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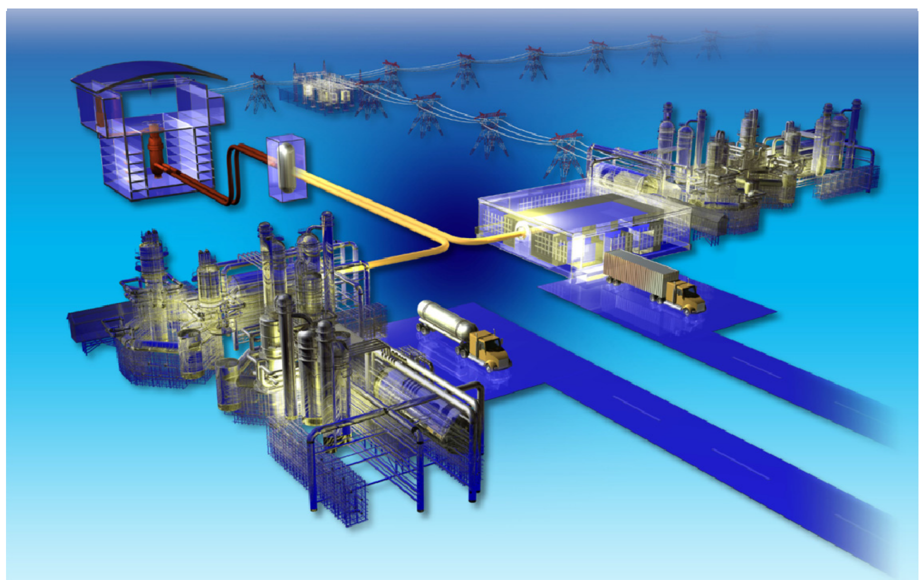
Technical Evaluation Study

Project No. 23843

Nuclear-Integrated Ammonia Production Analysis



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C For documented review and concurrence.

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REVISION LOG

Rev.	Date	Affected Pages	Revision Description
0	11/05/09	All	Newly issued document.
1	05/15/10	All	Comment resolution plus inclusion of economic results.
2	05/25/10	All	Correction to typographical error for average urea price in Tables 10-13.

EXECUTIVE SUMMARY

This technical evaluation (TEV) has been prepared as part of a study for the Next Generation Nuclear Plant (NGNP) Project to evaluate integration of high-temperature gas-cooled reactor (HTGR) technology with conventional chemical processes. This TEV addresses the integration of an HTGR with the production of ammonia and derivatives such as urea, nitric acid, and ammonium nitrate.

The HTGR can produce steam, high-temperature helium that can be used for process heat, and/or electricity. In conventional chemical processes these products are generated by the combustion of fossil fuels such as coal and natural gas, resulting in significant emissions of greenhouse gases (GHGs) such as carbon dioxide. Heat or electricity produced in an HTGR could be used to supply process heat or electricity to conventional chemical processes without generating any GHGs. This report describes how nuclear-generated heat or electricity could be integrated into conventional coal-to-ammonia and gas-to-ammonia processes and provides a preliminary economic analysis of the conventional and nuclear-integrated options.

For the gas-to-ammonia cases, results indicate that a single 450 MW_t HTGR is required to support production of 3,360 tons per day of ammonia. Substituting nuclear heat for natural gas combustion decreases natural gas consumption by 18%. As a result, CO_2 emissions are decreased by 70% from the conventional gas-to-ammonia case.

For the coal-to-ammonia case, a compelling need for HTGR heat integration was not identified. However, the coal-to-ammonia process is a net consumer of electricity; hence, production of electricity using an HTGR could be considered to supplement this process.

HTGR-integrated high-temperature steam electrolysis (HTSE) was also considered to supply hydrogen to the ammonia process. In this scenario, coal usage for hydrogen production can be completely eliminated; hence, the process configuration is radically altered and no longer resembles the conventional coal-to-ammonia case. Two such configurations were considered in this study, and results indicate that approximately four 600 MW_t HTGRs are required to support production of 3,360 tons per day of ammonia. Substituting hydrogen from water via nuclear-integrated HTSE instead of natural gas reforming or coal gasification (followed by shift conversion) eliminates fossil fuel consumption and the CO_2 emissions associated with these processes.

The following table outlines the urea prices necessary for a 12% internal rate of return (IRR) for the cases analyzed with and without a carbon tax as well as assessing the impact of reducing the HTGR capital cost by 30%. Average and historical high urea prices are also presented.

Table ES 1. Ammonia production economic results summary for a 12% IRR.

Technology	Urea Price ($/ton)*				
	No CO_2 Tax			$50/ton CO_2 Tax	$100/ton CO_2 Tax
	Low Nat. Gas	Avg. Nat. Gas	High Nat. Gas	Avg. Nat. Gas	Avg. Nat. Gas
Conventional Gas-to-Ammonia	244.36	272.23	348.85	302.09	331.96
HTGR Integrated Gas-to-Ammonia	263.32	286.12	348.81	306.73	327.34
HTGR Integrated Gas-to-Ammonia, -30% HTGR Cost	249.88	272.68	335.37	293.29	313.90
HTSE Integrated Gas-to-Ammonia, N_2 from Combustion	525.86	537.26	568.61	537.36	537.47
HTSE Integrated Gas-to-Ammonia, N_2 from ASU	482.14	493.54	524.89	493.64	493.75
US Urea Price, Average	396.42				
US Urea Price, High (Apr. 2008)	565.80				

*No carbon sequestration assumed for this analysis

Based on these results, integration of nuclear heat into a gas-to-ammonia process appears promising for the following reasons:

1. Urea selling price is comparable to the average U.S. selling price and the calculated production price for a conventional gas-to-ammonia process

2. Natural gas consumption can be reduced, thereby prolonging the availability of fossil resources in the U.S.

3. Carbon emissions can be significantly reduced.

Although the HTSE-integrated flowsheets offer even greater benefits in terms of reduced natural gas consumption and carbon emissions, at present these flowsheets do not appear to be economically viable.

CONTENTS

ACRONYMS AND NOMENCLATURE

AACE	Association for the Advancement of Cost Engineering
ASU	air separation unit
ATCF	after tax cash flow
BTCF	before tax cash flow
CEPCI	chemical engineering plant cost index
CMD	coal milling and drying
COPE	Claus oxygen-based process expansion
CTL	coal to liquids
DOE	Department of Energy
GT-MHR	gas turbine-modular helium reactor
HRSG	heat recovery steam generator
HTSE	high temperature steam electrolysis
HTGR	high temperature gas reactor
IHX	intermediate heat exchanger
INL	Idaho National Laboratory
IRR	internal rate of return
MACRS	modified accelerated cost recovery system
MARR	minimum annual rate of return
MMSCF	1,000,000 standard cubic feet
MSCF	1,000 standard cubic feet
NETL	National Energy Technology Laboratory
NGNP	Next Generation Nuclear Plant
NIBT	net income before taxes
O&M	operations and maintenance
PSA	pressure swing absorption
PW	present worth
SCOT	Shell Claus offgas treatment
SCR	Selective catalytic reduction
SGCP	Shell coal gasification process
SMR	steam methane reformer
TCI	total capital investment
TEV	Technical Evaluation

UAN	Urea/ammonium nitrate
USDA	United Stated Department of Agriculture
C_1	cost of equipment with capacity q_1
C_2	cost of equipment with capacity q_2
C_k	capital expenditures
d_k	depreciation
E_k	cash outflows
i'	IRR
k	year
n	exponential factor
q_1	equipment capacity
q_2	equipment capacity
R_k	revenues
t	tax rate
T_k	income taxes

1. INTRODUCTION

This technical evaluation (TEV) has been prepared as part of a study for the Next Generation Nuclear Plant (NGNP) Project to evaluate integration of high-temperature gas-cooled reactor (HTGR) technology with conventional chemical processes. The NGNP Project is being conducted under U.S. Department of Energy (DOE) direction to meet a national strategic need identified in the *Energy Policy Act* to promote reliance on safe, clean, economic nuclear energy and to establish a greenhouse-gas-free technology for the production of hydrogen. The NGNP represents an integration of high-temperature reactor technology with advanced hydrogen, electricity, and process heat production capabilities, thereby meeting the mission need identified by DOE. The strategic goal of the NGNP Project is to broaden the environmental and economic benefits of nuclear energy in the U.S. economy by demonstrating its applicability to market sectors not being served by light water reactors.

The HTGR produces steam, high-temperature helium that can be used for process heat, and/or electricity. A summary of these products and a brief description is shown in Table 1. For this study the HTGR outlet temperature is assumed to be 750°C; this reflects the initial HTGR design and assumes a more conservative outlet temperature. Eventually, temperatures of 950°C are anticipated. Additionally, a 50°C temperature approach is assumed between the primary and secondary helium loops, if helium is the delivered working fluid. As a result, the helium stream available for heat exchange is assumed to be at 700°C. In conventional chemical processes, these products are generated by the combustion of fossil fuels such as coal and natural gas, resulting in significant emissions of greenhouse gases (GHGs) such as carbon dioxide. Heat or electricity produced in an HTGR could be used to supply process heat or electricity to conventional chemical processes while generating minimal GHGs. The use of an HTGR to supply process heat or electricity to conventional processes is referred to as a nuclear-integrated process. This report describes how nuclear-generated heat or electricity could be integrated into conventional processes and provides a preliminary economic analysis to show which nuclear-integrated processes compare favorably with conventional processes.

Table 1. Projected outputs of the NGNP.

HTGR Product	Product Description
Steam	540°C and 17 MPa
High-Temperature Helium	Delivered at 700°C and 9.1 MPa
Electricity	Generated by Rankine cycle with 40% thermal efficiency

This TEV addresses potential integration opportunities for production of ammonia and derivatives such as urea, nitric acid, and ammonium nitrate. The HTGR would produce electricity, heat, and/or hydrogen and be located near the ammonia production facility. Details of the specific cases considered are described in Section 2 of this report. A separate study should be conducted to assess the optimal siting of the HTGR with respect to the ammonia facilities, balancing safety concerns associated with separation distance and heat losses associated with transporting high-temperature heat long distances.

The Advanced Process and Decision Systems Department at Idaho National Laboratory (INL) has spent several years developing detailed process simulations of chemical processes, typically utilizing fossil fuels such as coal, biomass, or natural gas as the feedstock. These simulations have been developed using Aspen Plus, a state-of-the-art, steady-state chemical process simulator (Aspen 2006). This study makes extensive use of these models and the modeling capability at INL to evaluate the integration of HTGR technology with commercial ammonia production methods. The outputs from the material and energy balances generated in Aspen Plus were utilized as inputs into the Excel economic models (Excel 2007).

This TEV assumes familiarity with Aspen Plus; hence, a detailed explanation of the software capabilities, thermodynamic packages, unit operation models, and solver routines is beyond the scope of this TEV. Similarly, a familiarity with gasification, steam methane reforming (SMR), ammonia synthesis, urea synthesis, nitric acid synthesis, and common gas purification technologies is assumed. Hence, a thorough explanation of these technologies is also considered to be beyond the scope of this TEV.

This TEV first presents an overview of the process modeling performed for each case. Afterwards, the results of the process modeling are discussed with emphasis placed on the impact of the HTGR integration. Next, an overview of the economic modeling is presented, followed by results of this modeling. Again, emphasis is placed on the impact of the HTGR integration. Finally, overall conclusions for the coal- and natural gas-based cases are discussed separately. These conclusions also focus on the impact of the HTGR integration.

2. PROCESS MODELING OVERVIEW

Plant models for the coal-to-ammonia, natural-gas-to-ammonia, and high-temperature steam electrolysis (HTSE) hydrogen-to-ammonia process were developed using Aspen Plus (Aspen 2006). Because of the size and complexity of the process modeled, the simulation was constructed using “hierarchy” blocks. A hierarchy block is a method for nesting one simulation within another simulation. In this manner, submodels for each major plant section were constructed separately and then combined to represent the entire process.

Significant emphasis in the models has been placed on heat integration between different parts of the plant. To facilitate energy tracking, Aspen’s “utility” blocks were used extensively. Utility blocks tracked electricity, steam, and cooling water usage. Aspen Plus Version 2006 (Build 20.0.3.4127), run under Windows XP SP3 on computer ID 410530 was used for all modeling calculations.

The original models for the fossil portion of the plant were developed using English units, which is common industrial practice in the United States. Nuclear plants typically use metric units; hence, this report contains both English and metric units depending on the context of the information presented.

Six cases were originally identified for modeling:

- Conventional coal-to-ammonia process
- Conventional natural gas-to-ammonia process
- Nuclear-integrated coal-to-ammonia process
- Nuclear-integrated natural-gas-to-ammonia process
- HTSE hydrogen-to-ammonia process using H_2/air combustion as the nitrogen source
- HTSE hydrogen-to-ammonia process using cryogenic air separation as the nitrogen source.

Producing ammonia from natural gas and coal are fundamentally different. Natural gas is hydrogen rich with a carbon-to-hydrogen ratio of 1:4. Coal is hydrogen limited with a carbon-to-hydrogen ratio of ~ 1:0.8. Coal requires significant capital to gasify and to clean up the resulting syngas of CO and H_2. CO is then used to produce additional H_2 via the water-gas shift reaction:

$$CO + H_2O \rightarrow CO_2 + H_2 \tag{1}$$

Excess heat is generated in the coal-based case due to carbon rejection (as CO_2).

For the natural-gas-to-ammonia process, nuclear heat was utilized, whereas for the coal-to-ammonia process, both nuclear heat and HTSE hydrogen were utilized. It is also possible to produce ammonia using HTSE as the hydrogen source rather than a fossil fuel. For this scenario, two distinct options were investigated for supplying nitrogen to the ammonia synthesis unit.

For the coal cases, a generic Illinois #6 coal was used as the feedstock due to its high availability and use within the United States. A dry-fed, entrained-flow, slagging gasifier (similar to a Shell, Uhde, or Siemens design) was selected as the gasification technology for this evaluation. For the natural gas cases, gas composition was taken from data published by Northwest Gas Association. Capacities for the coal cases were set based on a nominal gasifier throughput of 3,600 tonne/day, representing a commercial Shell gasifier. This resulted in an intermediate ammonia production rate of 3,360 ton/day. The capacity of a typical ammonia plant is between 1,500 and 2,000 ton/day; hence, this production rate is roughly equivalent to a 2-train ammonia plant configuration. For easy comparison of the cases, the natural gas cases were scaled to produce an equivalent amount of ammonia, with slight adjustments to compensate for differences in ammonia purity.

Because there are numerous derivative products that can be produced from ammonia, the models were constructed to include the most common derivative products: nitric acid, ammonium nitrate, and urea. By varying ammonia flow to each of these downstream processes, numerous end-product mixes could be simulated. For this study, the split of derivate products was adjusted to produce ammonium nitrate and urea in the right ratio for urea/ammonium nitrate (UAN) 32 synthesis (urea ammonium nitrate solution containing 32 wt% nitrogen—made by combining ammonium nitrate, urea, and water in a 45/35/20 blend). Hence, all modeled cases produced 2,939 ton/day of urea and 3,779 ton/day of ammonium nitrate.

Urea production requires CO_2 as a feed to the process. For the four fossil-based cases, CO_2 is readily available. For the two cases utilizing HTSE for the hydrogen supply, a CO_2 supply is not readily available and must be produced to support urea manufacture. For these cases, natural gas is burned to produce the required CO_2. While coal could have been selected for this purpose, natural gas is preferred for its much simpler gas cleanup requirements.

For the Aspen models described in this analysis, rigorous submodels of the nuclear power cycle and high-temperature electrolysis have not yet been integrated. Results for the nuclear power cycle and HTSE were calculated separately using the UNISIM modeling package. Cooling water requirements for HTSE were then estimated and added to the overall Aspen model results. Water consumption for the HTGR and associated power cycle has not been included, as a detailed water balance for the HTGR has not been completed at this time.

The general model descriptions for all cases are presented below. Although the method of producing syngas varies from case to case, production of ammonia, nitric acid, ammonium nitrate, and urea are essentially unchanged between cases.

2.1 Conventional Coal-to-Ammonia Case

Figure 1 shows the block flow diagram for the conventional coal-to-ammonia case. The proposed process includes unit operations for air separation, coal milling and drying, coal gasification, syngas cleaning and conditioning, sulfur recovery, CO_2 purification/compression, ammonia synthesis, nitric acid synthesis, ammonium nitrate synthesis, urea synthesis, power production, cooling towers, and water treatment. Each plant area is briefly described below. For each description, the name capitalized and enclosed in parentheses corresponds to the name of the hierarchy block within the Aspen process model.

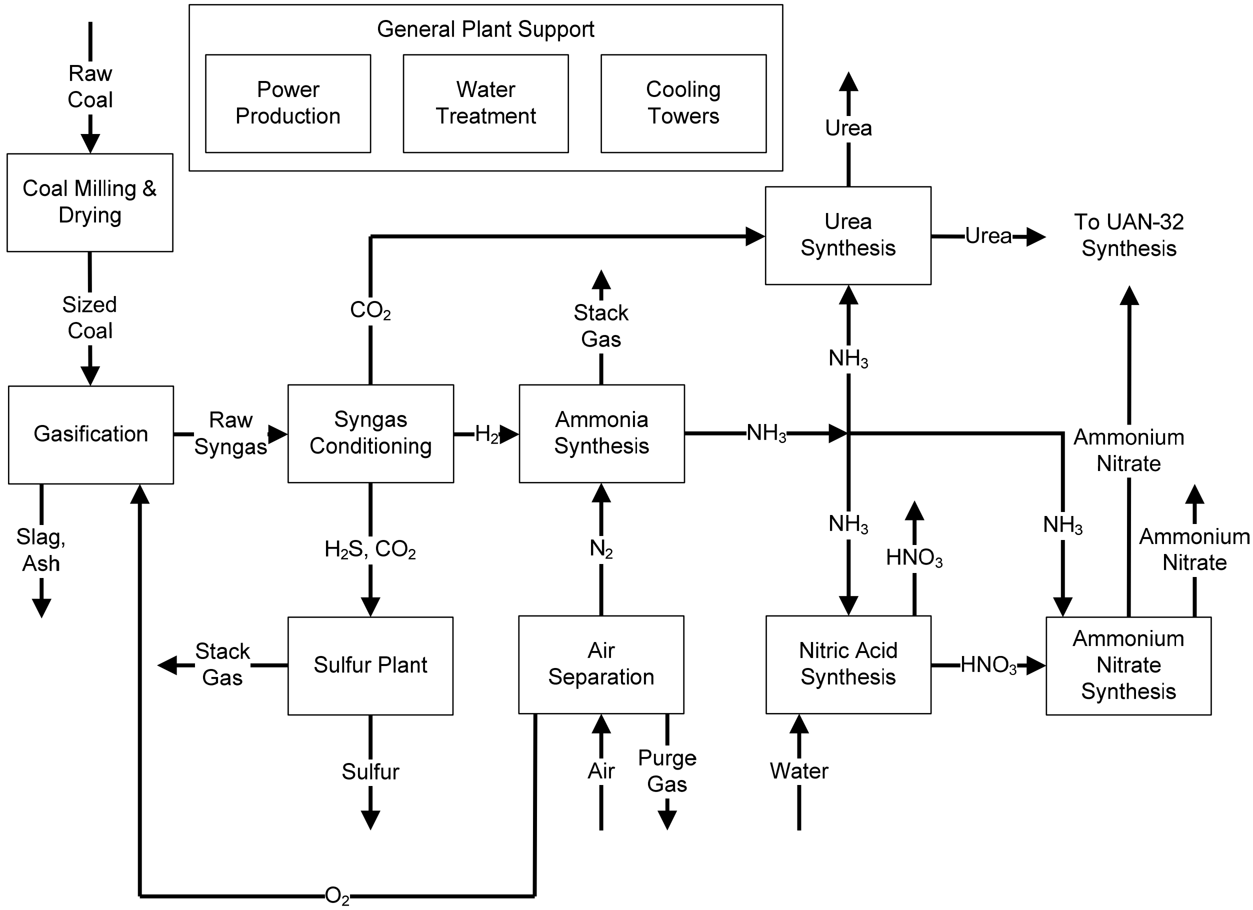

Figure 1. Block flow diagram for the conventional coal-to-ammonia process.

- **Air Separation (ASU) –** Oxygen is produced via a standard cryogenic Linde-type air separation unit (ASU) that utilizes two distillation columns and extensive heat exchange in a cold box (Linde 2008). The oxygen product is used for gasification. An O_2 purity of 99.5% is specified, although a 95% O_2 purity plant would also be suitable for this application. Oxygen demand for gasification drives the size of the ASU; hence, in addition to the primary nitrogen demand for ammonia synthesis, excess nitrogen is also available for other plant uses such as coal drying and transport.

- **Coal Milling and Drying (CMD) –** Coal is pulverized to below 90 µm using a roller mill to ensure efficient gasification. Currently, coal milling power consumption is modeled based on the power calculated by Aspen assuming a Hardgrove grindability index of 60. Drying is accomplished simultaneously using a heated inert gas stream. The gas stream removes evaporated water as it sweeps the pulverized coal through an internal classifier for collection in a baghouse. Inert nitrogen from the ASU is heated by firing clean syngas in a burner. This nitrogen is mixed with this hot gas to create a hot inert gas stream, which dries the Illinois coal down to 6% moisture (Shell 2005). Nitrogen is also used as transport gas for the coal from the baghouse to the lock hoppers. Pressurized nitrogen is used to transport the dry, sized coal into the gasifier. The transport gas is assumed to be 0.15 pounds of gas per pound of solids. The amount of N_2 vented during depressurization of the feed hopper is estimated using the ideal gas law.

- **Gasification (GASIFIER) –** The dry coal is gasified at 2,800°F using an entrained-flow, dry-fed, slagging, oxygen-blown gasifier. Oxygen is fed to the gasifier to achieve the outlet temperature of 2,800°F, while steam (700 psi) is fed such that the mass ratio of dry coal to steam is 7:1. This ratio was selected in order to inhibit methane formation in the gasifier. Although some heat is recovered in the membrane wall of the gasifier, the majority of the heat recovery is accomplished downstream of the gasifier in the syngas coolers, which cool the gas down to 464°F, generating high-pressure steam (Shell 2004). The syngas is further cooled by a water quench. A portion of the quenched syngas is returned to gasifier effluent to cool the particle-laden gas to below the ash-softening point. Makeup water is provided to the quench loop to achieve a blowdown rate of approximately 5% around the quench loop. This blowdown is then used in the slag quench loop. Of the water from the slag quench loop, 2.5% is assumed to be sent to water treatment to avoid any buildup of contaminants.

- **Syngas Conditioning (GAS-CLN)** – After gasification, the syngas is passed through sour shift reactors to maximize H_2 production via the water gas shift reaction as previously given in Equation 1. Steam (700 psi) is added to the syngas stream to maintain the water concentration necessary for the water gas shift reaction (steam-to-dry gas molar ratio of 1.2 is currently specified). To minimize the steam requirement, heat recuperation around the shift converters is employed in conjunction with a saturation/desaturation water recycle loop. Water recycle is limited in order to achieve a minimum inlet temperature to the second reactor of 450°F. Remaining condensed water is sent to water treatment to avoid high concentrations of ammonia and chloride compounds in the shift loop. Elemental mercury is then captured in a mercury guard bed. The syngas is further treated in an absorber with a solvent (a mixture of homologues of the dimethylether of polyethylene glycol) that acts as a chemical solvent for the removal of CO_2, H_2S, and COS (Selexol process). Absorption is carefully controlled to remove nearly all of the sulfur compounds while selectively removing only enough CO_2 to meet the downstream requirement for urea production. Additional removal of CO_2 is possible but would increase power consumption of the absorption process. The H_2S-rich stream is assumed to contain approximately 55% H_2S, with the remainder being CO_2 (Breckenridge 2000). H_2S containing gas from the sulfur reduction unit is also sent to the Selexol process for sulfur removal. Utility usage is calculated based on values presented in literature for the Selexol process (Burr 2008).

The sulfur-free gas is further purified using pressure swing adsorption (PSA) to produce a purified H_2 stream for ammonia synthesis. Nitrogen from the air separation unit is then mixed with the purified hydrogen to make a syngas suitable for ammonia synthesis (H_2-to-N_2 molar ratio of 3.0).

- **Sulfur Plant (CLAUS and S-REDUCT) –** Sulfur recovery is based on the Claus process. The Illinois coal has a sufficiently high sulfur content, which can create an acid gas stream with up to 60% H_2S. As a result, a straight-through Claus process can be used. In order to achieve optimal sulfur recovery, air flow to the Claus furnace is adjusted to achieve a molar ratio of 0.55:1 O_2 to H_2S. For this particular flow sheet, a Claus Oxygen-based Process Expansion (COPE) process was selected that uses pure oxygen as the oxidant rather than air. The advantage of this process over a conventional Claus process is that it minimizes inert gas recycle (N_2) to the Selexol unit, thus ensuring that a relatively pure stream of CO_2 can be produced for urea synthesis (Kohl 1997). Gas recycle is employed in the COPE process to limit the maximum temperature in the Claus furnace. Tail gas from the Claus unit is hydrogenated over a catalyst to convert the remaining sulfur species to H_2S, and this stream is recycled to the Selexol unit to maximize sulfur recovery. A small stream of clean syngas is used to fire and preheat the feed gas to the sulfur reduction unit.

- **CO_2 Purification/Compression (CO2-PUR)[a] –** Carbon dioxide is removed from the syngas in the Selexol process. By properly designing the solvent regeneration scheme, a relatively pure stream of CO_2 is produced. Final purification is accomplished using a zinc oxide sorbent bed to remove any remaining traces of sulfur, followed by a small catalytic oxidation unit to convert CO, H_2, and CH_4 to CO_2 and H_2O. Water is removed from the CO_2 using condensation, followed by compression to 2005 psi for urea synthesis. Eight stages are assumed for the CO_2 compression scheme, resulting in an overall efficiency of 84.4%. A small excess of CO_2 (above what is required for urea synthesis) is vented to the atmosphere.

- **Ammonia Synthesis (AMMONIA) –** Syngas feeding the ammonia synthesis unit has been previously adjusted to achieve the H_2/N_2 molar ratio 3.0. Incoming feed gas is compressed to 3,000 psi. Unreacted recycle gas is mixed with the fresh feed gas and preheated by cross exchanging

a. This block was modeled separately, but is included within the block labeled "Syngas Conditioning" in Figure 1.

with hot reactor effluent gases. Equilibrium conversion is assumed in the ammonia converters for the following reaction (Eggerman 2010):

$$N_2 + 3H_2 \rightarrow 2NH_3 \tag{2}$$

Effluent from the first ammonia converter is cooled by cross exchange with the reactor influent, followed by cooling in a steam generator. Additional steam is generated from the hot syngas downstream of the second and third ammonia conversion stages. Final cooling of the third stage effluent gas is accomplished using cooling water and recuperation with the cool recycle gas stream. Ammonia product is recovered in an ammonia separator. Effluent gas from this separator is further cooled using refrigeration. Additional ammonia is recovered in a second separator downstream of the refrigeration unit. Effluent gas from the second separator is recycled to the ammonia converters. Before entering the ammonia converters, the recycle gas is recompressed using a boost compressor and mixed with fresh syngas. Due to the very low concentrations of methane and argon entering the synthesis loop, inerts are removed from the synthesis loop with the ammonia product due to solubility alone.

Recovered ammonia is flashed to atmospheric pressure for storage. Ammonia in the flash gas is recovered in a wash column and subsequently distilled to remove water from the recovered product.

- **Nitric Acid Synthesis (HNO3) –** A monopressure process was selected for nitric acid synthesis due to capital cost advantages over a dual-pressure process. Cold ammonia is first preheated prior to mixing with compressed air. The ammonia-air mixture is reacted in an ammonia converter to produce N_2, NO, and H_2O. Due to the exothermic nature of the reactions, an exit temperature of 1,700°F is achieved (Clarke, 2005). Heat recuperation and steam generation are implemented downstream of the converter. As the gas cools, additional reaction with oxygen takes place to produce NO_2, N_2O_4, and finally HNO_3. Aqueous nitric is separated from the product stream while gaseous products are routed to an absorber to increase the yield of acid. A bleacher is included to further enhance reaction of NO with oxygen to improve acid yield. Unreacted gas exiting the absorber is routed to a selective catalytic reduction (SCR) unit for NO_x abatement prior to release to the atmosphere. This tailgas is also run through an expander, which significantly reduces the net electrical load required for air compression.

- **Ammonium Nitrate Synthesis (NH4NO3) –** Ammonia and nitric acid are preheated and then fed to a neutralizer where the following reaction occurs (Speight 2002):

$$NH_3 + HNO_3 \rightarrow NH_4NO_3 \tag{3}$$

The neutralizer products are then separated under slight negative pressure to remove excess water from the ammonium nitrate solution. Ninety-five percent of the solution is recycled back to the neutralizer, while 5% is withdrawn and further concentrated in a two-stage evaporator. Following evaporation, the concentrated ammonium nitrate solution is solidified using a prill tower. An Uhde-I. G. Farbenindustrie process was simulated. This particular process makes extensive use of process recycle and recuperation to minimize energy input requirements. Feeds to the process are controlled to maintain a pH of 1.5 in the neutralizer.

- **Urea Synthesis (UREA) –** The high-pressure Stamicarbon process was selected for urea synthesis. Principal feeds to this process are ammonia and CO_2. Reaction takes place at 2,005 psi; hence, compression of the feeds is required. The temperature is maintained around 360°F and NH_3/CO_2 molar ratio is maintained at 2.95:1. Ammonia and CO_2 react to form ammonium carbamate, which simultaneously dehydrates to form urea as (Mavrovic 2000):

$$\begin{aligned} 2NH_3 + CO_2 &\leftrightarrow NH_2CO_2NH_4 \\ NH_2CO_2NH_4 &\leftrightarrow NH_2CONH_2 + H_2O \end{aligned} \tag{4}$$

Liquid product is separated downstream of the reactor and fed to the CO_2 stripper. In the CO_2 stripper and downstream decomposer, unreacted carbamate is decomposed to ammonia and CO_2 to allow easy separation from the urea product. Recovered ammonia and CO_2 are cooled and condensed to reform ammonium carbamate. After stripping inert gases from the ammonium carbamate, this solution is recycled and mixed with fresh feed gases to the high-pressure carbamate condenser prior to introduction to the reactor.

Liquid urea product is concentrated using evaporation under vacuum conditions. The concentrated urea solution is then solidified using a fluidized bed granulator.

- **Power Production (HRSG-ST) –** The high- (700 psi) and low-pressure (60 psi) steams generated throughout the plant are sent to the power production block where they are passed through three steam turbines to generate power. The efficiencies of the turbines for the various steam pressures were calculated using Steam Pro, steam turbine modeling software from Thermoflow (Thermoflow 2009). It was found that even given low-quality steam at 60 psi, efficiencies for the steam turbines remain constant at approximately 81% combined thermodynamic and mechanical efficiency. It is recognized that the piping size for the 60 psia steam may be impractical and other methods for recovering the duty contained in this stream may be required. The condensed steams from the turbine outlets are mixed with condensate return from the plant, and makeup water is added to provide the necessary flow to the boiler feedwater pumps. Low-pressure steam is added to the deaerator. Purge gas produced in the plant is fired in the heat recovery steam generator (HRSG) to: (1) superheat and reheat steam feeds to the turbines, (2) raise additional steam to bolster power production, and (3) preheat boiler feed water prior to deaeration. Aspen Utility blocks are used to track all steam generation and use in the plant. This information is used as input to the power production section of the model, allowing reconciliation of the entire plant steam balance.

- **Cooling Towers (COOL-TWR) –** Conventional cooling towers are modeled in Aspen Plus using literature data. Air cooling could potentially be used in certain areas of the plant to decrease water consumption; however, for simplicity, cooling water only was assumed. The evaporation rate, drift, and blowdown are based on a rule-of-thumb guide for the design and simulation of wet cooling towers (Leeper 1981). Aspen Utility blocks are used to track all cooling water used in the plant. This information is used as input to the cooling tower section of the model, allowing reconciliation of the entire plant cooling water balance.

- **Water Treatment (H2O-TRTM) –** Water treatment is simplistically modeled in Aspen Plus using a variety of separation blocks. INL is currently collaborating with a major water treatment vendor to develop the water treatment portion of the model. The existing water treatment scenario is a place holder and will be revised as information is received from the water treatment vendor. Hence, it is anticipated that energy consumption for the water treatment portion of the plant could change considerably based on water treatment vendor feedback. Aspen Transfer blocks are used to reconcile water in and out flows from various parts of the plant, allowing reconciliation of the entire plant water balance.

2.2 Conventional Natural Gas-to-Ammonia Case

Figure 2 shows the block flow diagram for the conventional natural gas-to-ammonia case. The proposed process includes unit operations for natural gas reforming, syngas cleaning and conditioning, CO_2 compression, ammonia synthesis, nitric acid synthesis, ammonium nitrate synthesis, urea synthesis, power production, cooling towers, and water treatment. Each plant area is briefly described below. For each description, the name capitalized and enclosed in parentheses corresponds to the name of the hierarchy block within the Aspen process model. Because many of the unit operations downstream of syngas production remain unchanged from the conventional coal-to-ammonia flow sheet, emphasis for these operations is placed on differences in configuration between the cases.

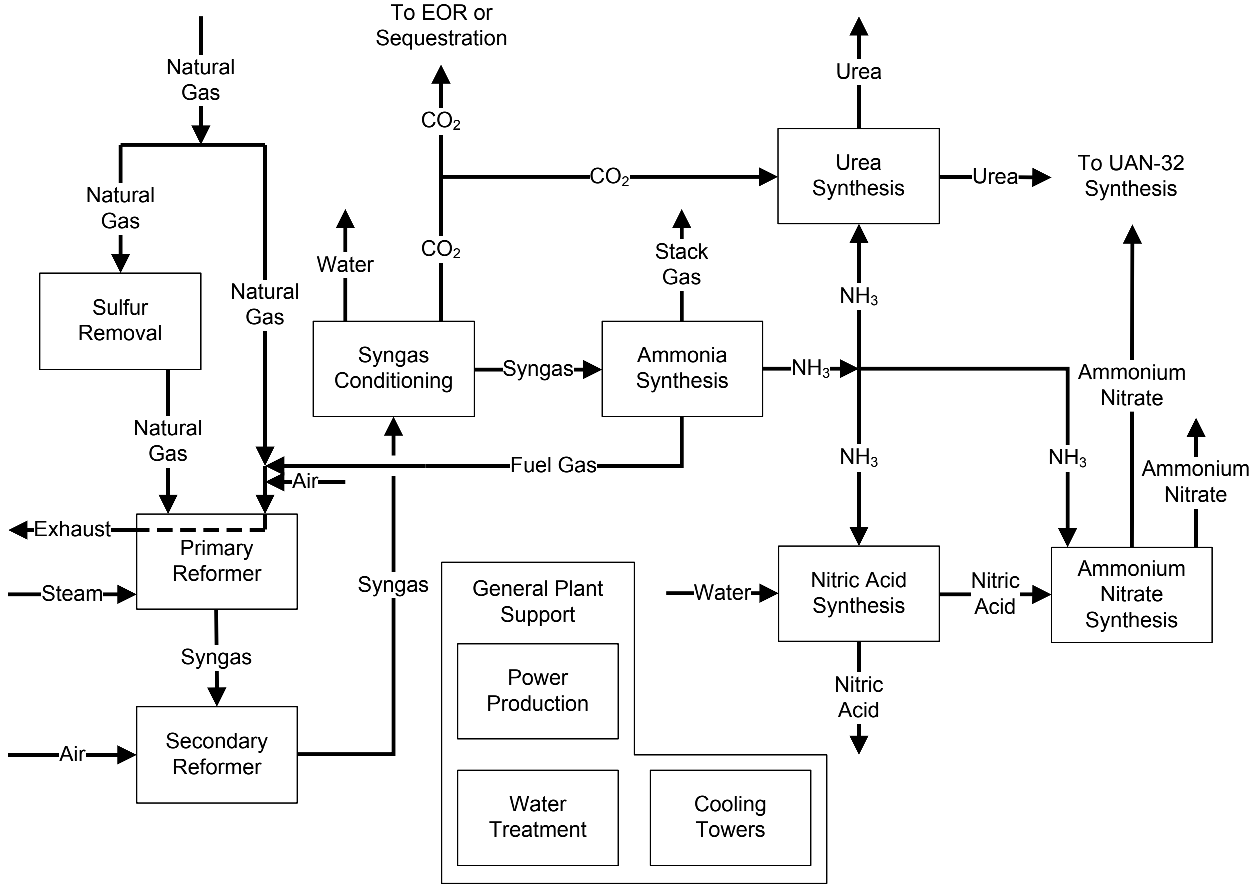

Figure 2. Block flow diagram for the conventional natural gas-to-ammonia process.

- **Natural Gas Purification and Reforming (REFORMER)**[b] – Two-step reforming consisting of primary steam reforming followed by secondary autothermal reforming was selected for syngas generation. Air is used as the oxidant in the autothermal reforming step, as this provides nitrogen to the process for downstream ammonia synthesis (Eggerman 2010). By carefully controlling process parameters, such as the steam-to-carbon inlet molar ratio, primary reformer temperature, amount of preheat to the secondary reformer, and secondary reformer temperature, a syngas containing the appropriate H_2/N_2 ratio for ammonia synthesis can be produced. Additionally, if the steam-to-carbon ratio is set high enough, additional water will not be required prior to downstream shift conversion. For this case, key parameters were set as follows:

b. The blocks labeled "Sulfur Removal," "Primary Reformer," and "Secondary Reformer" shown in Figure 2 are all included within this Aspen hierarchy block.

Preformer steam-to-carbon ratio	3.30
Primary reformer exit temperature	1,454°F (790°C)
Autothermal reformer outlet temperature	1,750°F (954°C)

Natural gas is split into two streams. Of the total natural gas flow, 22.7% is burned to provide heat for the primary reformer. The remaining 77.3% of the natural gas flow is compressed to 615 psi and then preheated to 329°F and saturated with hot water. After saturation, the gas is further heated to 662°F and mixed with a small amount of hydrogen. Sulfur is removed from the gas and then mixed with steam to achieve the desired steam-to-carbon molar ratio of 3.3. Because the resulting natural gas/steam mixture is preheated to only 1000°F, a preformer is not included in this flowsheet.

The natural gas/steam mixture is fed to the primary reformer, where methane is converted over a catalyst to CO, H_2, and CO_2. Methane conversion in this reactor is approximately 53%. A separate feed of the natural gas is mixed with fuel gas and burned to provide heat for the endothermic reforming reactions. The hot offgas from the reformer is exchanged with inlet syngas, water, air, and steam to provide preheat for these streams.

The effluent from the primary reformer and a preheated air stream are fed to the autothermal reformer where conversion of the remaining methane to syngas is accomplished. The oxygen-to-carbon molar ratio is set at 0.28, resulting in an exit temperature of 1,750°F. The hot syngas is cooled rapidly by exchange with boiler feed water to create high-pressure steam. The resulting syngas has a H_2/CO ratio of 4.4 and contains 5 mol% CO_2 and 0.5 mol% unreacted CH_4.

- **Syngas Conditioning (GAS-CLN)** – Syngas cleaning and conditioning for the natural gas-fed case is similar to the coal-fed case. However, the following changes are made for the natural gas-fed case:

  1. Water saturation is not necessary prior to shift conversion. By introducing sufficient steam to the reforming section of the plant, the syngas entering the high-temperature shift converter contains sufficient moisture for shift conversion. The steam-to-dry gas ratio of the syngas is 0.54, which is above the minimum requirement for the catalyst (0.40).

  2. The mercury sorbent bed is not required.

  3. Because sulfur has been removed prior to reforming, Selexol is used for CO_2 removal only. The Selexol unit is operated to remove the majority of the CO_2; hence, a PSA unit is not required.

The resulting syngas contains H_2 and N_2 in a molar ratio of 3:1. Small amounts of argon (0.3 mol%) and methane (1.1%) are also present in the cleaned syngas.

- **CO_2 Compression (CO2-COMP)[c] –** CO_2 compression in the natural gas-fed case is identical to the coal-fed case. However, because sulfur has been removed from the natural gas prior to reforming, sulfur removal from the CO_2 is not required.

- **Ammonia Synthesis (AMMONIA) –** Ammonia synthesis in the natural gas-fed case is identical to the coal-fed case. However, due the higher concentration of inerts in the feed gas (argon and methane), a purge is required. For this case, the purge rate is set at 0.54% of the recycle gas flow, resulting in an inert content in the recycle gas of about 15 mol%. Because the purge flow is significant, 90% of the H_2 and 30% of the N_2 are recovered using membrane separation. Recovered H_2 and N_2 are recompressed and remixed into the synthesis loop. The waste stream from the membrane is burned as fuel gas in the primary reformer.

- **Nitric Acid Synthesis (HNO3) –** Nitric acid synthesis in the natural gas-fed case is identical to the coal-fed case.

- **Ammonium Nitrate Synthesis (NH4NO3) –** Ammonium nitrate synthesis in the natural gas-fed case is identical to the coal-fed case.

- **Urea Synthesis (UREA) –** Urea synthesis in the natural gas-fed case is identical to the coal-fed case.

- **Power Production (HRSG-ST) –** The steam system in the natural gas-fed case is identical to the coal-fed case. However, power production is accomplished solely from steam generated throughout the plant. Purge gas from the ammonia synthesis loop is burned in the primary reformer; hence, it is not available for firing a HRSG.

- **Cooling Towers (COOL-TWR) –** The cooling water system in the natural gas-fed case is identical to the coal-fed case.

- **Water Treatment (H2O-TRTM) –** Water treatment for the natural gas-fed case is similar to the coal-fed case. Additional detail for the water treatment system will be added when vendor feedback has been received.

2.3 Nuclear-Integrated Coal-to-Ammonia Case

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c. This block was modeled separately but is included within the block labeled "Syngas Conditioning" in Figure 2.

Figure 3 shows the block flow diagram for the nuclear-integrated coal-to-ammonia case. As this diagram shows, a need for high-temperature heat (>300°C) integration was not identified. Additionally, modeling results for the conventional coal-to-ammonia case indicate that heat for coal drying, water treatment, and urea synthesis can be supplied by existing unit operations in the conventional configuration. Hence, electricity is the best integration need for this flow sheet. Because light water reactors could supply this need, there is no driving reason to integrate this flow sheet with nuclear HTGRs. Hence, process modeling was not performed for this case.

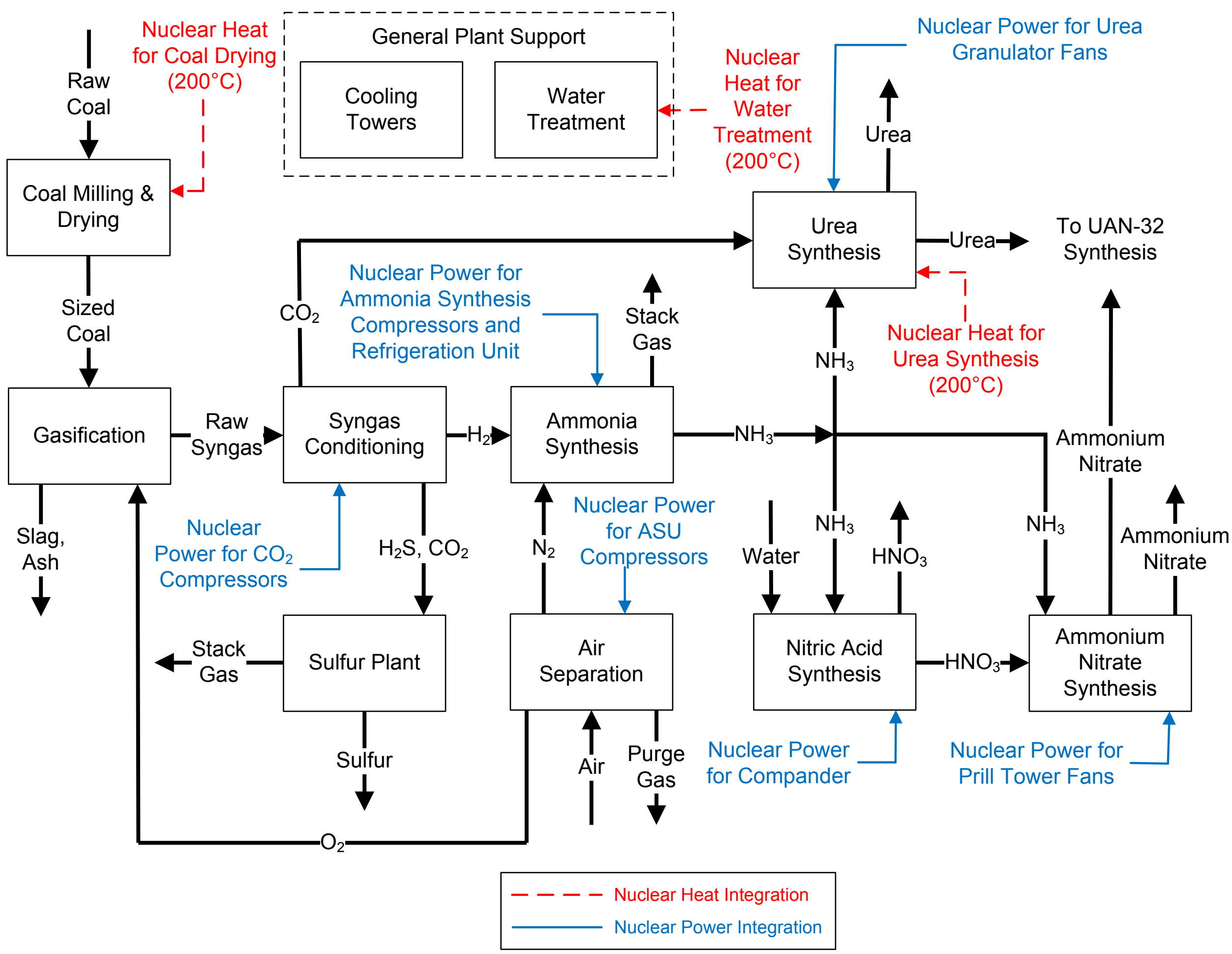

Figure 3. Block flow diagram for the nuclear-integrated coal-to-ammonia process.

2.4 Nuclear-Integrated Natural Gas-to-Ammonia Case

Figure 4 shows the block flow diagram for the nuclear-integrated natural gas-to-ammonia case. The proposed process includes the same unit operations as the conventional natural gas process except nuclear heat is used in the primary reformer rather than burning natural gas. Nuclear heat is also used to preheat all streams entering the primary reformer. Although nuclear heat was also considered for use in the urea synthesis and water treatment plants, sufficient low-temperature heat was produced elsewhere in the plant to meet these requirements. Because higher temperature heat is desired for steam reforming of natural gas, this configuration assumes the use of an intermediate heat exchanger (IHX) to provide 700°C helium gas as the heat transfer fluid to the reformer.

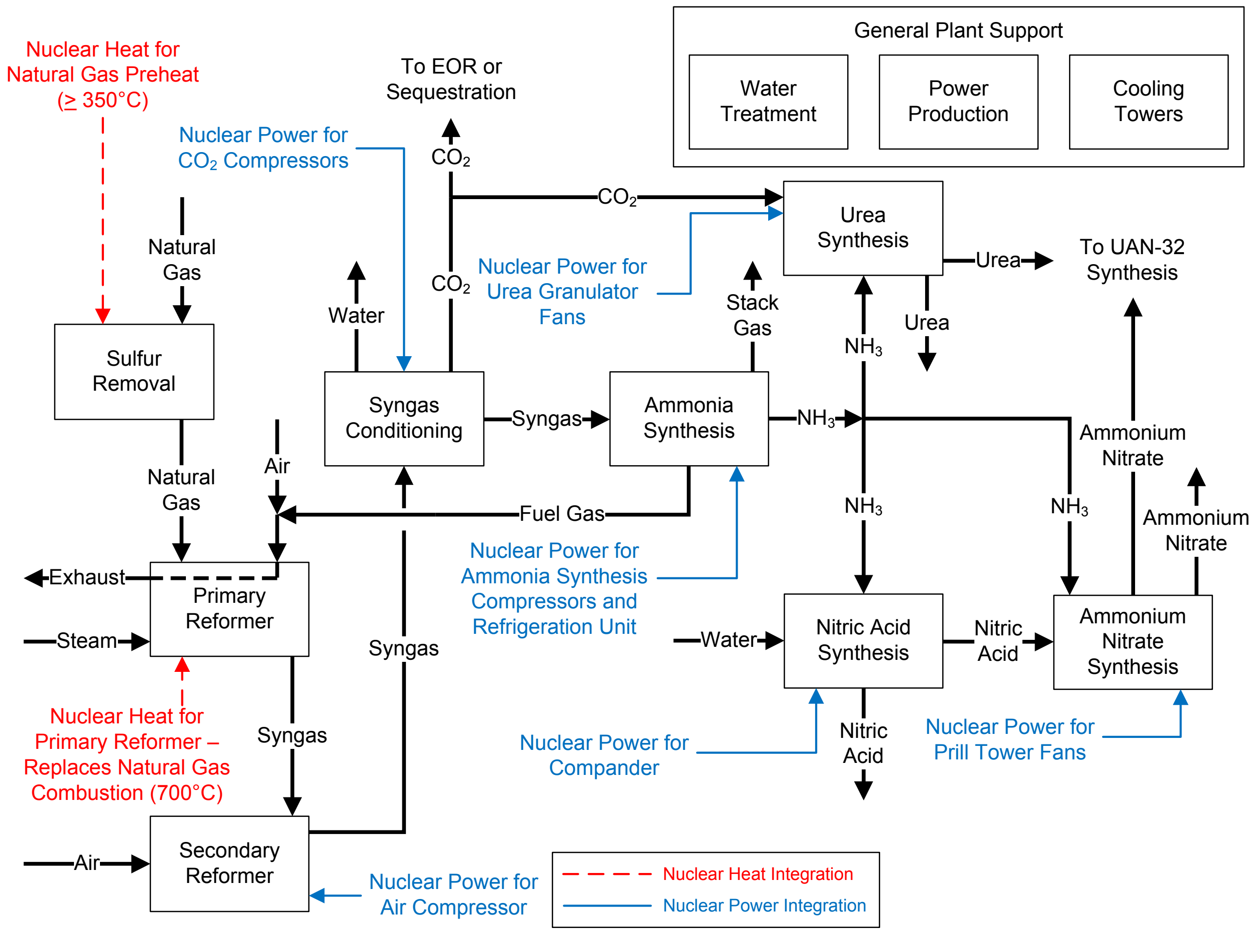

Figure 4. Block flow diagram for the nuclear-integrated natural gas-to-ammonia process.

