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May 13, 2020

Delivery Under Contract:
89243319CFE000016

U.S. Department of Energy (DOE)
National Energy Technology Laboratory (NETL)

Coal-Based Power Plants of the Future: Electricity and Ammonia Polygeneration Design Basis Report

Submitted To:

Brent Burns, Contracting Officer
National Energy Technology Lab
U.S. Department of Energy
3610 Collins Ferry Road
P.O. Box 880
Morgantown, WV 26507-088

Submitted By:

AST
Allegheny Science & Technology

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1. Concept Background

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

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

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

Table 1-1. Summary of Operating Modes

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

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

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

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

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

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

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

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

2. Process Description

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

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

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

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

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[2] 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

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

Figure 2-1. Polygeneration Plant Block Flow Diagram

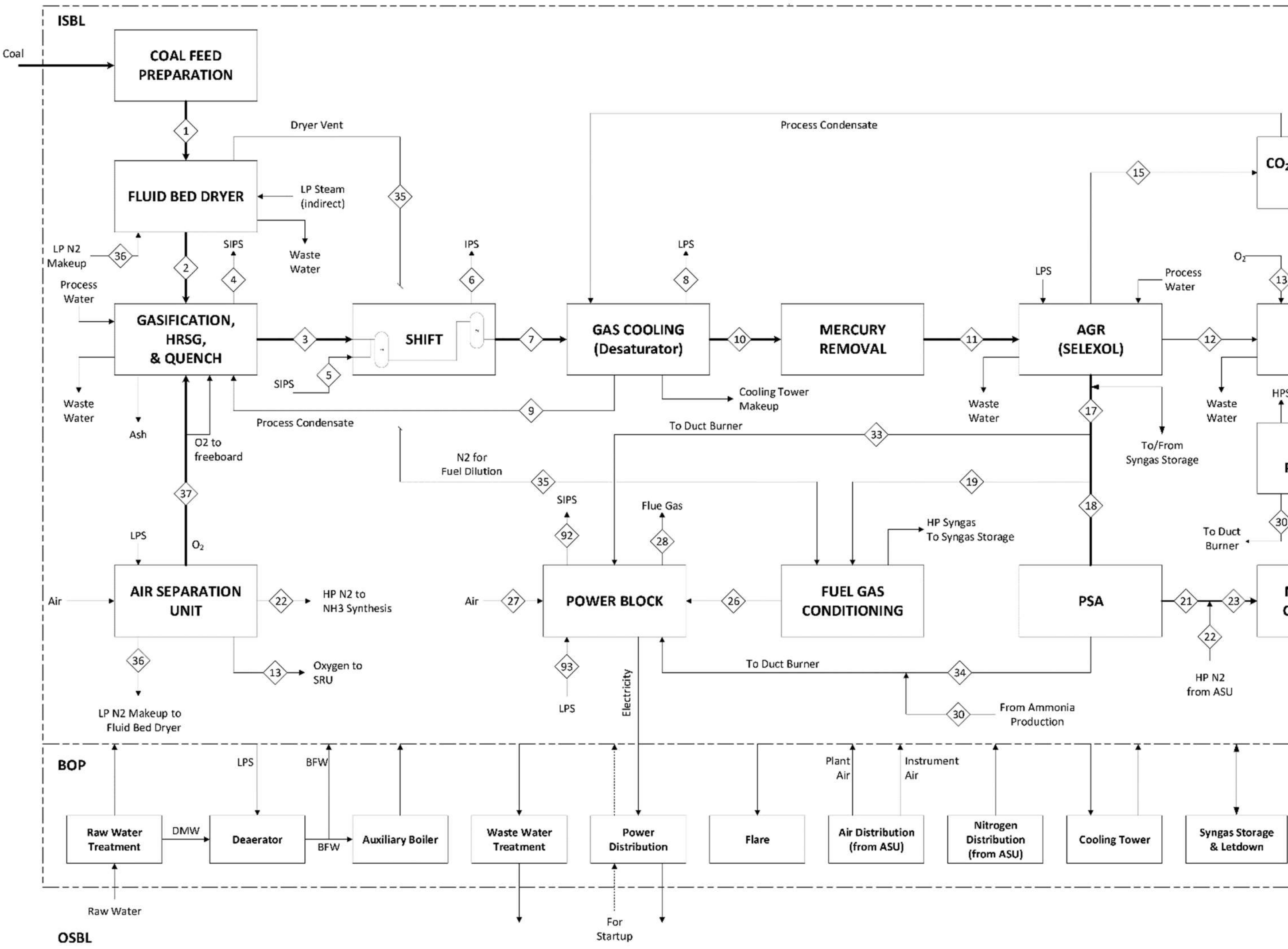

Table 2-1. Balanced Generation, 3 GTs Stream Table/Heat and Mass Balance

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

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

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

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

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

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

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

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

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

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

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

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

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

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

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

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

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

2.1.1 *Coal Receiving and Handling*

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

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

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

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

2.1.2 *Coal Preparation and Feed Systems*

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

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

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

2.1.3 *Coal Fluid Bed Drying System*

The primary purpose of the fluid bed dryer is to facilitate drying of the coal and releasing any hydrocarbons that are adsorbed in the pores of the crushed coal.[4] 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

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

- 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 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:

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.

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[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.

[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).

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.[7]

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

