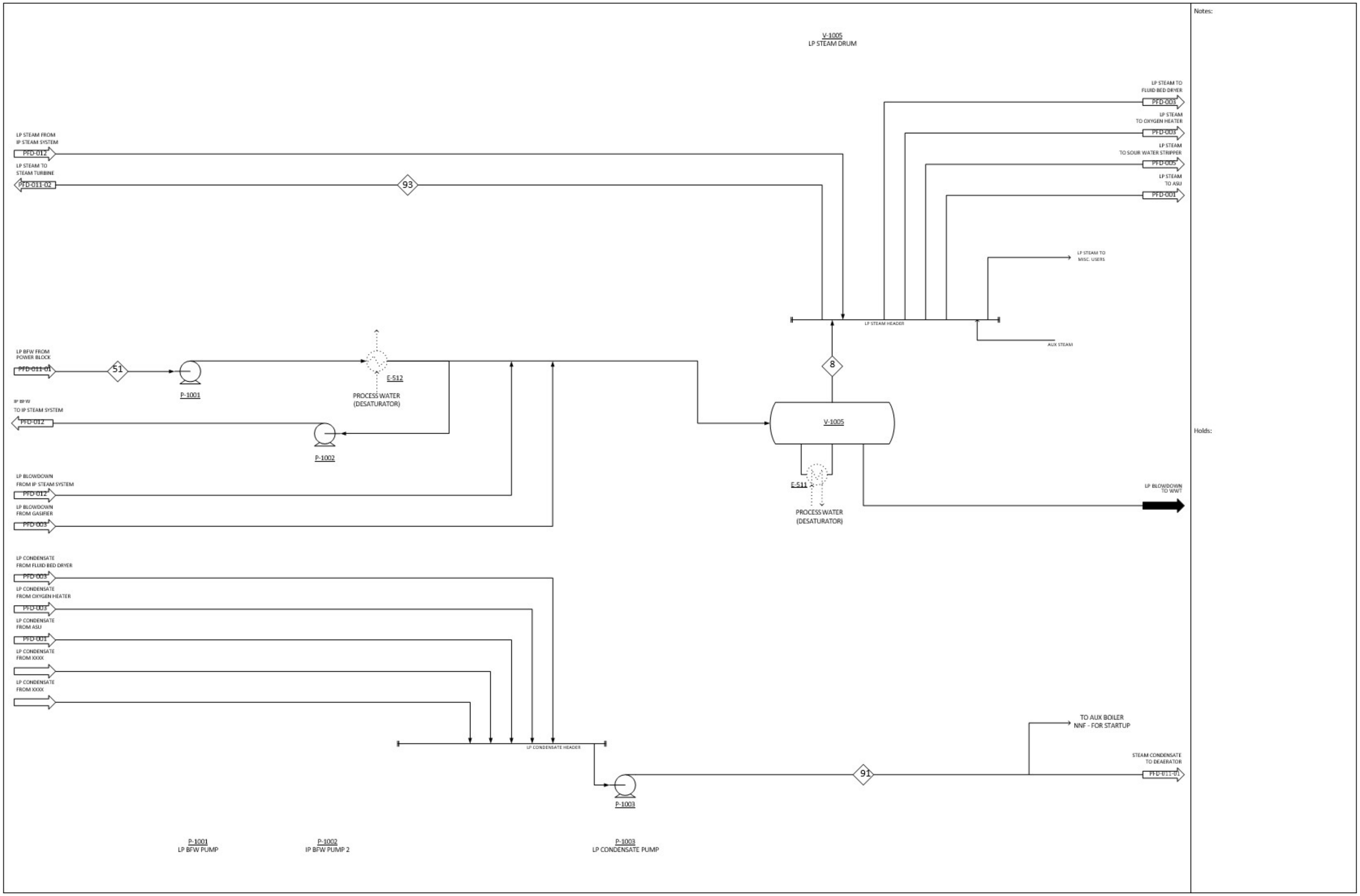


Exhibit 3-22: LP Steam System Process Flow Diagram



3.3 Major Equipment List

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

4. Ability of the Proposed Plant to Meet Coal First Design Criteria

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%[25] 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[26] 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 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

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[25] 38.3% HHV efficiency at the *Balanced Generation, 3 GT* operating point and 38.8% efficiency at the *Balanced Generation, 2 GT* operating point.

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

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.[27] 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).[28][29]

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 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

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

[28] 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)

[29] 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.

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

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.

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.[30]

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.

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.[31] 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 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

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[30] 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.

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

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 acts 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.

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.

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.

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 is based on 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.

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.

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.[32] 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.

[32] 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.

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[33]

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	75.005
O_2	22.998
Ar	1.280
H_2O	0.616
CO_2	0.050
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 and 8.5°F otherwise.

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

[33] This is consistent with the air composition for a "Midwest ISO" location as defined by NETL's *Quality Guideline for Energy System Studies: Process Modeling Design Parameters* (*National Energy Technology Laboratory, " Quality Guideline for Energy System Studies: Process Modeling Design Parameters " U.S. Department of Energy, Pittsburgh, PA, 2019.*).

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

The surrogate site selected for the financial modeling conforms to all generic site characteristics described above.

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[34]

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				

[34] 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	DIMENSIONS LENGTH m	SURFACE AREA m^2	FLUID T = Tube S = Shell		MATERIAL	DESIGN CONDITIONS PRES BAR(G)	DESIGN CONDITIONS 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					
								S					
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Exhibit C-5: Pumps[35]

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

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

Exhibit C-6: Pressure Vessels

ITEM NO.	No OFF	DESCRIPTION	ORI.	SECTION	DIMENSIONS		DESIGN CONDITION		MATERIAL		Remarks
					DIAMETER m	LENGTH m	PRES BAR(G)	TEMP °C	SHELL	INTERNALS	
R-401	1	Shift Reactor 1	V		3.4	7.30	40.0	475	CS SS Lined		Catalyst Bed Volume = 46 m³
R-402	1	Shift Reactor 2	V		3.5	7.30	40.0	330	CS SS Lined		Catalyst Bed Volume = 48 m³
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 m³/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; O2 Flow: 50,000 kg/h; O2 Press: 44 barg; O2 Purity 99.5%; Nitirogen Product: 44 barg; Power Cons (est): 20.62 MW	Includes N2 (NH3 / 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	CO2 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	
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
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	
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	
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	
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	
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	
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
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	
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	
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	
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	
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	
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	
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	
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	24562.68	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
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	
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	
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	
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	
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	100.0 0
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	
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
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	
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
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 Balances

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)