The proposed process includes unit operations for natural gas reforming, syngas cleaning and conditioning, CO_2 compression, ammonia synthesis, nitric acid synthesis, ammonium nitrate synthesis, urea synthesis, power production, cooling towers, and water treatment. Each unit operation is briefly described below. For each description, the name capitalized and enclosed in parentheses corresponds to the name of the hierarchy block within the Aspen process model. Because the majority of unit operations remain unchanged from the conventional natural gas-to-ammonia flowsheet, emphasis is placed on differences in configuration between cases.

- **Natural Gas Purification and Reforming (REFORMER)**[d] – Conditions in the reforming section of the plant had to be modified slightly because of the upper temperature limit of helium that can be delivered from the nuclear plant (700°C). Because of this constraint, primary reforming was split into two stages. In the first stage, nuclear heat is provided by hot helium exchange. The temperature of this reforming stage was set at 1,274°F (690°C). In the second stage of the primary reformer, purge gas is burned as the heat source. The temperature of this reforming stage was set at 1,400°F (760°C). Key reforming parameters for this case are summarized below:

Preformer steam-to-carbon ratio	3.30
Primary reformer stage 1 exit temperature	1,274°F (690°C)
Primary reformer stage 2 exit temperature	1,400°F (760°C)
Autothermal reformer outlet temperature	1,650°F (899°C)

  With these changes, methane conversion in the primary reformer decreased from 53% to 46%. Feed O_2/C and effluent syngas H_2/CO ratios for the autothermal reformer were 0.27 and 4.81, respectively.

- **Syngas Cleaning and Conditioning (GAS-CLN)** – Syngas cleaning and conditioning for the nuclear-integrated natural gas-fed case is identical to the conventional natural gas-fed case.

- **CO_2 Compression (CO2-COMP)**[e] **–** CO_2 compression for the nuclear-integrated natural gas-fed case is identical to the conventional natural gas-fed case.

- **Ammonia Synthesis (AMMONIA) –** Ammonia synthesis for the nuclear-integrated natural gas-fed case is identical to the conventional natural gas-fed case. Due to slightly higher inerts in the synthesis feed gas, a slightly higher purge rate is required.

d. The blocks labeled “Sulfur Removal,” “Primary Reformer,” and “Secondary Reformer” shown in Figure 4 are all included within this Aspen hierarchy block.

e. This block was modeled separately but is included within the block labeled “Syngas Conditioning” in Figure 4.

- **Nitric Acid Synthesis (HNO3) –** Nitric acid synthesis for the nuclear-integrated natural gas-fed case is identical to the conventional natural gas-fed case.

- **Ammonium Nitrate Synthesis (NH4NO3) –** Ammonium nitrate synthesis for the nuclear-integrated natural gas-fed case is identical to the conventional natural gas-fed case.

- **Urea Synthesis (UREA) –** Urea synthesis for the nuclear-integrated natural gas-fed case is identical to the conventional natural gas-fed case.

- **Power Production (HRSG-ST) –** The steam system for the nuclear-integrated natural gas-fed case is identical to the conventional natural gas-fed case.

- **Cooling Towers (COOL-TWR) –** The cooling water system for the nuclear-integrated natural gas-fed case is identical to the conventional natural gas-fed case.

- **Water Treatment (H2O-TRTM) –** Water treatment for the nuclear-integrated natural gas-fed case is identical to the conventional natural gas-fed case.

2.5 HTSE Hydrogen-to-Ammonia Case (Option 1: H_2/Air Combustion as the N_2 Source)

Figure 5 shows the block flow diagram for the HTSE hydrogen-to-ammonia case, which uses H_2/air combustion as the nitrogen source. The proposed process includes the same downstream unit operations as the nuclear-integrated and conventional ammonia processes. Syngas generation, however, is considerably modified. Hydrogen for ammonia synthesis is produced by HTSE. Nuclear HTGRs supply heat and power for the electrolysis units. Burning a fraction of the electrolysis H_2 with air generates a nitrogen-rich combustion gas. Water is subsequently condensed out of the combustion gas to yield nitrogen for ammonia synthesis.

Urea production requires CO_2 as a feed to the process. For this case, a CO_2 supply is not readily available and must be produced to support urea manufacture; hence, natural gas is burned with oxygen to produce the required CO_2. While coal could have been selected for this purpose, natural gas is preferred because of its much simpler gas cleanup requirements.

An opportunity was identified to use heat from the natural gas burner as topping heat for the electrolysis unit. As HTGR technology matures and reactor outlet temperatures increase, the nuclear reactors may be able to supply electrolysis topping heat. However, because of the upper limit of 700°C deliverable heat assumed in this study, supplying topping heat from the burner to the electrolyzers is an attractive means of increasing electrolyzer efficiency.

The proposed process includes unit operations for H_2 generation via high-temperature electrolysis, N_2 generation via H_2/air combustion, syngas purification, CO_2 generation via oxy-combustion of natural gas, CO_2 compression, ammonia synthesis, nitric acid synthesis, ammonium nitrate synthesis, urea synthesis, power production, cooling towers, and water treatment. Each unit operation is briefly described below. For each description, the name capitalized and enclosed in parentheses corresponds to the name of the hierarchy block within the Aspen process model. Because the majority of unit operations remain unchanged from the previously described ammonia flow sheets, emphasis is placed on differences in configuration between cases.

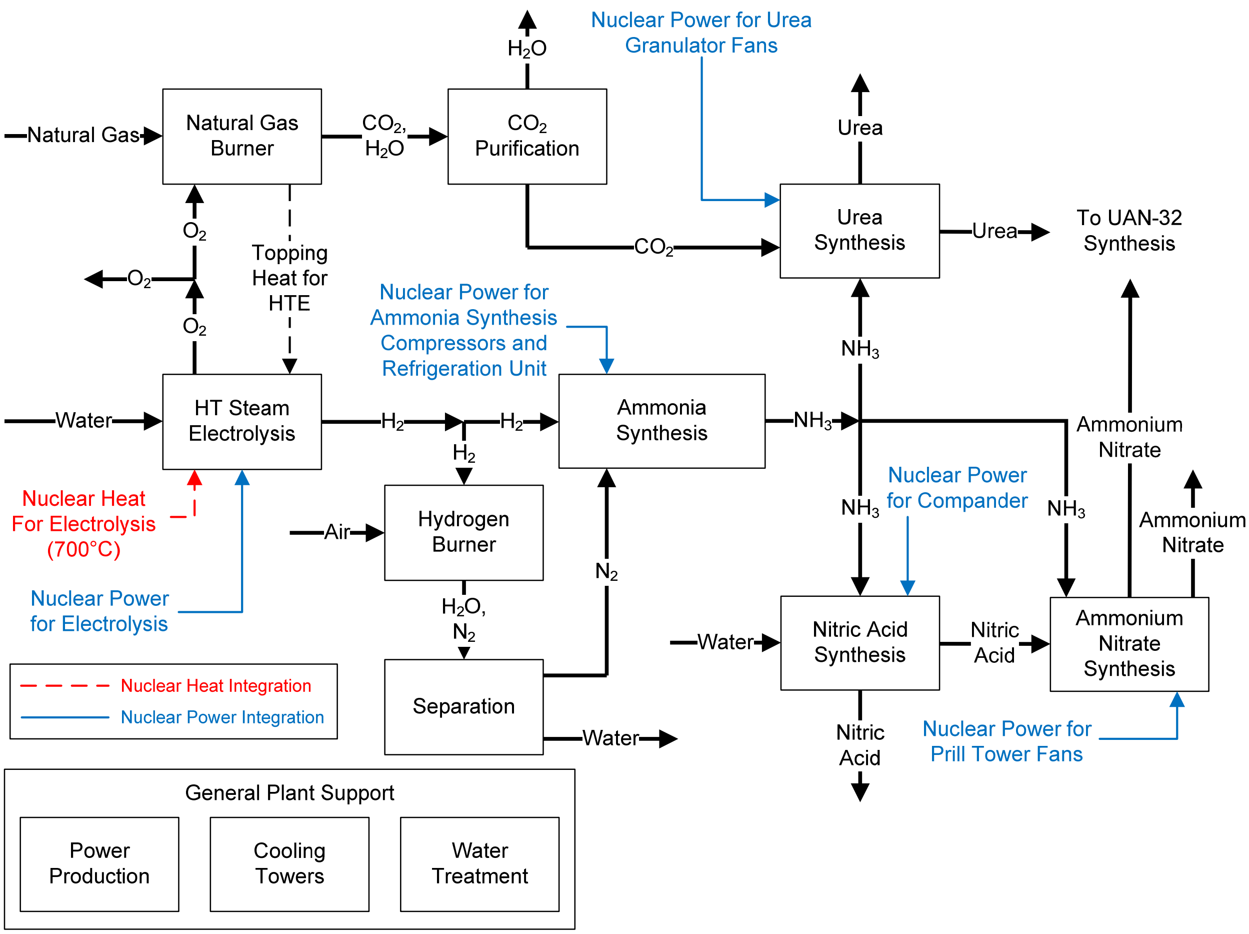

Figure 5. Block flow diagram for the HTSE hydrogen-to-ammonia process (N_2 from H_2/air combustion).

- **HT Steam Electrolysis (HTE)** – Hydrogen is generated from water via electrolysis. INL's detailed model for high-temperature electrolysis has not yet been converted to Aspen Plus. Hence, a very simplistic model is included and has been tuned using INL's detailed electrolysis model. Efficiency of the electrolysis process is improved by operating at elevated temperatures; 800°C was selected as the operating temperature in this study. Because of the presently assumed 700°C limit for heat delivery from the HTGRs, topping heat is supplied to the electrolysis process from the combustor in the CO_2 generation plant. Products from the electrolyzer include purified H_2 and O_2. The excess O_2 produced can be sold as an end product from this process.

- **Hydrogen Burner (N2-GEN)** – Nitrogen is generated by burning a fraction of the hydrogen product from electrolysis with air. For this case, 15.4% of the total electrolysis H_2 production is burned to produce N_2. Air is pressurized to 720 psi prior to being fed to the burner. Steam injection is used to moderate temperature within the burner; outlet temperature from the burner is controlled to 1,800°F. The hot exhaust gas from the burner is used to raise steam. Water is then condensed from the gas to produce a nitrogen stream for ammonia synthesis.

- **CO_2 Purification (CO2-GEN) –** Carbon dioxide for urea synthesis is generated by combustion of natural gas. The gas is preheated against hot combustion gas and sulfur is removed using a bed of zinc oxide. Oxygen from the electrolyzers is used as the oxidant in order to produce a nitrogen-free stream of CO_2. The hot combustion gas is used first to supply topping heat to the electrolyzer, followed by raising high-pressure steam in a heat-recovery boiler. The hot combustion gas is then used for recuperation with the natural gas feed stream and finally to raise low-pressure steam. A large fraction of the combustion gas is recycled to the combustor to moderate temperature within the burner. Water is then condensed from the gas stream prior to compression to 2,005 psi in preparation for urea synthesis.

- **Syngas Conditioning (GAS-CLN)**[f] – Syngas cleaning is similar to the conventional natural gas-fed case, but significantly simpler—only methanation and water removal are required.

- **Ammonia Synthesis (AMMONIA) –** Ammonia synthesis for this case is identical to the conventional natural gas-fed case.

- **Nitric Acid Synthesis (HNO3) –** Nitric acid synthesis for this case is identical to the conventional natural gas-fed case.

- **Ammonium Nitrate Synthesis (NH4NO3) –** Ammonium nitrate synthesis for this case is identical to the conventional natural gas-fed case.

f. This block was modeled separately but is included within the block labeled "Ammonia Synthesis" in Figure 5.

- **Urea Synthesis (UREA) –** Urea synthesis for this case is identical to the conventional natural gas-fed case.
- **Power Production (HRSG-ST) –** The steam system for this case is identical to the conventional natural gas-fed case.
- **Cooling Towers (COOL-TWR) –** The cooling water system for the nuclear-integrated natural gas-fed case is identical to the conventional natural gas-fed case.
- **Water Treatment (H2O-TRTM) –** Water treatment for this case is identical to the conventional natural gas-fed case.

2.6 HTSE Hydrogen-to-Ammonia Case (Option 2: Cryogenic ASU as the N_2 Source)

Figure 6 shows the block flow diagram for the HTSE hydrogen-to-ammonia case that uses cryogenic air separation as the nitrogen source. The proposed process includes similar unit operations as in the HTSE case with H_2/air combustion for N_2 generation. Hydrogen for ammonia synthesis is produced by HTSE. Nuclear HTGRs supply heat and power for the electrolysis units. However, a cryogenic air separation unit is used for N_2 generation instead of H_2 combustion.

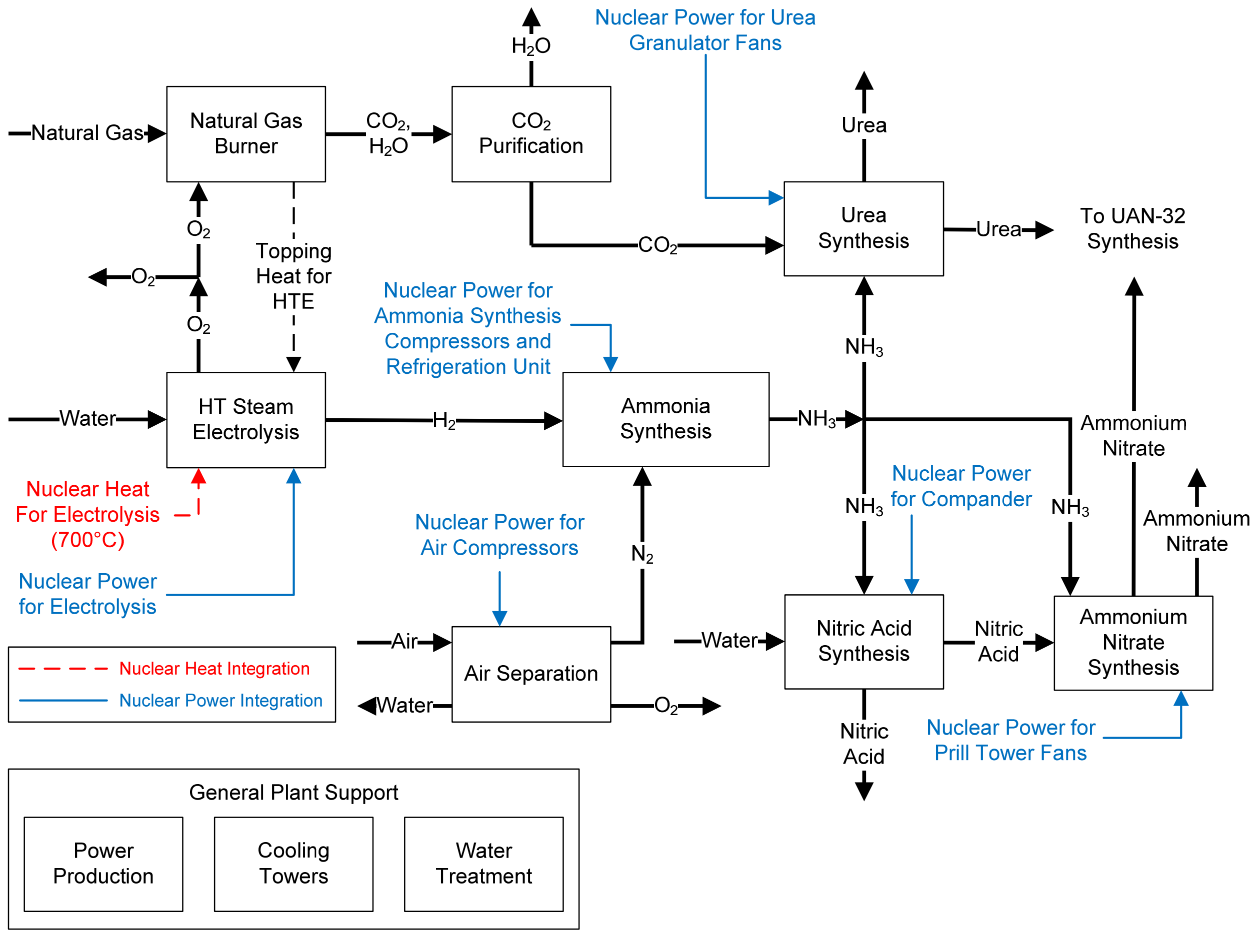

Figure 6. Block flow diagram for the HTSE hydrogen-to-ammonia process (N_2 from cryogenic ASU).

Urea production requires CO_2 as a feed to the process. For this case, a CO_2 supply is not readily available and must be produced to support urea manufacture; hence, natural gas is burned with oxygen to produce the required CO_2. While coal could have been selected for this purpose, natural gas is preferred because of the much simpler gas cleanup requirements.

An opportunity was identified to use heat from the natural gas burner as topping heat for the electrolysis unit. As HTGR technology matures and reactor outlet temperatures increase, the nuclear reactors may be able to supply electrolysis topping heat. However, because of the upper limit of 700°C deliverable heat assumed in this study, supplying topping heat from the burner to the electrolyzers is an attractive means of increasing electrolyzer efficiency.

The proposed process includes unit operations for H_2 generation via high-temperature electrolysis, N_2 generation via cryogenic air separation, syngas purification, CO_2 generation via oxy-combustion of natural gas, CO_2

compression, ammonia synthesis, nitric acid synthesis, ammonium nitrate synthesis, urea synthesis, power production, cooling towers, and water treatment. Each unit operation is briefly described below. For each description, the name capitalized and enclosed in parentheses corresponds to the name of the hierarchy block within the Aspen process model. Because the majority of unit operations remain unchanged from the previously described ammonia flow sheets, emphasis is placed on differences in configuration between cases.

- **HT Steam Electrolysis (HTE)** – Hydrogen generation for this case is identical to the HTSE hydrogen-to-ammonia case using H_2/air combustion for the N_2 source.

- **Air Separation (ASU)** – Nitrogen generation for this case is identical to the conventional coal-fed case. However, in that scenario, the throughput of the ASU was driven by oxygen requirements for gasification. In this case, a much smaller unit can be utilized because throughput is driven by N_2 demand for ammonia synthesis. The excess O_2 produced could potentially be sold as an end product from this process.

- **CO_2 Purification (CO2-GEN) –** Carbon dioxide generation for this case is identical to the HTSE hydrogen-to-ammonia case using H_2/air combustion for the N_2 source.

- **Syngas Conditioning (GAS-CLN)[g]** – Syngas cleaning for this case is identical to the HTSE hydrogen-to-ammonia case using H_2/air combustion for the N_2 source.

- **Ammonia Synthesis (Ammonia) –** Ammonia synthesis for this case is identical to the HTSE hydrogen-to-ammonia case using H_2/air combustion for the N_2 source.

- **Nitric Acid Synthesis (HNO3) –** Nitric acid synthesis for this case is identical to the HTSE hydrogen-to-ammonia case using H_2/air combustion for the N_2 source.

- **Ammonium Nitrate Synthesis (NH4NO3) –** Ammonium nitrate synthesis for this case is identical to the HTSE hydrogen-to-ammonia case using H_2/air combustion for the N_2 source.

- **Urea Synthesis (UREA) –** Urea synthesis for this case is identical to the HTSE hydrogen-to-ammonia case using H_2/air combustion for the N_2 source.

g. This block was modeled separately but is included within the block labeled "Ammonia Synthesis" in Figure 6.

- **Power Production (STEAM) –** The steam system for this case is similar to the HTSE hydrogen-to-ammonia case using H_2/air combustion for the N_2 source.

- **Cooling Towers (COOL-TWR) –** The cooling water system for this case is identical to the HTSE hydrogen-to-ammonia case using H_2/air combustion for the N_2 source.

- **Water Treatment (H2O-TRTM) –** Water treatment for this case is identical to the HTSE hydrogen-to-ammonia case using H_2/air combustion for the N_2 source.

3. PROCESS MODELING RESULTS

Based on the results for the conventional coal-to-ammonia case, a need for high-temperature (>300°C) heat integration was not identified for the coal-fed case. In addition, because light water reactors can supply power in this scenario, a strong need for integration with HTGRs is not apparent. Hence, modeling of a nuclear-integrated coal-to-ammonia case was not pursued. Note, however, that other gasification technologies (hydrogasification) could be considered, and these technologies may be more complementary for HTGR integration.

Analysis of the conventional natural-gas-to-ammonia case indicates a strong potential heat integration opportunity for a HTGR. In the conventional case, 22.7% of the natural gas feed to the process is burned to provide heat in the primary reformer. The operating temperature in the primary reformer is 790°C, which is somewhat higher than the assumed 700°C temperature that can be supplied by an HTGR in this analysis. Analysis seems to indicate, however, that primary reforming can be accomplished in two stages—the first using nuclear heat at 690°C, followed by a second stage reformer fueled entirely by purge gas that operates at a somewhat higher temperature.

A summary of the modeling results for all cases is presented in Table 2. The conventional coal and natural gas cases serve as a basis for comparison with the nuclear-integrated cases. Figure 7 graphically presents a high-level material and energy balance summary for each case. For the complete Aspen stream results for all cases, see Appendixes C, D, E, F, and G.

Results for the nuclear-integrated natural gas-to-ammonia case look promising. One 450 MW_t HTGR would be required to support this configuration. The split of heat to power required from the nuclear plant is 42%/58%. By substituting nuclear heat for natural gas combustion in the primary reformer, natural gas consumption can be reduced by 18%. In addition, CO_2 emissions can be reduced by 70%.

Results for the HTSE hydrogen cases also look promising. CO_2 emissions are cut to very low levels, and the need to sequester carbon is eliminated. In addition, natural gas consumption is reduced to only the amount required to supply carbon for urea production.

These configurations would require roughly four 600 MW_t HTGRs to support plant operation. It is apparent that, of the two HTSE hydrogen cases considered, the option utilizing cryogenic air separation for nitrogen production is more attractive from the perspectives of capital equipment cost.

Table 2. Ammonia modeling case study results.

	Conventional Coal-To-Ammonia Derivatives Process	Conventional Gas-To-Ammonia Derivatives Process	Nuclear-Integrated Gas-to-Ammonia Derivatives Process	HTSE-to-Ammonia Derivatives Process – N_2 from Combustion	HTSE-to-Ammonia Derivatives Process – N_2 from ASU
Inputs					
Coal Feed Rate (ton/day)	4,591	0	0	0	0
Natural Gas Feed Rate (MMSCFD)	N/A	88	72	36	36
# 600 MW_t HTGRs Required	N/A	N/A	0.75	4.10	3.58
Intermediate Products					
Ammonia (ton/day)	3,358	3,361	3,360	3,360	3,360
Nitric Acid (ton/day)	5,186	5,192	5,192	5,192	5,192
Outputs					
Urea (ton/day)	2,939	2,939	2,939	2,939	2,939
Ammonium Nitrate (ton/day)	3,779	3,779	3,779	3,779	3,779
Oxygen (ton/day)	0	0	0	2,508	2,435
Utility Summary					
Total Power (MW_e)	-141.8	-108.8	-106.2	-879.0	-769.5
Electrolyzers	N/A	N/A	N/A	-918.1	-778.1
Air Separation	-73.5	N/A	N/A	N/A	-23.2
Nitrogen Generation (non-ASU)	N/A	N/A	N/A	-18.8	N/A
Coal Milling and Drying	-2.5	N/A	N/A	N/A	N/A
Gasification Island	-2.5	N/A	N/A	N/A	N/A
Natural Gas Reforming	N/A	-21.6	-19.9	N/A	N/A
Syngas Purification	-5.4	-4.4	-4.5	N/A	N/A
Claus Plant	0.0	N/A	N/A	N/A	N/A
Sulfur Reduction (SCOT)	-0.7	N/A	N/A	N/A	N/A
Non-Nuclear Power Island	44.7	26.7	29.5	156.0	128.0
CO_2 Production/Purification/Compression	-7.4	-13.0	-13.1	-5.6	-5.6

Table 2. Ammonia modeling case study results.

	Conventional Coal-To-Ammonia Derivatives Process	Conventional Gas-To-Ammonia Derivatives Process	Nuclear-Integrated Gas-to-Ammonia Derivatives Process	HTSE-to-Ammonia Derivatives Process – N_2 from Combustion	HTSE-to-Ammonia Derivatives Process – N_2 from ASU
Ammonia Synthesis	-43.2	-45.2	-46.8	-39.0	-38.5
Nitric Acid Synthesis	-15.1	-15.1	-15.1	-15.1	-15.1
Ammonium Nitrate Synthesis	-24.9	-24.9	-24.9	-24.9	-24.9
Urea Synthesis	-4.4	-4.4	-4.4	-4.4	-4.4
Cooling Towers	-2.1	-1.5	-1.5	-2.5	-2.3
Water Treatment	-4.8	-5.4	-5.6	-6.6	-5.6
Fossil Plant Water Balance (gpm)*	-4,466	-3,928	-3,984	-6,334	-5,555
Evaporation Rate (gpm)	-4,891	-3,601	-3,657	-5,979	-5,386
CO_2 Emissions					
Captured (ton/day CO_2)	0	1,762	1,786	0	0
Emitted (ton/day CO_2)	7,987	1,449	430	11	11
Nuclear Integration Summary					
Electricity Demand (MW_e)	N/A	N/A	106.2	879.0	769.5
HTSE	N/A	N/A	N/A	918.1	778.1
Balance of Plant	N/A	N/A	106.2	-39.1	-8.6
Nuclear Heat Demand (MMBTU/hr)	N/A	N/A	649.2	894.6	758.2
Electrolyzers	N/A	N/A	N/A	894.6	758.2
Natural Gas Reforming	N/A	N/A	649.2	N/A	N/A
Fossil Topping Heat Supplied (MMBTU/hr)	N/A	N/A	N/A	50.4	42.7
Electrolysis Products					
Total Hydrogen (ton/day)	N/A	N/A	N/A	715	606
Total Oxygen (ton/day)	N/A	N/A	N/A	5,637	4,778

*Detailed water balance can be found in Appendix A.

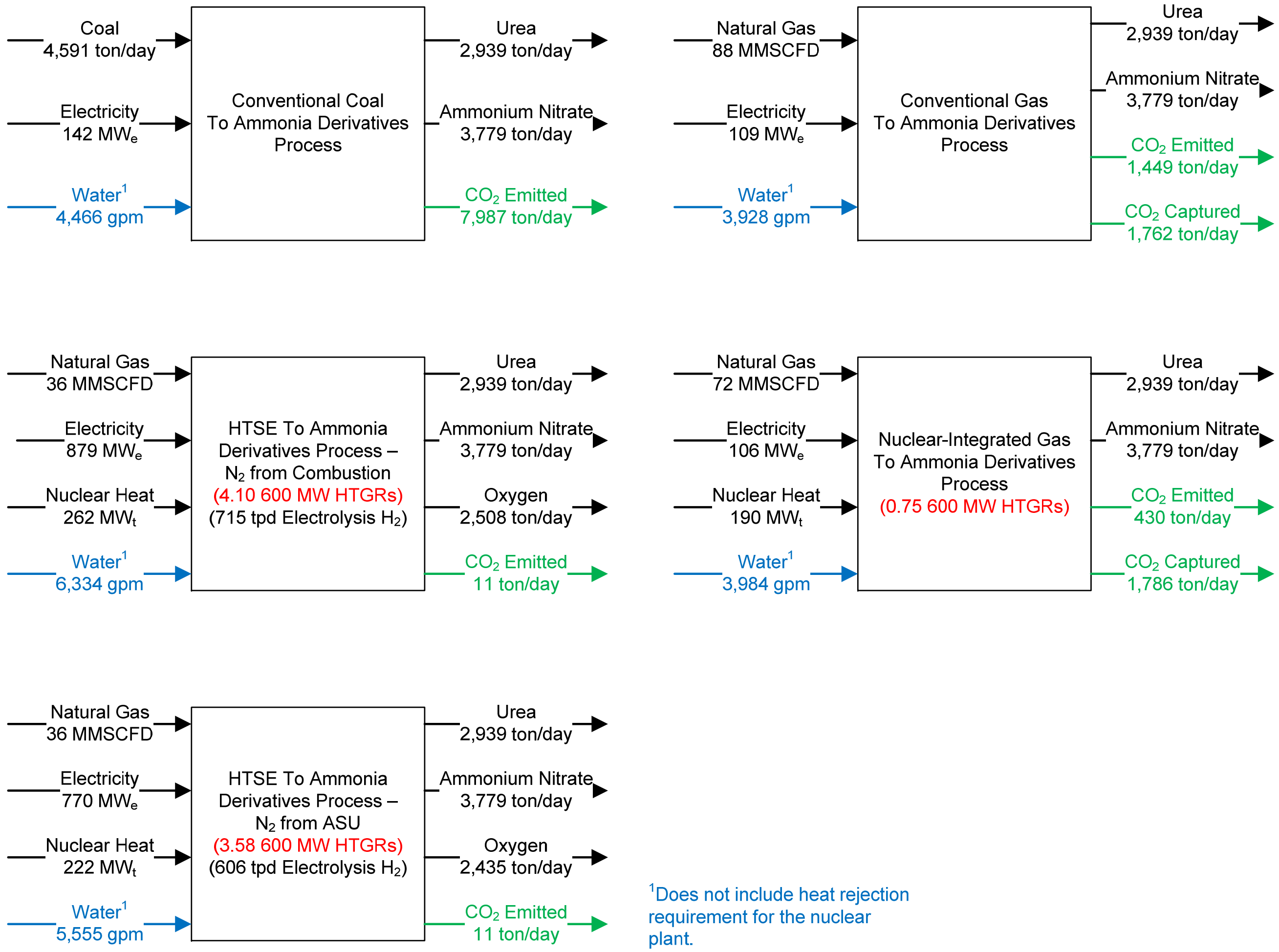

Figure 7. Ammonia modeling case material balance summary.[h]

[h] Water balance breakdown can be found in Appendix A.

4. ECONOMIC MODELING OVERVIEW

The economic viability of the ammonia processes was assessed using standard economic evaluation methods. The economics were evaluated for the conventional and nuclear-integrated options described in the previous sections. The total capital investment (TCI), based on the total equipment costs, annual revenues, and annual manufacturing costs were first calculated for the cases. The present worth (PW) of the annual cash flows (after taxes) was then calculated for the TCI, as well as the TCI at + 50% and – 30% of the HTGR cost, with the debt to equity ratios equal to 80%/20% and 55%/45%. The following sections describe the methods used to calculate the capital costs, annual revenues, annual manufacturing costs, and the resulting economic results.

4.1 Capital Cost Estimation

Equipment items for this study were not individually priced. Rather, cost estimates were based on scaled costs for major plant processes from published literature.

For the coal-to-ammonia scenarios, cost estimates were generated for:

- Coal preparation
- The ASU
- Gasification
- Gas cleanup
- Ammonia synthesis
- Nitric acid synthesis
- Urea synthesis
- Ammonium nitrate synthesis
- Steam turbines
- The HRSG
- Cooling towers.

For the gas-to-ammonia scenarios, cost estimates were generated for:

- SMR
- Gas cleanup
- Ammonia synthesis
- Nitric acid synthesis
- Urea synthesis
- Ammonium nitrate synthesis
- Steam turbines
- The HRSG
- Cooling towers
- The HTGR.

For the HTSE-to-ammonia scenarios, cost estimates were generated for:

- HTSE
- Gas cleanup
- Ammonia synthesis
- Nitric acid synthesis
- Urea synthesis
- Ammonium nitrate synthesis
- Steam turbines
- The HRSG
- Cooling towers
- The HTGR.

In some instances, several costs were averaged. Gas cleanup includes costs for water gas shift reactors, the Selexol process, sulfur recovery, pressure swing absorption, methanation, and CO_2 compression/liquefaction as appropriate for the various process configurations. Appendix B presents the detailed breakdown for the equipment item costs. It is assumed that there is no impact on the capital cost of the ammonia facility when sequestration is not assumed, as the Selexol process is required for gas cleanup, and though the last stage of the CO_2 compressor would not be required, this cost is negligible when compared to the TCI required for the ammonia process. The estimate presented is a Class 5 estimate and has a probable error of +50% and –30% (AACE 2005).

The capital costs presented are for inside the battery limits and exclude costs for administrative offices, storage areas, utilities, and other essential and nonessential auxiliary facilities. Fixed capital costs were estimated from literature estimates and scaled estimates (capacity, year, and material) from previous quotes. Capacity adjustments were based on the six-tenths factor rule:

$$C_2 = C_1\left(\frac{q_2}{q_1}\right)^n \tag{5}$$

where C_1 is the cost of the equipment item at capacity q_1, C_2 is the cost of the equipment at capacity q_2, and n is the exponential factor, which typically has a value of 0.6 (Peters 2002). It was assumed that the number of trains did not have an impact on cost scaling. Cost indices were used to adjust equipment prices from previous years to values in July of 2009 using the Chemical Engineering Plant Cost Index (CEPCI). Costs for HTGRs and HTSE were scaled directly based on capacity—the six-tenths factor rule was not used.

Table 3. CEPCI data.

Year	CEPCI	Year	CEPCI
1990	357.6	2000	394.1
1991	361.3	2001	394.3
1992	358.2	2002	395.6
1993	359.2	2003	402
1994	368.1	2004	444.2
1995	381.1	2005	468.2
1996	381.7	2006	499.6
1997	386.5	2007	525.4
1998	389.5	2008	575.4
1999	390.6	July 2009	512

For the nuclear-integrated cases, the estimates of capital costs and operating and maintenance costs assumed the nuclear plant was an “nth of a kind.” In other words, the estimates were based on the costs expected after the HTGR technology is integrated into an industrial application more than 10 times. The economic modeling calculations were based on two capital cost scenarios: a current best estimate of \$2,000/kW_t (INL 2007) and a target of \$1,400/kW_t (Demick 2009) where kW_t is the thermal rating of the plant. In comparison, light water nuclear reactor costs are approximately \$1,333/kW_t (NEI 2008). Based on the two capital cost scenarios for HTGR technology, the nominal capital cost for a 600 MW_t HTGR would be \$1.2 billion; the target capital cost would be \$840 million.

After cost estimates were obtained for each of the process areas, the costs for water systems, piping, instrumentation and control, electrical systems, and buildings and structures were added based on scaling factors for the total installed equipment costs (NETL 2000). These factors were not added to the cost of the HTGR, as the cost basis for the HGTR was assumed to represent a complete and operable system. Table 4 presents the factors utilized in this study:

Table 4. Capital cost adjustment factors.

Year	Factor
Water Systems	7.1%
Piping	7.1%
Instrumentation and Control	2.6%
Electrical Systems	8.0%
Buildings and Structures	9.2%

Finally, an engineering fee of 10% and a project contingency of 18% were assumed to determine the TCI. The capital cost provided for the HTGR represents a complete and operable system; the total value represents all inside battery limits (ISBL) and outside battery limits (OSBL) elements as well as contingency and owner's costs; therefore, engineering fees and contingencies were not applied to this cost.

The Association for the Advancement of Cost Engineering (AACE) International recognizes five classes of estimates. The level of project definition for this study was determined to be an AACE International Class 5 estimate. Though the baseline case is actually more in line with the AACE International Class 4 estimate, which is associated with equipment factoring, parametric modeling, historical relationship factors, and broad unit cost data, the HTGR project definition falls under an AACE International Class 5 estimate, associated with less than 2 percent project definition, and based on preliminary design methodology (AACE 2005). Since the HTGR is a larger portion of the total capital investment, an overall Class 5 estimate was assumed.

Based on the AACE International contingency guidelines as presented in DOE/FETC-99/1100 it would appear that the overall project contingency for the non-nuclear portion of the capital should be in the range of 30 to 50 percent, 30 to 40 percent for Class 4, and 50% for Class 5 (Parsons 1999). However, because the cost estimates were scaled based on estimated, quoted, and actual project costs, the overall non-nuclear project contingency should be more in the range of 15 to 20 percent. Eighteen percent was selected based on similar studies conducted by NETL (2007). Again, contingency was not applied to the HTGR as project contingency was accounted for in the basis for the capital cost estimate.

Table 5 and Figure 8 present the capital cost estimate breakdown for the conventional coal-to-ammonia case, Table 6 and Figure 9 for the conventional gas-to-ammonia case, Table 7 and Figure 10 for the nuclear-integrated gas-to-ammonia case, Table 8 and Figure 11 for the HTSE-to-ammonia (N_2 from combustion) case, and Table 9 and Figure 12 for the HTSE-to-ammonia (N_2 from ASU) case. Varying only the cost of the nuclear facility was an adequate assumption, as the HTGR accounts for a large percentage of the overall cost for the nuclear-integrated cases. In addition, there is a greater level of uncertainty in the nuclear plant price given the nascency of HTGR development.

Table 5. Total capital investment, conventional coal-to-ammonia case.

	Installed Cost	Engineering Fee	Contingency	Total Capital Cost
Coal Preparation	$43,520,480	$4,352,048	$8,617,055	$56,489,583
ASU	$86,335,052	$8,633,505	$17,094,340	$112,062,898
Gasification	$139,987,967	$13,998,797	$27,717,617	$181,704,381
Gas Cleaning	$162,917,578	$16,291,758	$32,257,680	$211,467,016
Ammonia Synthesis	$221,682,591	$22,168,259	$43,893,153	$287,744,004
Nitric Acid Synthesis	$202,970,888	$20,297,089	$40,188,236	$263,456,213
Ammonium Nitrate Synthesis	$129,812,295	$12,981,230	$25,702,834	$168,496,359
Urea Synthesis	$215,184,342	$21,518,434	$42,606,500	$279,309,276
ST, HRSG & CT	$31,249,924	$3,124,992	$6,187,485	$40,562,401
Water Systems	$87,589,939	$8,758,994	$17,342,808	$113,691,741
Piping	$87,589,939	$8,758,994	$17,342,808	$113,691,741
I&C	$32,075,189	$3,207,519	$6,350,887	$41,633,595
Electrical Systems	$98,692,889	$9,869,289	$19,541,192	$128,103,370
Buildings and Structures	$113,496,823	$11,349,682	$22,472,371	$147,318,876
Total Capital Investment				**$2,145,731,455**

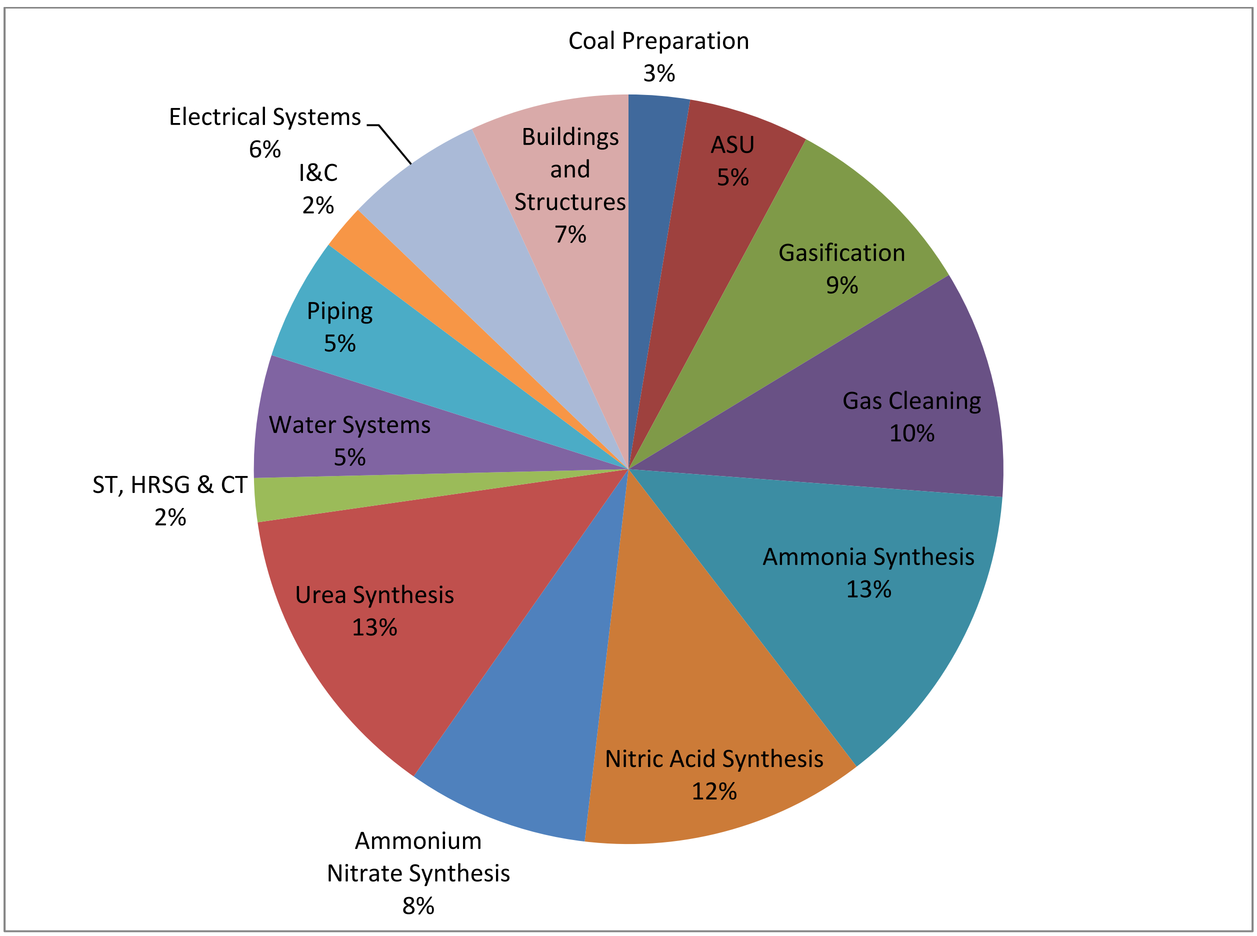

Figure 8. Total capital investment, conventional coal-to-ammonia case.

Table 6. Total capital investment, conventional gas-to-ammonia case.

	Installed Cost	Engineering Fee	Contingency	Total Capital Cost
Natural Gas Reforming	\$88,732,286	\$8,873,229	\$17,568,993	\$115,174,507
Shift Conversion	\$29,319,270	\$2,931,927	\$5,805,215	\$38,056,412
Selexol	\$34,653,118	\$3,465,312	\$6,861,317	\$44,979,748
Methanation	\$7,104,506	\$710,451	\$1,406,692	\$9,221,648
CO_2 Compression	\$9,312,091	\$931,209	\$1,843,794	\$12,087,094
Ammonia Synthesis	\$221,801,399	\$22,180,140	\$43,916,677	\$287,898,216
Nitric Acid Synthesis	\$203,111,753	\$20,311,175	\$40,216,127	\$263,639,056
Ammonium Nitrate Synthesis	\$129,812,295	\$12,981,230	\$25,702,834	\$168,496,359
Urea Synthesis	\$215,184,342	\$21,518,434	\$42,606,500	\$279,309,276
ST, HRSG and CT	\$15,579,691	\$1,557,969	\$3,084,779	\$20,222,440
Water Systems	\$67,777,363	\$6,777,736	\$13,419,918	\$87,975,018
Piping	\$67,777,363	\$6,777,736	\$13,419,918	\$87,975,018
I&C	\$24,819,880	\$2,481,988	\$4,914,336	\$32,216,204
Electrical Systems	\$76,368,860	\$7,636,886	\$15,121,034	\$99,126,780
Buildings and Structures	\$87,824,189	\$8,782,419	\$17,389,189	\$113,995,798
Total Capital Investment				**\$1,660,373,573**

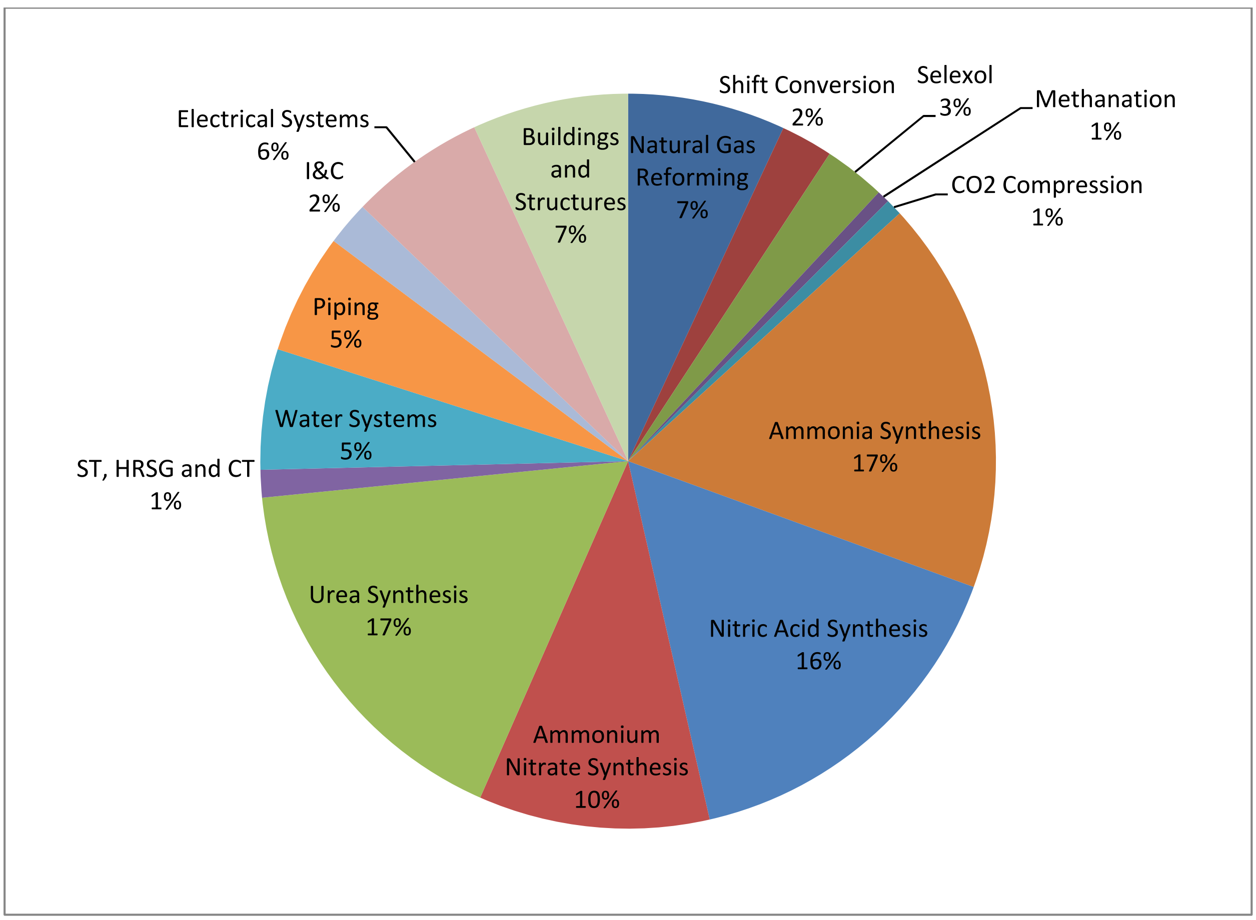

Figure 9. Total capital investment, conventional gas-to-ammonia case.

Table 7. Total capital investment, nuclear-integrated gas-to-ammonia case.