While not formally part of our current design basis or Pre-FEED objectives, it is important to note that this specific technology (i.e., a bubbling fluid bed) was selected for coal drying out of a desire to ensure that deployed capital equipment would allow for increased operational flexibility and additional option value opportunities throughout the plant's lifecycle. Specifically, the inclusion of this fluid bed vessel and system offers the opportunity to handle coals with sulfur content beyond that of the design basis coal while minimizing the need for future plant modifications and capital outlay. To this end, the specified vessel is designed such that it could accommodate limestone injection for sulfur scavenging if the plant operator determines that this is a desired process implementation. This additional sulfur mitigation opportunity can enable the use of high sulfur coal sources at some point in the plant's lifecycle without the need to expand the fixed capacity of the acid gas removal system beyond the size of the originally installed system. Similar to the ability of refineries to accept various qualities of crude oil feedstocks, this unit operation increases overall plant flexibility and supports potential future arbitrage opportunities among different available coal feedstocks[8].

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.

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

[8] 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.

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 make up for heat gained in the cold box during operation.

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

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

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

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

2.1.5 *Gasifier*

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

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

The devolatilized and dried coal is conveyed to the top of the lock hopper system where it is pressurized using N_2 before being fed to bottom of the fluidized bed gasifier. In the gasifier, the coal reacts with a sub-stoichiometric quantity of oxygen and steam to convert to a synthesis gas which contains primarily CO, H_2, CO_2, steam (H_2O), lesser amounts of N_2, CH_4 and a small amount of Ar. As this gasifier operates at about 1000°C, the syngas exiting the fluidized bed in standard SES U-gas operations contains roughly 7% methane. Methane content at this level can significantly reduce the effectiveness of pre-combustion carbon capture efforts. To address this concern, the design basis utilizes partial oxidation occurring in the freeboard of the gasifier to reduce methane content to roughly 1%. The WGS (Eq. 2.1) and steam methane reforming[9] (Eq. 2-2) reactions operate according to the following equations:

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

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

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

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

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

2.1.6 *Water Gas Shift*[10]

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

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[10] 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.

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

2.1.7 *Syngas Cooling*

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

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

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

1. About 20 MTPH is fed to the gas turbine (GT) feed preheater. This exchanger preheats the fuel to the gas turbines to 121°C after the fuel has been compressed in the GT fuel compressor. The outlet from the GT feed preheater is fed to the ammonia stripper.
2. About 170 MTPH of process water from the LP boiler is used to preheat LP BFW to 150°C. Part of the preheated BFW is fed to the LP boiler while the majority is pumped to 55 bar(a) and fed to the syngas cooler in the gasifier island and the HP boiler downstream of the first shift reactor.
3. The balance of the process water is used to re-boil the Selexol stripper column in the AGR. The hot water exiting the stripper reboiler is split into two streams.
    a. ~760 MTPH is returned to the top of the lower section of the desaturator at 149°C.
    b. The balance is used to produce low-low pressure steam (LLPS) at 2 bar(a) in the LLPS boiler, which is used exclusively as stripping steam in the deaerator.