	Installed Cost	Engineering Fee	Contingency	Total Capital Cost
Nuclear (Includes all costs)				768,613,333
Natural Gas Reforming	$78,666,814	$7,866,681	$15,576,029	$102,109,525
Shift Conversion	$29,941,964	$2,994,196	$5,928,509	$38,864,669
Selexol	$34,950,194	$3,495,019	$6,920,138	$45,365,351
Methanation	$7,103,237	$710,324	$1,406,441	$9,220,002
CO_2 Compression	$9,348,529	$934,853	$1,851,009	$12,134,390
Ammonia Synthesis	$221,761,801	$22,176,180	$43,908,837	$287,846,818
Nitric Acid Synthesis	$203,111,753	$20,311,175	$40,216,127	$263,639,056
Ammonium Nitrate Synthesis	$129,812,295	$12,981,230	$25,702,834	$168,496,359
Urea Synthesis	$215,184,342	$21,518,434	$42,606,500	$279,309,276
ST, HRSG and CT	$16,614,823	$1,661,482	$3,289,735	$21,566,040
Nuclear Power Cycle	$90,964,386	$9,096,439	$18,010,948	$118,071,773
Water Systems	$67,201,198	$6,720,120	$13,305,837	$87,227,156
Piping	$67,201,198	$6,720,120	$13,305,837	$87,227,156
I&C	$24,608,890	$2,460,889	$4,872,560	$31,942,339
Electrical Systems	$75,719,660	$7,571,966	$14,992,493	$98,284,119
Buildings and Structures	$87,077,609	$8,707,761	$17,241,367	$113,026,737
Total Capital Investment				**$2,532,944,099**
Total Capital Investment (+50% HTGR)				**$2,917,250,765**
Total Capital Investment (-30% HTGR)				**$2,302,360,099**

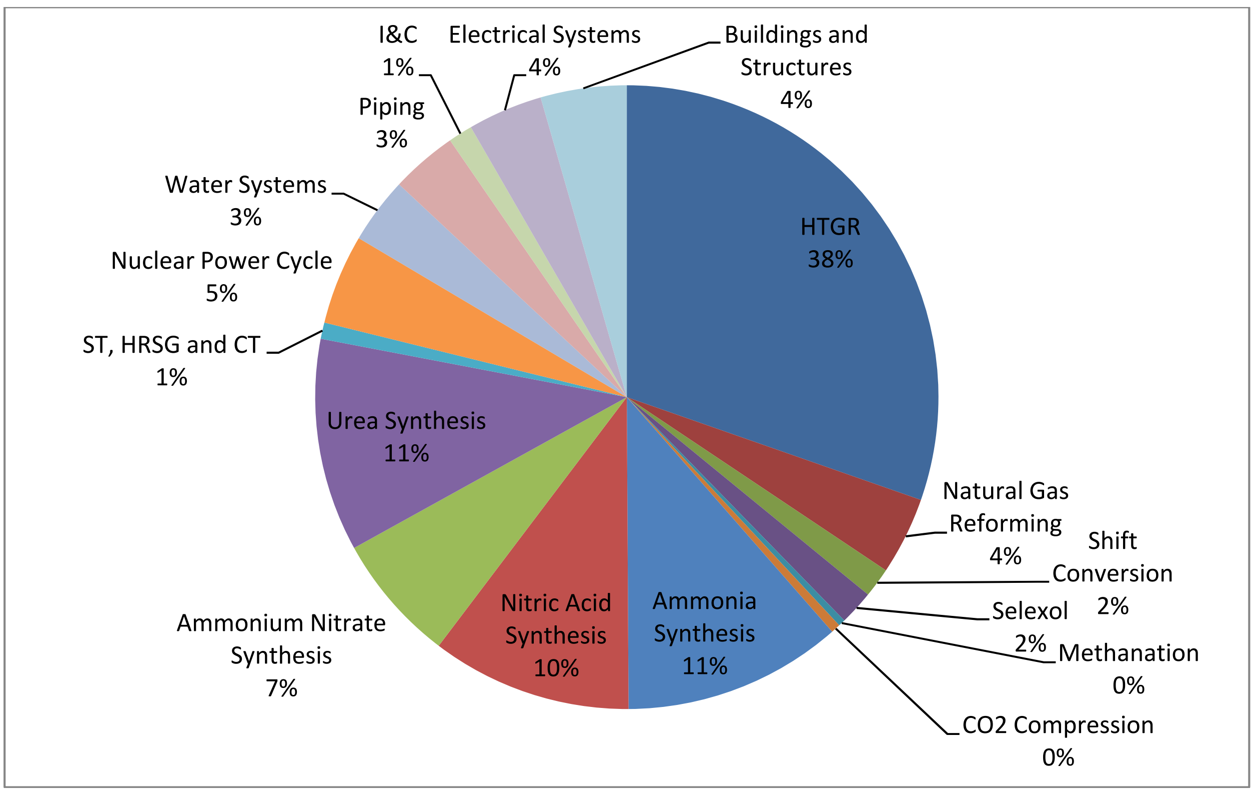

Figure 10. Total capital investment, nuclear-integrated gas-to-ammonia case.

Table 8. Total capital investment, HTSE-to-ammonia (N_2 from combustion) case.

	Installed Cost	Engineering Fee	Contingency	Total Capital Cost
Nuclear (Includes all costs)				4,201,101,416
HTSE	$271,216,232	$27,121,623	$53,700,814	$352,038,669
N_2 Generation (Combustion)	$12,900,791	$1,290,079	$2,554,357	$16,745,226
Methanation	$7,103,237	$710,324	$1,406,441	$9,220,002
CO_2 Gen. & Compression	$11,210,720	$1,121,072	$2,219,722	$14,551,514
Ammonia Synthesis	$221,761,801	$22,176,180	$43,908,837	$287,846,818
Nitric Acid Synthesis	$203,111,753	$20,311,175	$40,216,127	$263,639,056
Ammonium Nitrate Synthesis	$129,812,295	$12,981,230	$25,702,834	$168,496,359
Urea Synthesis	$215,184,342	$21,518,434	$42,606,500	$279,309,276
ST, HRSG and CT	$40,856,623	$4,085,662	$8,089,611	$53,031,897
Nuclear Power Cycle	$615,345,050	$61,534,505	$121,838,320	$798,717,875
Water Systems	$79,034,203	$7,903,420	$15,648,772	$102,586,396
Piping	$79,034,203	$7,903,420	$15,648,772	$102,586,396
I&C	$28,942,103	$2,894,210	$5,730,536	$37,566,849
Electrical Systems	$89,052,624	$8,905,262	$17,632,419	$115,590,305
Buildings and Structures	$102,410,517	$10,241,052	$20,277,282	$132,928,851
Total Capital Investment				**$6,935,956,906**
Total Capital Investment (+50% HTGR)				**$9,036,507,614**
Total Capital Investment (-30% HTGR)				**$5,675,626,481**

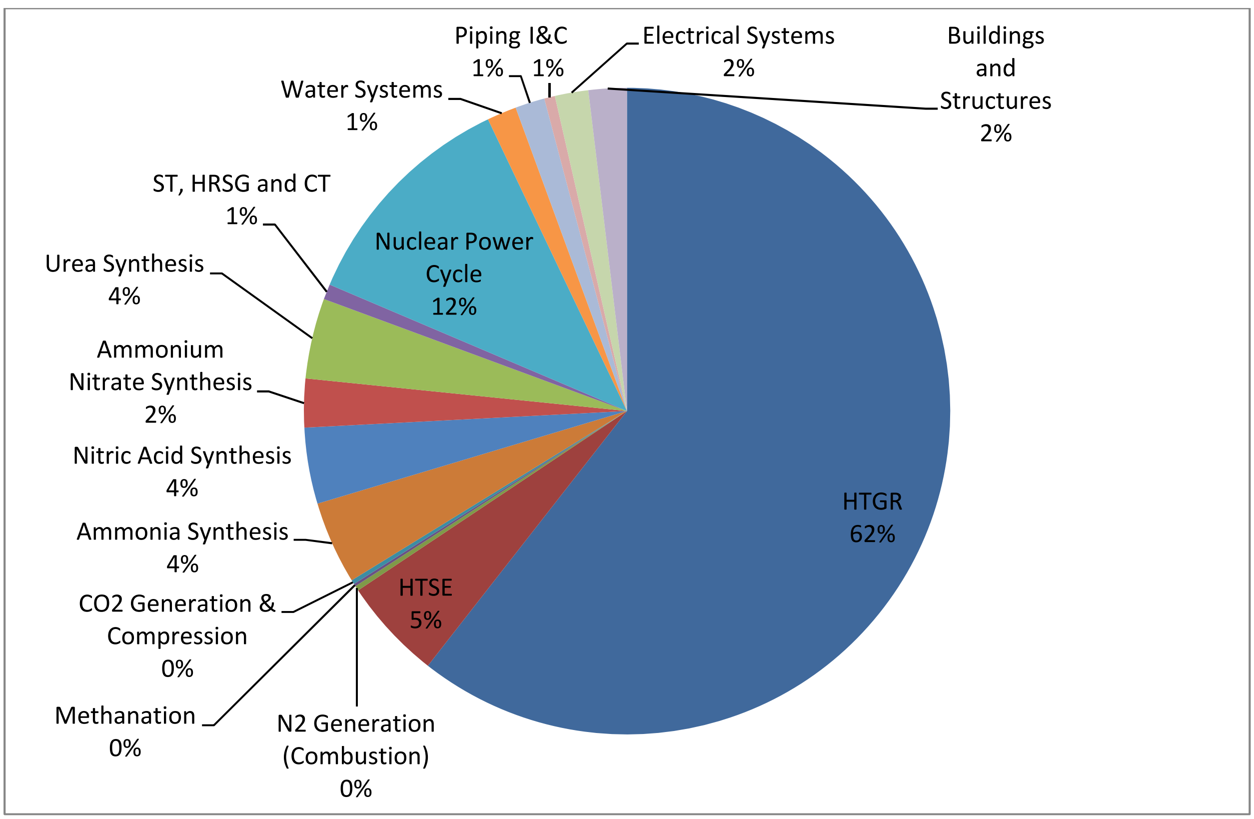

Figure 11. Total capital investment, HTSE-to-ammonia (N_2 from combustion) case.

Table 9. Total capital investment, HTSE-to-ammonia (N_2 from ASU) case.

	Installed Cost	Engineering Fee	Contingency	Total Capital Cost
Nuclear (Includes all costs)				3,666,008,905
HTSE	$229,865,245	$22,986,525	$45,513,319	$298,365,088
N_2 Generation (Cryo. ASU)	$36,589,053	$3,658,905	$7,244,633	$47,492,591
Methanation	$7,103,237	$710,324	$1,406,441	$9,220,002
CO_2 Gen. & Compression	$11,210,617	$1,121,062	$2,219,702	$14,551,380
Ammonia Synthesis	$221,761,801	$22,176,180	$43,908,837	$287,846,818
Nitric Acid Synthesis	$203,111,753	$20,311,175	$40,216,127	$263,639,056
Ammonium Nitrate Synthesis	$129,812,295	$12,981,230	$25,702,834	$168,496,359
Urea Synthesis	$215,184,342	$21,518,434	$42,606,500	$279,309,276
ST and CT	$36,974,182	$3,697,418	$7,320,888	$47,992,488
Nuclear Power Cycle	$568,211,288	$56,821,129	$112,505,835	$737,538,251
Water Systems	$77,504,489	$7,750,449	$15,345,889	$100,600,827
Piping	$77,504,489	$7,750,449	$15,345,889	$100,600,827
I&C	$28,381,926	$2,838,193	$5,619,621	$36,839,740
Electrical Systems	$87,329,002	$8,732,900	$17,291,142	$113,353,045
Buildings and Structures	$100,428,352	$10,042,835	$19,884,814	$130,356,001
Total Capital Investment				**$6,302,210,656**
Total Capital Investment (+50% HTGR)				**$8,135,215,109**
Total Capital Investment (-30% HTGR)				**$5,202,407,985**

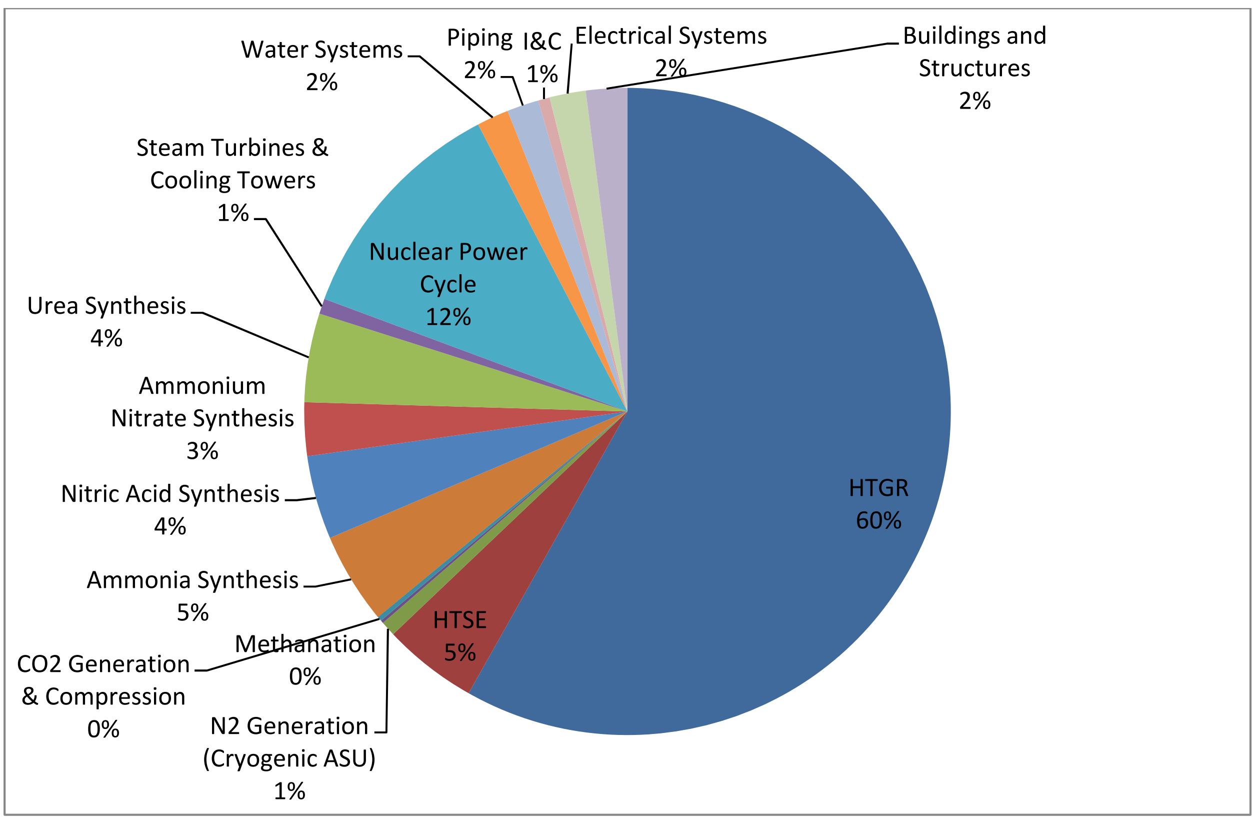

Figure 12. Total capital investment, HTSE-to-ammonia (N_2 from ASU) case.

4.2 Estimation of Revenue

Yearly revenues were estimated for all cases based on recent price data for the various products generated. Revenues were estimated for low, average, and high prices for urea and ammonium nitrate. High prices correspond to prices from April 2008, low prices are from April 2002, and average prices were the average of the high and low values. Product prices were gathered from the U. S. Department of Agriculture (USDA) and represent actual purchase prices for farm use (USDA 2010). For the coal-based case, selling prices for slag and sulfur were not varied in the study. A stream factor of 92% is assumed for both the fossil and nuclear plants. Table 10 presents the revenues for the conventional coal-to-ammonia case. Table 11 presents the revenues for the conventional gas-to-ammonia case, the nuclear-integrated gas-to-ammonia case, and both HTSE-to-ammonia cases (i.e., revenues are the same for all of these cases).

Table 10. Annual revenues, conventional coal-to-ammonia case.

	Price		Generated		Annual Revenue
Slag	25.63	$/ton	328	ton/day	$2,821,332
Sulfur	38.13	$/ton	153	ton/day	$1,960,719
Urea – Low	227.04	$/ton	2,939	ton/day	$224,068,446
Urea – Avg.	396.42	$/ton	2,939	ton/day	$391,232,816
Urea – High	565.80	$/ton	2,939	ton/day	$558,397,186
AN – Low	231.79	$/ton	3,779	ton/day	$294,143,498
AN – Avg.	376.76	$/ton	3,779	ton/day	$478,103,183
AN – High	521.73	$/ton	3,779	ton/day	$662,062,869
Annual Revenue, low					**$522,993,995**
Annual Revenue, average					**$874,118,051**
Annual Revenue, high					**$1,225,242,106**

Table 11. Annual revenues, conventional and nuclear-integrated gas-to-ammonia cases.

	Price		Generated		Annual Revenue
Urea – Low	227.04	$/ton	2,939	ton/day	$224,068,446
Urea – Avg.	396.42	$/ton	2,939	ton/day	$391,232,816
Urea – High	565.80	$/ton	2,939	ton/day	$558,397,186
AN – Low	231.79	$/ton	3,779	ton/day	$294,143,498
AN – Avg.	376.76	$/ton	3,779	ton/day	$478,103,183
AN – High	521.73	$/ton	3,779	ton/day	$662,062,869
Annual Revenue, low					**$518,211,944**
Annual Revenue, average					**$869,335,999**
Annual Revenue, high					**$1,220,460,055**

Table 12. Annual revenues, HTSE-to-ammonia (N_2 from combustion) case.

	Price		Generated		Annual Revenue
Oxygen	41.60	$/ton	2,506	ton/day	$35,006,192
Urea – Low	227.04	$/ton	2,939	ton/day	$224,068,446
Urea – Avg.	396.42	$/ton	2,939	ton/day	$391,232,816
Urea – High	565.80	$/ton	2,939	ton/day	$558,397,186
AN – Low	231.79	$/ton	3,779	ton/day	$294,143,498
AN – Avg.	376.76	$/ton	3,779	ton/day	$478,103,183
AN – High	521.73	$/ton	3,779	ton/day	$662,062,869
Annual Revenue, low					**$553,218,136**
Annual Revenue, average					**$904,342,191**
Annual Revenue, high					**$1,255,466,246**

Table 13. Annual revenues, HTSE-to-ammonia (N_2 from ASU) case.

	Price		Generated		Annual Revenue
Oxygen	41.60	$/ton	2,435	ton/day	$34,018,469
Urea – Low	227.04	$/ton	2,939	ton/day	$224,068,446
Urea – Avg.	396.42	$/ton	2,939	ton/day	$391,232,816
Urea – High	565.80	$/ton	2,939	ton/day	$558,397,186
AN – Low	231.79	$/ton	3,779	ton/day	$294,143,498
AN – Avg.	376.76	$/ton	3,779	ton/day	$478,103,183
AN – High	521.73	$/ton	3,779	ton/day	$662,062,869
Annual Revenue, low					**$552,230,413**
Annual Revenue, average					**$903,354,468**
Annual Revenue, high					**$1,254,478,523**

4.3 Estimation of Manufacturing Costs

Manufacturing cost is the sum of direct and indirect manufacturing costs. Direct manufacturing costs for this project include the cost of raw materials, utilities, and operating labor and maintenance. Indirect manufacturing costs include estimates for the cost of overhead and insurance and taxes (Perry 2008).

Labor costs are assumed to be 1.15% of the total capital investment for the fossil portion of each case (for a discussion of how this value was determined, see TEV-672). Maintenance costs were assumed to be 3% of the total capital investment per the *Handbook of Petroleum Processing*. The power cycles and HTSE were not included in the TCI for operation and maintenance costs, as they were calculated separately. Taxes and insurance were assumed to be 1.5% of the TCI, excluding the HTGR, an overhead of 65% of the labor and maintenance costs was assumed, and royalties were assumed to be 1% of the coal or natural gas cost based on information presented in the *Handbook of Petroleum Processing* (Jones 2006). Table 14 provides the manufacturing costs for the conventional coal-to-ammonia case. Availability of both the fossil and nuclear plants was assumed to be 92%.

For the gas-to-ammonia and nuclear-integrated gas-to-ammonia cases, manufacturing costs were calculated for two scenarios: with sequestration and without sequestration. For these cases, only the sequestration results are presented below.

For the gas-to-ammonia and HTSE-to-ammonia cases, natural gas prices were varied to account for the large fluctuations seen in the market. Costs were calculated for low ($4.50/MSCFD), average ($6.50/MSCFD), and high ($12.00/MSCFD) industrial natural gas prices. High prices correspond to prices from June 2008, low prices are from September 2009, and the average price was chosen to reflect current natural gas price (EIA 2010). Only average natural gas prices are presented in the gas-to-ammonia tables below.

Operating and maintenance costs for the nuclear plant were based on data from General Atomics for the GT-MHR HTGR published in 2002; these costs were inflated to 2009 dollars (GA 2002). HTSE cell-replacement costs were calculated assuming cell replacement every eight years based on vendor input; see TEV-693, "Nuclear-Integrated Hydrogen Production Analysis," for detailed information regarding calculation of cell replacements costs (McKellar 2010).

Table 14. Annual manufacturing costs, conventional coal-to-ammonia case.

	Price		Consumed		Annual Cost
Direct Costs					
Materials					
Coal	36.16	\$/ton	4,591	ton/day	\$55,747,791
Fly Ash Disposal	33.20	\$/ton	138	ton/day	\$1,533,551
Selexol Solvent	2.57	\$/gal	8.13	gal/day	\$7,031
Makeup H_2O Treatment	0.02	\$/k-gal	6,431	k-gal/day	\$52,819
Wastewater Treatment	1.32	\$/k-gal	2,379	k-gal/day	\$1,050,381
Claus Catalyst	21.00	\$/ft^3	0.545	ft^3/day	\$3,841
SCOT Catalyst	275.00	\$/ft^3	0.083	ft^3/day	\$7,663
Carbon, Hg Guard Bed	5.57	\$/lb	6.000	lb/day	\$11,193
Zinc Oxide	300.00	\$/ft^3	0.035	ft^3/day	\$3,515
WGS Catalyst	825.00	\$/ft^3	0.674	ft^3/day	\$186,648
Methanation Catalyst	700.00	\$/ft^3	0.025	ft^3/day	\$5,865
Ammonia Synthesis Catalyst	775.00	\$/ft^3	0.029	ft^3/day	\$7,547
Platinum (HNO_3 Catalyst)	23,118	\$/lb	0.622	lb/day	\$4,831,085
SCR Catalyst	424.93	\$/ft^3	0.135	ft^3/day	\$19,298
Utilities					
Electricity	1.67	\$/kW-d	141,800	kW	\$79,538,501
Water	0.05	\$/k-gal	6,431	k-gal/day	\$99,337
Royalties					\$557,478
Labor and Maintenance					\$89,047,855
Indirect Costs					
Overhead					\$57,881,106
Insurance and Taxes					\$32,185,972
Manufacturing Costs					**\$322,778,478**

Table 15. Annual manufacturing costs, conventional gas-to-ammonia case.

	Price		Consumed		Annual Cost
Direct Costs					
Materials					
Average Natural Gas	6.50	\$/MSCF	88	MSCFD	\$192,077,600
Makeup H_2O Treatment	0.02	\$/k-gal	5,656	k-gal/day	\$46,452
Wastewater Treatment	1.32	\$/k-gal	2,474	k-gal/day	\$1,092,218
HDS Catalyst	700.00	\$/ft^3	0.04	ft^3/day	\$8,613
Zinc Oxide	300.00	\$/ft^3	0.344	ft^3/day	\$34,681
Preforming Catalyst	2,350	\$/ft^3	0.00	ft^3/day	\$0
Primary SMR Catalyst	750.00	\$/ft^3	0.16	ft^3/day	\$39,585
Secondary SMR Catalyst	650.00	\$/ft^3	0.04	ft^3/day	\$9,372
HTS Catalyst	380.00	\$/ft^3	0.19	ft^3/day	\$23,927
LTS Catalyst	600.00	\$/ft^3	0.15	ft^3/day	\$30,611
Selexol Solvent	2.57	\$/gal	8.14	gal/day	\$7,037
Methanation Catalyst	700.00	\$/ft^3	0.042	ft^3/day	\$9,906
Ammonia Synthesis Catalyst	775.00	\$/ft^3	0.029	ft^3/day	\$7,498
Platinum (HNO_3 Catalyst)	23,118	\$/lb	0.62	lb/day	\$4,836,674
SCR Catalyst	424.93	\$/ft^3	0.135	ft^3/day	\$19,321
CO_2 Sequestration	14.54	\$/ton	1,762	ton/day	\$8,601,852
Utilities					
Electricity	1.67	\$/kW-day	108,800	kW	\$61,028,131
Water	0.05	\$/k-gal	5,656	k-gal/day	\$87,363
Royalties					\$1,920,776
Labor and Maintenance					\$68,905,503
Indirect Costs					
Overhead					\$44,788,577
Insurance and Taxes					\$24,905,604
Manufacturing Costs, Average Natural Gas					**\$408,481,302**

Table 16. Annual manufacturing costs, nuclear-integrated gas-to-ammonia case.

	Price		Consumed		Annual Cost
Direct Costs					
Materials					
Average Natural Gas	6.50	\$/MSCF	72	MSCFD	\$157,154,400
Makeup H_2O Treatment	0.02	\$/k-gal	5,737	k-gal/day	\$47,122
Wastewater Treatment	1.32	\$/k-gal	2,560	k-gal/day	\$1,130,241
HDS Catalyst	700.00	\$/ft^3	0.04	ft^3/day	\$9,077
Zinc Oxide	300.00	\$/ft^3	0.363	ft^3/day	\$36,592
Preforming Catalyst	2,350	\$/ft^3	0.00	ft^3/day	\$0
Primary SMR Catalyst	750.00	\$/ft^3	0.40	ft^3/day	\$101,298
Secondary SMR Catalyst	650.00	\$/ft^3	0.04	ft^3/day	\$8,638
HTS Catalyst	380.00	\$/ft^3	0.20	ft^3/day	\$24,891
LTS Catalyst	600.00	\$/ft^3	0.16	ft^3/day	\$31,810
Selexol Solvent	2.57	\$/gal	8.14	gal/day	\$7,035
Methanation Catalyst	700.00	\$/ft^3	0.043	ft^3/day	\$10,122
Ammonia Synthesis Catalyst	775.00	\$/ft^3	0.030	ft^3/day	\$7,770
Platinum (HNO_3 Catalyst)	23,118	\$/lb	0.62	lb/day	\$4,836,674
SCR Catalyst	424.93	\$/ft^3	0.135	ft^3/day	\$19,319
Nuclear Fuel	8.80	\$/MW-h	180	MW_e	\$12,758,100
CO_2 Sequestration	14.54	\$/ton	1,786	ton/day	\$8,719,017
Utilities					
Electricity	1.67	\$/kW-day	0	kW	\$0
Water	0.05	\$/k-gal	5,737	k-gal/day	\$88,622
Royalties					\$1,571,544
O&M, Nuclear	3.57	\$/MW-h	180	MW_e	\$5,172,203
Labor and Maintenance, Fossil					\$68,319,748
Indirect Costs					
Overhead					\$44,407,836
Insurance and Taxes					\$26,464,961
Manufacturing Costs, Average Natural Gas					**\$330,927,021**

Table 17. Annual manufacturing costs, HTGR-to-ammonia (N_2 from combustion).

	Price		Consumed		Annual Cost
Direct Costs					
Materials					
Average Natural Gas	6.50	\$/MSCF	36	MSCFD	\$78,577,200
Makeup H_2O Treatment	0.02	\$/k-gal	9,121	k-gal/day	\$74,913
Wastewater Treatment	1.32	\$/k-gal	3,038	k-gal/day	\$1,341,588
HDS Catalyst	700.00	\$/ft^3	0.02	ft^3/day	\$5,693
Zinc Oxide	300.00	\$/ft^3	0.181	ft^3/day	\$18,254
Methanation Catalyst	700.00	\$/ft^3	0.031	ft^3/day	\$7,331
Ammonia Synthesis Catalyst	775.00	\$/ft^3	0.030	ft^3/day	\$7,697
Platinum (HNO_3 Catalyst)	23,118	\$/lb	0.62	lb/day	\$4,836,674
SCR Catalyst	424.93	\$/ft^3	0.135	ft^3/day	\$19,318
HTSE Cell Replacements	0.02	\$/lb H_2	1,430	k-lb H_2/day	\$11,646,491
Nuclear Fuel	8.80	\$/MW-h	984	MW_e	\$69,733,466
CO_2 Sequestration	14.54	\$/ton	0	ton/day	\$0
Utilities					
Electricity	1.67	\$/kW-day	0	kW	\$0
Water	0.05	\$/k-gal	9,121	k-gal/day	\$140,890
Royalties					\$785,772
O&M, Nuclear	3.57	\$/MW-h	984	MW_e	\$28,270,324
Labor and Maintenance, Fossil					\$74,118,627
Indirect Costs					
Overhead					\$48,177,107
Insurance and Taxes					\$41,022,832
Manufacturing Costs, Average Natural Gas					**\$358,784,178**

Table 18. Annual manufacturing costs, HTGR-to-ammonia (N_2 from ASU).

	Price		Consumed		Annual Cost
Direct Costs					
Materials					
Average Natural Gas	6.50	\$/MSCF	36	MSCFD	\$78,577,200
Makeup H_2O Treatment	0.02	\$/k-gal	7,999	k-gal/day	\$65,696
Wastewater Treatment	1.32	\$/k-gal	2,648	k-gal/day	\$1,169,280
HDS Catalyst	700.00	\$/ft^3	0.02	ft^3/day	\$5,693
Zinc Oxide	300.00	\$/ft^3	0.181	ft^3/day	\$18,254
Methanation Catalyst	700.00	\$/ft^3	0.031	ft^3/day	\$7,181
Ammonia Synthesis Catalyst	775.00	\$/ft^3	0.029	ft^3/day	\$7,514
Platinum (HNO_3 Catalyst)	23,118	\$/lb	0.62	lb/day	\$4,836,674
SCR Catalyst	424.93	\$/ft^3	0.135	ft^3/day	\$19,318
HTSE Cell Replacements	0.02	\$/lb H_2	1,212	k-lb H_2/day	\$9,870,809
Nuclear Fuel	8.80	\$/MW-h	858	MW$_e$	\$60,851,544
CO_2 Sequestration	14.54	\$/ton	0	ton/day	\$0
Utilities					
Electricity	1.67	\$/kW-day	0	kW	\$0
Water	0.05	\$/k-gal	7,999	k-gal/day	\$123,555
Royalties					\$785,772
O&M, Nuclear	3.57	\$/MW-h	858	MW$_e$	\$24,669,545
Labor and Maintenance, Fossil					\$73,513,473
Indirect Costs					
Overhead					\$47,783,758
Insurance and Taxes					\$39,543,026
Manufacturing Costs, Average Natural Gas					**\$341,848,293**

4.4 Economic Comparison

To assess the economic desirability of the coal-to-ammonia and gas-to-ammonia cases, several economic indicators were calculated for each case. For all cases the internal rate of return (IRR) for low, average, and high product selling price was calculated, as well as low, average, and high natural gas prices for the gas-to-ammonia cases. In addition, the product price necessary for a return of 12% was calculated for all cases for the baseline coal cost as well as low, average, and high natural gas prices. The following assumptions were made for the economic analyses:

- The plant startup year is 2014.
- A construction period of three years for the fossil plant and five years for the nuclear plant. In order to easily perform the analysis in 2009 dollars, the following simplifying assumptions were made:
  - Fossil plant construction begins in 2011

- Nuclear plant construction begins in 2009

- It is assumed that all reactors come online at the same time. A study was conducted for CTL to determine the impact of six-month and three-month reactor staging versus all reactors coming online at one time. It was determined that the simplification of assuming all reactors online at once does not impact the economic results significantly enough to warrant the complexity of creating multiple staging trains for each scenario. These results were assumed to be equally valid for all nuclear-integrated ammonia cases.
  - Percent capital invested for the fossil plant is 33% per year
  - Percent capital invested for the HTGR is 20% per year
- Plant startup time is one year.
  - Operating costs are 85% of the total value during startup
  - Revenues are 60% of the total value during startup
- The analysis period for the economic evaluation assumes an economic life of 30 years, excluding construction time (the model is built to accommodate up to 40 years).
- An availability of 92% was assumed for both the fossil and nuclear plants; the plants are assumed to operate 365 days a year, 24 hours per day.
- An inflation rate of 2.5% is assumed.
- Debt to equity ratios of 80%/20% and 55%/45% are calculated; however, results are only presented for 80%/20%.
  - The interest rate on debt is assumed to be 8%
  - The repayment term on the loan is assumed to be 15 years
- The effective income tax rate is 38.9%.
  - State tax is 6%
  - Federal tax is 35%
- MACRS depreciation is assumed.
- A CO_2 tax of $0/ton to $100/ton for coal-to-ammonia and $0/ton to $200/ton for the gas-to-ammonia cases is investigated.

4.4.1 Cash Flow

To assess the IRR and PW of each scenario, it is necessary to calculate the after tax cash flow (ATCF). To calculate the ATCF it is necessary to first calculate the revenues (R_k); cash outflows (E_k); sum of all noncash, or book, costs such as depreciation (d_k); net income before taxes (NIBT); the effective income tax rate (t); and the income taxes (T_k), for each year (k). The taxable income is revenue minus the sum of all cash outflow and noncash costs. Therefore the income taxes per year are defined as follows (Sullivan 2003):

$$T_k = t(R_k - E_k - d_k) \tag{6}$$

Depreciation for the economic calculations was calculated using a standard MACRS depreciation method with a property class of 15 years. Depreciation was assumed for the total capital investment over the 5-year construction schedule, including inflation. Table 19 presents the recovery rates for a 15-year property class (Perry 2008):

Table 19. MACRS depreciation.

Year	Recovery Rate	Year	Recovery Rate
1	0.05	9	0.0591
2	0.095	10	0.059
3	0.0855	11	0.0591
4	0.077	12	0.059
5	0.0693	13	0.0591
6	0.0623	14	0.059
7	0.059	15	0.0591
8	0.059	16	0.0295

The ATCF is then the sum of the before tax cash flow (BTCF) minus the income taxes owed. Note that the expenditures for capital are not taxed but are included in the BTCF each year there is a capital expenditure (C_k); this includes the equity capital and the debt principle. The BTCF is defined as follows (Sullivan 2003):

$$BTCF_k = R_k - E_k - C_k \tag{7}$$

The ATCF can then be defined as:

$$ATCF_k = BTCF_k - T_k \tag{8}$$

It should be noted that when a CO_2 tax credit is included in the economic analysis, the tax would be treated essentially as a manufacturing cost, decreasing the yearly revenue.

4.4.2 Internal Rate of Return

The IRR method is the most widely used rate of return method for performing engineering economic analyses. This method solves for the interest rate that equates the equivalent worth of an alternative's cash inflows to the equivalent worth of cash outflows (after tax cash flow), i.e., the interest rate at which the PW is zero. The resulting interest is the IRR (i'). For the project to be economically viable, the calculated IRR must be greater than the desired minimum annual rate of return (MARR) (Sullivan 2003).

$$PW(i'\%) = \sum_{k=0}^{N} ATCF_k \left(1 + i'\right)^{-k} = 0 \tag{9}$$

IRR calculations were performed for a 80%/20% debt to equity ratio for the calculated fossil TCI and at +50% TCI and -30% TCI for the HTGR for all cases at low, average, and high product prices for coal-to-ammonia and gas-to-ammonia as well as low, average, and high natural gas purchase prices for gas-to-ammonia. In addition, the price of urea and ammonium nitrate necessary for an IRR of 12% and a PW of zero was calculated for each case at each debt to equity ratio. The IRR and product price required (for an IRR of 12%) was solved for using the Goal Seek function in Excel (Excel 2007).

Finally, a CO_2 tax was included into the calculations to determine the price of urea and ammonium nitrate necessary in all cases for a 12% IRR and a CO_2 tax of $0/ton to $100/ton for coal-to-ammonia and $0/ton to $200/ton for gas-to-ammonia. These cases were calculated for 80%/20% and 55%/45% debt to equity ratios for the TCI and +50% TCI and -30% TCI of the HTGR. Additionally, the gas-to-ammonia cases were calculated for either sequestering or not sequestering the CO_2. The tax calculated was added to the existing tax liability for each year.

5. ECONOMIC MODELING RESULTS

5.1 Coal-to-Ammonia Economic Results

Table 20 presents the results for an 80%/20% debt to equity ratio for the conventional coal-to-ammonia case, listing the IRR for low, average, and high urea selling prices, and the urea selling price required for a 12% IRR. Figure 13 depicts the associated IRR results for the conventional coal-to-ammonia case. Figure 14 shows the impact of a carbon tax on the selling price of urea and ammonium nitrate. Because a suitable nuclear-integrated coal-to-ammonia case was not identified, these results are presented for information only. Results indicate that a return of 12% can be achieved for a urea production price of $259.07/ton. Note that carbon sequestration is not considered in this analysis. Without sequestration, the urea production price increases sharply as a function of the carbon tax imposed; a urea selling price of $407.64/ton is required to achieve a 12% return when a $100/ton carbon tax is imposed.

Table 20. Conventional coal-to ammonia IRR results.

	TCI	
	IRR	$/ton urea
	$2,087,531,131	
Conventional Coal-to-Ammonia	9.04	$227.04
	27.61	$396.42
	42.55	$565.80
	12.00	$259.07

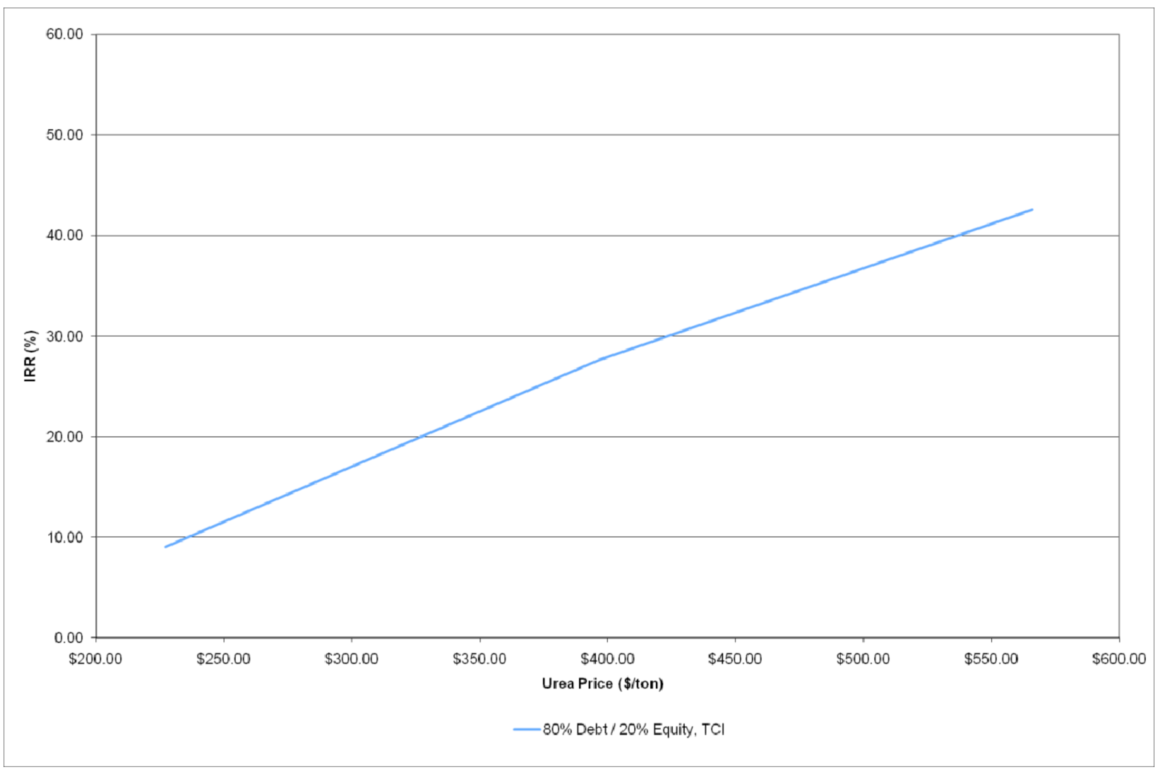

Figure 13. Conventional coal-to-ammonia IRR economic results.

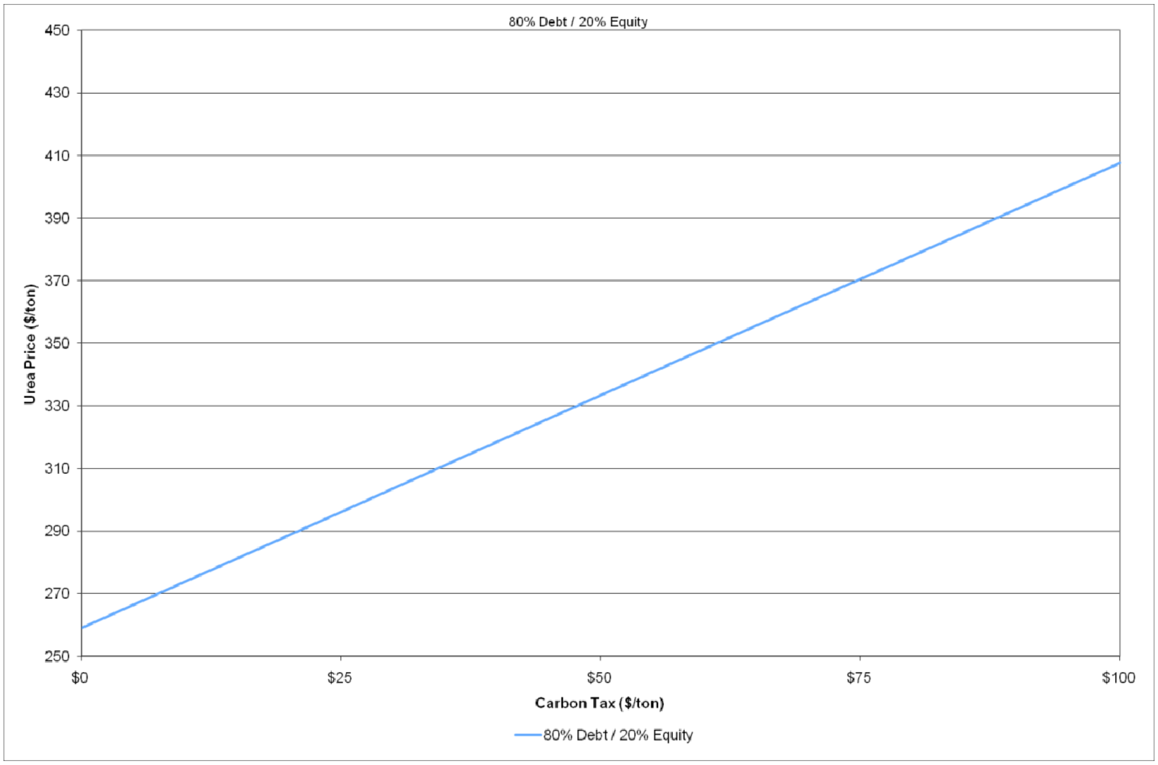

Figure 14. Conventional coal-to-ammonia carbon tax results (12% IRR).

5.2 Gas-to-Ammonia and HTSE-to-Ammonia Economic Results

Table 21 presents the results for an 80%/20% debt to equity ratio for the conventional gas-to-ammonia, nuclear-integrated gas-to-ammonia, and HTSE-to-ammonia cases, listing the IRR for low, average, and high urea selling prices, and the urea selling price required for a 12% IRR. These results do not assume carbon sequestration. Figure 15 depicts the associated IRR results for these cases at the average natural gas price of $6.50/MSCF. These results indicate that a favorable 12% IRR is achievable for the HTGR-integrated case when urea price rises above $286/ton; hence, such an option appears to be economically viable given current fertilizer prices. The HTSE-to-ammonia cases do not look as favorable, as urea prices would need to be around $500/ton to achieve a 12% return.

Table 21. Gas-to-ammonia and HTSE-to-ammonia IRR results (no sequestration).

	TCI -30% HTGR		**TCI**		**TCI +50% HTGR**	
	IRR	$/ton urea	IRR	$/ton urea	IRR	$/ton urea
Conventional Gas-to-Ammonia Low NG: $4.50/MSCF			*$1,660,373,573*			
			10.69	$227.04		
			33.29	$396.42		
			50.69	$565.80		
			12.00	$244.36		
HTGR-Integrated Gas-to-Ammonia Low NG: $4.50/MSCF	*$2,302,360,099*		*$2,532,944,099*		*$2,917,250,765*	
	10.47	$227.04	9.24	$227.04	7.59	$227.04
	26.19	$396.42	23.68	$396.42	20.44	$396.42
	38.46	$565.80	35.00	$565.80	30.56	$565.80
	12.00	$249.88	12.00	$263.32	12.00	$285.71
HTSE-to-Ammonia N_2 from Comb. Low NG: $4.50/MSCF	*$5,675,626,481*		*$6,935,956,906*		*$9,036,507,614*	
	0.72	$227.04	-0.69	$227.04	-$2.41	$227.04
	9.57	$396.42	7.30	$396.42	4.74	$396.42
	16.14	$565.80	13.02	$565.80	9.59	$565.80
	12.00	$452.42	12.00	$525.86	12.00	$648.27
HTSE-to-Ammonia N_2 from ASU Low NG: $4.50/MSCF	*$5,202,407,985*		*$6,302,210,656*		*$8,135,215,109*	
	1.92	$227.04	0.48	$227.04	-$1.27	$227.04
	10.99	$396.42	8.64	$396.42	5.97	$396.42
	17.93	$565.80	14.67	$565.80	11.07	$565.80
	12.00	$418.05	12.00	$482.14	12.00	$588.96
Conventional Gas-to-Ammonia Average NG: $6.50/MSCF			*$1,660,373,573*			
			5.84	$227.04		
			29.67	$396.42		
			47.77	$565.80		
			12.00	$272.23		
HTGR-Integrated Gas-to-Ammonia Average NG: $6.50/MSCF	*$2,302,360,099*		*$2,532,944,099*		*$2,917,250,765*	
	7.78	$227.04	6.73	$227.04	5.31	$227.04
	24.15	$396.42	21.81	$396.42	18.80	$396.42
	36.81	$565.80	33.48	$565.80	29.20	$565.80
	12.00	$272.68	12.00	$286.12	12.00	$308.51
HTSE-to-Ammonia N_2 from Comb. Average NG: $6.50/MSCF	*$5,675,626,481*		*$6,935,956,906*		*$9,036,507,614*	
	-0.15	$227.04	-1.50	$227.04	-$3.15	$227.04
	9.05	$396.42	6.85	$396.42	4.34	$396.42
	15.70	$565.80	12.64	$565.80	9.28	$565.80
	12.00	$463.82	12.00	$537.26	12.00	$659.67
HTSE-to-Ammonia N_2 from ASU Average NG: $6.50/MSCF	*$5,202,407,985*		*$6,302,210,656*		*$8,135,215,109*	
	1.06	$227.04	-0.31	$227.04	-$2.00	$227.04
	10.45	$396.42	8.16	$396.42	5.56	$396.42
	17.46	$565.80	14.27	$565.80	10.74	$565.80
	12.00	$429.45	12.00	$493.54	12.00	$600.36
Conventional Gas-to-Ammonia High NG: $12.00/MSCF			*$1,660,373,573*			
			N/A	$227.04		
			19.06	$396.42		
			39.25	$565.80		
			12.00	$348.85		

Table 21. Gas-to-ammonia and HTSE-to-ammonia IRR results (no sequestration).

	TCI -30% HTGR		TCI		TCI +50% HTGR	
	IRR	$/ton urea	IRR	$/ton urea	IRR	$/ton urea
HTGR-Integrated Gas-to-Ammonia High NG: $12.00/MSCF	*$2,302,360,099*		*$2,532,944,099*		*$2,917,250,765*	
	-2.07	$227.04	-2.66	$227.04	-3.50	$227.04
	18.24	$396.42	16.41	$396.42	14.03	$396.42
	32.02	$565.80	29.07	$565.80	25.28	$565.80
	12.00	$335.37	12.00	$348.81	12.00	$371.20
HTSE-to-Ammonia N_2 from Comb. High NG: $12.00/MSCF	*$5,675,626,481*		*$6,935,956,906*		*$9,036,507,614*	
	-3.00	$227.04	-4.19	$227.04	-$5.66	$227.04
	7.58	$396.42	5.55	$396.42	3.22	$396.42
	14.48	$565.80	11.59	$565.80	8.40	$565.80
	12.00	$495.17	12.00	$568.61	12.00	$691.01
HTSE-to-Ammonia N_2 from ASU High NG: $12.00/MSCF	*$5,202,407,985*		*$6,302,210,656*		*$8,135,215,109*	
	-1.68	$227.04	-2.88	$227.04	-$4.39	$227.04
	8.91	$396.42	6.81	$396.42	4.39	$396.42
	16.17	$565.80	13.16	$565.80	9.81	$565.80
	12.00	$460.80	12.00	$524.89	12.00	$631.70

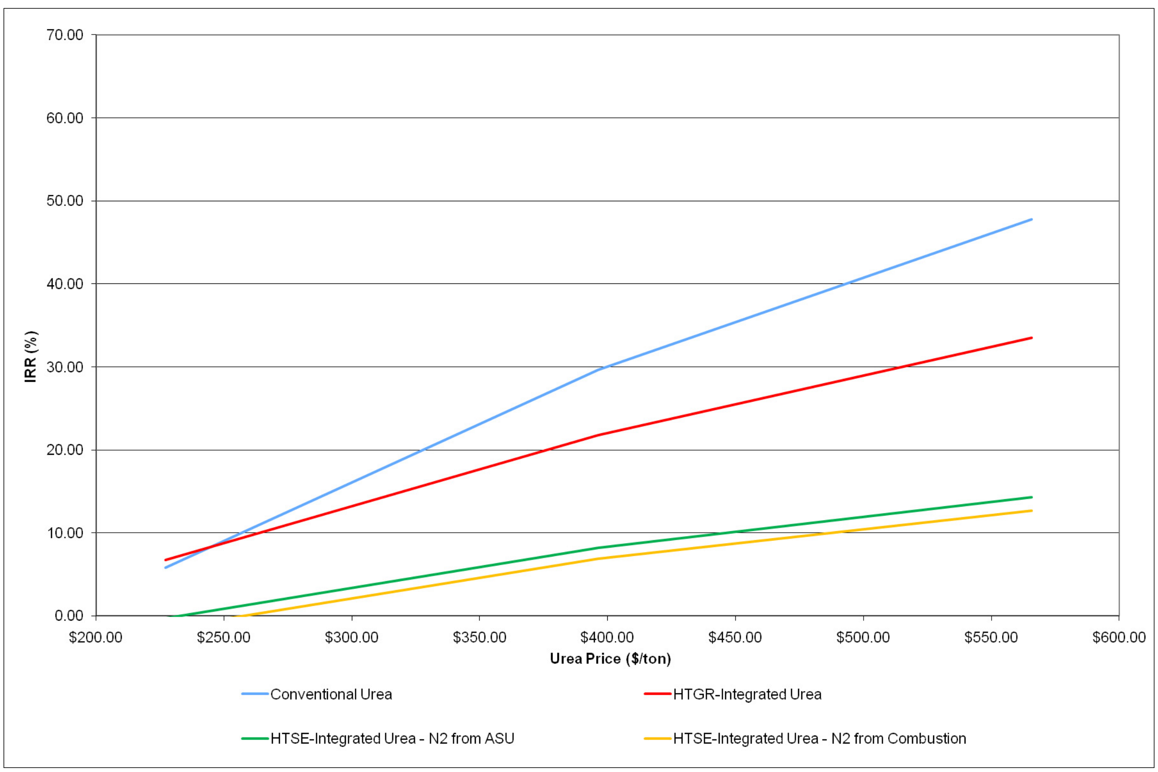

Figure 15. Gas-to-ammonia and HTSE-to-ammonia IRR economic results (average natural gas price, no sequestration).

Because the nuclear-integrated gas-to-ammonia case does not combust natural gas in the primary reformer, CO_2 emissions are reduced in this configuration compared to the conventional gas-to-ammonia case. Hence, implementation of a carbon tax will improve the economics of this case in comparison to conventional case. As seen in Table 22 and Figure 16, the selling price of urea is equal for both cases with a CO_2 tax of approximately $75/ton. For a tax in excess of this amount, the nuclear-integrated case economically outperforms the conventional case. Note that this result is valid independent of whether CO_2 is sequestered downstream of the water-gas shift converters (i.e., sequestration in this location is equally applicable to either the conventional or nuclear-integrated gas-to-ammonia case).

Table 22. Gas-to-ammonia and HTSE-to-ammonia carbon tax results (average natural gas price, no sequestration).

Carbon Tax $/ton		**TCI -30% HTGR**	**TCI**	**TCI +50% HTGR**
			Urea Price ($/ton)	
Conventional Gas-to-Ammonia	0		272.23	
	50		302.09	
	100		331.96	
	150		361.82	
	200		391.69	
HTGR-Integrated Gas-to-Ammonia	0	272.68	286.12	308.51
	50	293.29	306.73	329.12
	100	313.90	327.34	349.73
	150	334.51	347.95	370.34
	200	355.12	368.56	390.96
HTSE-to-Ammonia N_2 from Comb.	0	463.82	537.26	659.67
	50	463.92	537.36	659.77
	100	464.02	537.47	659.87
	150	464.13	537.57	659.98
	200	464.23	537.67	660.08
HTSE-to-Ammonia N_2 from ASU	0	429.45	493.54	600.36
	50	429.56	493.64	600.46
	100	429.66	493.75	600.56
	150	429.76	493.85	600.66
	200	429.86	493.95	600.77

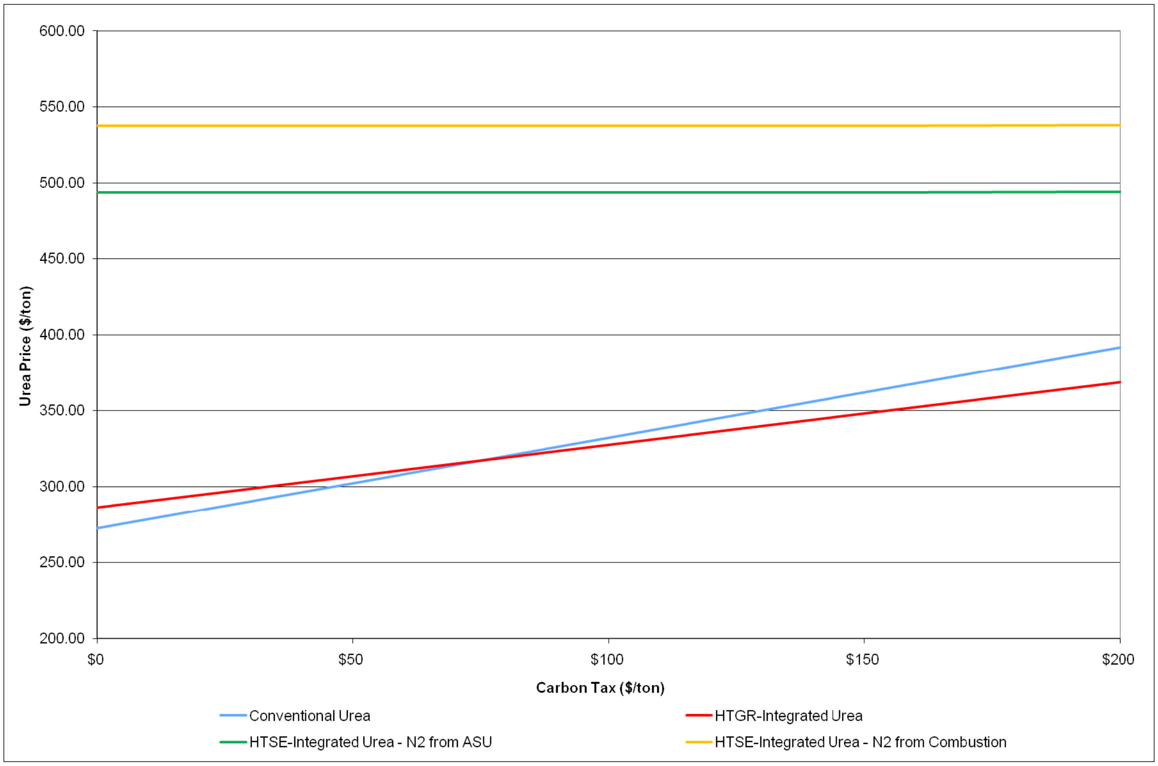

Figure 16. Impact of a carbon tax on gas-to-ammonia and HTSE-to-ammonia economics (average natural gas price, no sequestration).

Figure 17 presents results from the sensitivity analysis which varies natural gas price. For this analysis, the IRR is fixed by varying the selling prices of urea and ammonium nitrate. As the cost of natural gas increases, the HTGR-integrated case compares more favorably with the conventional gas-to-ammonia case. This is due to a sharper increase in manufacturing costs for the conventional case due to higher natural gas usage than for the HTGR-integrated case. However, in the absence of a carbon tax, economics of the nuclear-integrated case do not outperform the conventional case until natural gas prices rise to about $12/MSCF. If the cost of implementing HTGR technology can be reduced by 30%, these results change considerably. As shown in Figure 18, with a reduced HTGR implementation cost, the nuclear-integrated case outperforms the conventional gas-to-ammonia case when natural gas prices rise above $6.75/MSCF.

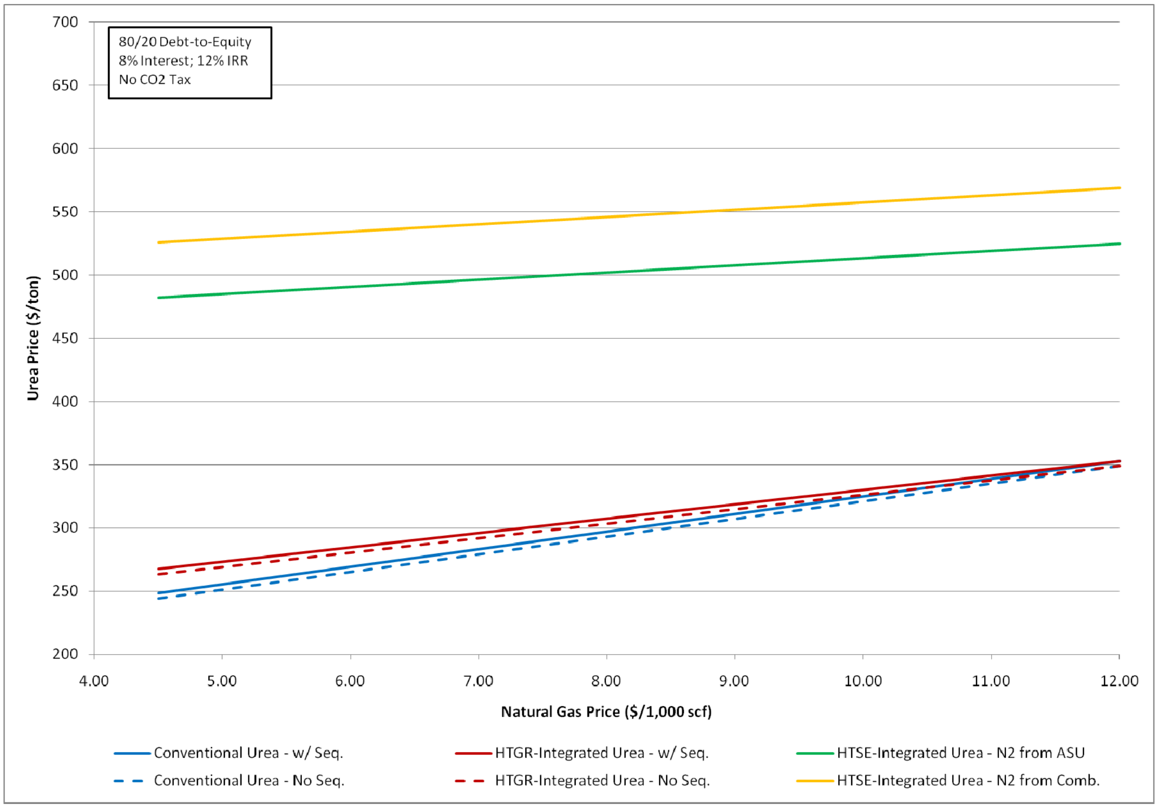

Figure 17. Gas-to-ammonia and HTSE-to-ammonia, urea price as a function of natural gas price (HTGR TCI, no CO_2 tax).

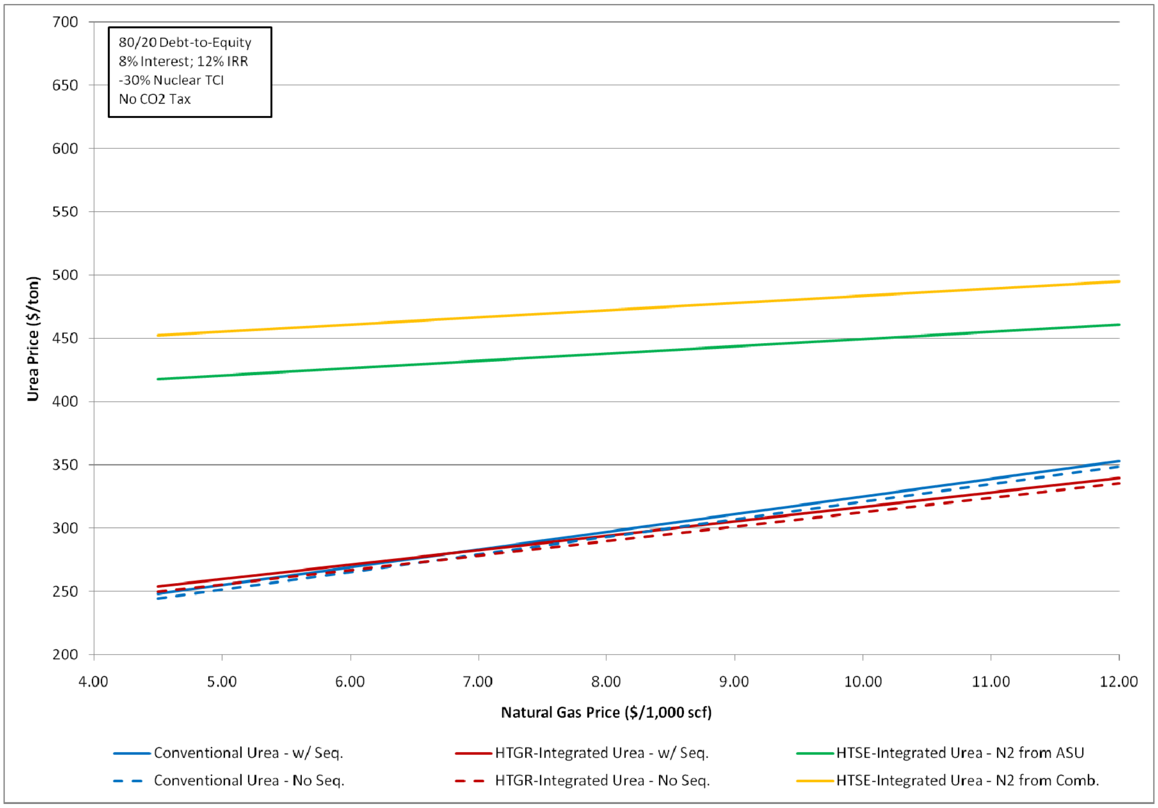

Figure 18. Gas-to-ammonia and HTSE-to-ammonia, urea price as a function of natural gas price (-30% HTGR TCI, no CO_2 tax).

To further emphasize the impact of a carbon tax, Figure 19 and Figure 20 plot the required selling price of urea as a function of natural gas price for a 12% IRR. Figure 19 assumes a \$50/ton CO_2 tax while Figure 20 assumes a \$100/ton CO_2 tax. For a CO_2 tax of \$50/ton, economics of the nuclear-integrated case outperform the conventional case only for natural gas prices above \$8.25/MSCF. However, for a CO_2 tax of \$100/ton, economics of the nuclear-integrated case outperform the conventional case for all natural gas prices above \$4.50/MSCF. Hence, it can be concluded that if carbon taxes are mandated in the future, favorability of the nuclear-integrated case improves significantly in comparison to the conventional gas-to-ammonia case.

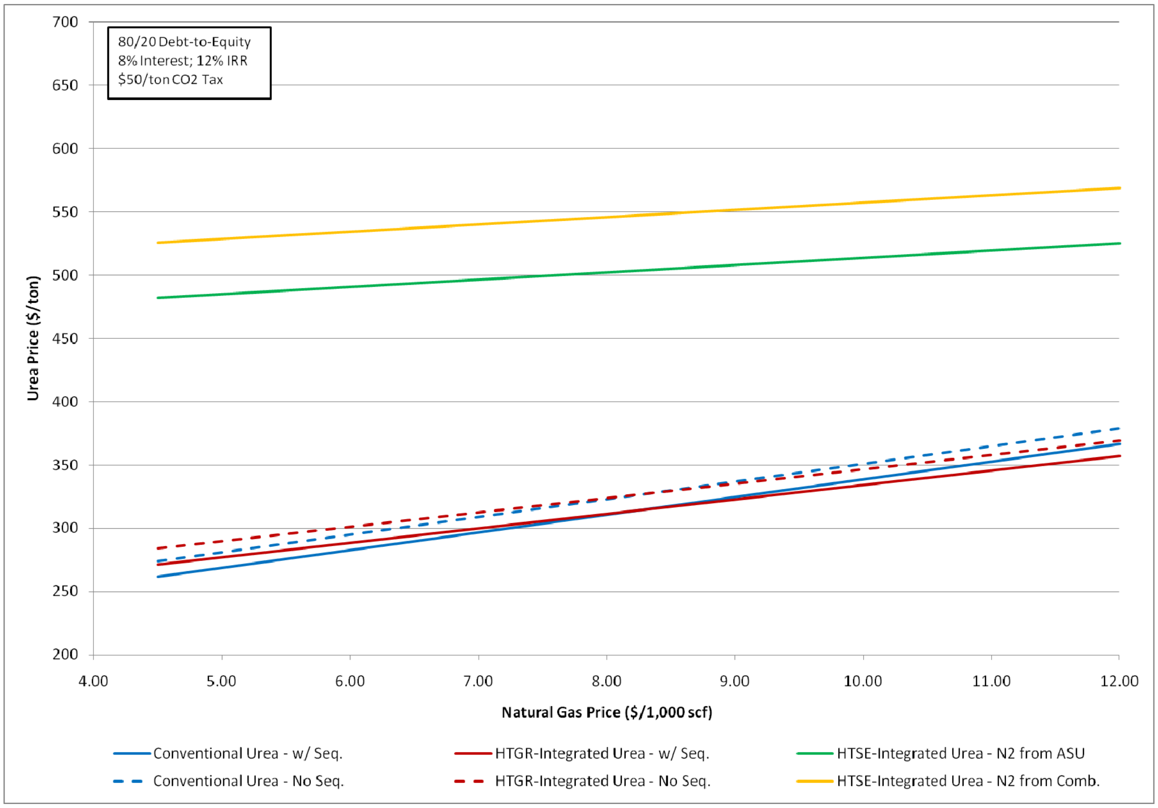

Figure 19. Gas-to-ammonia and HTSE-to-ammonia, urea price as a function of natural gas price (HTGR TCI, \$50/ton CO_2 tax).

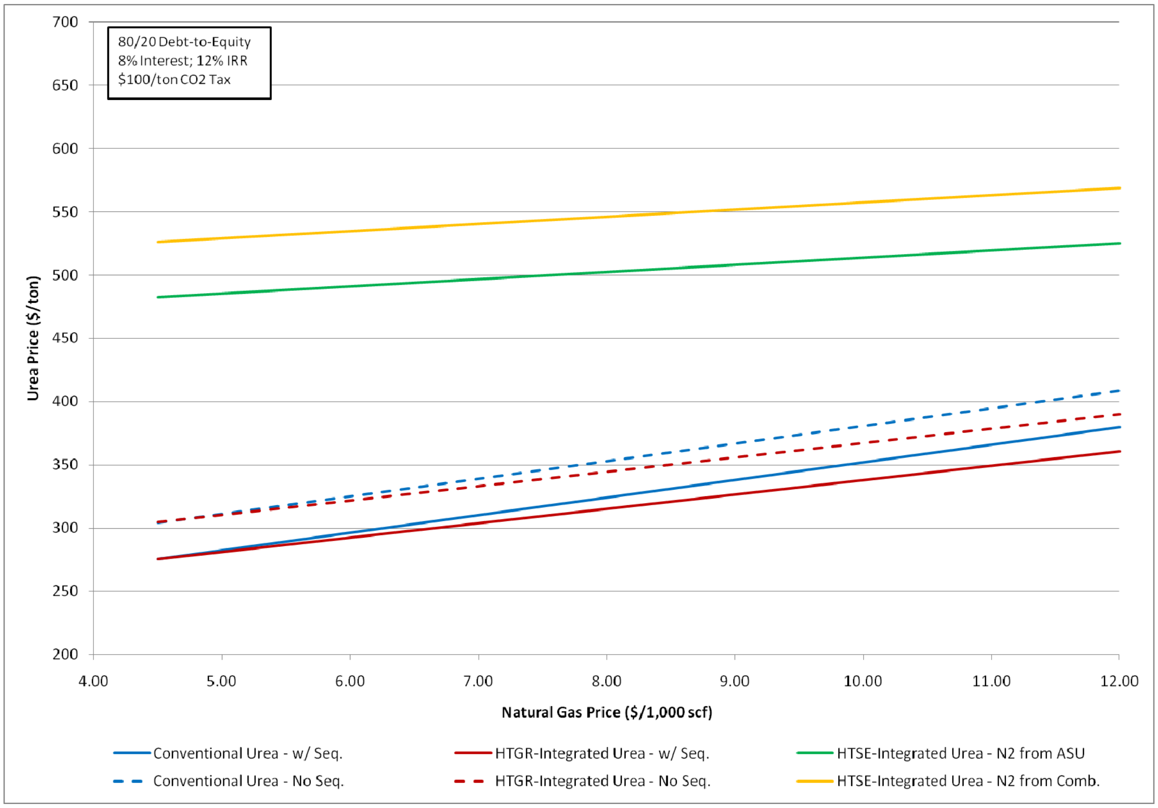

Figure 20. Gas-to-ammonia and HTSE-to-ammonia, urea price as a function of natural gas price (HTGR TCI, $100/ton CO_2 tax).

6. COAL-TO-AMMONIA CONCLUSIONS

Based on analysis of the conventional coal-to-ammonia case, a need for high-temperature heat integration was not identified for the coal-fed case. In addition, because light water reactors can supply power in this scenario, a strong need for integration with HTGRs is not apparent. Hence, modeling of a nuclear-integrated coal-to-ammonia case was not pursued. Note, however, that other gasification technologies (hydrogasification) could be considered, and these technologies may be more complementary for HTGR integration.

Because a suitable nuclear-integrated coal-to-ammonia case was not identified, an economic analysis was performed only for the conventional coal-to-ammonia case. Results indicate that a return of 12% can be achieved for a urea production price of $259.07/ton. Without sequestration, the urea production price increases sharply as a function of the carbon tax imposed; a urea selling price of $407.64/ton is required to achieve a 12% return when a $100/ton carbon tax is imposed.

7. GAS-TO-AMMONIA CONCLUSIONS

Results for the nuclear-integrated natural gas-to-ammonia case look promising. One 450 MW_t HTGR would be required to support this configuration. By substituting nuclear heat for natural gas combustion in the primary reformer, natural gas consumption is decreased by 18%. CO_2 emissions decrease by 70% compared to the conventional gas-to-ammonia case.

Results for the HTSE hydrogen cases also look promising from an emissions perspective. CO_2 emissions are cut to very low levels, and the need to sequester carbon is eliminated. In addition, natural gas consumption is reduced to only the amount required to supply carbon for urea production. These configurations would require roughly four 600 MW_t HTGRs to support plant operation. It is apparent that, of the two HTSE hydrogen cases considered, the option utilizing cryogenic air separation for nitrogen production is more attractive from the perspective of capital equipment cost.

Economically, the nuclear-integrated gas-to-ammonia option provides economic stability with respect to fluctuations in natural gas prices. Although the IRR is slightly lower than for conventional gas-to-ammonia, it is still significantly above 12% for all but the lowest fertilizer prices seen over the last decade. Furthermore, as the cost of natural gas increases, and if a carbon tax is implemented, the nuclear-integrated case compares more favorably with the conventional gas-to-ammonia case. For a modest carbon tax of \$50/ton CO_2, economics for the nuclear-integrated case outperform the conventional case when natural gas prices rise above \$8.25/MSCF. For a more substantial carbon tax of \$100/ton CO_2, economics for the nuclear-integrated case outperform the conventional case when natural gas prices rise above \$4.50/MSCF. In addition, if the cost of the HTGR can be reduced by 30%, the HTGR-integrated case outperforms the conventional gas-to-ammonia case when the price of natural gas rises above \$6.75/MSCF (regardless of a carbon tax). Hence, it is clear that the economic performance of the nuclear-integrated case can outperform the conventional gas-to-ammonia case in a number of scenarios including: (1) a combination of elevated natural gas prices and adoption of a modest carbon tax, and (2) further development and commercialization of HTGR technology are able to reduce the price of implementation by 30%.

Economics for the HTSE-to-ammonia cases do not look as favorable. Although these options are attractive due to simplification of gas cleanup requirements, they significantly increase the capital cost of the plant. This is due principally to the need for four nuclear reactors to support both hydrogen and electrical demands of the plant. In order to be economically attractive, urea prices would need to be around \$500/ton to achieve a 12% return. Even if further development and commercialization of HTGR technology are able to reduce the price of implementation by 30%, the selling price of urea would still need to be in excess of \$400/ton to achieve a 12% return for the HTSE-to-ammonia cases.

8. FUTURE WORK AND RECOMMENDATIONS

The following items should be performed in the future to further refine the process and economic modeling performed for the ammonia cases:

- Refined estimates of the HTGR capital cost, annual fuel costs, and annual O&M costs should be developed to refine the economic results.

- A separate study should be conducted to assess the optimal siting of the HTGR with respect to the ammonia facilities, balancing safety concerns associated with separation distance and heat losses associated with transporting high-temperature heat long distances.

- Rigorous Aspen Plus submodels of the HTGR and HTSE units should be developed to fully couple heat and power integration from the HTGR.

- The simplified water treatment hierarchy should be replaced with more rigorous water treatment models based on vendor input.

- It is likely that process and economic results for the nuclear-integrated cases could be improved if the HTGR temperature could be increased from 750 to 950°C. Hence, a study to quantify the performance improvement is recommended.

- Due to differences in how utility companies and ammonia producers currently evaluate economics of proposed projects, and due to the high capital costs associated with integrating these technologies, additional work is recommended to identify and evaluate more realistic business models for this integration.

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10. APPENDIXES

Appendix A, Detailed Modeling Results and Flowsheets

Appendix B, Ammonia Capital Cost Estimates

Appendix C, [Electronic] Coal to Ammonia Conventional Stream Results.xls

Appendix D, [Electronic] NG to Ammonia Conventional Stream Results.xls

Appendix E, [Electronic] NG to Ammonia Nuclear Integration Stream Results.xls

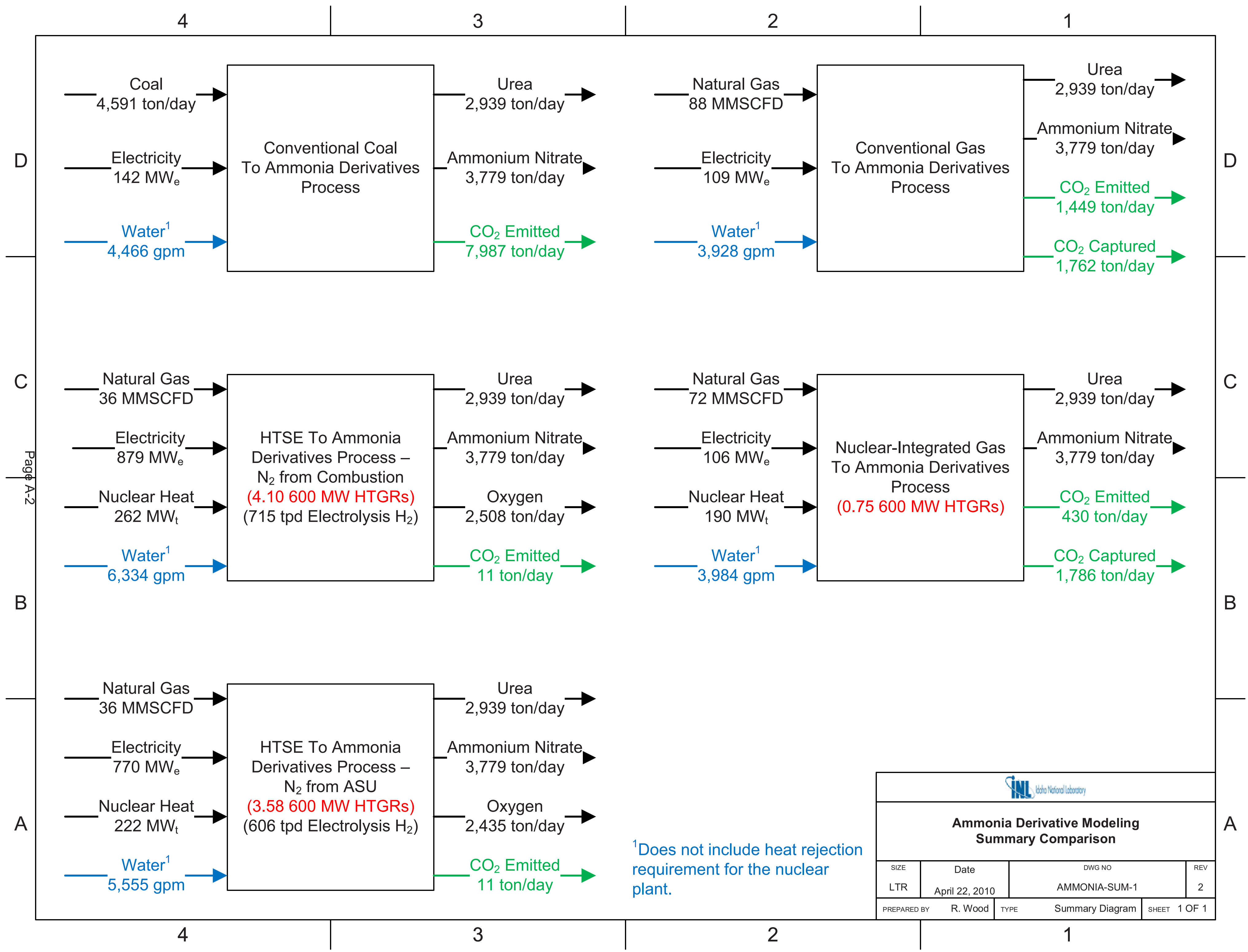
Appendix F, [Electronic] HTSE to Ammonia – N2 from Combustion Stream Results.xls

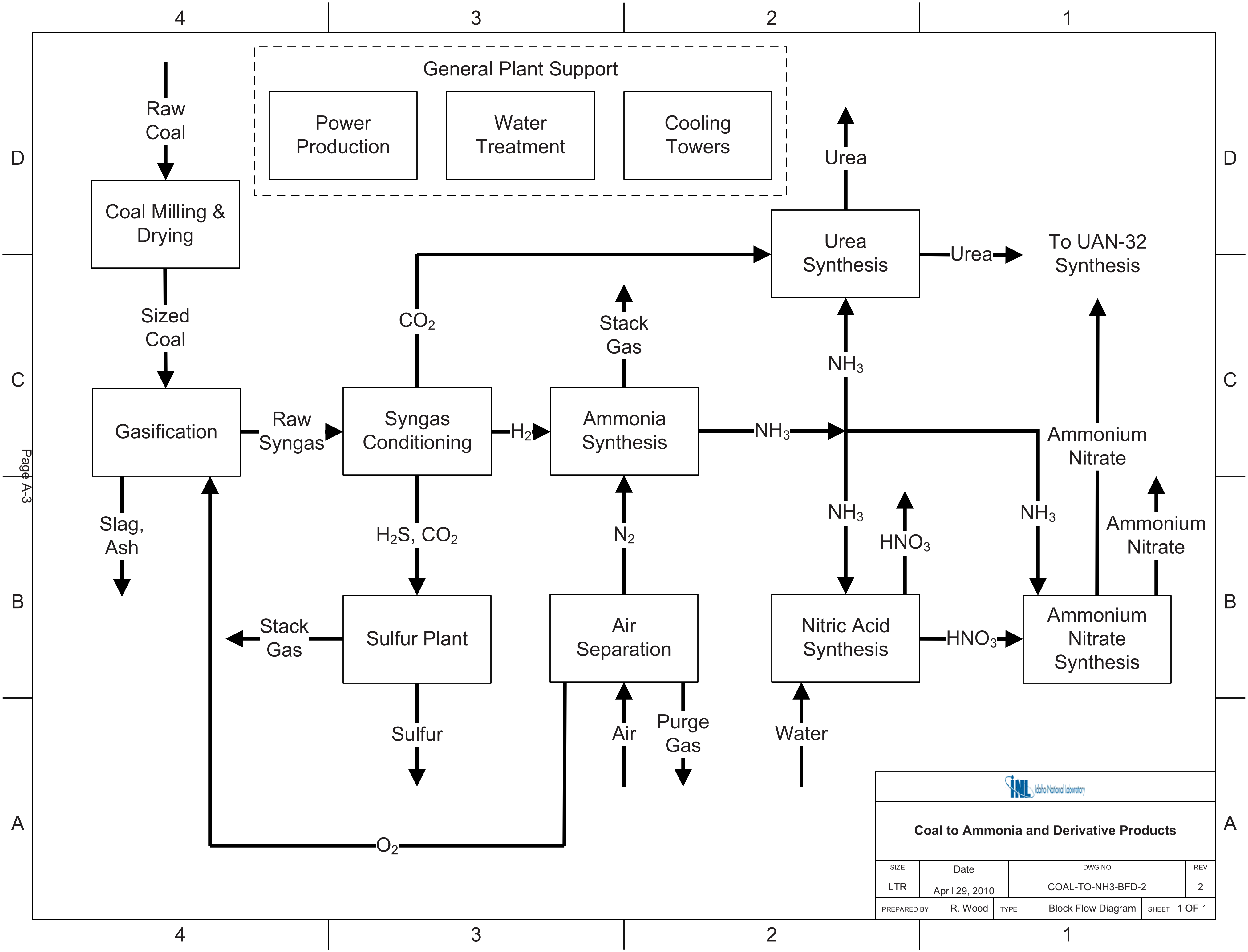
Appendix G, [Electronic] HTSE to Ammonia – N2 from Cryo ASU Stream Results.xls

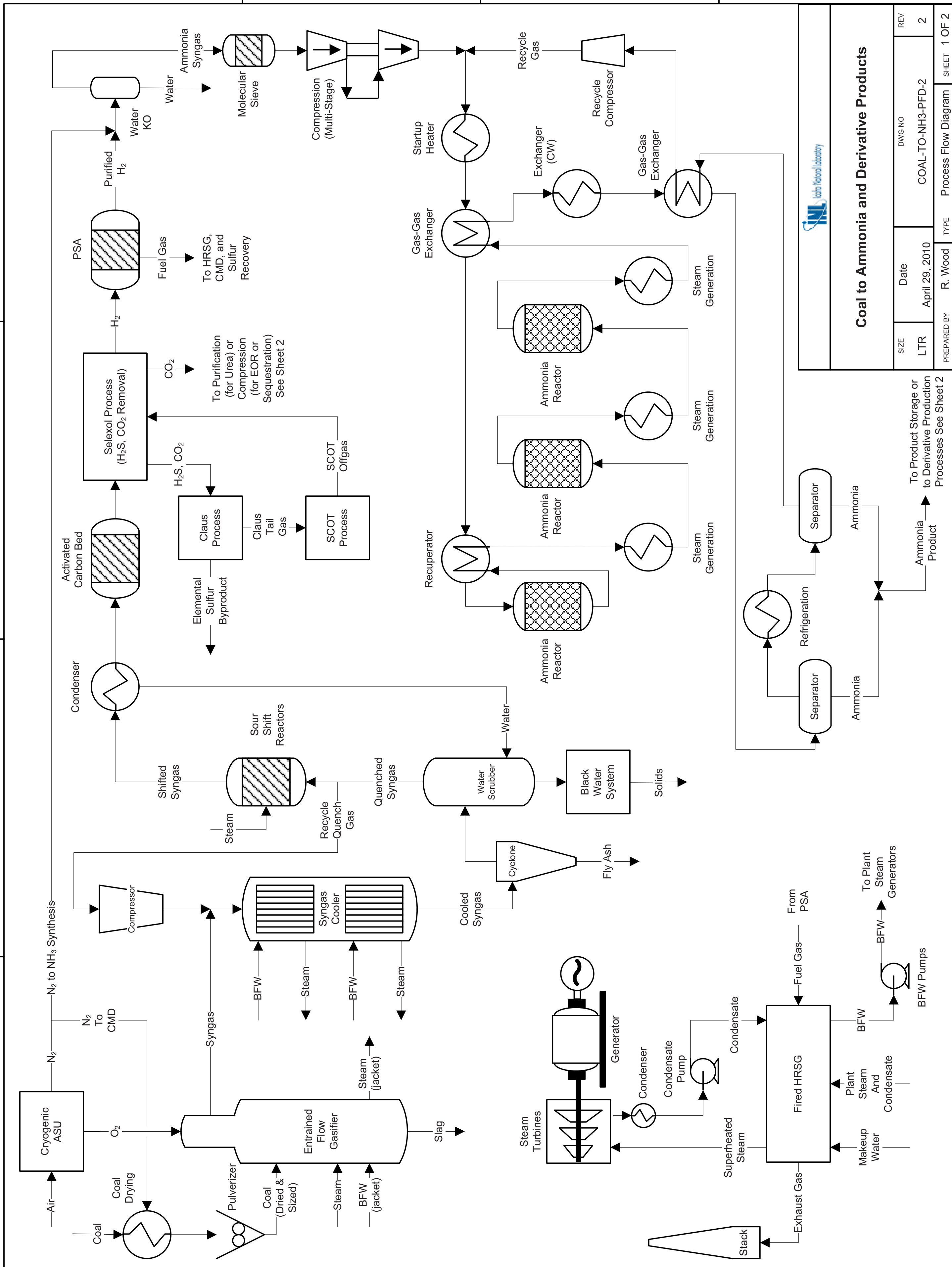
Appendix A
Detailed Modeling Results and Flowsheets

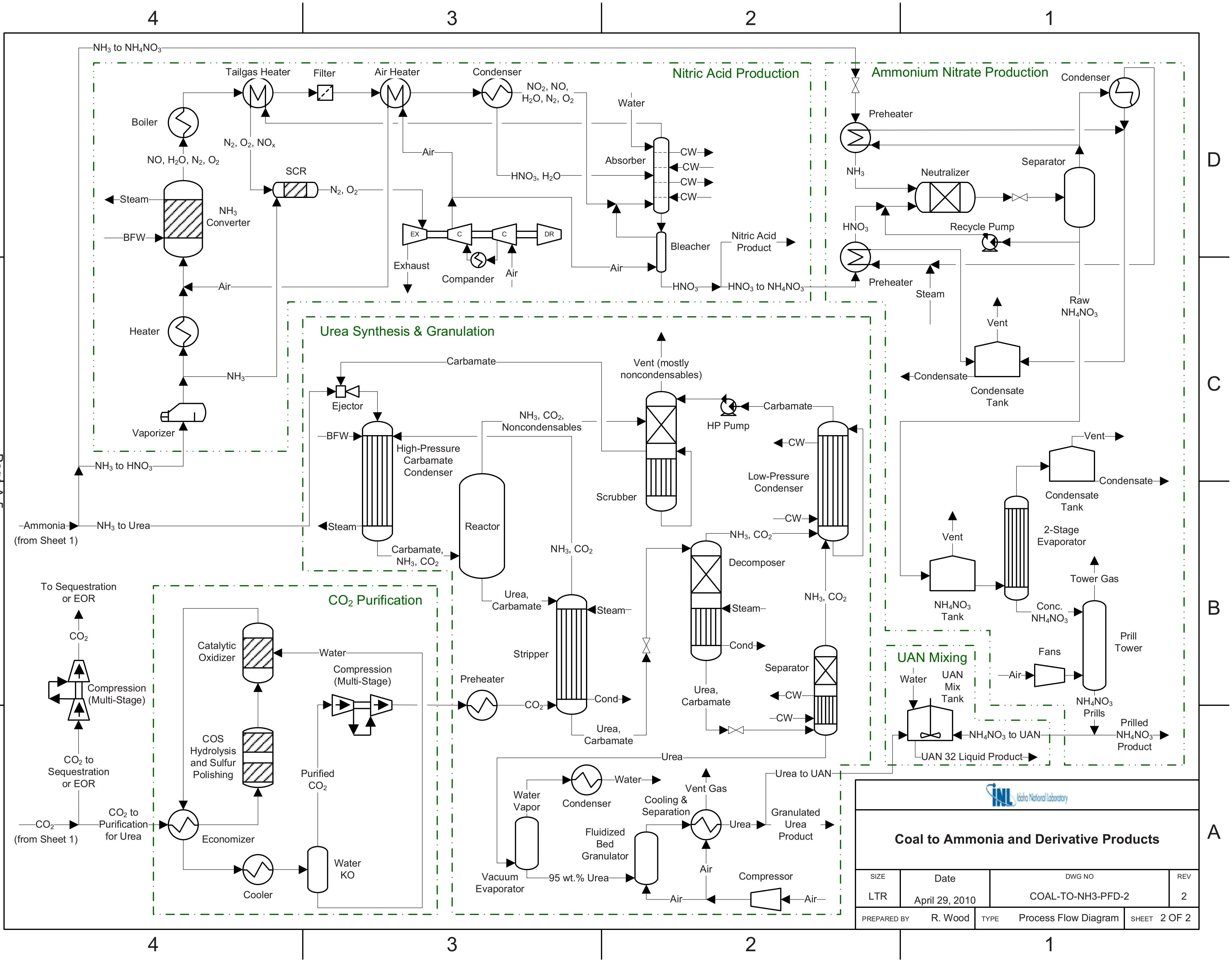
Ammonia Derivative Case Summary

	Coal Conventional	Natural Gas Conventional	Natural Gas Nuclear Assisted	HTSE N2 from Combustion	HTSE N2 from ASU
Inputs					
Coal Feed Rate (ton/day)	4,591	0	0	0	0
Natural Gas Feed Rate (MMSCFD)	n/a	88	72	36	36
# 600 MWt HTGRs Required	n/a	n/a	0.75	4.10	3.58
Intermediate Products					
Ammonia (ton/day)	3,358	3,361	3,360	3,360	3,360
Nitric Acid (ton/day)	5,186	5,192	5,192	5,192	5,192
Outputs					
Urea (ton/day)	2,939	2,939	2,939	2,939	2,939
Ammonium Nitrate (ton/day)	3,779	3,779	3,779	3,779	3,779
Oxygen (ton/day)	0	0	0	2,508	2,435
Utility Summary					
Total Power (MW)	-141.8	-108.8	-106.2	-879.0	-769.5
Electrolyzers	n/a	n/a	n/a	-918.1	-778.1
Air Separation	-73.5	n/a	n/a	n/a	-23.2
Nitrogen Generation (non-ASU)	n/a	n/a	n/a	-18.8	n/a
Coal Milling & Drying	-2.5	n/a	n/a	n/a	n/a
Gasification Island	-2.5	n/a	n/a	n/a	n/a
Natural Gas Reforming	n/a	-21.6	-19.9	n/a	n/a
Syngas Purification	-5.4	-4.4	-4.5	n/a	n/a
Claus Plant	0.0	n/a	n/a	n/a	n/a
Sulfur Reduction (SCOT)	-0.7	n/a	n/a	n/a	n/a
Non-Nuclear Power Island	44.7	26.7	29.5	156.0	128.0
CO_2 Production/Purification/Compression	-7.4	-13.0	-13.1	-5.6	-5.6
Ammonia Synthesis	-43.2	-45.2	-46.8	-39.0	-38.5
Nitric Acid Synthesis	-15.1	-15.1	-15.1	-15.1	-15.1
Ammonium Nitrate Synthesis	-24.9	-24.9	-24.9	-24.9	-24.9
Urea Synthesis	-4.4	-4.4	-4.4	-4.4	-4.4
Cooling Towers	-2.1	-1.5	-1.5	-2.5	-2.3
Water Treatment	-4.8	-5.4	-5.6	-6.6	-5.6
Total Water Balance (gpm)	-4,466	-3,928	-3,984	-6,334	-5,555
Evaporation Rate (gpm)	-4,891	-3,601	-3,657	-5,979	-5,386
CO_2 Emissions					
Captured (ton/day CO_2)	0	1,762	1,786	0	0
Emitted (ton/day CO_2)	7,987	1,449	430	11	11
Nuclear Integration Summary					
Electricity Demand (MW)	n/a	n/a	106.2	879.0	769.5
HTSE	n/a	n/a	n/a	918.1	778.1
Balance of Plant	n/a	n/a	106.2	-39.1	-8.6
Nuclear Heat Demand (MMBTU/hr)	n/a	n/a	649.2	894.6	758.2
Electrolyzers	n/a	n/a	n/a	894.6	758.2
Natural Gas Reforming	n/a	n/a	649.2	n/a	n/a
Fossil Topping Heat Supplied (MMBTU/hr)	n/a	n/a	n/a	50.4	42.7
Electrolysis Products					
Total Hydrogen (ton/day)	n/a	n/a	n/a	715	606
Total Oxygen (ton/day)	n/a	n/a	n/a	5,637	4,778


Coal
4,591 ton/day
Electricity
142 MWe
Water1
4,466 gpm
Conventional Coal
To Ammonia Derivatives
Process
Urea
2,939 ton/day
Ammonium Nitrate
3,779 ton/day
CO2 Emitted
7,987 ton/day
Natural Gas
88 MMSCFD
Electricity
109 MWe
Water1
3,928 gpm
Conventional Gas
To Ammonia Derivatives
Process
Urea
2,939 ton/day
Ammonium Nitrate
3,779 ton/day
CO2 Emitted
1,449 ton/day
CO2 Captured
1,762 ton/day
Natural Gas
36 MMSCFD
Electricity
879 MWe
Nuclear Heat
262 MWt
Water1
6,334 gpm
HTSE To Ammonia
Derivatives Process –
N2 from Combustion
(4.10 600 MW HTGRs)
(715 tpd Electrolysis H2)
Urea
2,939 ton/day
Ammonium Nitrate
3,779 ton/day
Oxygen
2,508 ton/day
CO2 Emitted
11 ton/day
Natural Gas
72 MMSCFD
Electricity
106 MWe
Nuclear Heat
190 MWt
Water1
3,984 gpm
Nuclear-Integrated Gas
To Ammonia Derivatives
Process
(0.75 600 MW HTGRs)
Urea
2,939 ton/day
Ammonium Nitrate
3,779 ton/day
CO2 Emitted
430 ton/day
CO2 Captured
1,786 ton/day
Natural Gas
36 MMSCFD
Electricity
770 MWe
Nuclear Heat
222 MWt
Water1
5,555 gpm
HTSE To Ammonia
Derivatives Process –
N2 from ASU
(3.58 600 MW HTGRs)
(606 tpd Electrolysis H2)
Urea
2,939 ton/day
Ammonium Nitrate
3,779 ton/day
Oxygen
2,435 ton/day
CO2 Emitted
11 ton/day
1Does not include heat rejection
requirement for the nuclear
plant.
Idaho National Laboratory
Ammonia Derivative Modeling
Summary Comparison
SIZE
LTR
Date
April 22, 2010
DWG NO
AMMONIA-SUM-1
REV
2
PREPARED BY
R. Wood
TYPE
Summary Diagram
SHEET
1 OF 1


General Plant Support
Power Production
Water Treatment
Cooling Towers
Raw Coal
Coal Milling & Drying
Sized Coal
Gasification
Slag, Ash
Raw Syngas
Syngas Conditioning
CO_2
H_2
H_2S, CO_2
Sulfur Plant
Stack Gas
Sulfur
Ammonia Synthesis
Stack Gas
N_2
Air Separation
Air
Purge Gas
O_2
NH_3
Urea Synthesis
Urea
Urea
To UAN-32 Synthesis
Nitric Acid Synthesis
Water
HNO_3
HNO_3
Ammonium Nitrate Synthesis
Ammonium Nitrate
Ammonium Nitrate
INL Idaho National Laboratory
Coal to Ammonia and Derivative Products
SIZE LTR
Date April 29, 2010
DWG NO COAL-TO-NH3-BFD-2
REV 2
PREPARED BY R. Wood
TYPE Block Flow Diagram
SHEET 1 OF 1


Coal to Ammonia and Derivative Products
Idaho National Laboratory
SIZE LTR
Date April 29, 2010
DWG NO COAL-TO-NH3-PFD-2
REV 2
PREPARED BY R. Wood
TYPE Process Flow Diagram
SHEET 1 OF 2
Air
Cryogenic ASU
N_2
N_2 to NH_3 Synthesis
N_2 To CMD
O_2
Coal
Coal Drying
Pulverizer
Coal (Dried & Sized)
Steam
BFW (jacket)
Steam (jacket)
Entrained Flow Gasifier
Slag
Syngas
Compressor
Syngas Cooler
BFW
Steam
BFW
Steam
Cooled Syngas
Cyclone
Fly Ash
Water Scrubber
Black Water System
Solids
Quenched Syngas
Recycle Quench Gas
Steam
Sour Shift Reactors
Shifted Syngas
Water
Condenser
Activated Carbon Bed
Selexol Process (H_2S, CO_2 Removal)
H_2S, CO_2
Claus Process
Elemental Sulfur Byproduct
Claus Tail Gas
SCOT Process
SCOT Offgas
CO_2
To Purification (for Urea) or Compression (for EOR or Sequestration) See Sheet 2
H_2
PSA
Fuel Gas
To HRSG, CMD, and Sulfur Recovery
Purified H_2
Water KO
Water
Ammonia Syngas
Molecular Sieve
Compression (Multi-Stage)
Recycle Gas
Recycle Compressor
Startup Heater
Gas-Gas Exchanger
Exchanger (CW)
Gas-Gas Exchanger
Steam Generation
Ammonia Reactor
Steam Generation
Ammonia Reactor
Steam Generation
Recuperator
Ammonia Reactor
Separator
Ammonia
Refrigeration
Separator
Ammonia
Ammonia Product
To Product Storage or to Derivative Production Processes See Sheet 2
Steam Turbines
Generator
Condenser
Condensate Pump
Condensate
Superheated Steam
Fired HRSG
Fuel Gas
From PSA
BFW
BFW Pumps
BFW
To Plant Steam Generators
Plant Steam And Condensate
Makeup Water
Exhaust Gas
Stack
1
2
3
4
A
B
C
D



Coal to Ammonia and Derivative Products
SIZE LTR
Date April 29, 2010
DWG NO COAL-TO-NH3-PFD-2
REV 2
PREPARED BY R. Wood
TYPE Process Flow Diagram
SHEET 2 OF 2
Nitric Acid Production
Ammonium Nitrate Production
Urea Synthesis & Granulation
CO2 Purification
UAN Mixing
Tailgas Heater
Filter
Air Heater
Condenser
NO2, NO, H2O, N2, O2
Water
Absorber
CW
Boiler
N2, O2, NOx
NO, H2O, N2, O2
SCR
N2, O2
Air
HNO3, H2O
Steam
NH3 Converter
BFW
EX
C
DR
Exhaust
Compander
Bleacher
Nitric Acid Product
HNO3
HNO3 to NH4NO3
Heater
NH3
Vaporizer
NH3 to HNO3
NH3 to NH4NO3
Ammonia
(from Sheet 1)
NH3 to Urea
Preheater
Neutralizer
Separator
Recycle Pump
Raw NH4NO3
Vent
Condensate
Condensate Tank
Carbamate
Ejector
Vent (mostly noncondensables)
HP Pump
NH3, CO2, Noncondensables
High-Pressure Carbamate Condenser
Scrubber
Low-Pressure Condenser
Reactor
Carbamate, NH3, CO2
NH3, CO2
Decomposer
Urea, Carbamate
Stripper
Cond
Separator
2-Stage Evaporator
Tower Gas
Conc. NH4NO3
NH4NO3 Tank
Prill Tower
Fans
NH4NO3 Prills
Prilled NH4NO3 Product
UAN Mix Tank
NH4NO3 to UAN
UAN 32 Liquid Product
To Sequestration or EOR
CO2
Compression (Multi-Stage)
CO2 to Sequestration or EOR
CO2 to Purification for Urea
Catalytic Oxidizer
COS Hydrolysis and Sulfur Polishing
Purified CO2
Economizer
Cooler
Water KO
Urea
Urea to UAN
Water Vapor
Vent Gas
Cooling & Separation
Granulated Urea Product
Fluidized Bed Granulator
Vacuum Evaporator
95 wt.% Urea
Compressor