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

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

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

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

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

2.1.8 *Syngas Clean Up*

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

2.1.8.1 Ammonia Removal

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

2.1.8.2 Mercury Removal

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

2.1.8.3 Acid Gas Removal

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

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

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

Shifted, cooled syngas from the mercury guard beds enters the AGR unit where it is blended with a cooled stream of recycle gas from the H_2S concentrator. The gas blend is fed into the H_2S absorber where it is contacted with cooled, loaded, Selexol solution. "Loaded solution" is defined as Selexol solution that has been through the CO_2 absorber and, consequently, is loaded with CO_2.

H_2S, COS, some CO_2, and small quantities of other gases (primarily hydrogen) are absorbed into the solution.

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

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

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

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

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

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

2.1.8.4 CO_2 Compression and Drying

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

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

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

2.1.8.5 Sulfur Recovery

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

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

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

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

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

2.1.8.6 Tail Gas Treatment Unit

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

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

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

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

2.1.9 *Syngas Management*

The purpose of the syngas management operation is to monitor and regulate the distribution of syngas (as well as relevant ancillary streams such as nitrogen, steam, etc.) between the various operating sections. This includes managing storage capacity to respond to changes in electrical load and extraction of hydrogen for ammonia synthesis. Primarily, this involves routing clean syngas between one of three possible dispositions: (1) a tank for temporary storage[11], (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

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

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

operating point transition, and this storage time can be extended to 60 – 80 minutes by performing other operational adjustments during the transition period.

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

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

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

2.1.10 *Ammonia Generation*

2.1.10.1 Hydrogen Purification

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

2.1.10.2 Ammonia Synthesis and Refrigeration

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

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

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

1. Raising steam at 105 bar(a)
2. Heat exchange to the syngas feed in the feed/product interchanger
3. Heat exchange against cooling water
4. Heat exchange against the recycle gas from the knock out pot
5. A refrigeration unit

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

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

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

2.1.11 *Power Block*

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

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

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

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

2.1.11.1 Fuel Gas Conditioning

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

2.1.11.2 LM2500+ Gas Turbine

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

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

2.1.11.3 Heat Recovery and Steam Generation

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

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

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

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

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

2.1.11.4 Selective Catalytic Reduction (SCR)

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

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

2.2 Key System Assumptions

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

Table 2-2. Key System Assumptions

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

2.3 Five Operating Points for Insight into Operational Performance and Flexibility

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

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

2.3.1 *Balanced Ammonia and Electricity Generation, Three Turbines*

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

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

2.3.2 *Balanced Ammonia and Electricity Generation, Two Turbines*

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

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

This ability to achieve roughly the same net plant outputs from different combinations of operating equipment characteristics allows for greater flexibility for the plant operator to efficiently and intelligently meet real-world demands. For example, if two turbines are already on-line, it is possible to quickly ramp up to the *Balanced Generation* point without the need to start the third turbine. If it is anticipated that no additional grid demand beyond the 51 MW of export will be requested in the near future, the plant can continue to operate on just the two turbines[12]. 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

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[12] 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[13].

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

2.3.5 *Maximum Electricity Production*

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

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

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

2.3.6 *Summary of Operating Points*

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

Figure 2-2. Summary Description of Defined Operating Points

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

Table 2-3. Summary Table of Defined Operating Points

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

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

3.1 *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%[14] 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[15] 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

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[14] 38.3% HHV efficiency at the *Balanced Generation, 3 GT* operating point and 38.8% efficiency at the *Balanced Generation, 2 GT* operating point.