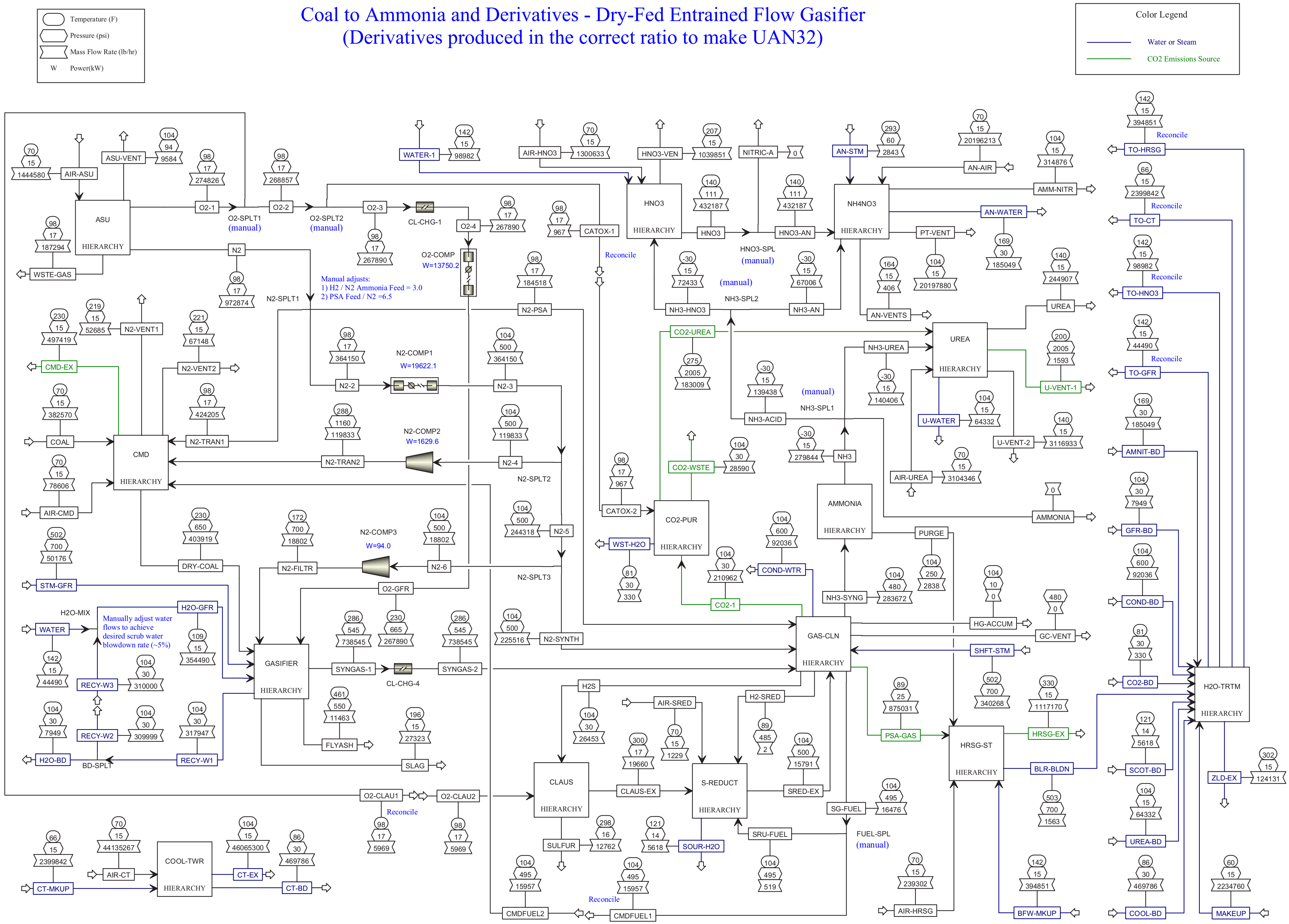
```
Calculator Block SUMMARY

   FEED & PRODUCT SUMMARY:

     FEEDS:

       RAW COAL FEED RATE =                 4590.8 TON/DY
         COAL HHV AS FED =                 10934. BTU/LB
         COAL MOISTURE AS FED =               13.70 %

       PROXIMATE ANALYSIS (DRY BASIS):
          MOISTURE                            13.70 %
          FIXED CARBON                        40.12 %
          VOLATILE MATTER                     49.28 %
          ASH                                 10.60 %

       ULTIMATE ANALYSIS (DRY BASIS):
          ASH                                 10.60 %
          CARBON                              70.27 %
          HYDROGEN                             4.84 %
          NITROGEN                             1.36 %
          CHLORINE                             0.11 %
          SULFUR                               3.72 %
          OXYGEN                               9.10 %

       SULFANAL ANALYSIS (DRY BASIS):
          PYRITIC                              1.94 %
          SULFATE                              0.08 %
          ORGANIC                              1.70 %

     INTERMEDIATES:

       COAL FEED RATE AFTER DRYING =        4214.8 TON/DY
         COAL HHV AFTER DRYING =           11910. BTU/LB
         COAL MOISTURE AFTER DRYNG =           6.00 %

       RAW SYNGAS MASS FLOW =             686715. LB/HR
       RAW SYNGAS VOLUME FLOW =                7. MMSCFD @ 60˚F
       GASIFIER COLD GAS EFFICIENCY =         80.36 %
       RAW SYNGAS COMPOSITION:
          H2                                  29.0 MOL.%
          CO                                  53.7 MOL.%
          CO2                                  4.0 MOL.%
          N2                                   6.1 MOL.%
          H2O                                  5.6 MOL.%
          H2S                              10519. PPMV
          CH4                                 79. PPMV

       QUENCHED SYNGAS MASS FLOW =        738545. LB/HR
       QUENCHED SYNGAS VOLUME FLOW =           9. MMSCFD @ 60˚F
       QUENCHED SYNGAS COMPOSITION:
          H2                                  26.9 MOL.%
          CO                                  49.8 MOL.%
          CO2                                  3.8 MOL.%
          N2                                   7.6 MOL.%
          H2O                                 10.7 MOL.%
          H2S                               9766. PPMV
          CH4                                 71. PPMV

       CLEANED SYNGAS MASS FLOW =         283672. LB/HR
       CLEANED SYNGAS VOLUME FLOW =            9. MMSCFD @ 60˚F
       CLEANED SYNGAS HHV =                 7863. BTU/SCF
       CLEANED SYNGAS COMPOSITION:
          H2                                  75.0 MOL.%
```

```
     CO                                       0.0 MOL.%
     CO2                                      0.0 MOL.%
     N2                                      25.0 MOL.%
     H2O                                      0.0 MOL.%
     H2S                                      0. PPMV
     CH4                                      0. PPMV

  AMMONIA MASS FLOW =                  279845. LB/HR
  AMMONIA MASS FLOW =                    3358. TON/DY
     AMMONIA PURITY =                    100.00 MOL.%
  AMMONIA PRODUCED / COAL FED =            0.73 LB/LB

  NITRIC ACID MASS FLOW =              432186. LB/HR
  NITRIC ACID MASS FLOW =                5186. TON/DY
     NITRIC ACID PURITY =                 57.93 WT.%
  HNO3 PRODUCED / NH3 FED =                5.97 LB/LB

FINAL PRODUCTS:

  AMMONIA MASS FLOW =                        0. LB/HR
  AMMONIA MASS FLOW =                        0. TON/DY
     AMMONIA PURITY =                    100.00 MOL.%

  NITRIC ACID MASS FLOW =                    0. LB/HR
  NITRIC ACID MASS FLOW =                    0. TON/DY
     NITRIC ACID PURITY =                 57.93 WT.%

  UREA MASS FLOW =                     244907. LB/HR
  UREA MASS FLOW =                       2939. TON/DY
  UREA PRODUCED / AMMONIA FED =            1.74 LB/LB
  UREA PRODUCED / CO2 FED =                1.34 LB/LB

  AMMONIUM NITRATE MASS FLOW =         314876. LB/HR
  AMMONIUM NITRATE MASS FLOW =           3779. TON/DY
  NH4NO3 PRODUCED / NITRIC ACID FED =      0.73 LB/LB
  NH4NO3 PRODUCED / AMMONIA FED =          4.70 LB/LB


POWER SUMMARY:

  ELECTRICAL GENERATORS:
    STEAM TURBINE POWER GENERATION =      46.5 MW
  GENERATOR SUBTOTAL =                    46.5 MW

  ELECTRICAL CONSUMERS:
    COAL PROCESSING POWER CONSUMPTION =    2.5 MW
    ASU POWER CONSUMPTION =               73.5 MW
    GASIFIER POWER CONSUMPTION =           2.5 MW
    GAS CLEANING POWER CONSUMPTION =       5.4 MW
    CLAUS POWER CONSUMPTION =              0.0 MW
    SCOT POWER CONSUMPTION =               0.7 MW
    POWER BLOCK POWER CONSUMPTION =        1.8 MW
    CO2 PROCESSING POWER CONSUMPTION =     7.4 MW
    AMMONIA SYNTH. POWER CONSUMPTION =    43.2 MW
    UREA SYNTHESIS POWER CONSUMPTION =     4.4 MW
    HNO3 SYNTHESIS POWER CONSUMPTION =    15.1 MW
    NH4NO3 SYNTH. POWER CONSUMPTION =     24.9 MW
    COOLING TOWER POWER CONSUMPTION =      2.1 MW
    WATER TREATMENT POWER CONSUMPTION =    4.8 MW
  CONSUMER SUBTOTAL =                    188.3 MW

  NET PLANT POWER CONSUMPTION =          141.8 MW
```

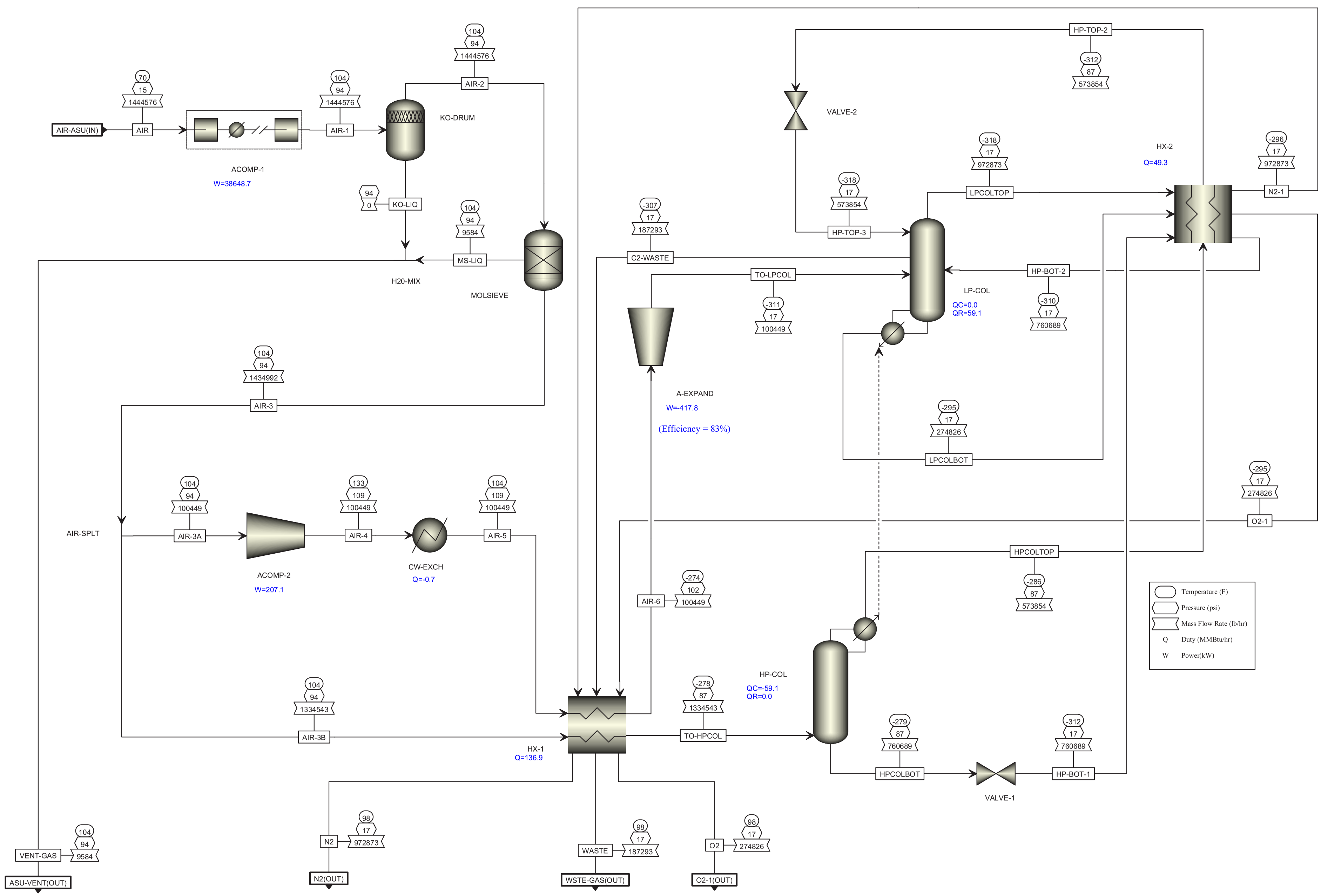
Conventional Coal-to-Ammonia Results

```
WATER BALANCE:

  EVAPORATIVE LOSSES:
    CMD WATER NOT RECOVERED =              84.1 GPM
    COOLING TOWER EVAPORATION =          4558.6 GPM
    ZLD SYSTEM EVAPORATION =              248.1 GPM
  TOTAL EVAPORATIVE LOSSES =             4890.7 GPM

  WATER CONSUMED:
    GASIFIER ISLAND MAKEUP =               88.9 GPM
    BOILER FEED WATER MAKEUP =            789.1 GPM
    NITRIC ACID FEED WATER =              197.8 GPM
    COOLING TOWER MAKEUP =               4795.8 GPM
  TOTAL WATER CONSUMED =                 5871.6 GPM

  WATER GENERATED:
    GASIFIER ISLAND BLOWDOWN =             15.9 GPM
    SYNGAS CONDENSER BLOWDOWN =           183.9 GPM
    SOUR WATER BLOWDOWN=                   11.2 GPM
    BOILER BLOWDOWN =                       3.1 GPM
    CO2 PURIFICATION BLOWDOWN =             0.7 GPM
    UREA PRODUCTION BLOWDOWN =            128.6 GPM
    NH4NO3 PRODUCTION BLOWDOWN =          369.8 GPM
    COOLING TOWER BLOWDOWN =              938.8 GPM
  TOTAL WATER GENERATED =                1652.0 GPM

  PLANT WATER SUMMARY:
    NET MAKEUP WATER REQUIRED =          4465.9 GPM
    WATER CONSUMED / COAL FED =             5.84 LB/LB


CO2 BALANCE:

  CO2 INCORPORATED INTO UREA PRODUCT =  2177. TON/DY
  CO2 INCORPORATED INTO UREA PRODUCT =    38. MMSCFD @ 60˚F
    CO2 PURITY =                          99.5 %

  CO2 EMITTED (TOTAL) =                 7987. TON/DY
  CO2 EMITTED (TOTAL) =                  138. MMSCFD @ 60˚F
    FROM HRSG =                         7457. TON/DY
    FROM CMD =                           159. TON/DY
    FROM UREA VENTS =                     28. TON/DY
    EXCESS FROM SELEXOL =                343. TON/DY
  CO2 EMITTED / COAL FED =                 1.74 LB/LB


Calculator Block AIRPROPS

  HUMIDITY DATA FOR STREAM AIR-ASU:
    HUMIDITY RATIO =                      43.5 GRAINS/LB
    RELATIVE HUMIDITY =                   39.9 %


Calculator Block PH       Hierarchy: NH4NO3

  TARGET NEUTRALIZER PH =   1.500
  ACTUAL NEUTRALIZER PH =   1.499


Calculator Block RATIO    Hierarchy: UREA

  TARGET REACTOR NH3/CO2 RATIO =   2.950
  ACTUAL REACTOR NH3/CO2 RATIO =   2.923
```

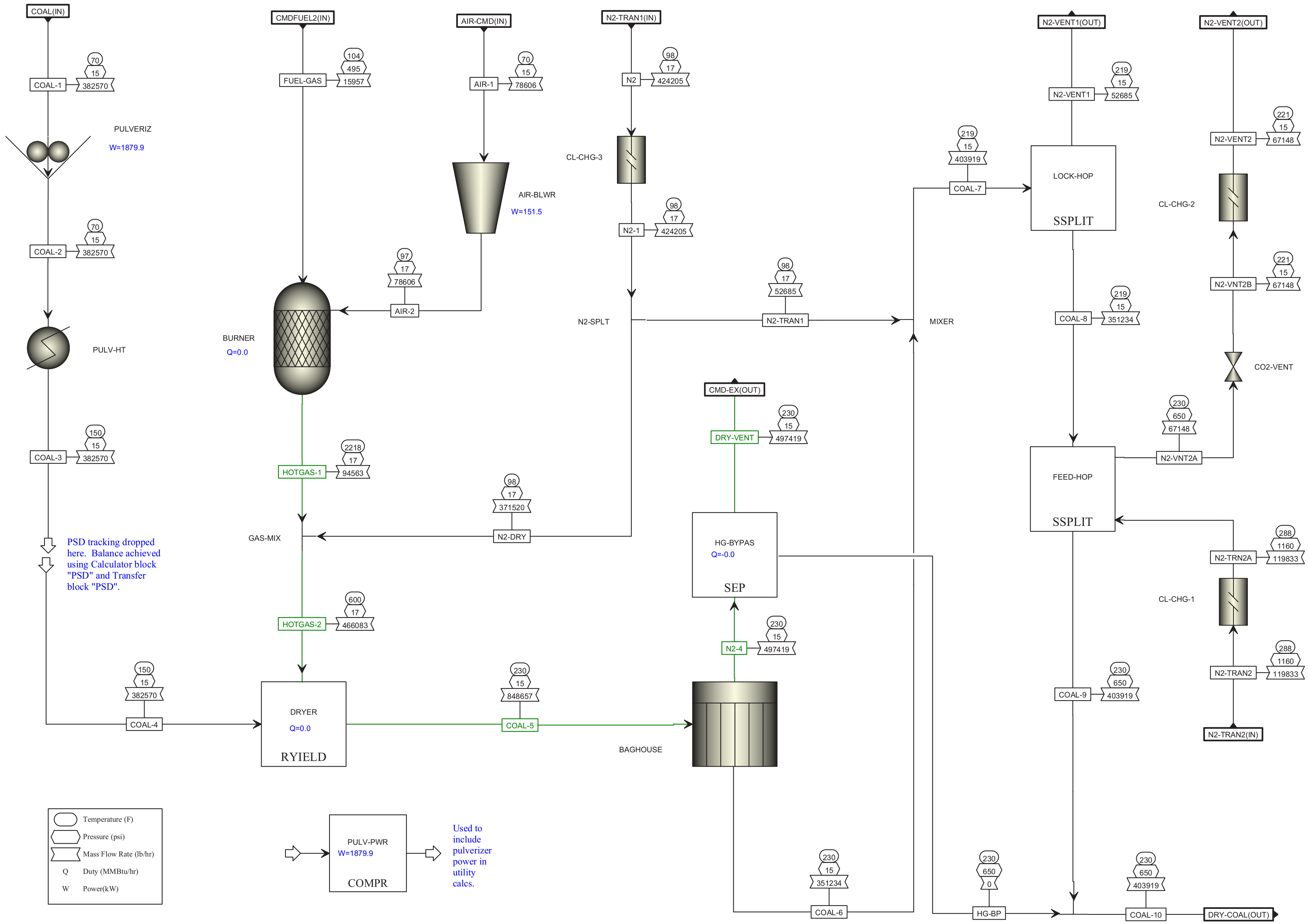
Coal to Ammonia and Derivatives - Dry-Fed Entrained Flow Gasifier
(Derivatives produced in the correct ratio to make UAN32)



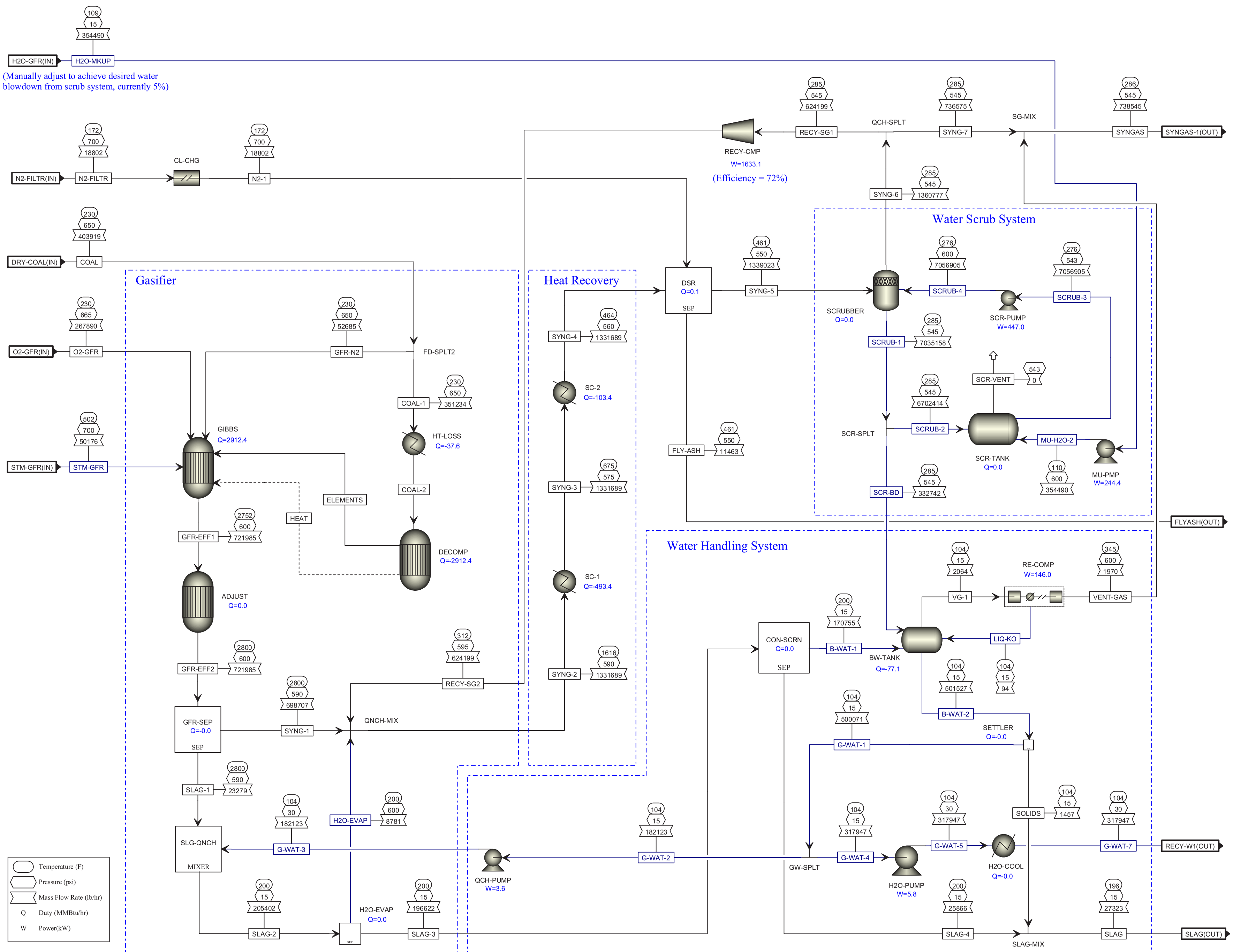
Air Separation Unit 99.5% O2 Purity



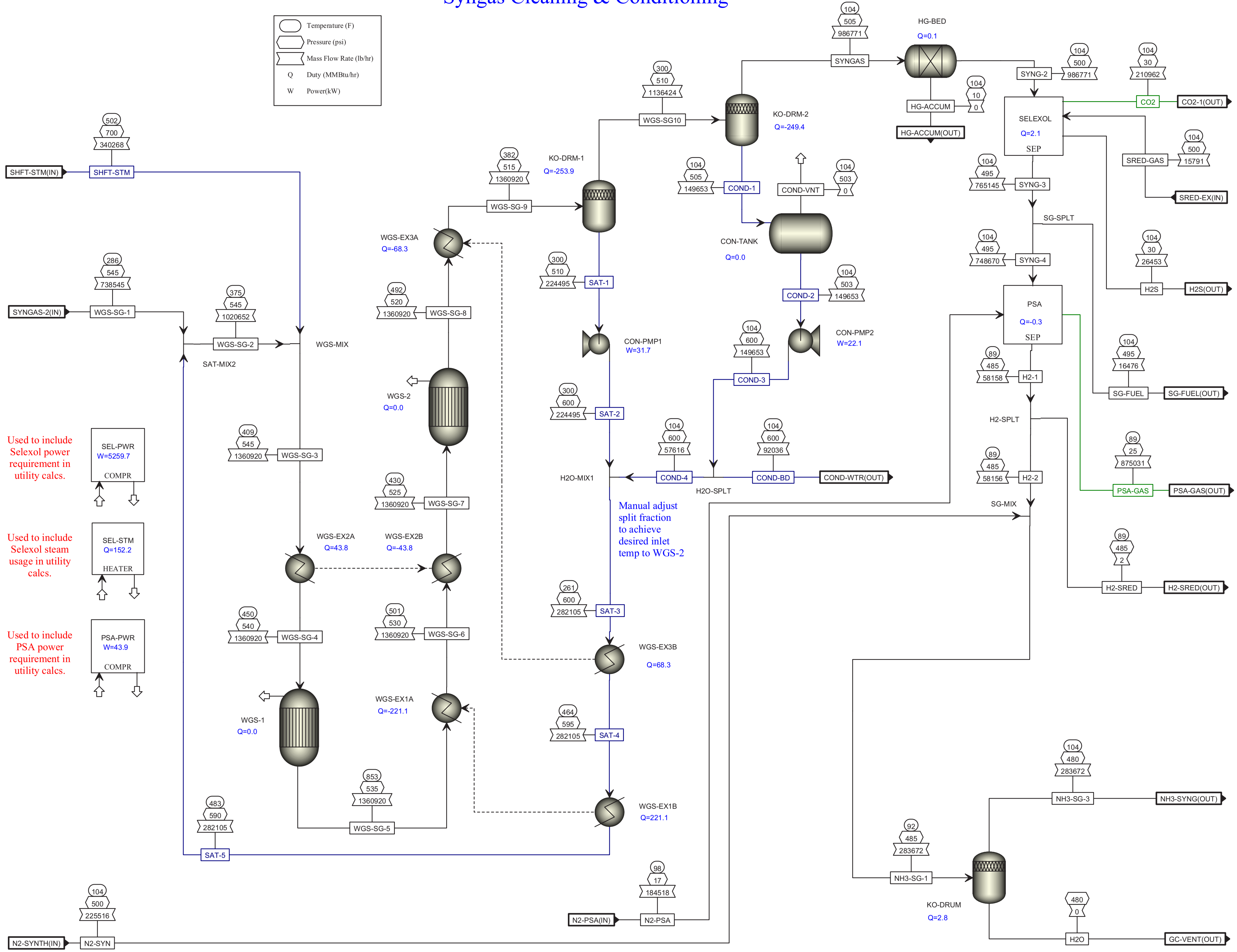
Coal Milling & Drying



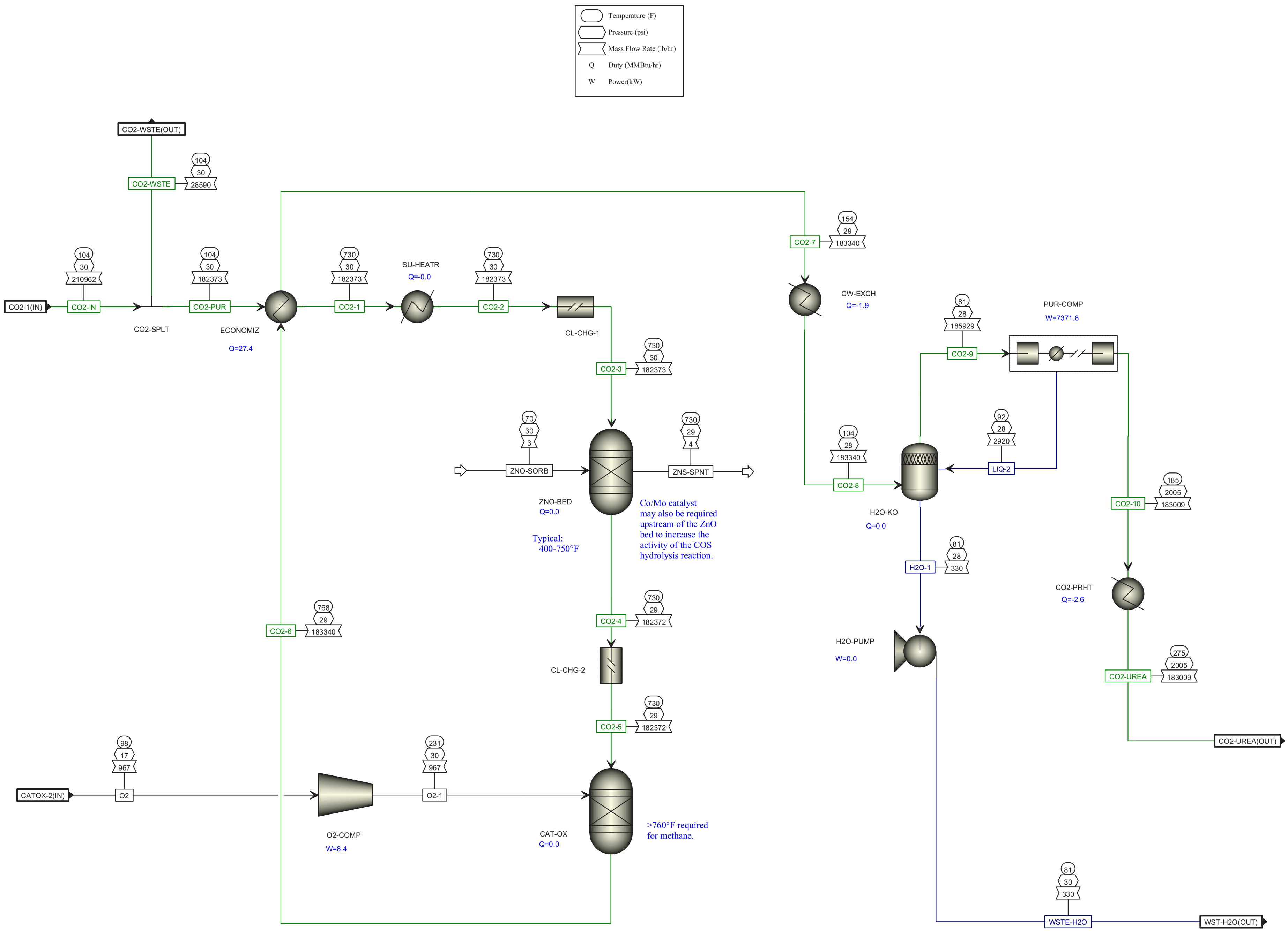
Dry-Fed Gasifier w/ Heat Recovery



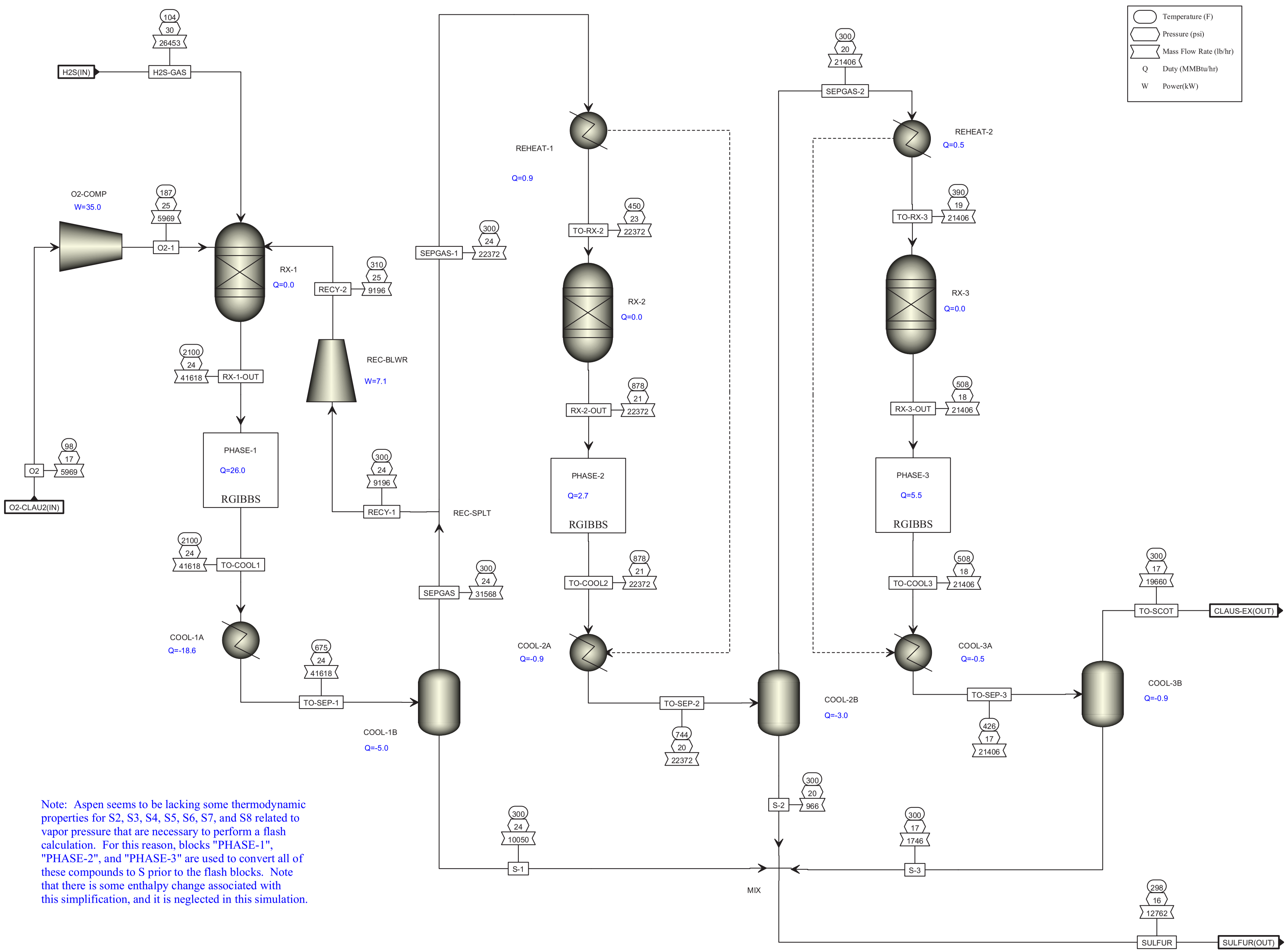
Syngas Cleaning & Conditioning



CO2 Purification

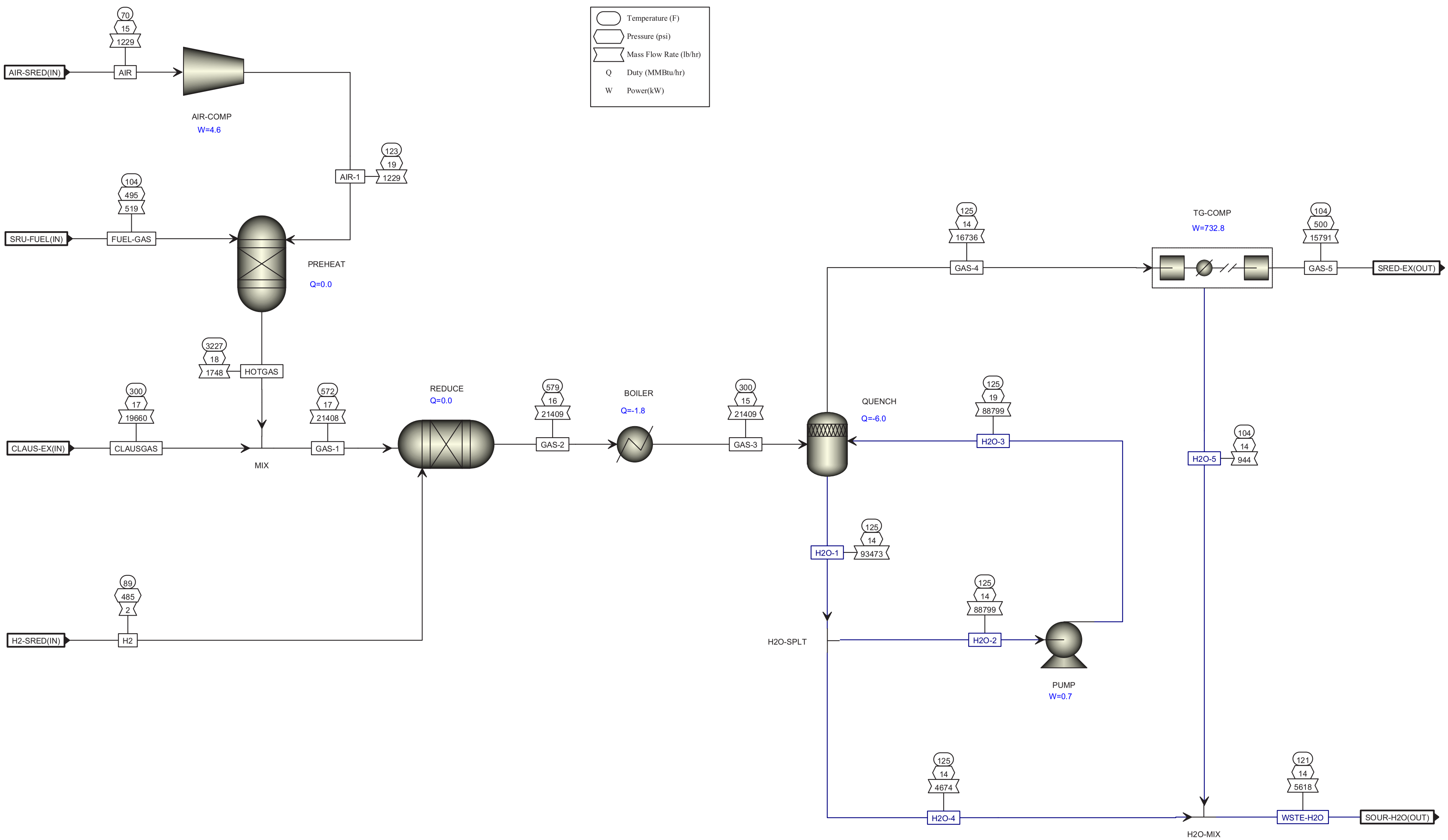


O2 Claus (COPE) Process

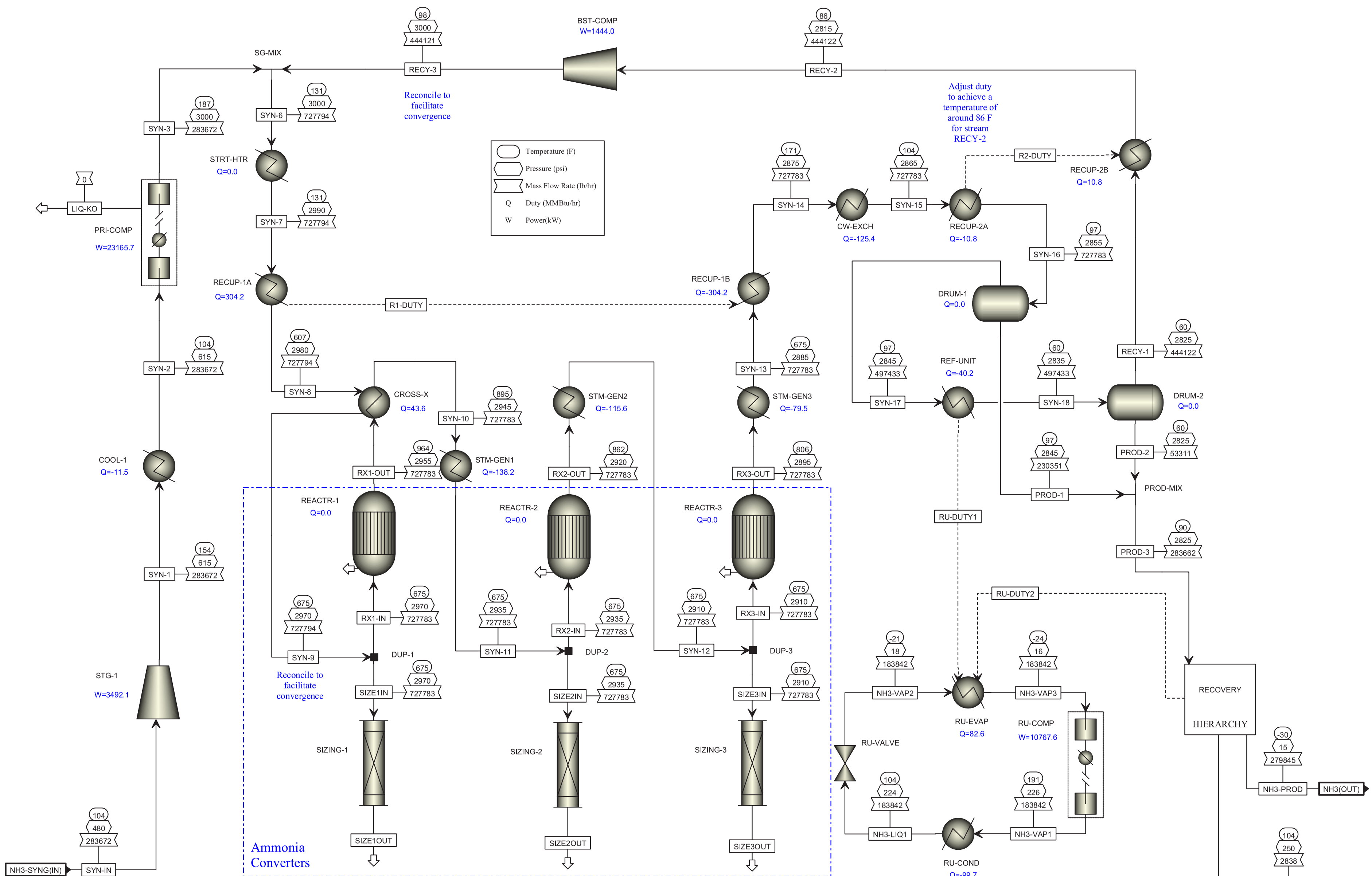


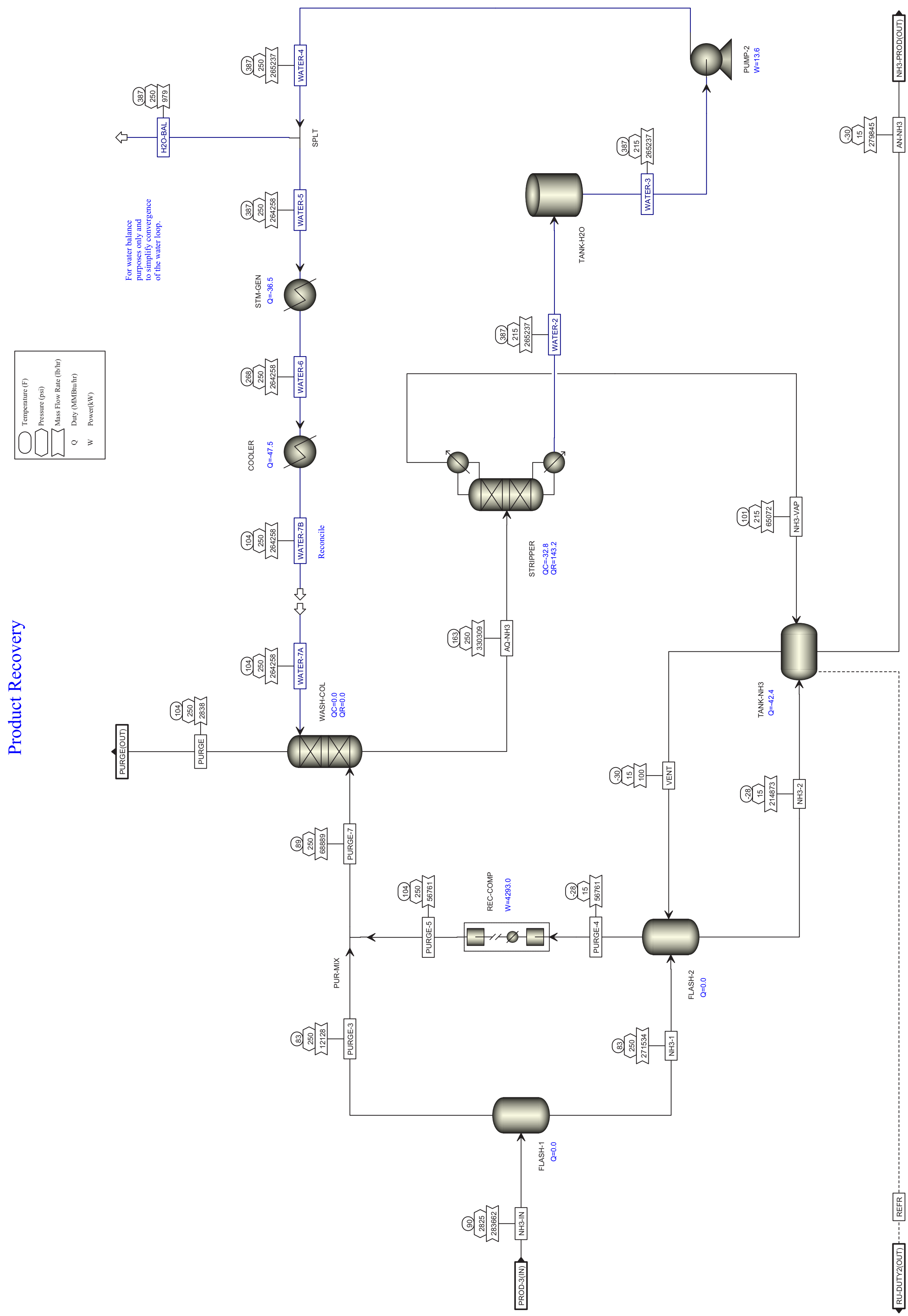
Note: Aspen seems to be lacking some thermodynamic properties for S2, S3, S4, S5, S6, S7, and S8 related to vapor pressure that are necessary to perform a flash calculation. For this reason, blocks "PHASE-1", "PHASE-2", and "PHASE-3" are used to convert all of these compounds to S prior to the flash blocks. Note that there is some enthalpy change associated with this simplification, and it is neglected in this simulation.

Catalytic Sulfur Reduction (SCOT or Beavon Process w/o H2S Absorber)

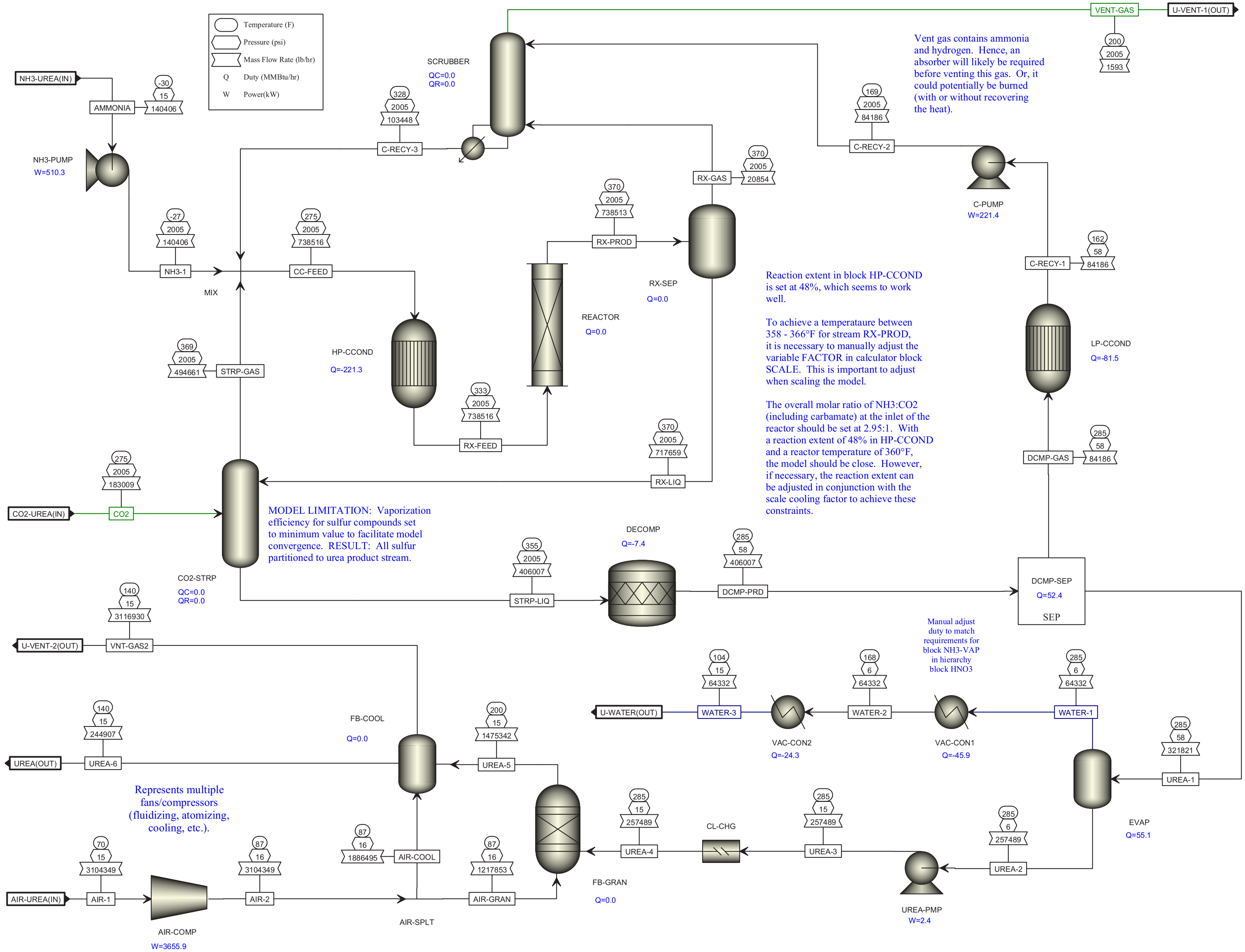


Ammonia Synthesis @ 3000 psi

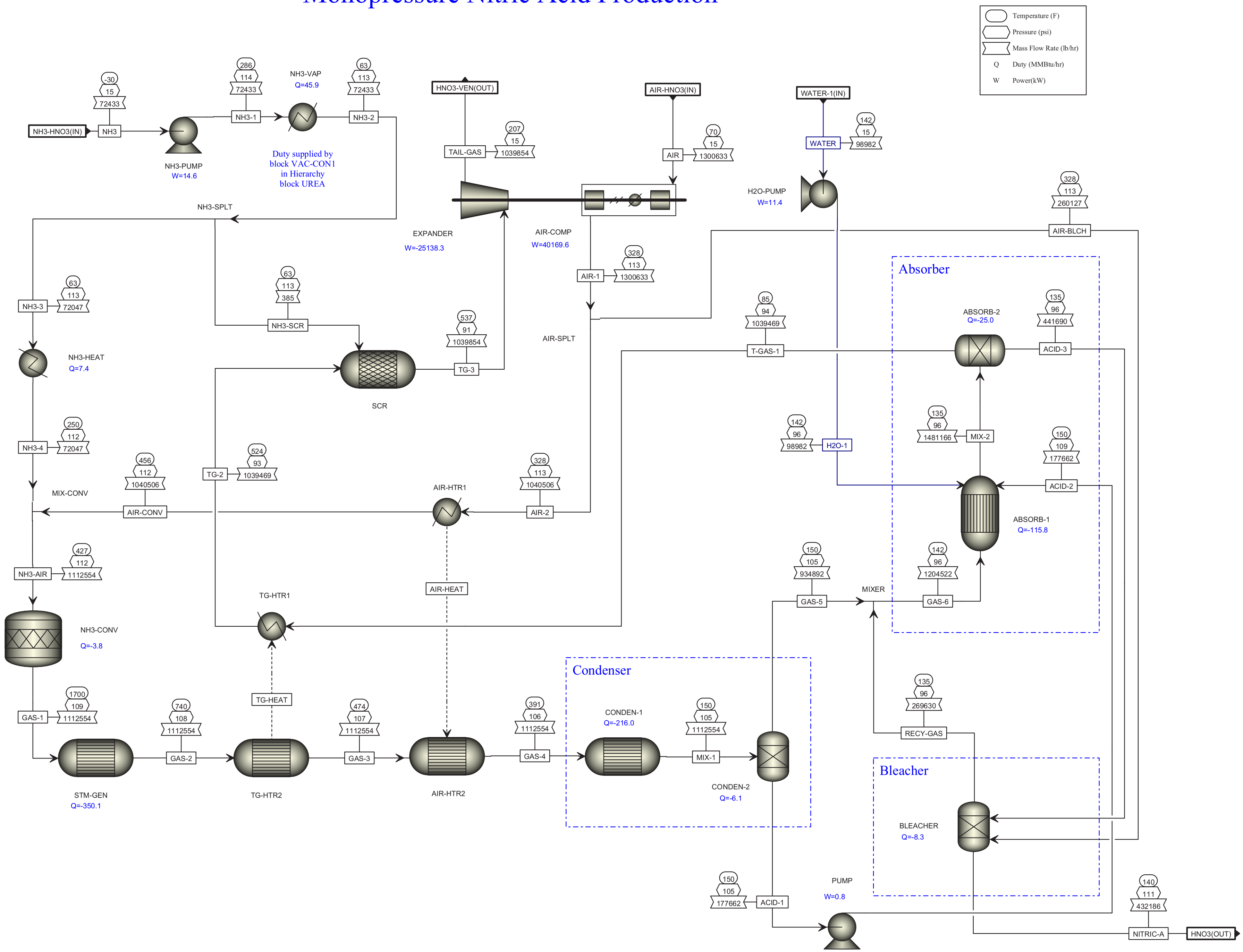



Product Recovery
Temperature (F)
Pressure (psi)
Mass Flow Rate (lb/hr)
Q Duty (MMBtu/hr)
W Power(kW)
For water balance purposes only and to simplify convergence of the water loop.
H2O-BAL 387 250 979
WATER-4 387 250 265237
SPLT
WATER-5 387 250 264258
STM-GEN Q=-36.5
WATER-6 268 250 264258
COOLER Q=-47.5
WATER-7B 104 250 264258
Reconcile
WATER-7A 104 250 264258
WASH-COL QC=0.0 QR=0.0
PURGE 104 250 2838
PURGE(OUT)
PURGE-7 89 250 68889
PUR-MIX
PURGE-5 104 250 56761
REC-COMP W=4293.0
PURGE-4 -28 15 56761
PURGE-3 83 250 12128
FLASH-1 Q=0.0
NH3-IN 90 2825 283662
PROD-3(IN)
NH3-1 83 250 271534
FLASH-2 Q=0.0
VENT -30 15 100
NH3-2 -28 15 214873
TANK-NH3 Q=-42.4
AQ-NH3 163 250 330309
STRIPPER QC=-32.8 QR=143.2
WATER-2 387 215 265237
TANK-H2O
WATER-3 387 215 265237
PUMP-2 W=13.6
NH3-VAP 101 215 65072
AN-NH3 -30 15 279845
NH3-PROD(OUT)
REFR
RU-DUTY2(OUT)

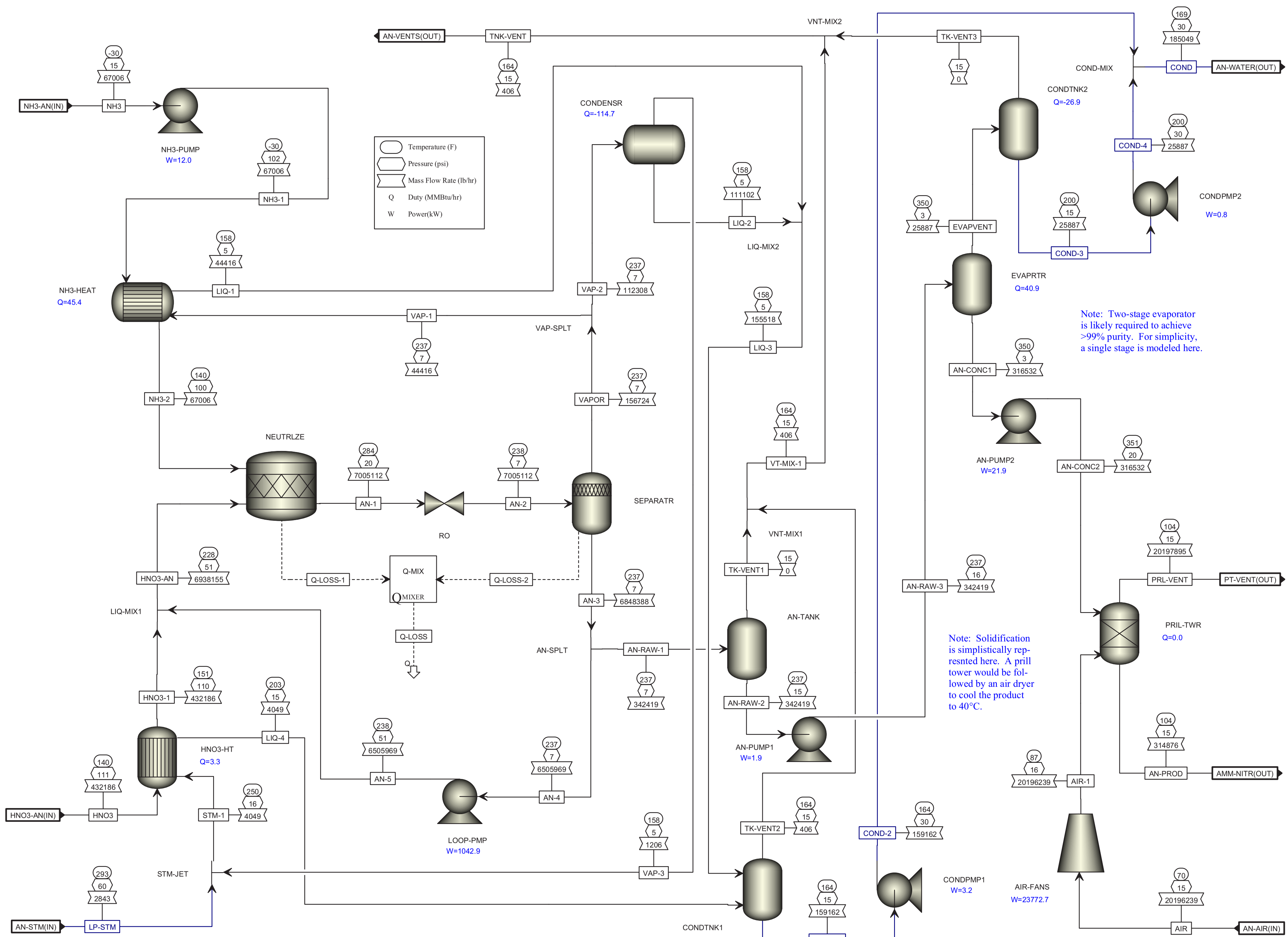
Urea Synthesis via the H.P. Stamicarbon CO2 Stripping Process

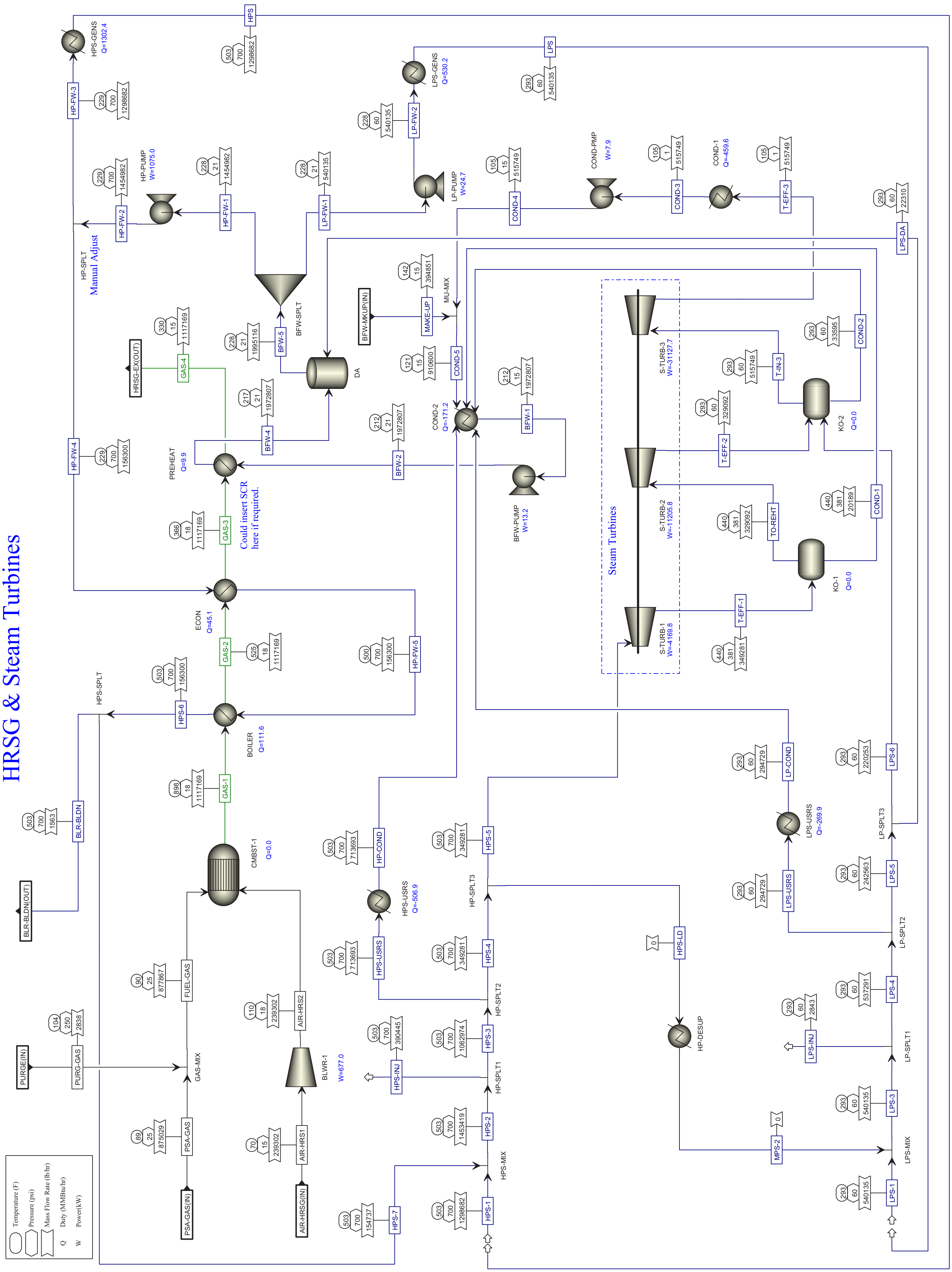


Monopressure Nitric Acid Production



Ammonium Nitrate Production (Uhde-I.G. Farbenindustrie Process)

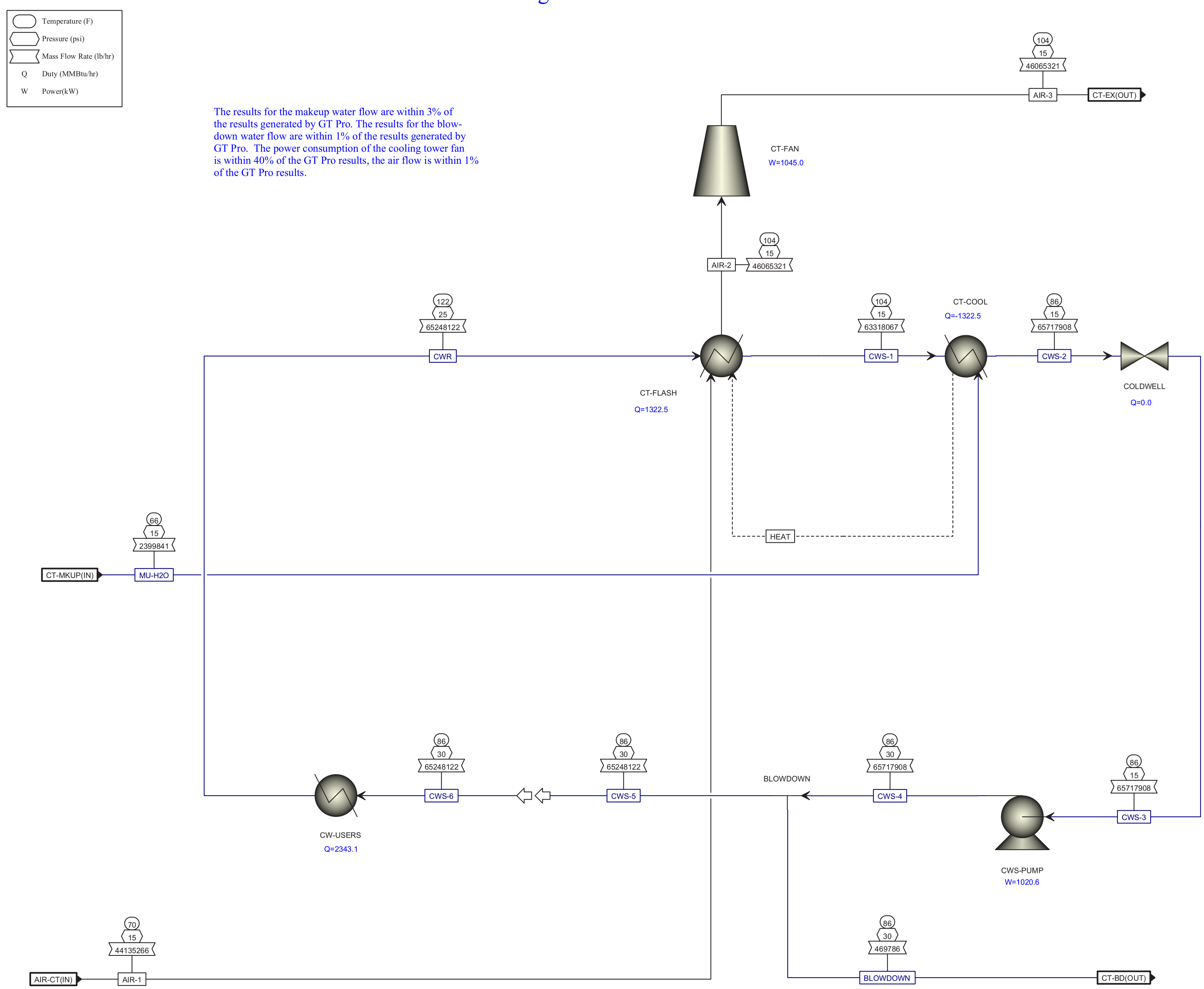




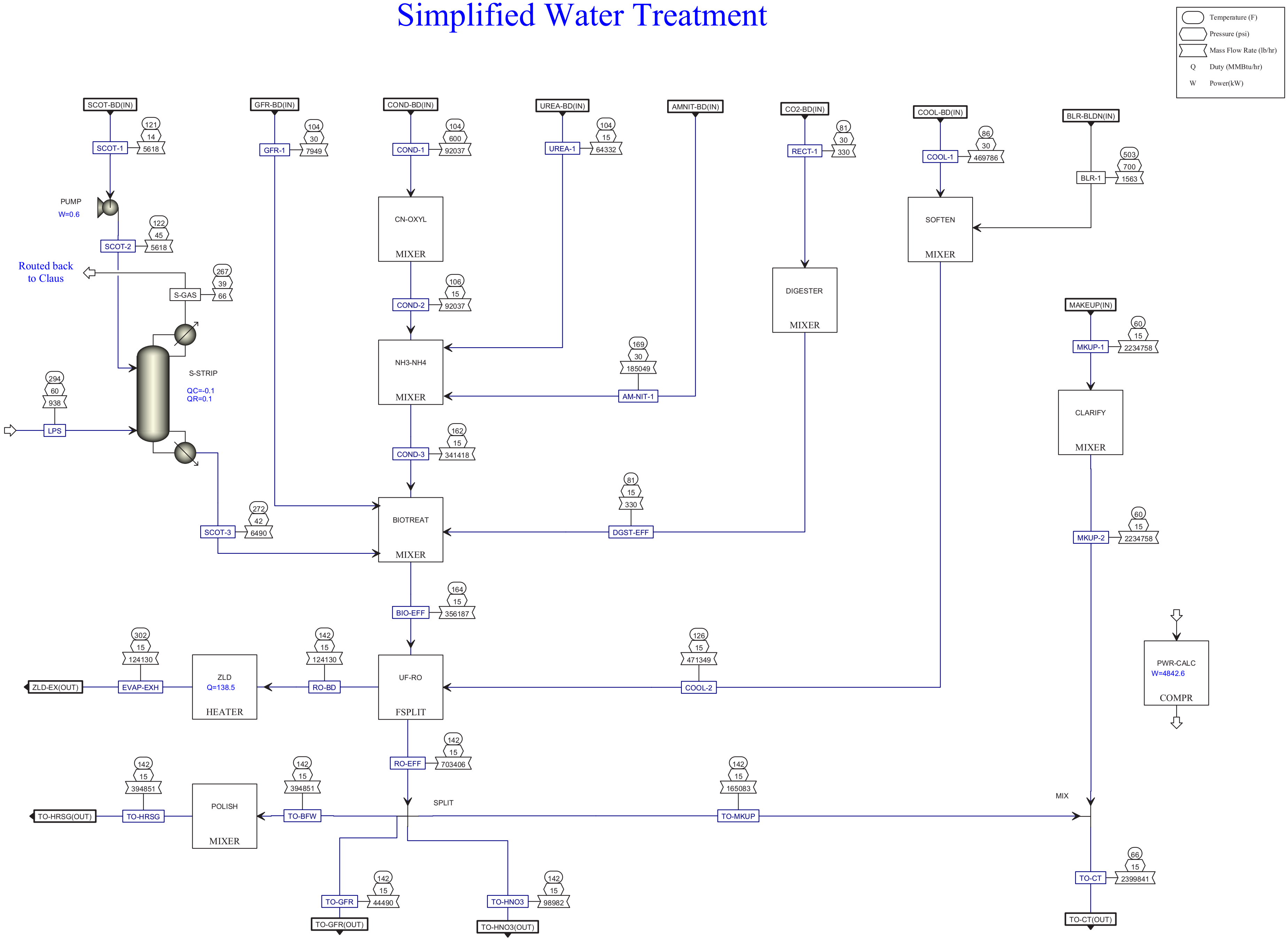
HRSG & Steam Turbines
Temperature (F)
Pressure (psi)
Mass Flow Rate (lb/hr)
Q Duty (MMBtu/hr)
W Power(kW)
PSA-GAS(IN)
PSA-GAS
PURGE(IN)
PURG-GAS
FUEL-GAS
GAS-MIX
AIR-HRSG(IN)
AIR-HRS1
BLWR-1
W=677.0
AIR-HRS2
CMBST-1
Q=0.0
GAS-1
BOILER
Q=111.6
GAS-2
ECON
Q=45.1
GAS-3
Could insert SCR here if required.
PREHEAT
Q=9.9
GAS-4
HRSG-EX(OUT)
BLR-BLDN(OUT)
BLR-BLDN
HPS-SPLT
HPS-6
HP-FW-5
HP-FW-4
HP-FW-3
HPS-GENS
Q=1302.4
HPS
HP-SPLT
Manual Adjust
HP-FW-2
HP-PUMP
W=1075.0
HP-FW-1
BFW-SPLT
BFW-5
DA
BFW-4
BFW-2
BFW-PUMP
W=13.2
BFW-1
COND-2
Q=-171.2
BFW-MKUP(IN)
MAKE-UP
MU-MIX
COND-5
LP-FW-1
LP-PUMP
W=24.7
LP-FW-2
LPS-GENS
Q=530.2
LPS
COND-4
COND-PMP
W=7.9
COND-3
COND-1
Q=-459.6
T-EFF-3
LPS-DA
Steam Turbines
S-TURB-1
W=-4169.8
S-TURB-2
W=-11205.8
S-TURB-3
W=-31127.7
T-EFF-1
KO-1
Q=0.0
TO-REHT
T-EFF-2
KO-2
Q=0.0
T-IN-3
COND-2
COND-1
HPS-7
HPS-1
HPS-MIX
HPS-2
HP-SPLT1
HPS-INJ
HPS-3
HP-SPLT2
HPS-USRS
HPS-USRS
Q=-506.9
HP-COND
HPS-4
HP-SPLT3
HPS-5
HPS-LD
HP-DESUP
MPS-2
LPS-1
LPS-MIX
LPS-3
LP-SPLT1
LPS-INJ
LPS-4
LP-SPLT2
LPS-USRS
LPS-USRS
Q=-269.9
LP-COND
LPS-5
LP-SPLT3
LPS-6

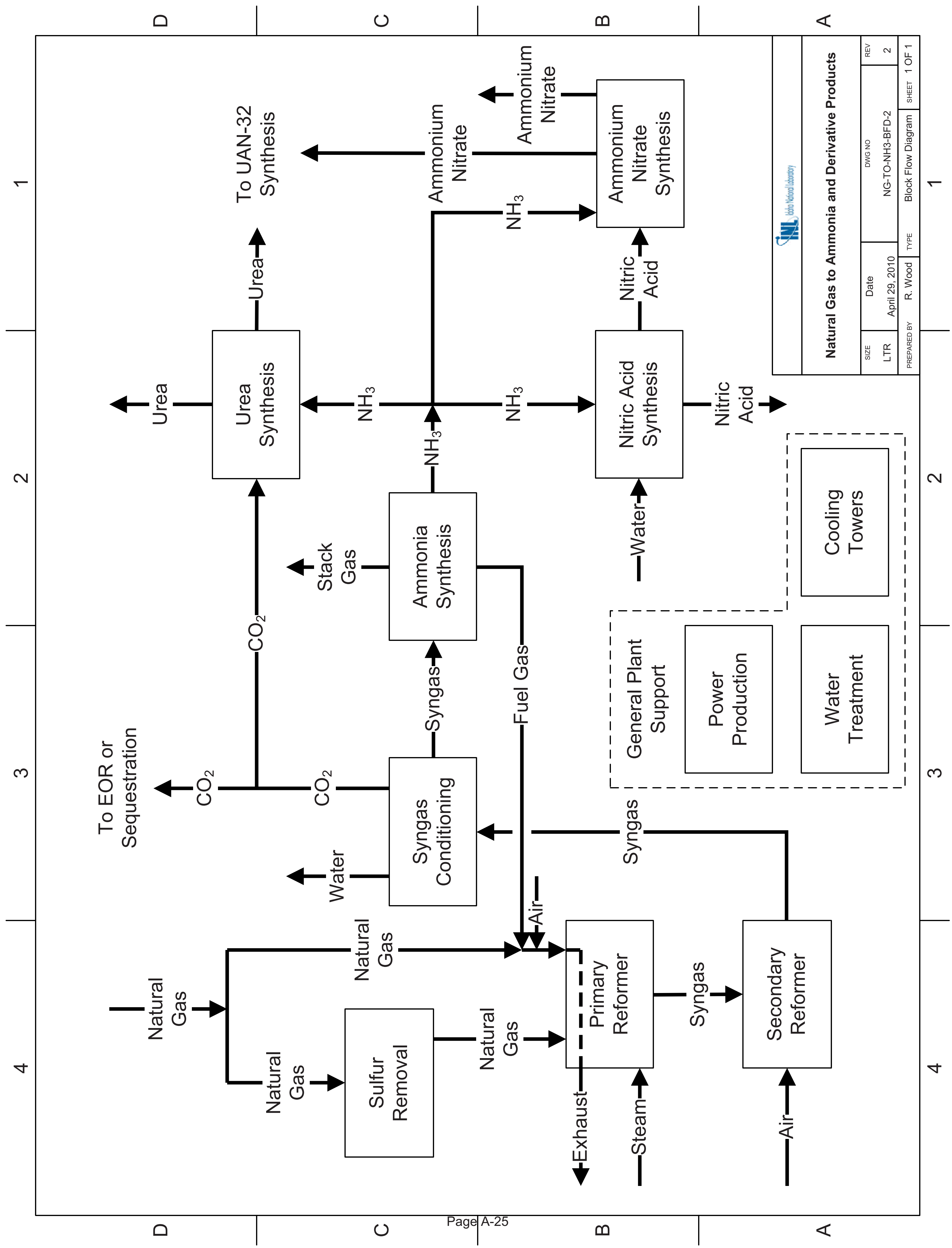
Cooling Tower

The results for the makeup water flow are within 3% of the results generated by GT Pro. The results for the blow-down water flow are within 1% of the results generated by GT Pro. The power consumption of the cooling tower fan is within 40% of the GT Pro results, the air flow is within 1% of the GT Pro results.

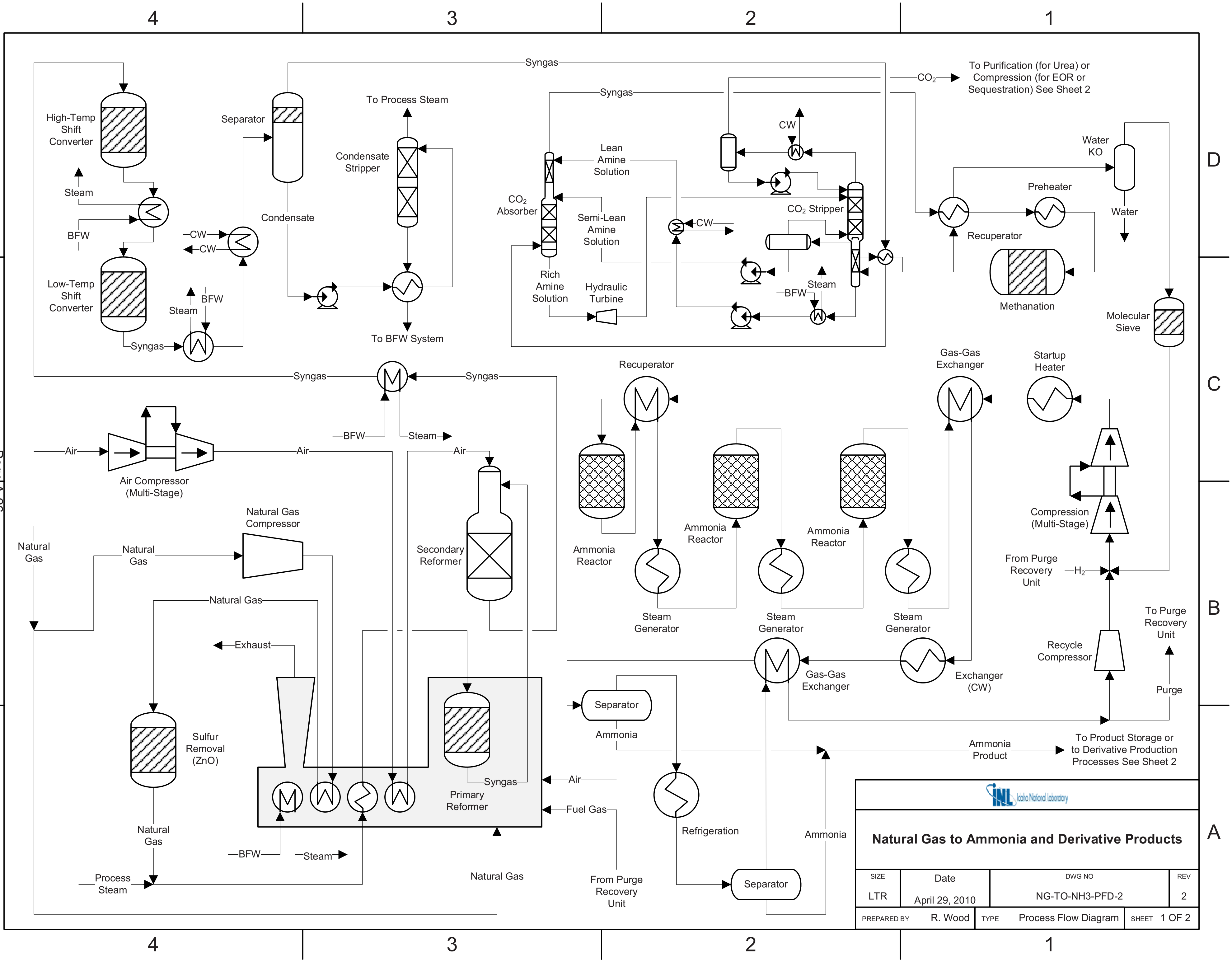


Simplified Water Treatment

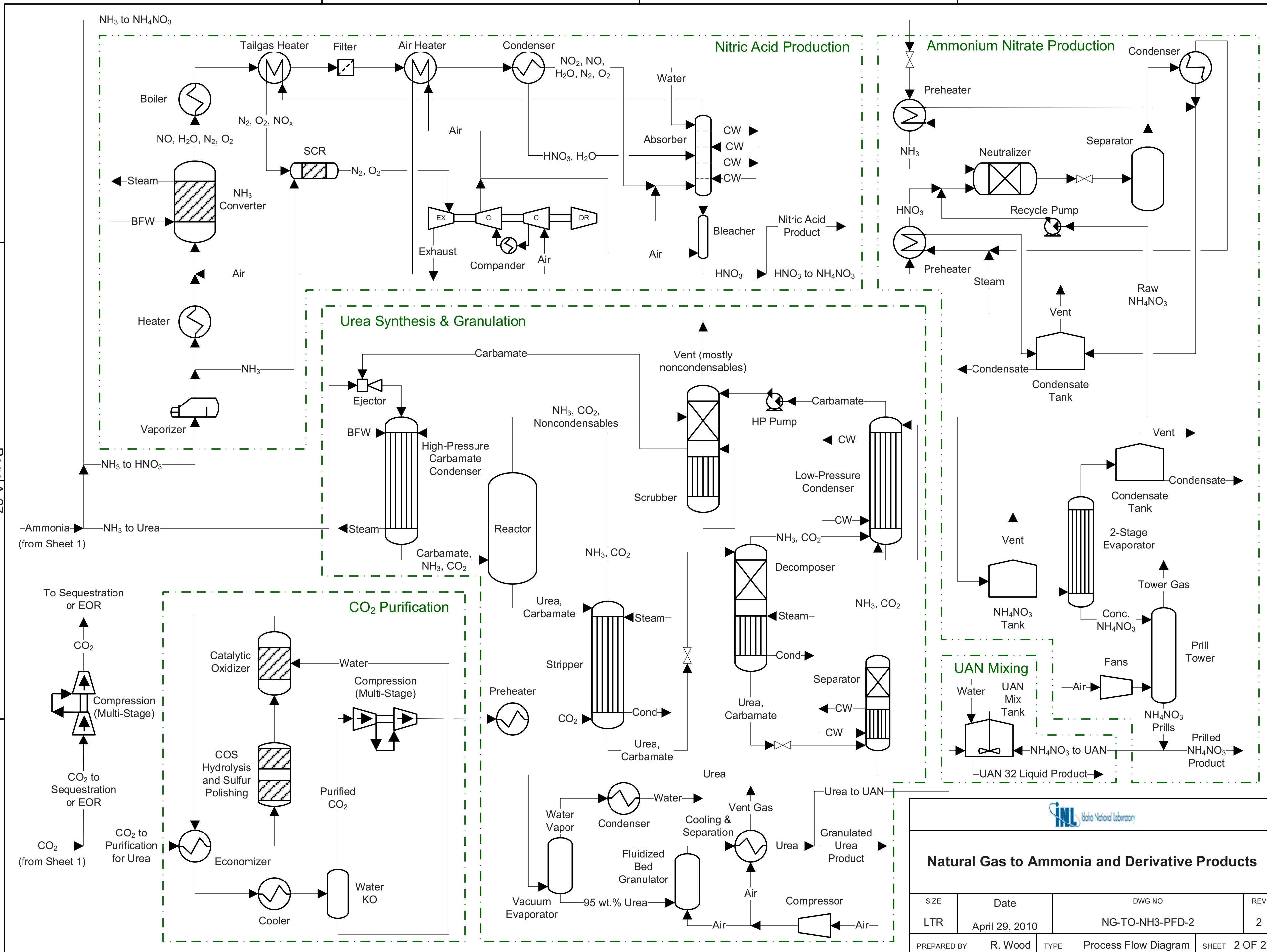




Natural Gas to Ammonia and Derivative Products
Idaho National Laboratory
SIZE LTR
Date April 29, 2010
DWG NO NG-TO-NH3-BFD-2
REV 2
PREPARED BY R. Wood
TYPE Block Flow Diagram
SHEET 1 OF 1
Natural Gas
Sulfur Removal
Primary Reformer
Secondary Reformer
Syngas
Air
Steam
Exhaust
Fuel Gas
Syngas Conditioning
Water
CO_2
To EOR or Sequestration
Ammonia Synthesis
Stack Gas
NH_3
Urea Synthesis
Urea
To UAN-32 Synthesis
Nitric Acid Synthesis
Nitric Acid
Ammonium Nitrate Synthesis
Ammonium Nitrate
General Plant Support
Power Production
Water Treatment
Cooling Towers
1
2
3
4
A
B
C
D



High-Temp Shift Converter
Low-Temp Shift Converter
Steam
BFW
Syngas
Separator
Condensate
CW
Condensate Stripper
To Process Steam
To BFW System
CO$_2$ Absorber
Lean Amine Solution
Semi-Lean Amine Solution
Rich Amine Solution
Hydraulic Turbine
CO$_2$ Stripper
CO$_2$
To Purification (for Urea) or Compression (for EOR or Sequestration) See Sheet 2
Recuperator
Preheater
Methanation
Water KO
Water
Molecular Sieve
Air
Air Compressor (Multi-Stage)
Natural Gas
Natural Gas Compressor
Secondary Reformer
Sulfur Removal (ZnO)
Exhaust
Primary Reformer
Fuel Gas
Process Steam
From Purge Recovery Unit
Ammonia Reactor
Steam Generator
Gas-Gas Exchanger
Startup Heater
Compression (Multi-Stage)
H_2
Recycle Compressor
To Purge Recovery Unit
Purge
Exchanger (CW)
Ammonia
Refrigeration
Ammonia Product
To Product Storage or to Derivative Production Processes See Sheet 2
Idaho National Laboratory
Natural Gas to Ammonia and Derivative Products
SIZE LTR
Date April 29, 2010
DWG NO NG-TO-NH3-PFD-2
REV 2
PREPARED BY R. Wood
TYPE Process Flow Diagram
SHEET 1 OF 2


Nitric Acid Production
Ammonium Nitrate Production
Urea Synthesis & Granulation
CO_2 Purification
UAN Mixing
NH_3 to NH_4NO_3
Tailgas Heater
Filter
Air Heater
Condenser
NO_2, NO, H_2O, N_2, O_2
Water
Absorber
CW
Boiler
N_2, O_2, NO_x
NO, H_2O, N_2, O_2
SCR
N_2, O_2
Air
HNO_3, H_2O
Steam
NH_3 Converter
BFW
EX
C
DR
Exhaust
Compander
Bleacher
Nitric Acid Product
HNO_3
HNO_3 to NH_4NO_3
Heater
NH_3
Vaporizer
NH_3 to HNO_3
Ammonia (from Sheet 1)
NH_3 to Urea
Preheater
Neutralizer
Separator
Recycle Pump
Raw NH_4NO_3
Vent
Condensate
Condensate Tank
2-Stage Evaporator
NH_4NO_3 Tank
Conc. NH_4NO_3
Tower Gas
Prill Tower
Fans
NH_4NO_3 Prills
Prilled NH_4NO_3 Product
UAN Mix Tank
NH_4NO_3 to UAN
UAN 32 Liquid Product
Carbamate
Ejector
Vent (mostly noncondensables)
HP Pump
NH_3, CO_2, Noncondensables
High-Pressure Carbamate Condenser
Reactor
Scrubber
Low-Pressure Condenser
Carbamate, NH_3, CO_2
NH_3, CO_2
Decomposer
Cond
Urea, Carbamate
Stripper
Separator
Urea
Urea to UAN
To Sequestration or EOR
CO_2
Compression (Multi-Stage)
CO_2 to Sequestration or EOR
CO_2 (from Sheet 1)
CO_2 to Purification for Urea
Catalytic Oxidizer
COS Hydrolysis and Sulfur Polishing
Purified CO_2
Economizer
Cooler
Water KO
Water Vapor
Vacuum Evaporator
95 wt.% Urea
Fluidized Bed Granulator
Cooling & Separation
Vent Gas
Granulated Urea Product
Compressor
Idaho National Laboratory
Natural Gas to Ammonia and Derivative Products
SIZE LTR
Date April 29, 2010
DWG NO NG-TO-NH3-PFD-2
REV 2
PREPARED BY R. Wood
TYPE Process Flow Diagram
SHEET 2 OF 2

Conventional Gas-to-Ammonia Results

```
Calculator Block SUMMARY

   FEED SUMMARY:

     NATURAL GAS PROPERTIES:

       MASS FLOW =                         1989. TON/DY
       VOLUME FLOW =                         88. MMSCFD @ 60°F
       HHV =                              23063. BTU/LB
       HHV =                               1044. BTU/SCF @ 60°F
       ENERGY FLOW =                      91733. MMBTU/DY

       COMPOSITION:
          METHANE =                          93.571 MOL.%
          ETHANE =                            3.749 MOL.%
          PROPANE =                           0.920 MOL.%
          BUTANE =                            0.260 MOL.%
          PENTANE =                           0.040 MOL.%
          HEXANE =                            0.010 MOL.%
          NITROGEN =                          1.190 MOL.%
          OXYGEN =                            0.010 MOL.%
          CO2 =                               0.250 MOL.%
          C4H10S =                            1. PPMV
          C2H6S =                             0. PPMV
          H2S =                               0. PPMV

   INTERMEDIATE PRODUCT SUMMARY:

     RAW SYNGAS MASS FLOW =             903998. LB/HR
     RAW SYNGAS VOLUME FLOW =              520. MMSCFD @ 60°F
     RAW SYNGAS COMPOSITION:
       H2                                    36.3 MOL.%
       CO                                     8.2 MOL.%
       CO2                                    5.1 MOL.%
       N2                                    14.7 MOL.%
       AR                                     0.2 MOL.%
       H2O                                   35.0 MOL.%
       CH4                                    0.5 MOL.%

     CLEANED SYNGAS MASS FLOW =         294757. LB/HR
     CLEANED SYNGAS VOLUME FLOW =          308. MMSCFD @ 60°F
     CLEANED SYNGAS COMPOSITION:
       H2                                    73.8 MOL.%
       N2                                    24.8 MOL.%
       AR                                     0.3 MOL.%
       CH4                                    1.1 MOL.%
       CO                                     0. PPMV
       CO2                                    0. PPMV
       H2O                                    1. PPMV

     AMMONIA MASS FLOW =                280047. LB/HR
     AMMONIA MASS FLOW =                  3361. TON/DY
       AMMONIA PURITY =                      99.99 MOL.%
     AMMONIA PRODUCED / NAT. GAS FED =        1.69 LB/LB

     NITRIC ACID MASS FLOW =            432702. LB/HR
     NITRIC ACID MASS FLOW =              5192. TON/DY
       NITRIC ACID PURITY =                  57.93 WT.%
     HNO3 PRODUCED / NH3 FED =                1.55 LB/LB

   FINAL PRODUCT SUMMARY:

     AMMONIA MASS FLOW =                      0. LB/HR
     AMMONIA MASS FLOW =                      0. TON/DY
```

Conventional Gas-to-Ammonia Results

```
    AMMONIA PURITY =                         99.99 MOL.%

  NITRIC ACID MASS FLOW =                     0. LB/HR
  NITRIC ACID MASS FLOW =                     0. TON/DY
    NITRIC ACID PURITY =                     57.93 WT.%

  UREA MASS FLOW =                       244925. LB/HR
  UREA MASS FLOW =                         2939. TON/DY
  UREA PRODUCED / AMMONIA FED =              1.74 LB/LB
  UREA PRODUCED / CO2 FED =                  1.34 LB/LB

  AMMONIUM NITRATE MASS FLOW =           314950. LB/HR
  AMMONIUM NITRATE MASS FLOW =             3779. TON/DY
  NH4NO3 PRODUCED / NITRIC ACID FED =        0.73 LB/LB
  NH4NO3 PRODUCED / AMMONIA FED =            4.69 LB/LB

  STEAM EXPORT AVAILABLE:
    SATURATED STEAM FLOW =               376229. LB/HR
      PRESSURE =                             30. PSIA


POWER SUMMARY:

  ELECTRICAL GENERATORS:
    STEAM TURBINE POWER GENERATION =      28.9 MW
  GENERATOR SUBTOTAL =                    28.9 MW

  ELECTRICAL CONSUMERS:
    NG REFORMER POWER CONSUMPTION =       21.6 MW
    GAS CLEANING POWER CONSUMPTION =       4.4 MW
    POWER BLOCK POWER CONSUMPTION =        2.2 MW
    CO2 PROCESSING POWER CONSUMPTION =    13.0 MW
    AMMONIA SYNTH. POWER CONSUMPTION =    45.2 MW
    UREA SYNTHESIS POWER CONSUMPTION =     4.4 MW
    HNO3 SYNTHESIS POWER CONSUMPTION =    15.1 MW
    NH4NO3 SYNTH. POWER CONSUMPTION =     24.9 MW
    COOLING TOWER POWER CONSUMPTION =      1.5 MW
    WATER TREATMENT POWER CONSUMPTION =    5.4 MW
  CONSUMER SUBTOTAL =                    137.7 MW

  NET PLANT POWER CONSUMPTION =          108.8 MW


WATER BALANCE:

  EVAPORATIVE LOSSES:
    COOLING TOWER EVAPORATION =         3270.0 GPM
    ZLD SYSTEM EVAPORATION =             330.5 GPM
  TOTAL EVAPORATIVE LOSSES =            3600.6 GPM

  WATER CONSUMED:
    BOILER FEED WATER MAKEUP =          1674.8 GPM
    AMMONIA WASH WATER =                   0.0 GPM
    NITRIC ACID FEED WATER =             198.0 GPM
    COOLING TOWER MAKEUP =              3441.9 GPM
  TOTAL WATER CONSUMED =                5314.8 GPM

  WATER GENERATED:
    NG REFORMER BLOWDOWN =               542.3 GPM
    UREA PRODUCTION BLOWDOWN =           128.6 GPM
    NH4NO3 PRODUCTION BLOWDOWN =         370.2 GPM
    COOLING TOWER BLOWDOWN =             676.8 GPM
  TOTAL WATER GENERATED =               1717.8 GPM
```

Conventional Gas-to-Ammonia Results

PLANT WATER SUMMARY:

NET MAKEUP WATER REQUIRED =	3927.6 GPM
WATER CONSUMED / NG FED =	11.86 LB/LB

CO_2 BALANCE:

CO_2 INCORPORATED INTO UREA PRODUCT =	2177. TON/DY
CO_2 INCORPORATED INTO UREA PRODUCT =	38. MMSCFD @ 60°F
CO_2 PURITY =	99.4 %
CO_2 CAPTURED (SEQ. OR EOR) =	1762. TON/DY
CO_2 CAPTURED (SEQ. OR EOR) =	30. MMSCFD @ 60°F
CO_2 PURITY =	99.4 %
CO_2 EMITTED (TOTAL) =	1449. TON/DY
CO_2 EMITTED (TOTAL) =	25. MMSCFD @ 60°F
FROM REFORMER =	1421. TON/DY
FROM UREA VENTS =	28. TON/DY
CO_2 EMITTED / NG FED =	0.73 LB/LB

Calculator Block AIRPROPS

HUMIDITY DATA FOR STREAM PRI-AIR:

HUMIDITY RATIO =	43.5 GRAINS/LB
RELATIVE HUMIDITY =	39.9 %

Calculator Block NG-RFMR Hierarchy: REFORMER

SULFUR REMOVAL CONDITIONS:

INLET BED TEMPERATURE =	662. °F

PRIMARY REFORMER CONDITIONS:

INLET TEMPERATURE =	1000. °F
STEAM TO CARBON MOLAR RATIO =	3.30
NATURAL GAS BURNED FOR HEAT =	22.70 %
OUTLET TEMPERATURE =	1454. °F
METHANE CONVERSION =	53.3 %

SECONDARY REFORMER CONDITIONS:

INLET GAS TEMPERATURE =	1454. °F
INLET AIR TEMPERATURE =	1000. °F
STEAM TO CARBON MOLAR RATIO =	2.40
OXYGEN TO CARBON MOLAR RATIO =	0.28
OUTLET TEMPERATURE =	1750. °F
H_2/CO =	4.40

OUTLET COMPOSITION:

H_2	36.2844 MOL.%
CO	8.2398 MOL.%
CO_2	5.0602 MOL.%
H_2O	35.0414 MOL.%
CH_4	0.4807 MOL.%
N_2	14.6614 MOL.%

Calculator Block PH Hierarchy: NH_4NO_3

TARGET NEUTRALIZER PH = 1.500

```
   ACTUAL NEUTRALIZER PH =   1.499


Calculator Block RATIO    Hierarchy: UREA

   TARGET REACTOR NH3/CO2 RATIO =   2.950
   ACTUAL REACTOR NH3/CO2 RATIO =   2.951
```

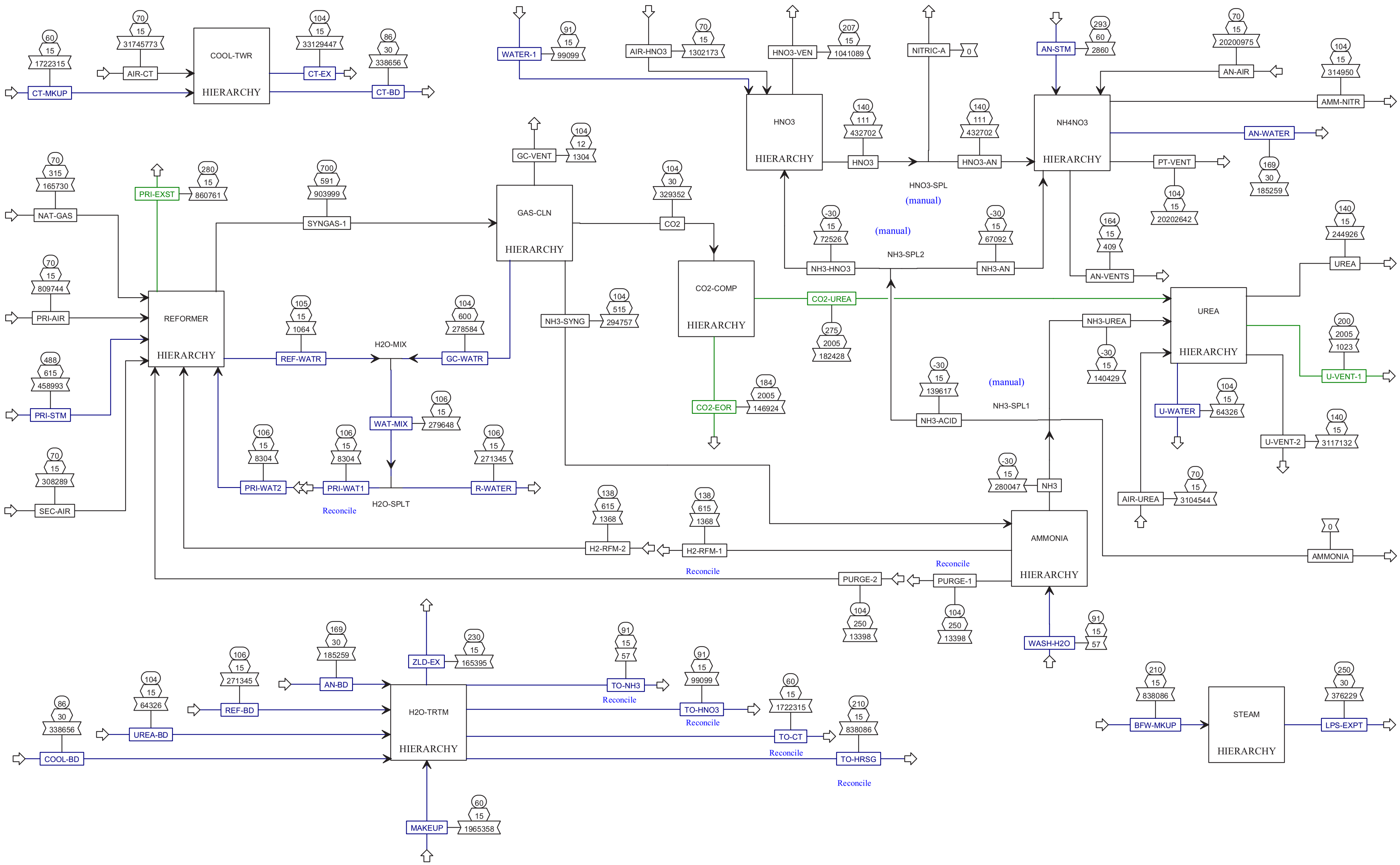
Natural Gas to Ammonia and Derivatives Baseline
(Derivatives produced in the correct ratio to make UAN32)