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

operator to select the combination (i.e., number of turbines engaged and at what load) to optimize for either efficiency or increased transient response.

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

The system currently leverages significant heat integration between unit operations to maximize the advantages offered by the various exothermic and endothermic chemical processes as well as the residual heat from the combustion turbine outlet.[16] 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).[17][18]

3.2 *System Modularity*

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

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

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

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

[17] 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)

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

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

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

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

3.3 *Carbon Capture and Low Emissions*

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

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

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

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

3.4 *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.[19]

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

3.5 *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

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[19] 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.

transients in the system.[20] 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 an ammonia loop) and lower quality heat recovery from a methanol-based system. Current synthesis process technology is known to scale down better for ammonia than methanol. Additionally, developments in the area of renewable energy-derived ammonia are driving process intensification innovations in ammonia synthesis that later generations of this technology platform may leverage. This also indicates that ammonia production is more complimentary to reduced design, construction, and commissioning efforts. Carbon is rejected at a point source in ammonia production allowing more efficient life-cycle carbon dioxide capture (compared to distributed carbon dioxide emissions after methanol end use). Methanol production requires more water than ammonia synthesis. Additionally, ammonia transport costs act as a protective buffer to potential disruptions caused by cheap natural gas-derived mega-plants (cf. methanol), making the ammonia market inherently distributed which is complimentary to a distributed power system.

3.6 *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

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

at higher cycles of concentration than otherwise. This is to be further refined at a later stage of the project once the disposition of the process condensate and the raw water make-up is known.

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

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

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

3.7 *Reduced Design, Construction, and Commission Schedules*

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

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

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

3.8 Improved Maintainability

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

Status: *Preliminarily met -*

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

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

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

3.9 Integration with Other Plant Value Streams

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

Status: *Met – Inherent in the polygeneration design*

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

3.10 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.[21] Additionally, natural gas can be fired in the duct burners to increase net power

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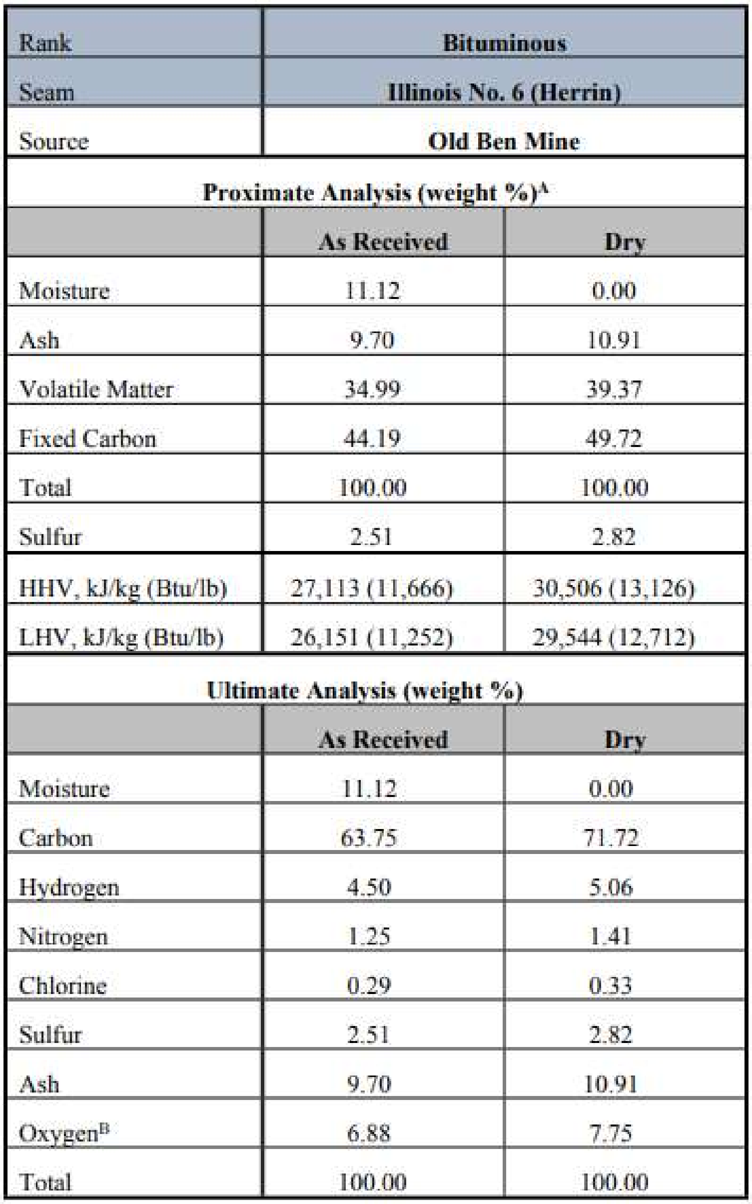
[21] As discussed in *Section 2.4.1, i*t is estimated that a maximum of 80 MMscfd of natural gas would be required to cover both transitions between operating points, as well as supplying additional power to assist in restarting the gasification plant, including the gasifier, shift unit, and utilities. 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.

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

Appendix A: Coal Feed Design Characteristics

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

Table 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:

Table 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%)

Table B-1.2 Site Ambient Conditions

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

^The cooling water temperature is the cooling tower water exit temperature.

This is set to 8.5°F above ambient wet bulb conditions in ISO cases.

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

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

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

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

B-2 Analysis of Project Specific Site Characteristics and Concerns

There are a number of key factors influencing the commercial and economic viability of a coal-based power plant with co-generation of ammonia product for export, including:

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

While determining an exact location that offers the best combination of these primary specific siting factors is a complex activity beyond the scope of this report, it is important to provide high-level thoughts and analysis related to these various factors.

B-2.1 Electricity Need and Demand

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

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

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

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

B-2.2 Access to a Reliable Coal Supply

As the production and delivery of coal within the Midwestern US is fairly mature, it is believed that this should not be an overly limiting factor in site selection, assuming that the general characteristics detailed in Section B-1 are met (e.g., sufficient available land, sufficient rail access, etc.).

B-2.3 Ammonia Need and Demand

One inherent advantage with virtually any site selected in the Midwestern US is the fact that inland production of NH3 can sell at a price premium over ammonia delivered to ports locations, typically situated on the US Gulf Coast (USGC), as it avoids many of the expensive costs associated with long overland transportation. It is this reality that allows relatively smaller scale ammonia cogeneration plants to compete with much larger operations while still maintaining the ability to charge prices that are closer to retail values.

An additional inland market of particular relevance to this study is the US Corn Belt as this market offers opportunities to sell ammonia at significant price premium over USGC sales. While this price premium typically is ~$150/MT, it can go as high as ~$300/MT.[22] The ability to sell ammonia at this Corn Belt price premium is an effective tool to help maximize plant revenue and profit even when operating at modes that exhibit a reduction in net power export.

B-2.4 Ability to Export Ammonia Product to Relevant Customer Segments

While not as extensive as the coal delivery and transport resources (essentially just available rail line), the existence of existing transportation assets such as the Magellan and Kaneb pipelines help to facilitate movement from the Midwestern generation site of the proposed polygeneration plant to other Midwestern locations and, more importantly for reasons detailed above, relevant Corn Belt locations. A map of these pipelines can be seen below in Figure B-2.1 with high-level details of relevant existing ammonia transportation and storage assets presented in Table B-2.1.

Figure B-2.1 Relevant Pipeline Asset Map

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

Table B-2.1 Existing Ammonia Transportation and Storage Asset Summary

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

Pre-FEED efforts going forward will continue to evaluate the optimal proximity to pipeline and other distribution infrastructure with the goal of simultaneously extracting a transportation premium relative to USGC prices by intelligently accessing the Corn Belt market while remaining protected from competing with world-scale, natural gas-based ammonia plants. This is a complex business optimization that has not been resolved at the time of setting the design basis.