Temperature (F)
Pressure (psi)
Mass Flow Rate (lb/hr)

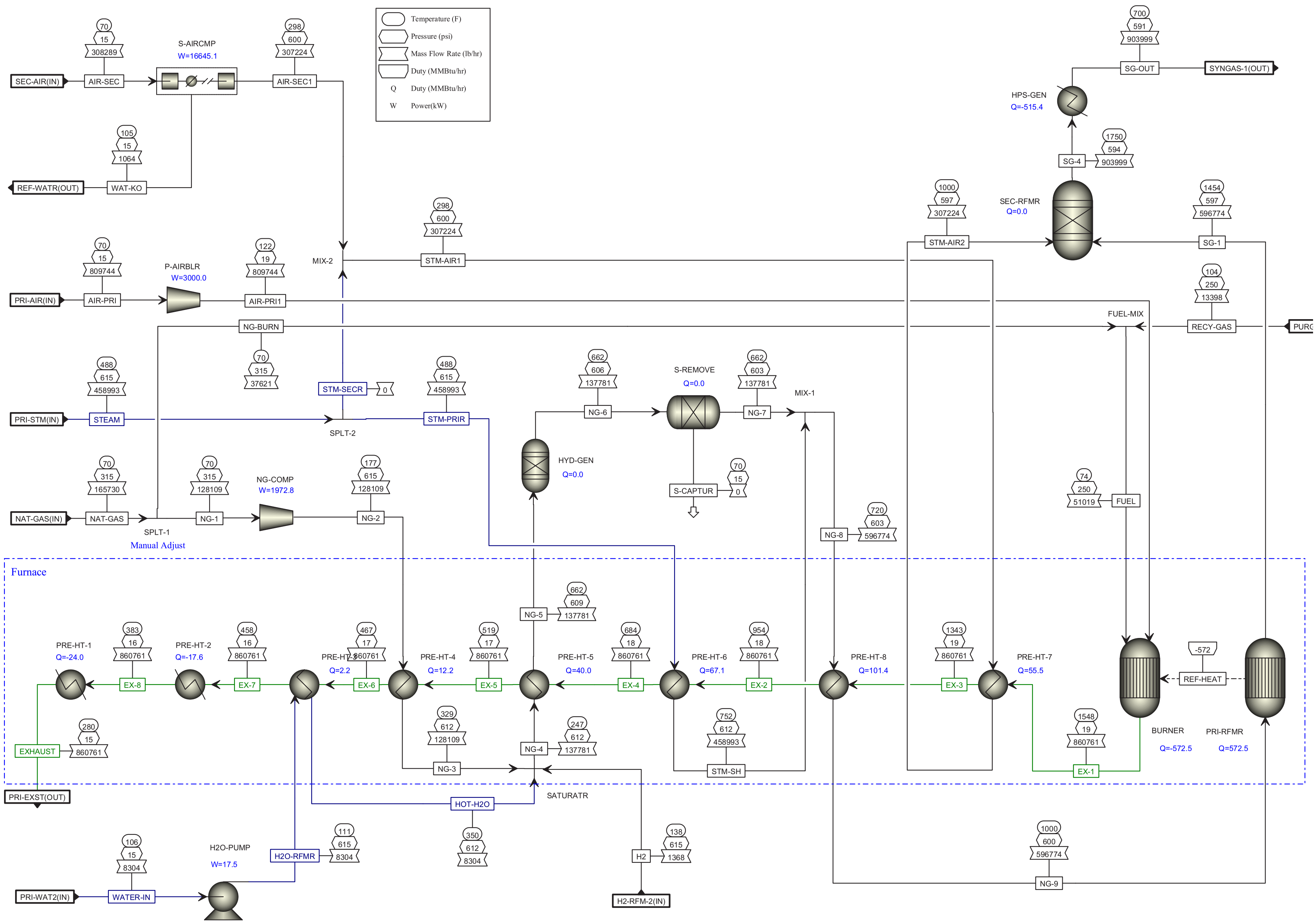
Color Legend

Water or Steam

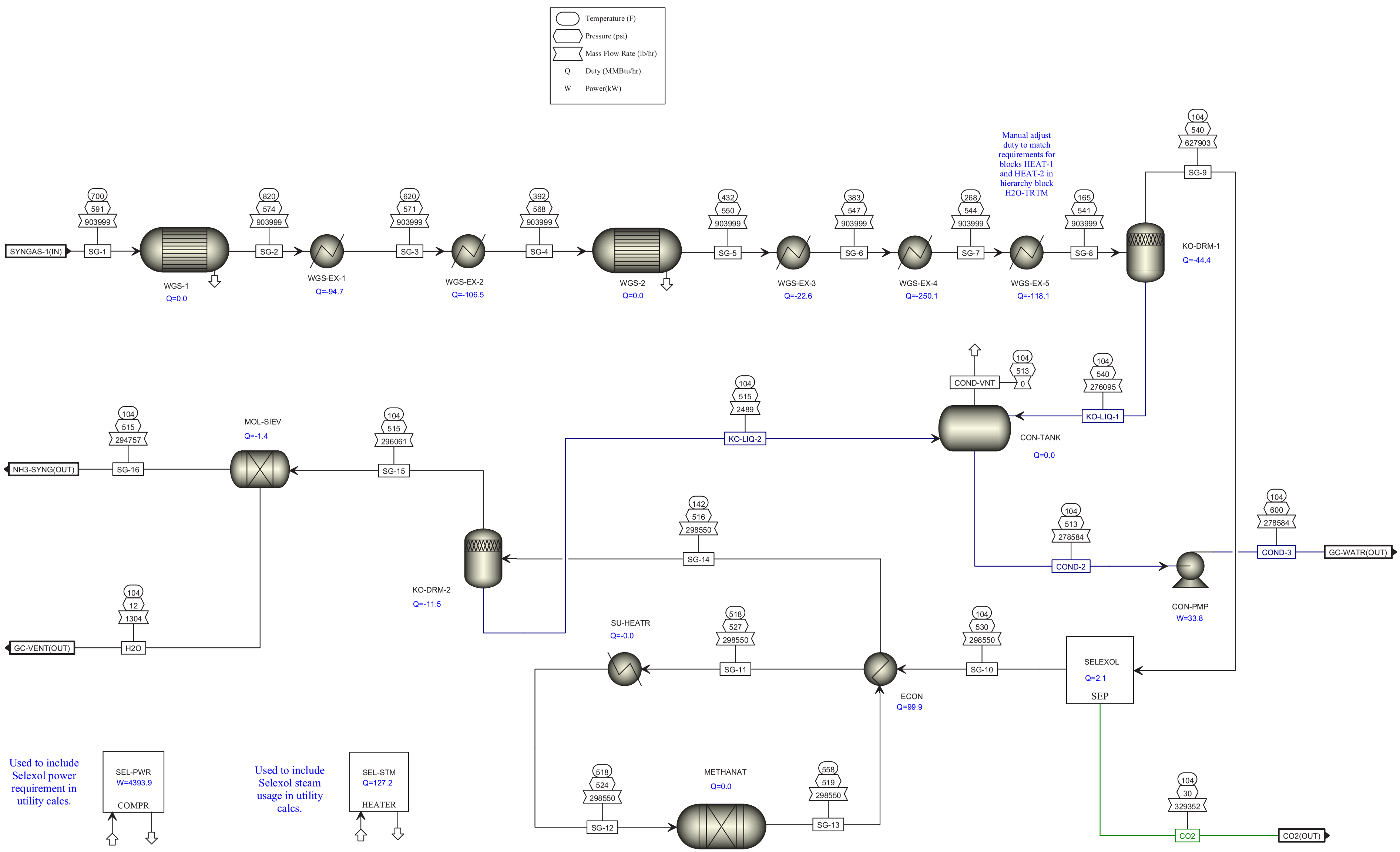
CO2 Emissions Source



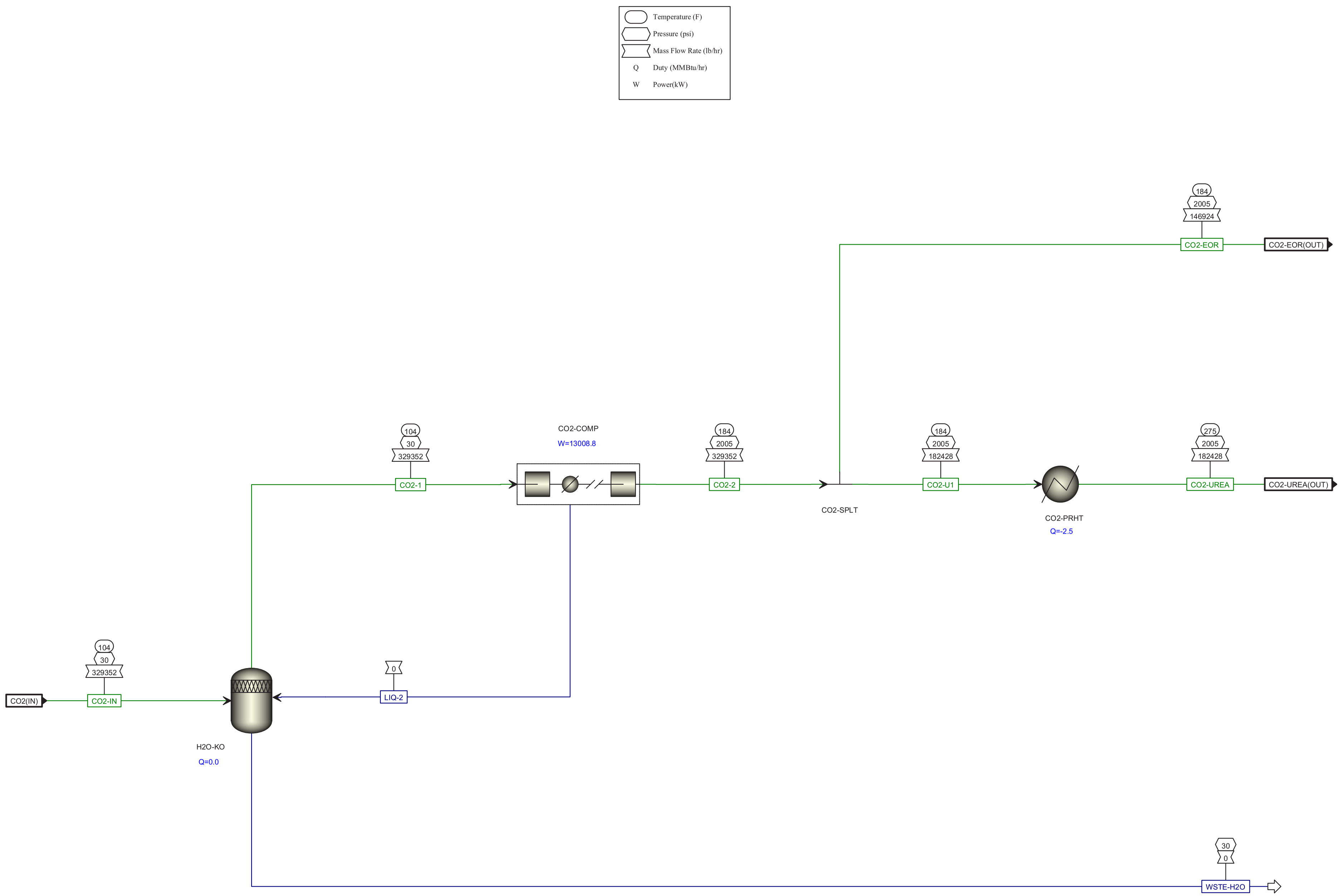
Natural Gas Two-Step Reforming



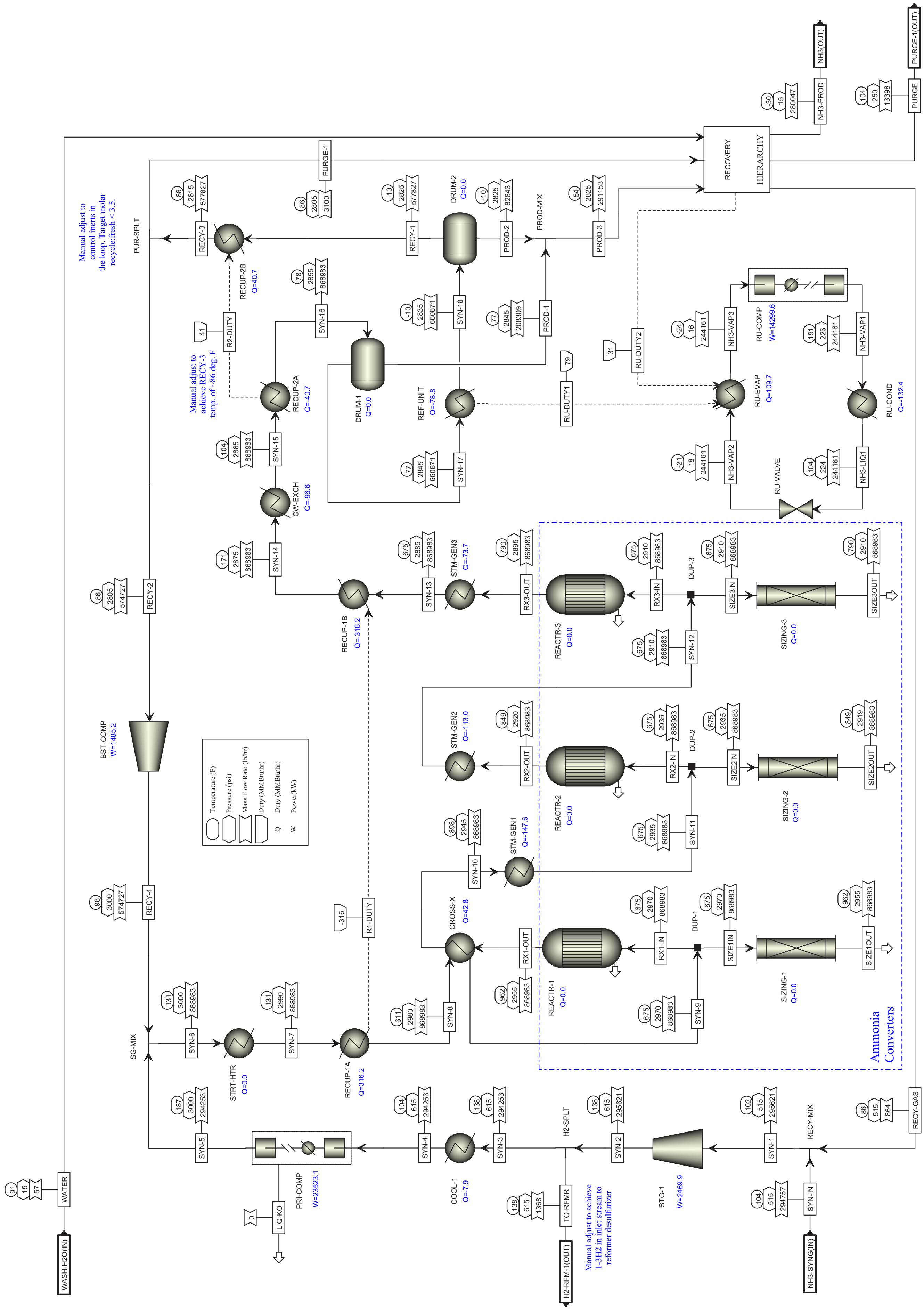
Syngas Cleaning & Conditioning



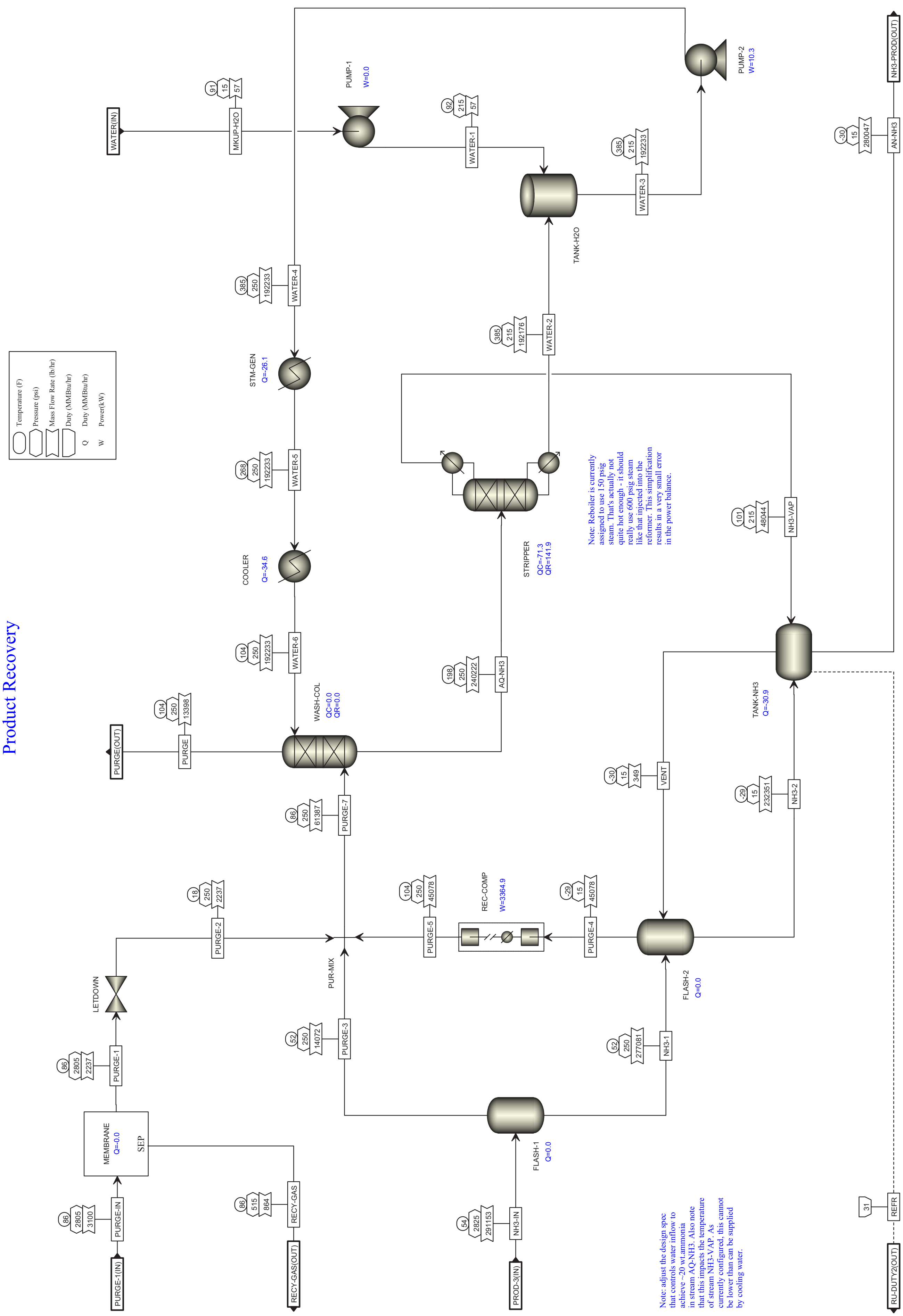
CO2 Compression



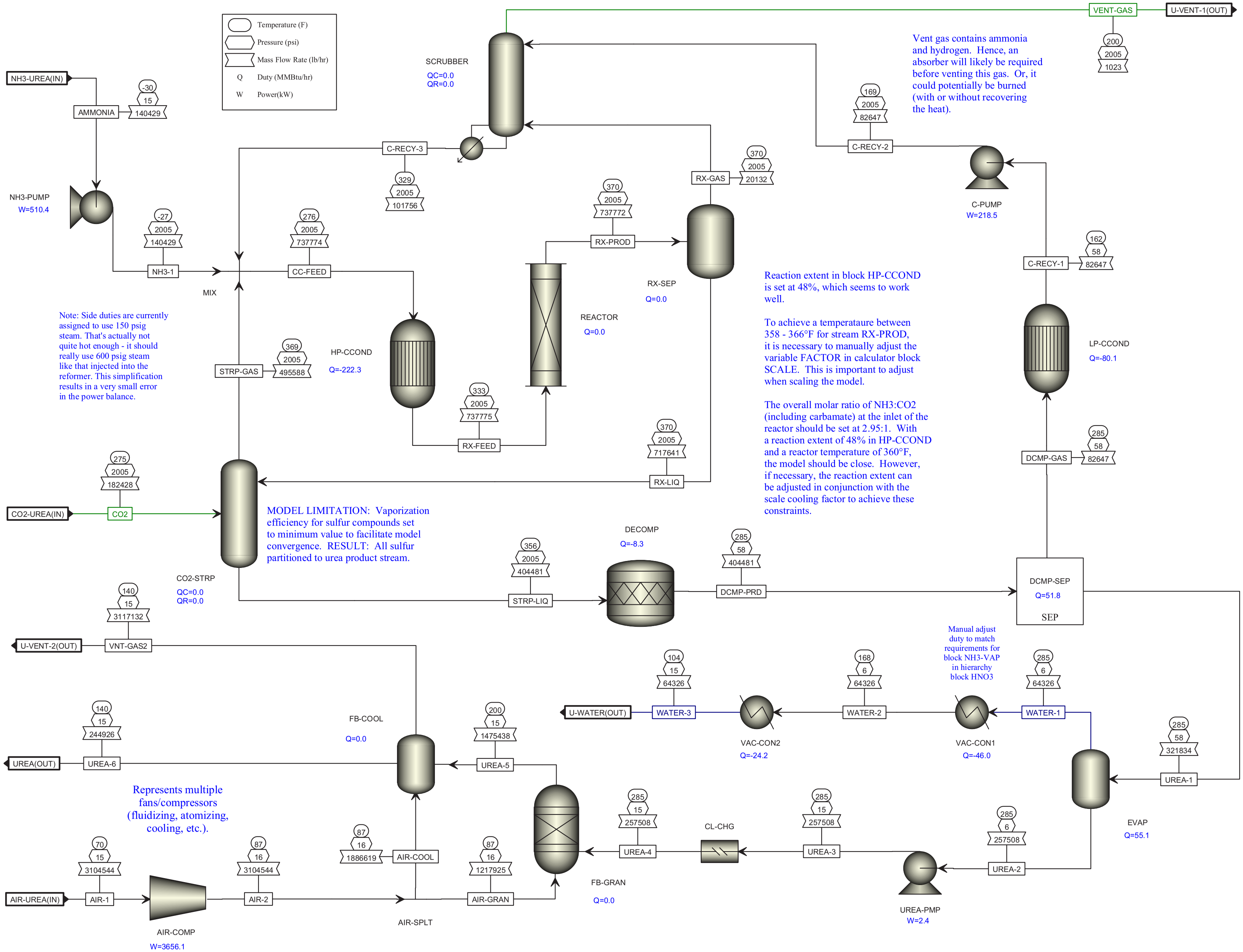
Ammonia Synthesis @ 3000 psi



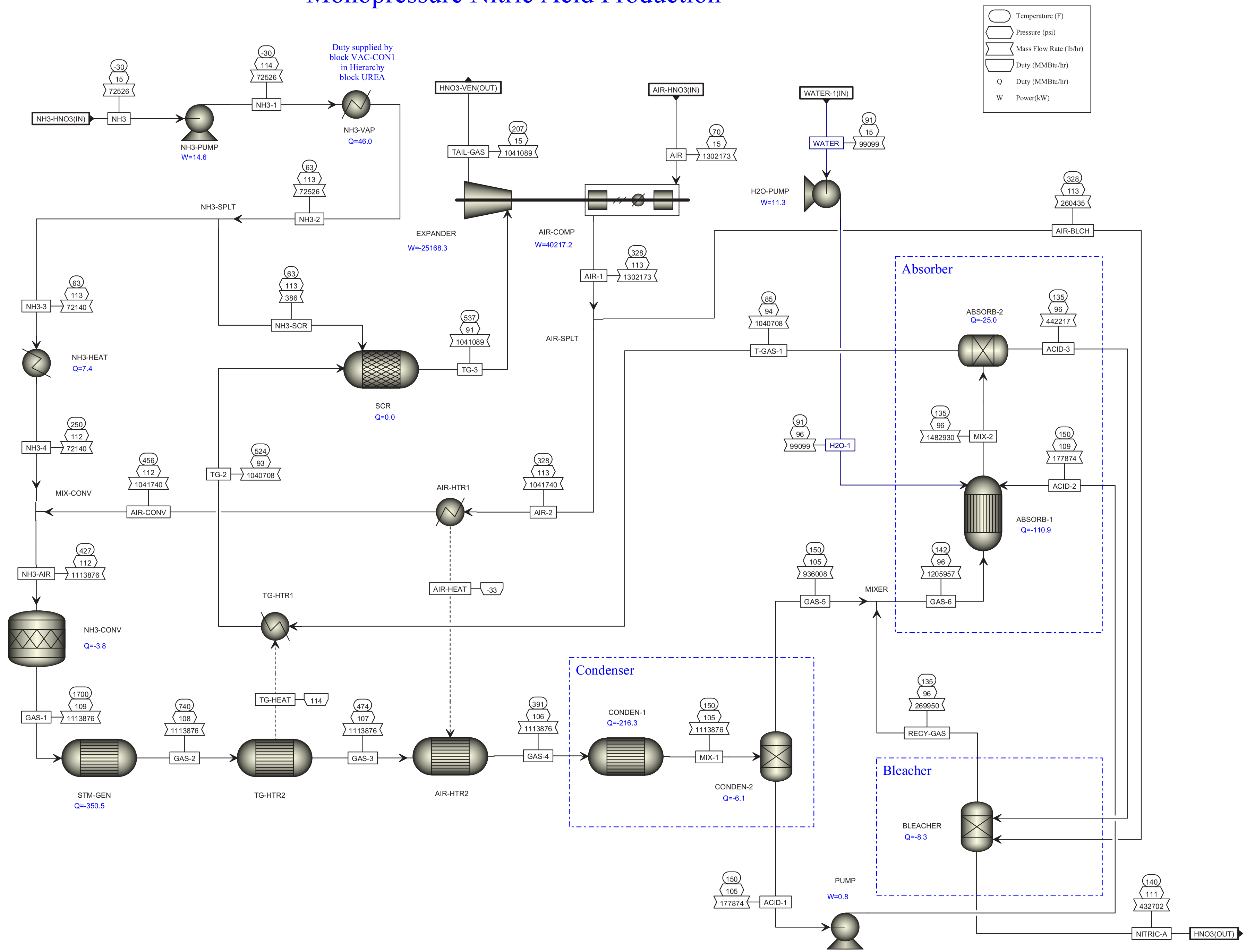
Product Recovery



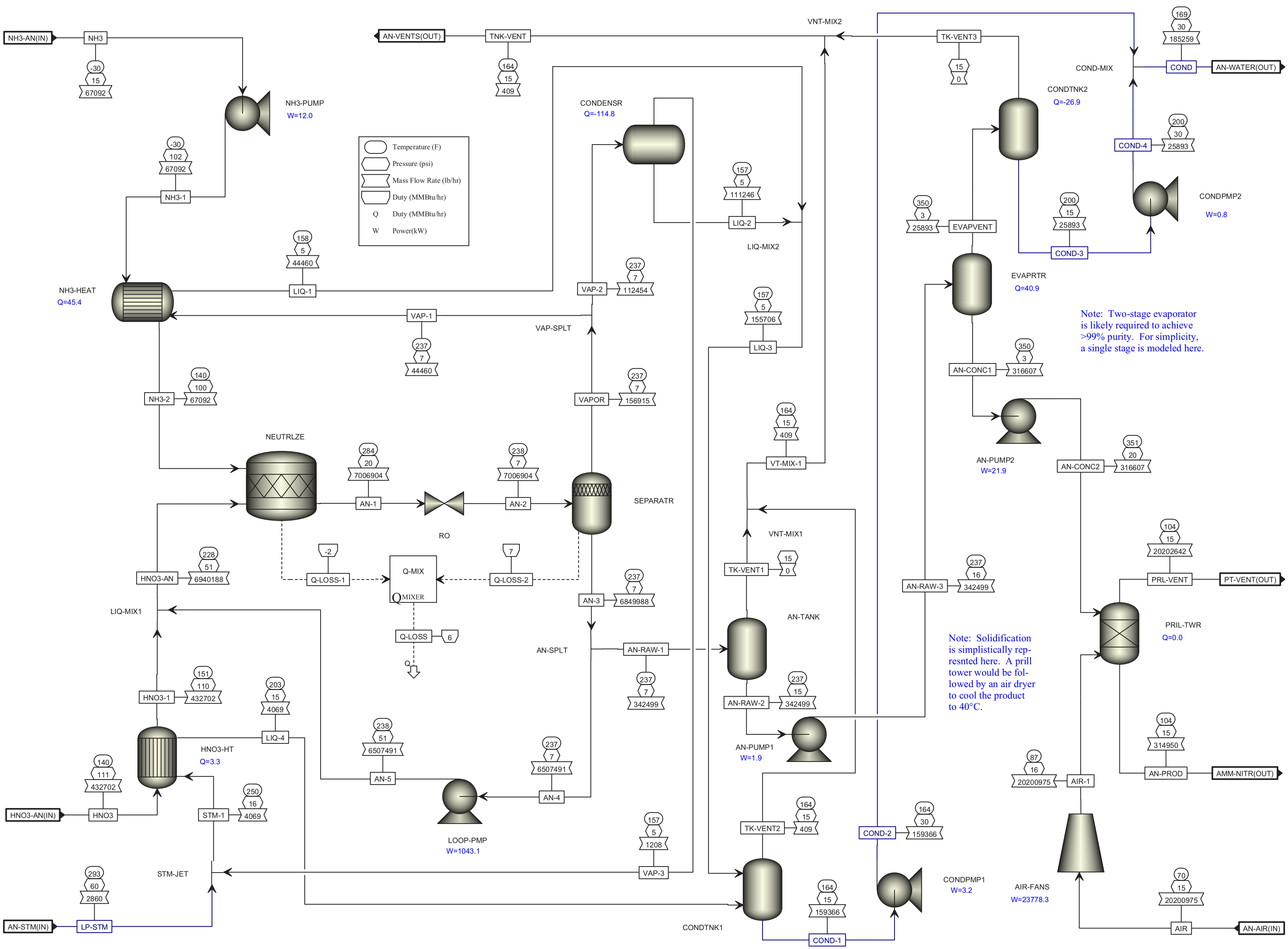
Urea Synthesis via the H.P. Stamicarbon CO2 Stripping Process



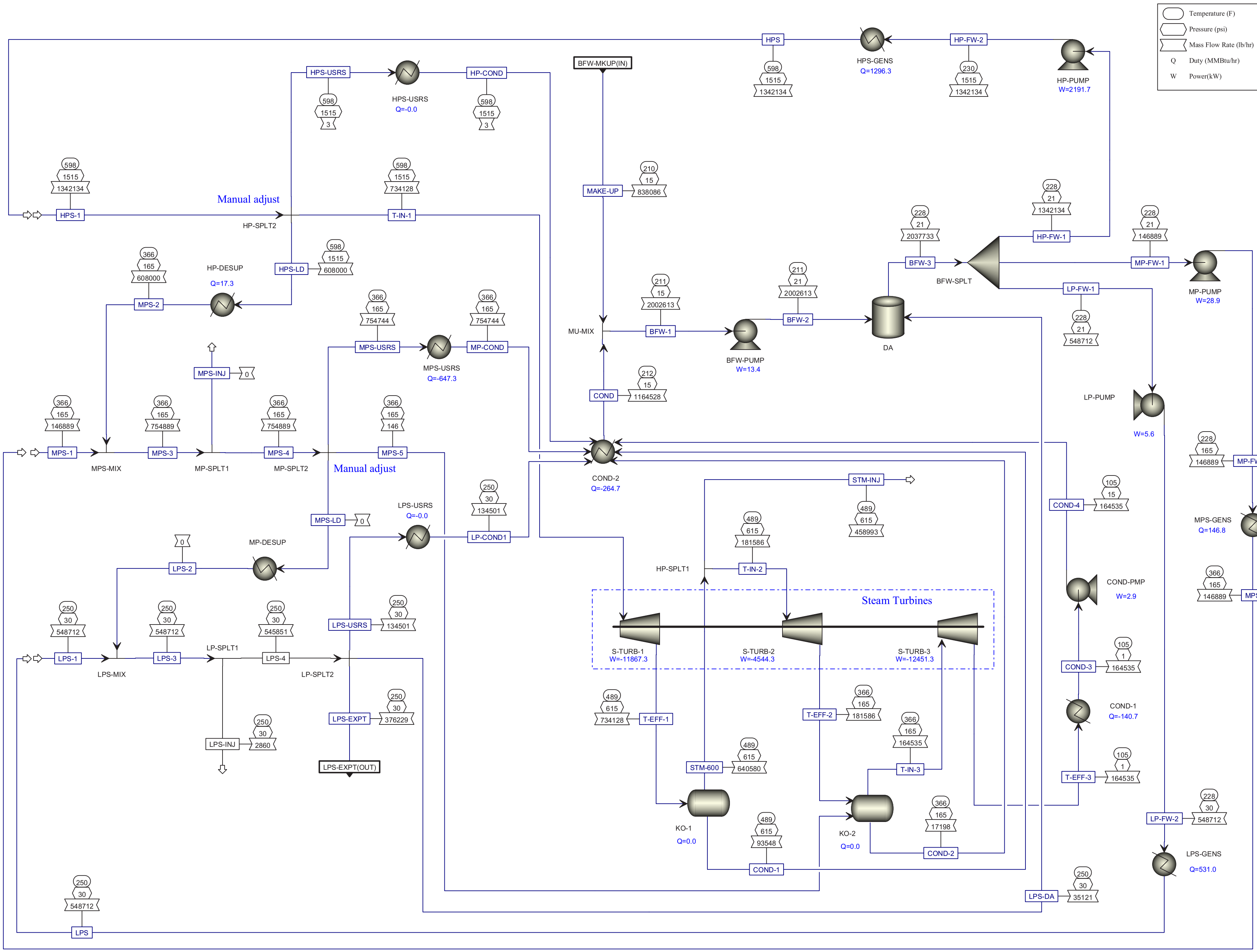
Monopressure Nitric Acid Production



Ammonium Nitrate Production (Uhde-I.G. Farbenindustrie Process)

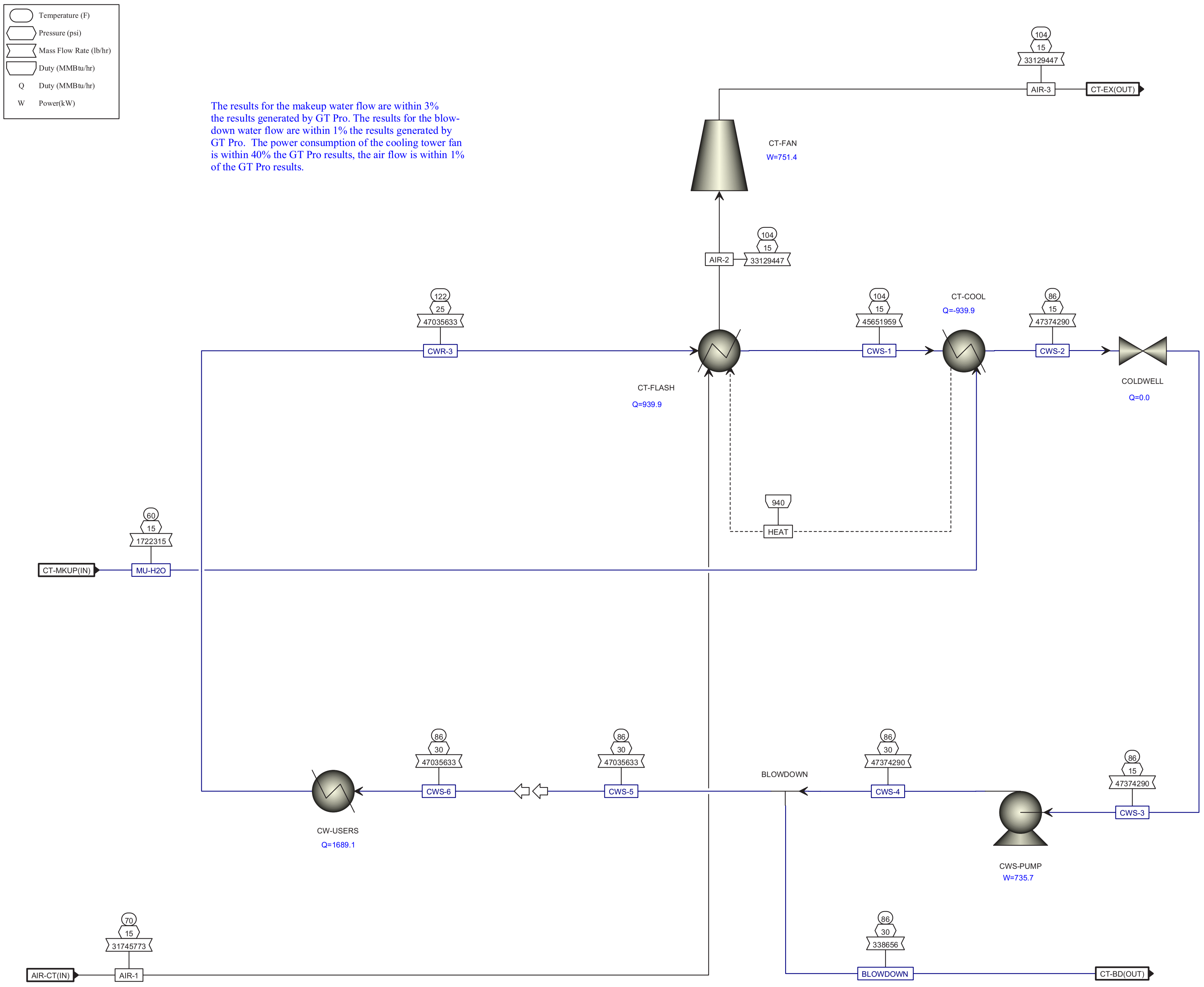


HRSG & Steam Turbines

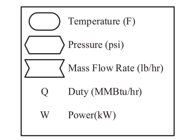


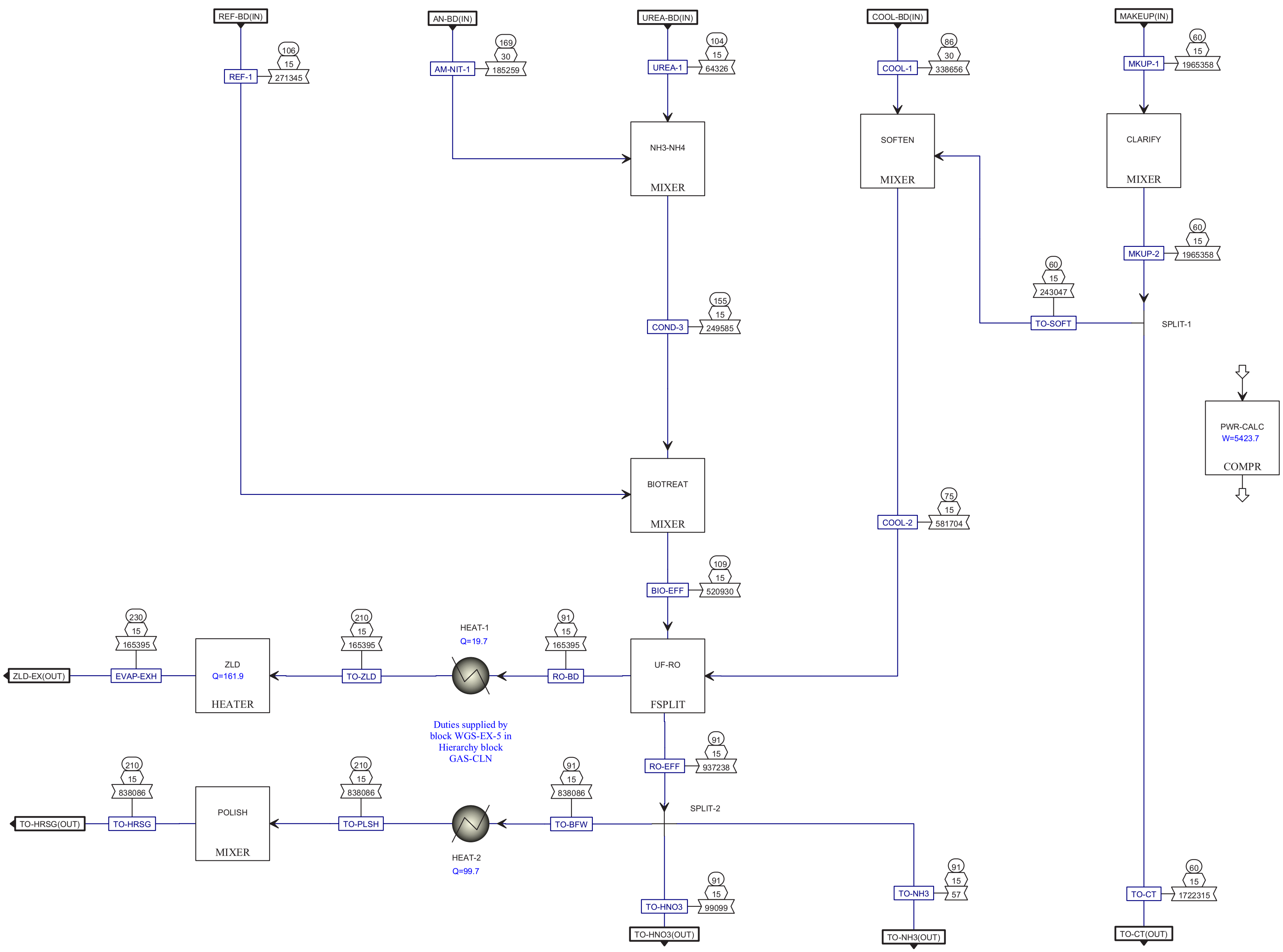
Cooling Tower

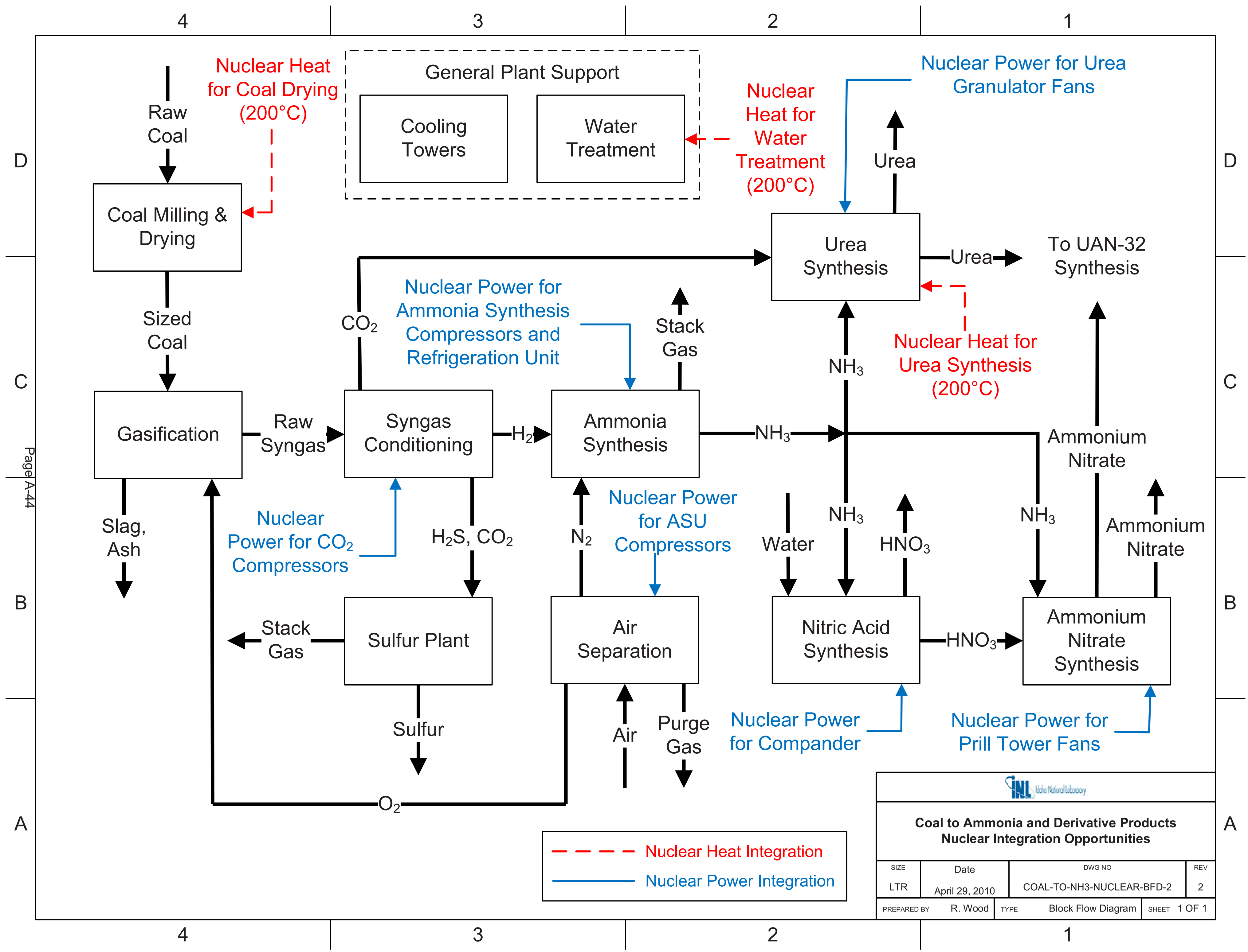
The results for the makeup water flow are within 3% the results generated by GT Pro. The results for the blow-down water flow are within 1% the results generated by GT Pro. The power consumption of the cooling tower fan is within 40% the GT Pro results, the air flow is within 1% of the GT Pro results.

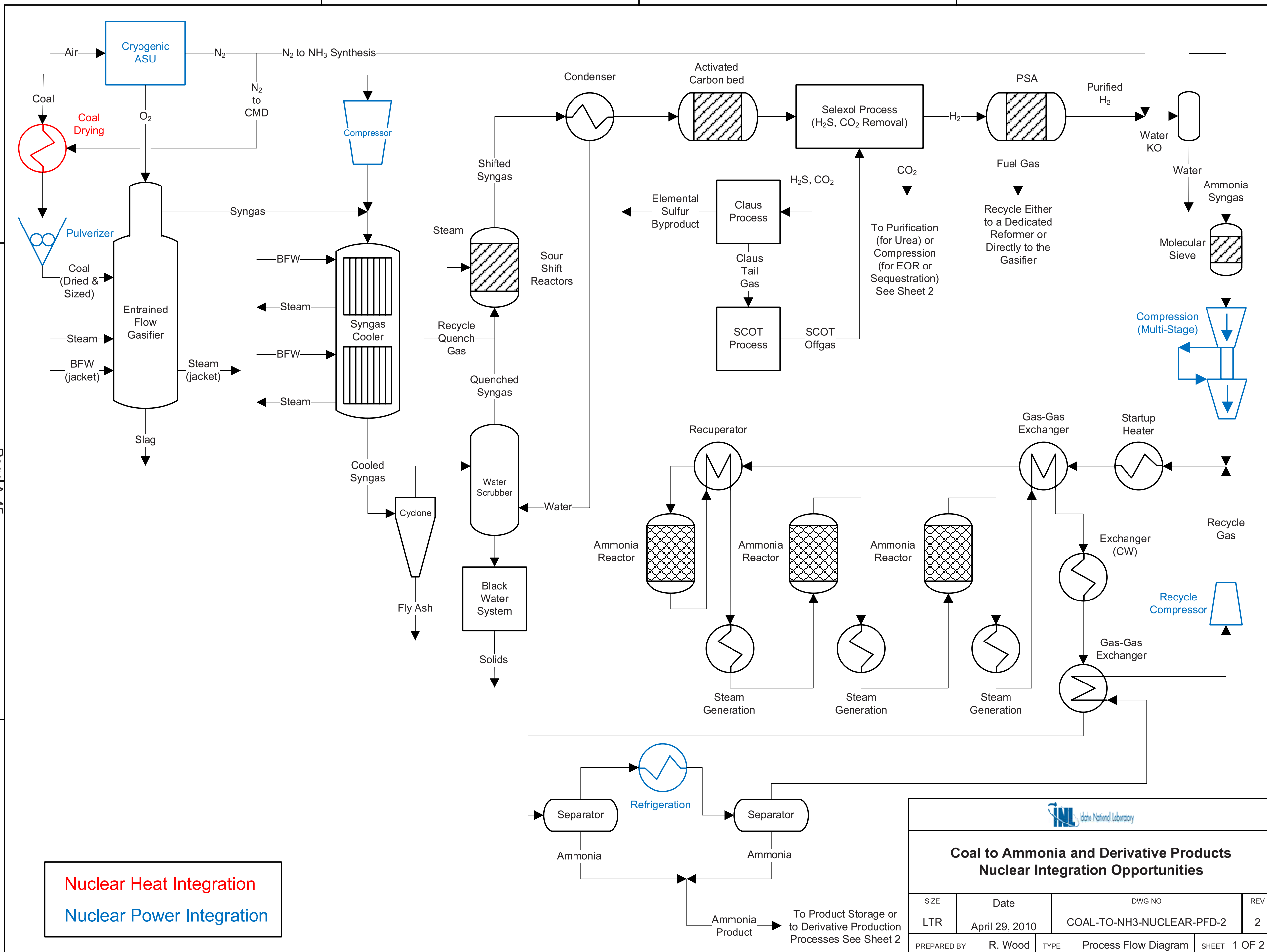


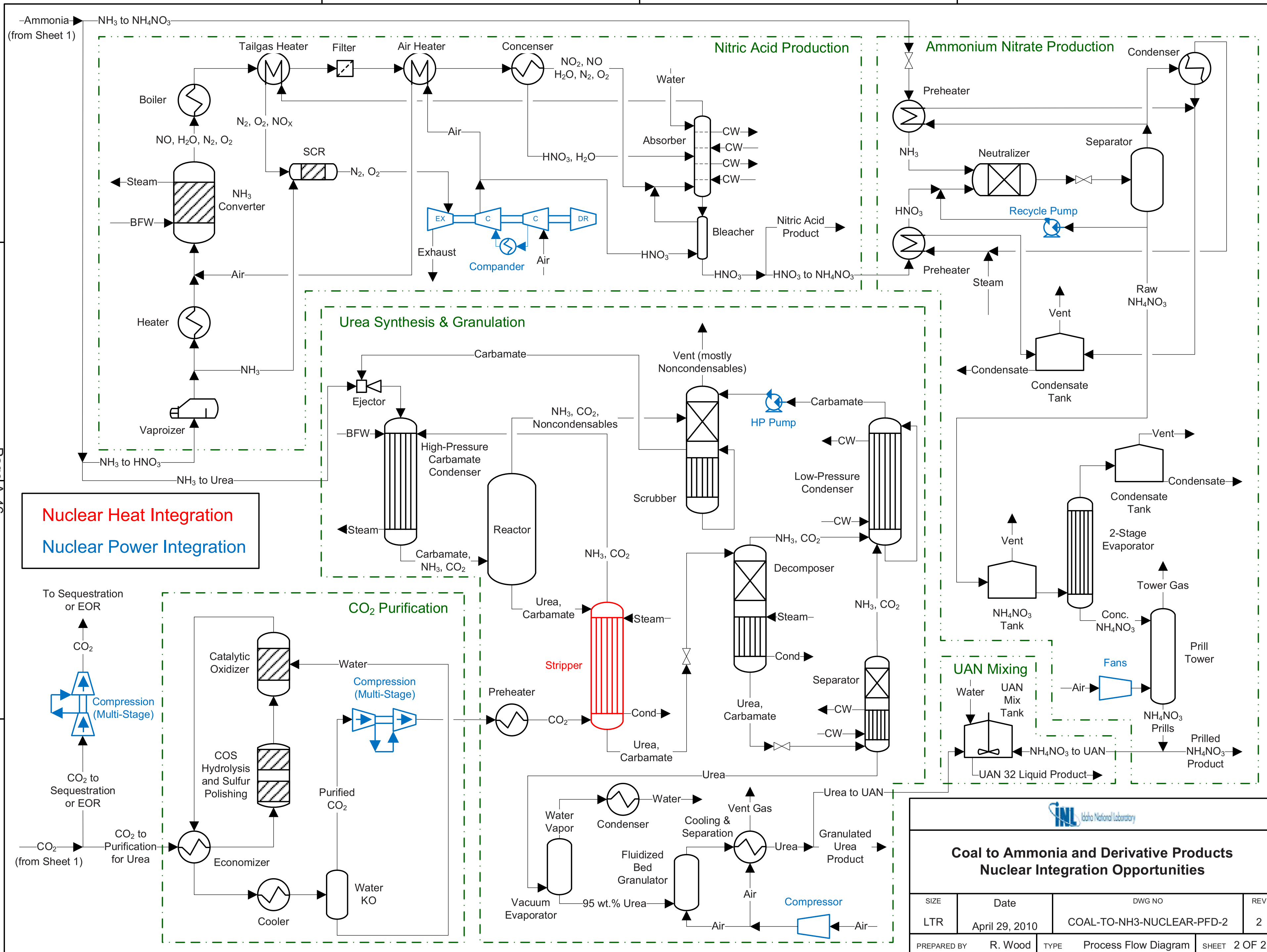
Simplified Water Treatment



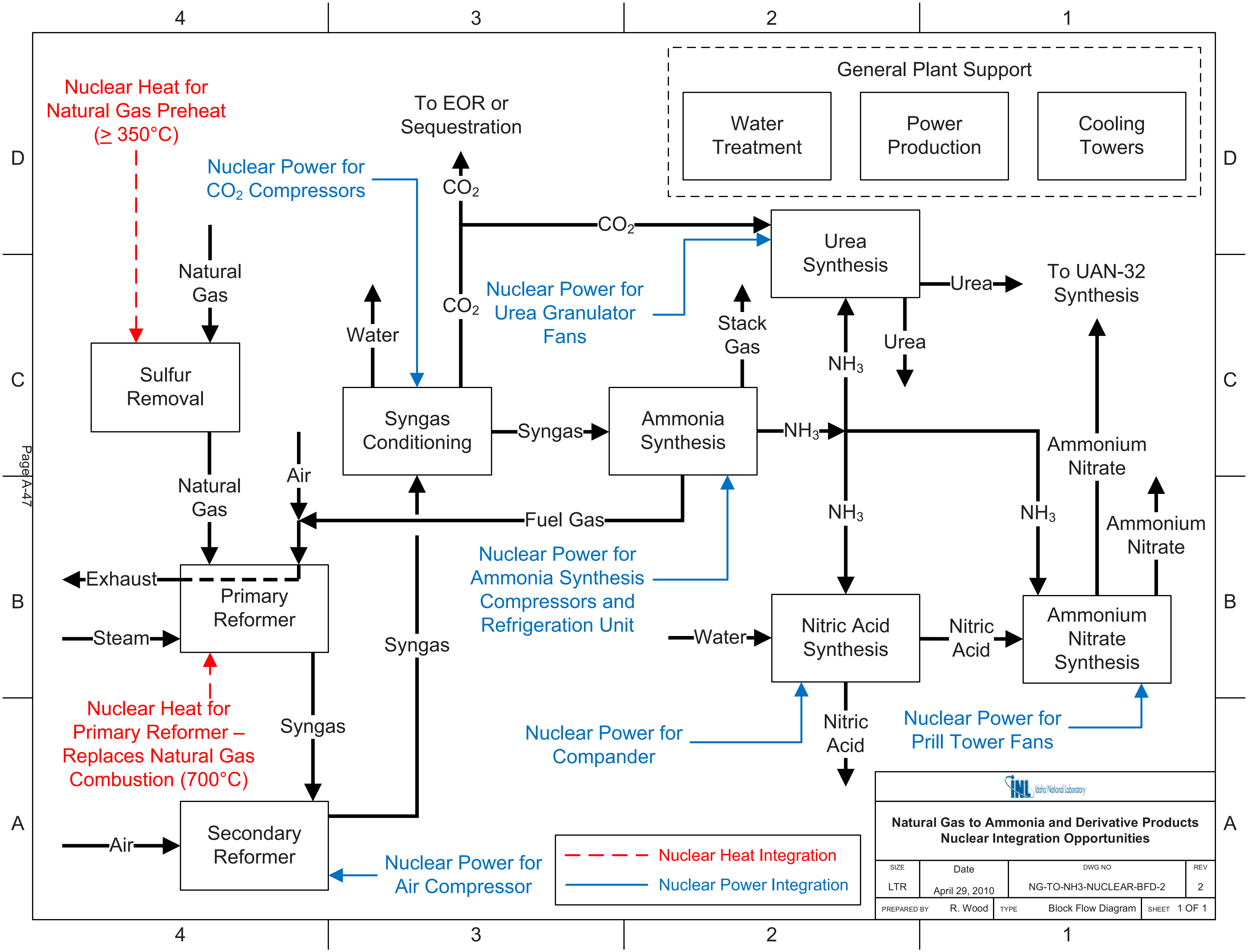


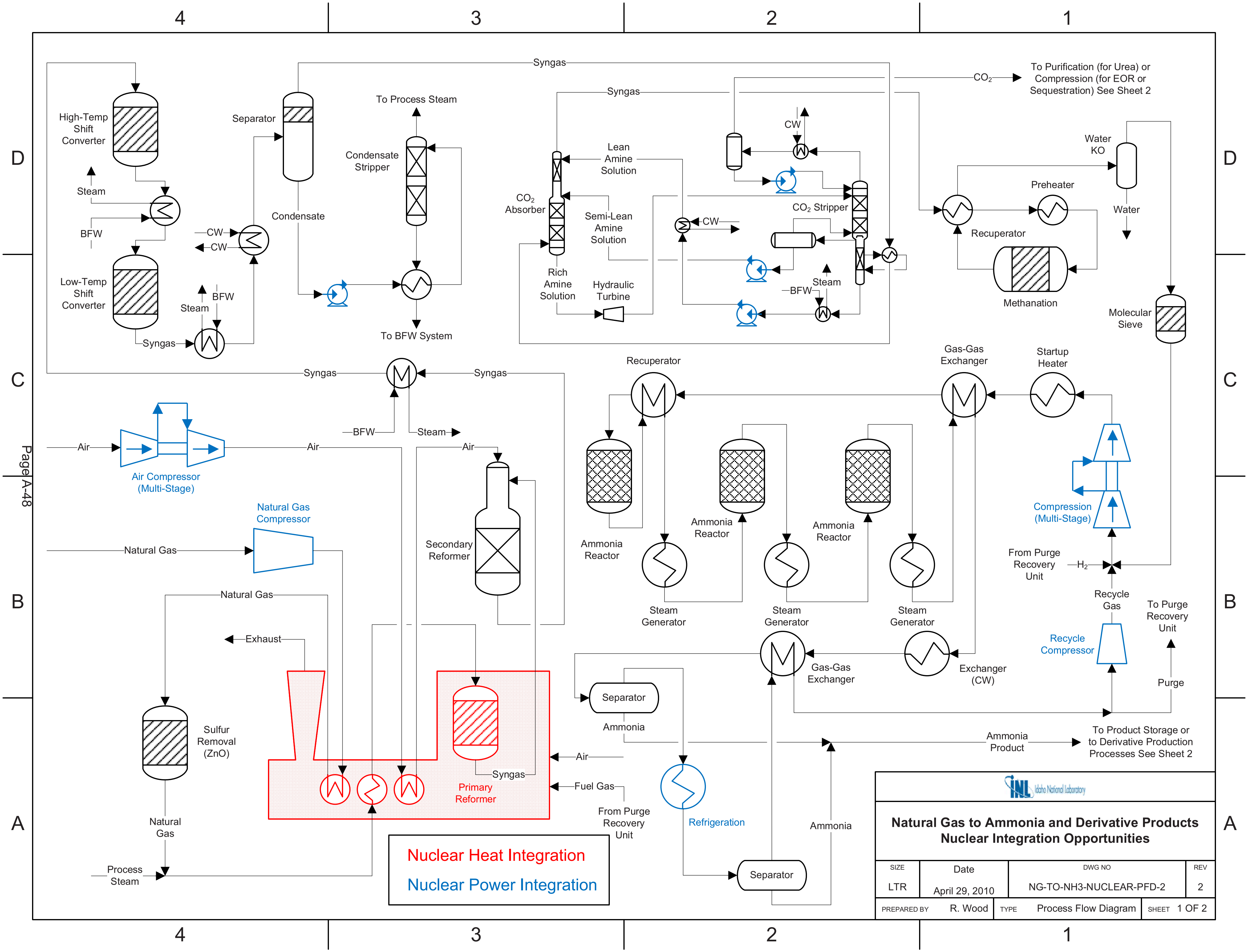

General Plant Support
Cooling Towers
Water Treatment
Raw Coal
Nuclear Heat for Coal Drying (200°C)
Coal Milling & Drying
Sized Coal
Gasification
Slag, Ash
Raw Syngas
Syngas Conditioning
CO2
Nuclear Power for CO2 Compressors
H2S, CO2
Sulfur Plant
Stack Gas
Sulfur
O2
H2
Nuclear Power for Ammonia Synthesis Compressors and Refrigeration Unit
Ammonia Synthesis
Stack Gas
N2
Nuclear Power for ASU Compressors
Air Separation
Air
Purge Gas
NH3
Nuclear Heat for Water Treatment (200°C)
Nuclear Power for Urea Granulator Fans
Urea Synthesis
Urea
To UAN-32 Synthesis
Nuclear Heat for Urea Synthesis (200°C)
Water
HNO3
Nitric Acid Synthesis
Nuclear Power for Compander
Ammonium Nitrate
Ammonium Nitrate Synthesis
Nuclear Power for Prill Tower Fans
Nuclear Heat Integration
Nuclear Power Integration
Idaho National Laboratory
Coal to Ammonia and Derivative Products
Nuclear Integration Opportunities
SIZE
LTR
Date
April 29, 2010
DWG NO
COAL-TO-NH3-NUCLEAR-BFD-2
REV
2
PREPARED BY
R. Wood
TYPE
Block Flow Diagram
SHEET
1 OF 1

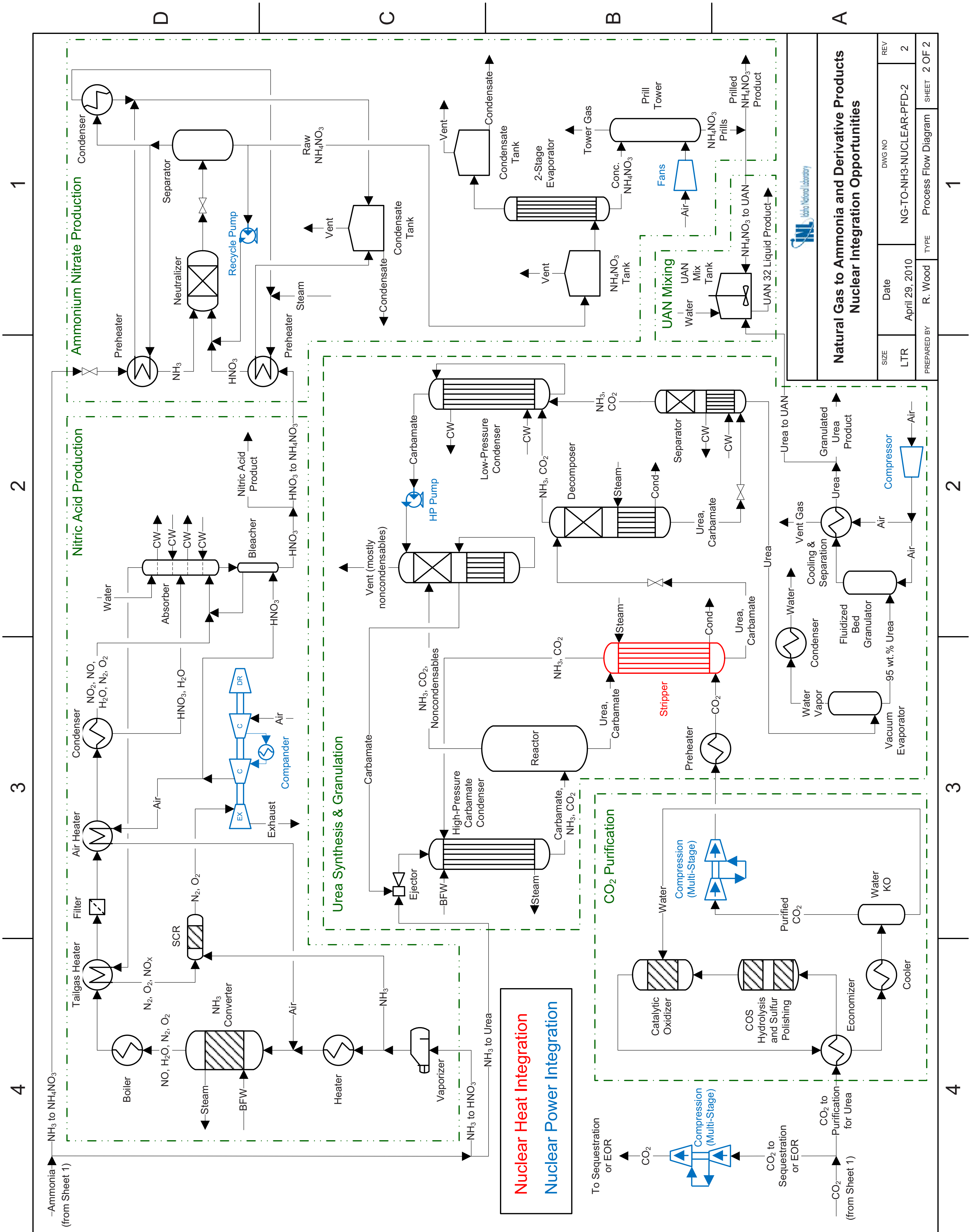

Air
Cryogenic ASU
N_2
N_2 to NH_3 Synthesis
N_2 to CMD
O_2
Coal
Coal Drying
Pulverizer
Coal (Dried & Sized)
Steam
BFW (jacket)
Entrained Flow Gasifier
Steam (jacket)
Slag
Syngas
Compressor
BFW
Steam
BFW
Steam
Syngas Cooler
Cooled Syngas
Cyclone
Fly Ash
Recycle Quench Gas
Steam
Sour Shift Reactors
Shifted Syngas
Quenched Syngas
Water Scrubber
Water
Black Water System
Solids
Condenser
Activated Carbon bed
Selexol Process (H_2S, CO_2 Removal)
H_2S, CO_2
CO_2
Claus Process
Elemental Sulfur Byproduct
Claus Tail Gas
SCOT Process
SCOT Offgas
To Purification (for Urea) or Compression (for EOR or Sequestration) See Sheet 2
H_2
PSA
Fuel Gas
Recycle Either to a Dedicated Reformer or Directly to the Gasifier
Purified H_2
Water KO
Water
Ammonia Syngas
Molecular Sieve
Compression (Multi-Stage)
Startup Heater
Gas-Gas Exchanger
Recuperator
Ammonia Reactor
Ammonia Reactor
Ammonia Reactor
Steam Generation
Steam Generation
Steam Generation
Exchanger (CW)
Recycle Gas
Recycle Compressor
Gas-Gas Exchanger
Separator
Refrigeration
Separator
Ammonia
Ammonia
Ammonia Product
To Product Storage or to Derivative Production Processes See Sheet 2
Nuclear Heat Integration
Nuclear Power Integration
INL Idaho National Laboratory
Coal to Ammonia and Derivative Products
Nuclear Integration Opportunities
SIZE
LTR
Date
April 29, 2010
DWG NO
COAL-TO-NH3-NUCLEAR-PFD-2
REV
2
PREPARED BY
R. Wood
TYPE
Process Flow Diagram
SHEET
1 OF 2



Ammonia (from Sheet 1)
NH3 to NH4NO3
NH3 to HNO3
NH3 to Urea
Nitric Acid Production
Tailgas Heater
Filter
Air Heater
Concenser
NO2, NO H2O, N2, O2
Water
Absorber
CW
Boiler
N2, O2, NOX
NO, H2O, N2, O2
SCR
N2, O2
Air
HNO3, H2O
Steam
BFW
NH3 Converter
EX
C
DR
Exhaust
Compander
Bleacher
Nitric Acid Product
HNO3
HNO3 to NH4NO3
Heater
NH3
Vaproizer
Ammonium Nitrate Production
Condenser
Preheater
Separator
Neutralizer
Recycle Pump
Raw NH4NO3
Vent
Condensate
Condensate Tank
Urea Synthesis & Granulation
Carbamate
Ejector
Vent (mostly Noncondensables)
HP Pump
NH3, CO2, Noncondensables
High-Pressure Carbamate Condenser
Scrubber
Low-Pressure Condenser
Reactor
Carbamate, NH3, CO2
NH3, CO2
Decomposer
Urea, Carbamate
Stripper
Cond
Separator
Urea
2-Stage Evaporator
Tower Gas
NH4NO3 Tank
Conc. NH4NO3
Prill Tower
Fans
NH4NO3 Prills
Prilled NH4NO3 Product
UAN Mixing
UAN Mix Tank
NH4NO3 to UAN
UAN 32 Liquid Product
Urea to UAN
Nuclear Heat Integration
Nuclear Power Integration
To Sequestration or EOR
CO2
Compression (Multi-Stage)
CO2 to Sequestration or EOR
CO2 (from Sheet 1)
CO2 to Purification for Urea
CO2 Purification
Catalytic Oxidizer
COS Hydrolysis and Sulfur Polishing
Economizer
Cooler
Water KO
Purified CO2
Preheater
Water Vapor
Condenser
Vent Gas
Cooling & Separation
Granulated Urea Product
Fluidized Bed Granulator
Vacuum Evaporator
95 wt.% Urea
Compressor
Idaho National Laboratory
Coal to Ammonia and Derivative Products
Nuclear Integration Opportunities
SIZE LTR
Date April 29, 2010
DWG NO COAL-TO-NH3-NUCLEAR-PFD-2
REV 2
PREPARED BY R. Wood
TYPE Process Flow Diagram
SHEET 2 OF 2


General Plant Support
Water Treatment
Power Production
Cooling Towers
Nuclear Heat for Natural Gas Preheat (≥ 350°C)
Natural Gas
Sulfur Removal
Natural Gas
Air
Exhaust
Steam
Primary Reformer
Nuclear Heat for Primary Reformer – Replaces Natural Gas Combustion (700°C)
Syngas
Air
Secondary Reformer
Nuclear Power for Air Compressor
Nuclear Power for CO_2 Compressors
To EOR or Sequestration
CO_2
CO_2
CO_2
Water
Syngas Conditioning
Syngas
Syngas
Fuel Gas
Ammonia Synthesis
Stack Gas
Nuclear Power for Urea Granulator Fans
Nuclear Power for Ammonia Synthesis Compressors and Refrigeration Unit
Urea Synthesis
Urea
Urea
To UAN-32 Synthesis
NH_3
NH_3
NH_3
NH_3
Water
Nitric Acid Synthesis
Nitric Acid
Nitric Acid
Nuclear Power for Compander
Ammonium Nitrate Synthesis
Ammonium Nitrate
Ammonium Nitrate
Nuclear Power for Prill Tower Fans
Nuclear Heat Integration
Nuclear Power Integration
Natural Gas to Ammonia and Derivative Products
Nuclear Integration Opportunities
SIZE LTR
Date April 29, 2010
DWG NO NG-TO-NH3-NUCLEAR-BFD-2
REV 2
PREPARED BY R. Wood
TYPE Block Flow Diagram
SHEET 1 OF 1


High-Temp Shift Converter
Steam
BFW
Low-Temp Shift Converter
Syngas
Steam
BFW
CW
CW
Separator
Condensate
Condensate Stripper
To Process Steam
To BFW System
Syngas
Syngas
CO_2 Absorber
Lean Amine Solution
Semi-Lean Amine Solution
Rich Amine Solution
Hydraulic Turbine
CW
CW
CO_2 Stripper
Steam
BFW
CO_2
To Purification (for Urea) or Compression (for EOR or Sequestration) See Sheet 2
Water KO
Preheater
Water
Recuperator
Methanation
Molecular Sieve
Syngas
Syngas
BFW
Steam
Air
Air Compressor (Multi-Stage)
Air
Air
Natural Gas Compressor
Natural Gas
Secondary Reformer
Natural Gas
Exhaust
Sulfur Removal (ZnO)
Natural Gas
Process Steam
Syngas
Primary Reformer
Air
Fuel Gas
From Purge Recovery Unit
Nuclear Heat Integration
Nuclear Power Integration
Recuperator
Ammonia Reactor
Ammonia Reactor
Ammonia Reactor
Steam Generator
Steam Generator
Steam Generator
Gas-Gas Exchanger
Startup Heater
Compression (Multi-Stage)
From Purge Recovery Unit
H_2
Recycle Gas
Recycle Compressor
To Purge Recovery Unit
Purge
Gas-Gas Exchanger
Exchanger (CW)
Separator
Ammonia
Refrigeration
Separator
Ammonia
Ammonia Product
To Product Storage or to Derivative Production Processes See Sheet 2
Idaho National Laboratory
Natural Gas to Ammonia and Derivative Products
Nuclear Integration Opportunities
SIZE
LTR
Date
April 29, 2010
DWG NO
NG-TO-NH3-NUCLEAR-PFD-2
REV
2
PREPARED BY
R. Wood
TYPE
Process Flow Diagram
SHEET
1 OF 2



Natural Gas to Ammonia and Derivative Products
Nuclear Integration Opportunities
SIZE LTR
Date April 29, 2010
DWG NO NG-TO-NH3-NUCLEAR-PFD-2
REV 2
PREPARED BY R. Wood
TYPE Process Flow Diagram
SHEET 2 OF 2
Nuclear Heat Integration
Nuclear Power Integration
Ammonium Nitrate Production
Nitric Acid Production
Urea Synthesis & Granulation
CO2 Purification
UAN Mixing
Ammonia (from Sheet 1)
CO2 (from Sheet 1)
NH3 to NH4NO3
NH3 to HNO3
NH3 to Urea
Vaporizer
Heater
NH3 Converter
Boiler
Tailgas Heater
Filter
Air Heater
Condenser
SCR
Compander
Exhaust
Absorber
Bleacher
Nitric Acid Product
HNO3 to NH4NO3
Preheater
Neutralizer
Recycle Pump
Separator
Condensate Tank
NH4NO3 Tank
2-Stage Evaporator
Prill Tower
Fans
Tower Gas
Prilled NH4NO3 Product
NH4NO3 to UAN
UAN Mix Tank
UAN 32 Liquid Product
Ejector
High-Pressure Carbamate Condenser
Reactor
Stripper
HP Pump
Low-Pressure Condenser
Decomposer
Vent (mostly noncondensables)
Vacuum Evaporator
Fluidized Bed Granulator
Cooling & Separation
Compressor
Granulated Urea Product
Urea to UAN
95 wt.% Urea
Compression (Multi-Stage)
Purified CO2
Water KO
Cooler
Economizer
COS Hydrolysis and Sulfur Polishing
Catalytic Oxidizer
CO2 to Purification for Urea
CO2 to Sequestration or EOR
To Sequestration or EOR

Nuclear Gas-to-Ammonia Results

Calculator Block SUMMARY

```
   FEED SUMMARY:

     NATURAL GAS PROPERTIES:

       MASS FLOW =                         1622. TON/DY
       VOLUME FLOW =                         72. MMSCFD @ 60˚F
       HHV =                              23063. BTU/LB
       HHV =                               1044. BTU/SCF @ 60˚F
       ENERGY FLOW =                      74818. MMBTU/DY

       COMPOSITION:
          METHANE =                          93.571 MOL.%
          ETHANE =                            3.749 MOL.%
          PROPANE =                           0.920 MOL.%
          BUTANE =                            0.260 MOL.%
          PENTANE =                           0.040 MOL.%
          HEXANE =                            0.010 MOL.%
          NITROGEN =                          1.190 MOL.%
          OXYGEN =                            0.010 MOL.%
          CO2 =                               0.250 MOL.%
          C4H10S =                            1. PPMV
          C2H6S =                             0. PPMV
          H2S =                               0. PPMV

   INTERMEDIATE PRODUCT SUMMARY:

     RAW SYNGAS MASS FLOW =              941756. LB/HR
     RAW SYNGAS VOLUME FLOW =               538. MMSCFD @ 60˚F
     RAW SYNGAS COMPOSITION:
       H2                                    35.6 MOL.%
       CO                                     7.4 MOL.%
       CO2                                    5.5 MOL.%
       N2                                    14.4 MOL.%
       AR                                     0.2 MOL.%
       H2O                                   35.7 MOL.%
       CH4                                    1.1 MOL.%

     CLEANED SYNGAS MASS FLOW =          304786. LB/HR
     CLEANED SYNGAS VOLUME FLOW =           313. MMSCFD @ 60˚F
     CLEANED SYNGAS COMPOSITION:
       H2                                    72.8 MOL.%
       N2                                    24.7 MOL.%
       AR                                     0.3 MOL.%
       CH4                                    2.2 MOL.%
       CO                                     0. PPMV
       CO2                                    0. PPMV
       H2O                                    1. PPMV

     AMMONIA MASS FLOW =                 280012. LB/HR
     AMMONIA MASS FLOW =                   3360. TON/DY
       AMMONIA PURITY =                      99.99 MOL.%
     AMMONIA PRODUCED / NAT. GAS FED =        2.07 LB/LB

     NITRIC ACID MASS FLOW =             432657. LB/HR
     NITRIC ACID MASS FLOW =               5192. TON/DY
       NITRIC ACID PURITY =                  57.93 WT.%
     HNO3 PRODUCED / NH3 FED =                1.55 LB/LB

   FINAL PRODUCT SUMMARY:

     AMMONIA MASS FLOW =                      0. LB/HR
     AMMONIA MASS FLOW =                      0. TON/DY
```

Nuclear Gas-to-Ammonia Results

```
    AMMONIA PURITY =                          99.99 MOL.%

  NITRIC ACID MASS FLOW =                        0. LB/HR
  NITRIC ACID MASS FLOW =                        0. TON/DY
    NITRIC ACID PURITY =                      57.93 WT.%

  UREA MASS FLOW =                          244897. LB/HR
  UREA MASS FLOW =                            2939. TON/DY
  UREA PRODUCED / AMMONIA FED =                1.74 LB/LB
  UREA PRODUCED / CO2 FED =                    1.34 LB/LB

  AMMONIUM NITRATE MASS FLOW =              314918. LB/HR
  AMMONIUM NITRATE MASS FLOW =                3779. TON/DY
  NH4NO3 PRODUCED / NITRIC ACID FED =          0.73 LB/LB
  NH4NO3 PRODUCED / AMMONIA FED =              4.69 LB/LB

  STEAM EXPORT AVAILABLE:
    SATURATED STEAM FLOW =                  379864. LB/HR
      PRESSURE =                                30. PSIA


POWER SUMMARY:

  ELECTRICAL GENERATORS:
    STEAM TURBINE POWER GENERATION =       31.8 MW
  GENERATOR SUBTOTAL =                     31.8 MW

  ELECTRICAL CONSUMERS:
    NG REFORMER POWER CONSUMPTION =        19.9 MW
    GAS CLEANING POWER CONSUMPTION =        4.5 MW
    POWER BLOCK POWER CONSUMPTION =         2.3 MW
    CO2 PROCESSING POWER CONSUMPTION =     13.1 MW
    AMMONIA SYNTH. POWER CONSUMPTION =     46.8 MW
    UREA SYNTHESIS POWER CONSUMPTION =      4.4 MW
    HNO3 SYNTHESIS POWER CONSUMPTION =     15.1 MW
    NH4NO3 SYNTH. POWER CONSUMPTION =      24.9 MW
    COOLING TOWER POWER CONSUMPTION =       1.5 MW
    WATER TREATMENT POWER CONSUMPTION =     5.6 MW
  CONSUMER SUBTOTAL =                     138.0 MW

  NET PLANT POWER CONSUMPTION =           106.2 MW


WATER BALANCE:

  EVAPORATIVE LOSSES:
    COOLING TOWER EVAPORATION =          3315.8 GPM
    ZLD SYSTEM EVAPORATION =              340.8 GPM
  TOTAL EVAPORATIVE LOSSES =             3656.5 GPM

  WATER CONSUMED:
    BOILER FEED WATER MAKEUP =           1732.8 GPM
    AMMONIA WASH WATER =                    0.0 GPM
    NITRIC ACID FEED WATER =              198.0 GPM
    COOLING TOWER MAKEUP =               3489.9 GPM
  TOTAL WATER CONSUMED =                 5421.0 GPM

  WATER GENERATED:
    NG REFORMER BLOWDOWN =                592.6 GPM
    UREA PRODUCTION BLOWDOWN =            128.5 GPM
    NH4NO3 PRODUCTION BLOWDOWN =          370.2 GPM
    COOLING TOWER BLOWDOWN =              686.2 GPM
  TOTAL WATER GENERATED =                1777.6 GPM
```

```
                   Nuclear Gas-to-Ammonia Results
     PLANT WATER SUMMARY:
       NET MAKEUP WATER REQUIRED =          3984.2 GPM
       WATER CONSUMED / NG FED =              14.75 LB/LB


   CO2 BALANCE:

     CO2 INCORPORATED INTO UREA PRODUCT =  2177. TON/DY
     CO2 INCORPORATED INTO UREA PRODUCT =    38. MMSCFD @ 60˚F
       CO2 PURITY =                          99.3 %

     CO2 CAPTURED (SEQ. OR EOR) =          1786. TON/DY
     CO2 CAPTURED (SEQ. OR EOR) =            31. MMSCFD @ 60˚F
       CO2 PURITY =                          99.3 %

     CO2 EMITTED (TOTAL) =                  430. TON/DY
     CO2 EMITTED (TOTAL) =                    7. MMSCFD @ 60˚F
       FROM REFORMER =                      403. TON/DY
       FROM UREA VENTS =                     28. TON/DY
     CO2 EMITTED / NG FED =                   0.27 LB/LB


   NUCLEAR INTEGRATION REQUIREMENTS:

     TOTAL ELECTRICITY DEMAND =             106.2 MW

     TOTAL HEAT DEMAND =                    190.3 MW
     TOTAL HEAT DEMAND =                    649.2 MMBTU/HR

       HELIUM FLOWRATE REQUIRED =       1627010. LB/HR
       HELIUM FLOWRATE REQUIRED =           205.00 KG/S

       HELIUM SUPPLY TEMPERATURE =         1292. DEG. F.
       HELIUM SUPPLY TEMPERATURE =          700. DEG. C.

       HELIUM RETURN TEMPERATURE =          970. DEG. F.
       HELIUM RETURN TEMPERATURE =          521. DEG. C.


Calculator Block AIRPROPS

   HUMIDITY DATA FOR STREAM PRI-AIR:
     HUMIDITY RATIO =                        43.5 GRAINS/LB
     RELATIVE HUMIDITY =                     39.9 %


Calculator Block PH       Hierarchy: NH4NO3

   TARGET NEUTRALIZER PH =   1.500
   ACTUAL NEUTRALIZER PH =   1.499


Calculator Block NG-RFMR  Hierarchy: REFORMER

   SULFUR REMOVAL CONDITIONS:

     INLET BED TEMPERATURE =               662. ˚F

   PRIMARY REFORMER CONDITIONS:

     INLET TEMPERATURE =                  1000. ˚F
     STEAM TO CARBON MOLAR RATIO =           3.30
     NATURAL GAS BURNED FOR HEAT =           0.00 %
     OUTLET TEMPERATURE =                 1400. ˚F
```

Nuclear Gas-to-Ammonia Results

METHANE CONVERSION = 45.6 %

SECONDARY REFORMER CONDITIONS:

INLET GAS TEMPERATURE =	1400. ˚F
INLET AIR TEMPERATURE =	1000. ˚F
STEAM TO CARBON MOLAR RATIO =	2.48
OXYGEN TO CARBON MOLAR RATIO =	0.27
OUTLET TEMPERATURE =	1650. ˚F
H2/CO =	4.81

OUTLET COMPOSITION:

H2	35.6158 MOL.%
CO	7.4062 MOL.%
CO2	5.4932 MOL.%
H2O	35.7229 MOL.%
CH4	1.1428 MOL.%
N2	14.3765 MOL.%

Calculator Block RATIO   Hierarchy: UREA

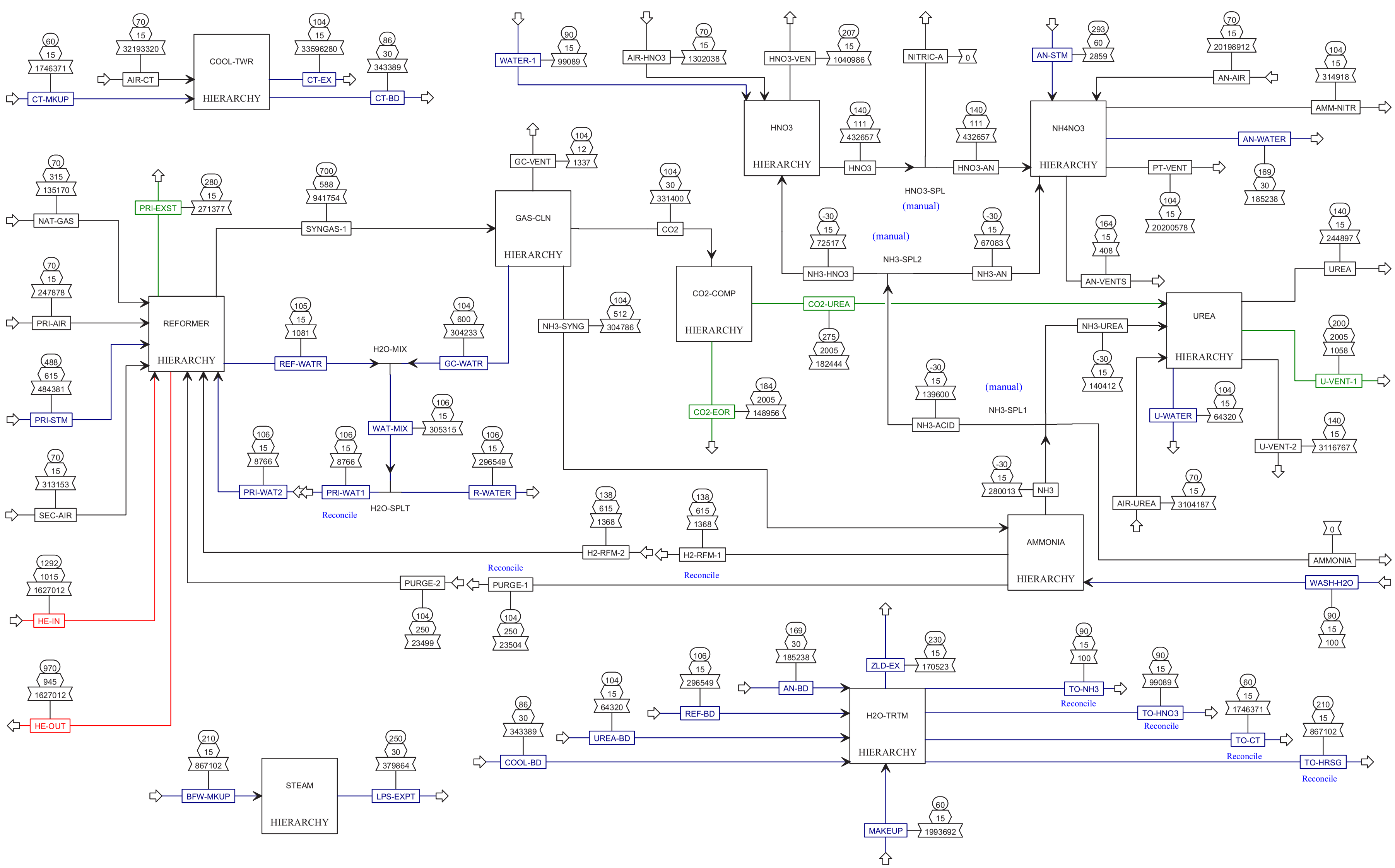
TARGET REACTOR NH3/CO2 RATIO = 2.950
ACTUAL REACTOR NH3/CO2 RATIO = 2.953

Natural Gas to Ammonia and Derivatives - Nuclear Integration
(Derivatives produced in the correct ratio to make UAN32)

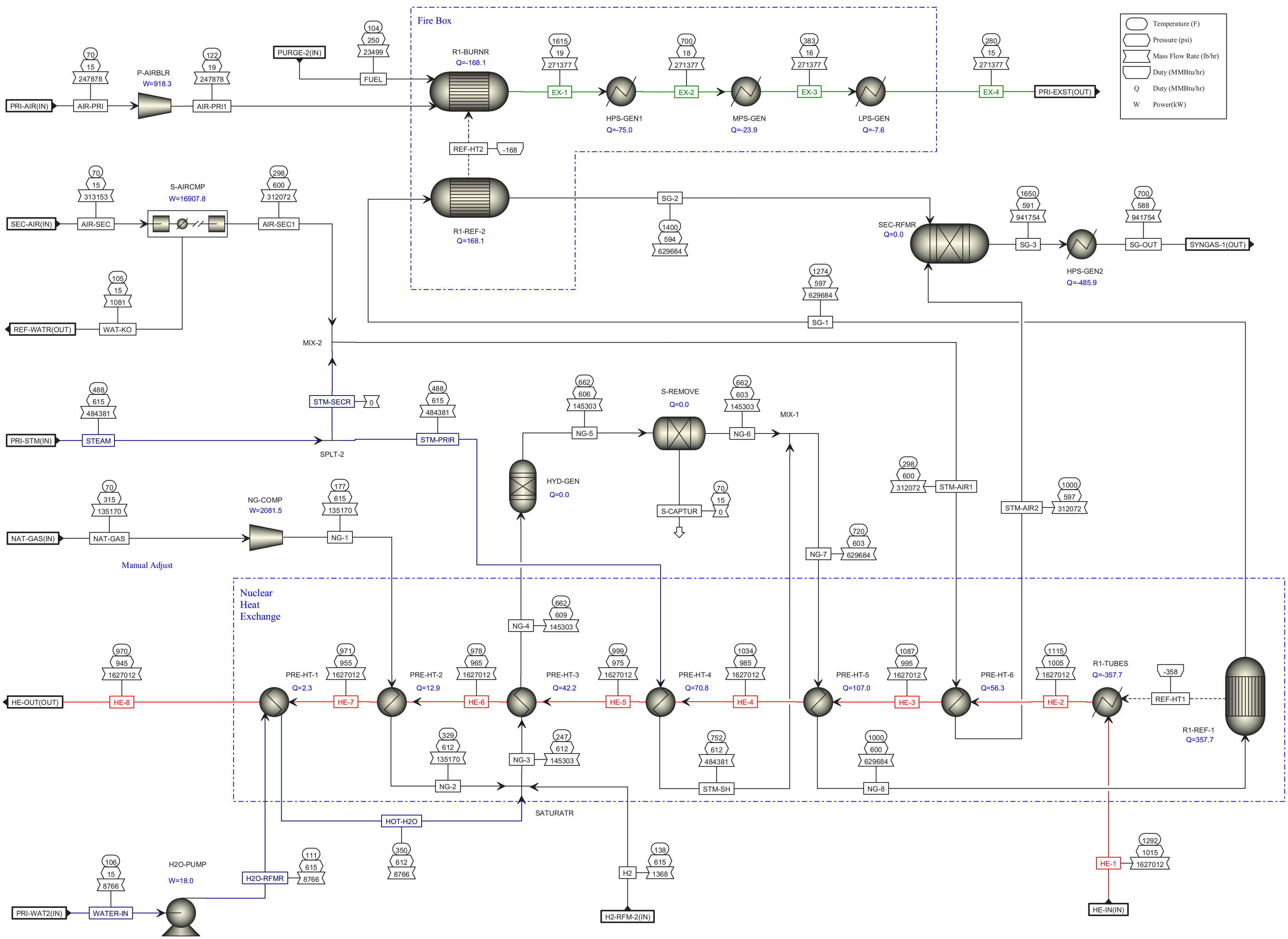
Temperature (F)
Pressure (psi)
Mass Flow Rate (lb/hr)

Color Legend

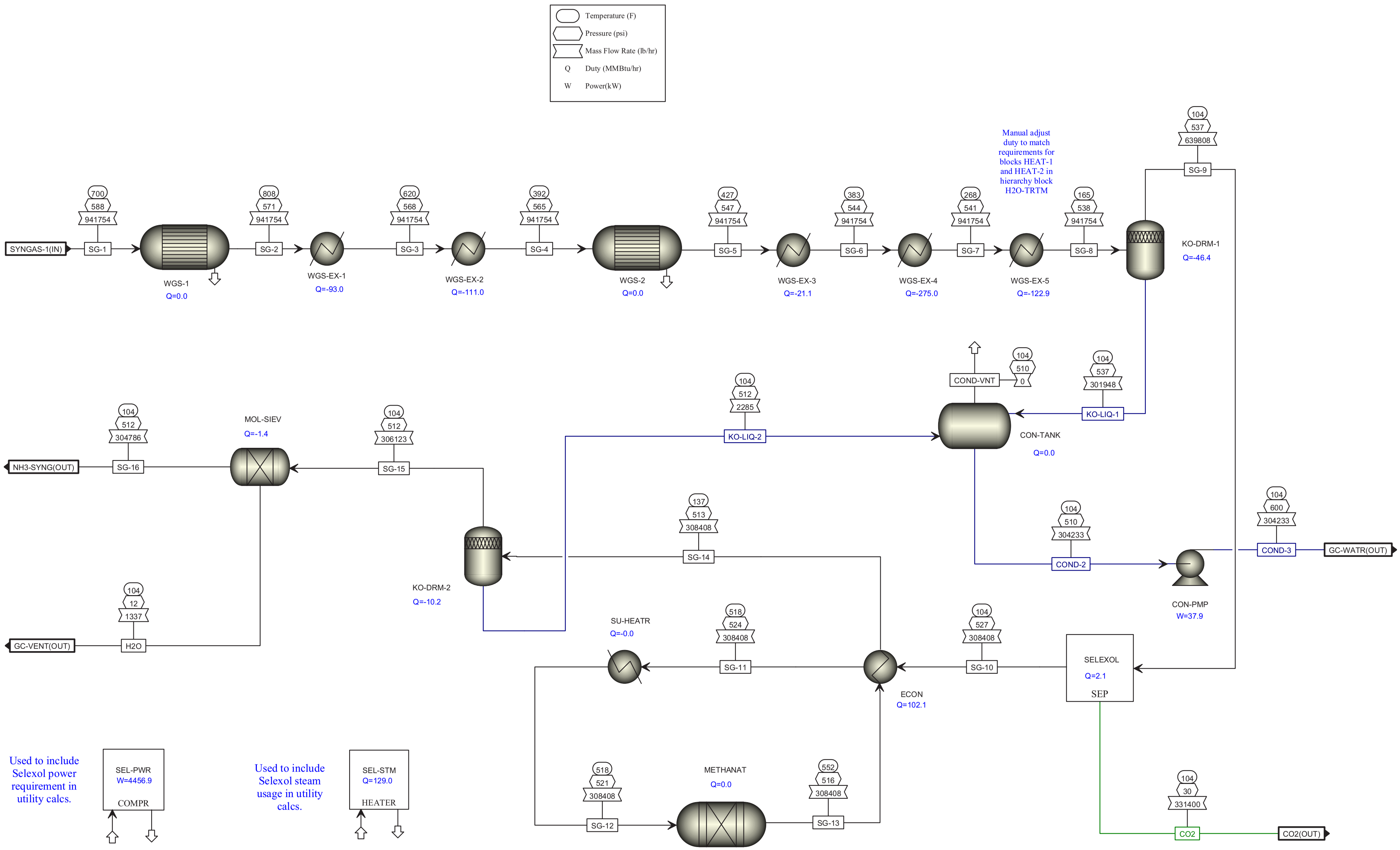
Water or Steam
CO2 Emissions Source
Helium from HTGR



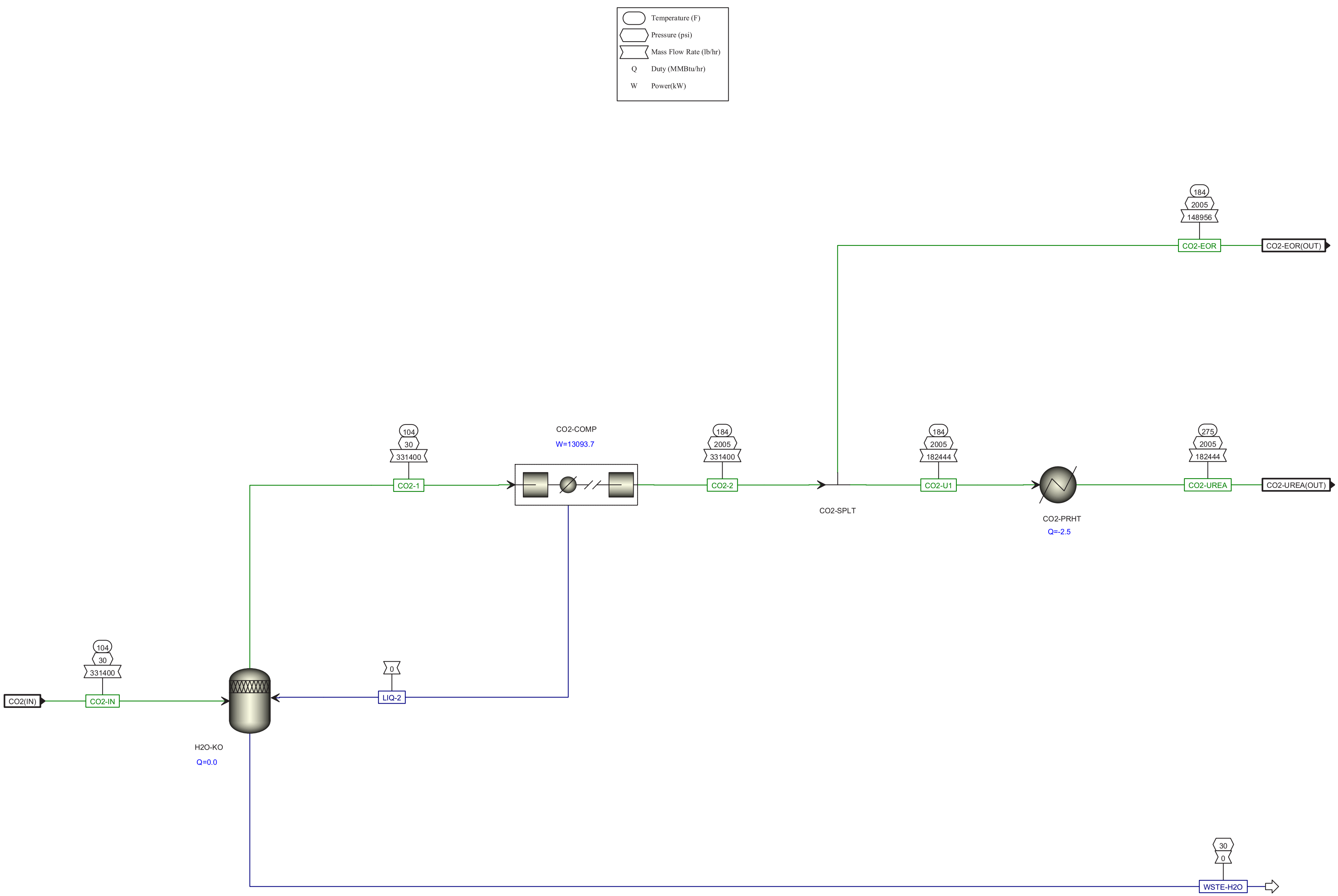
Natural Gas Reforming



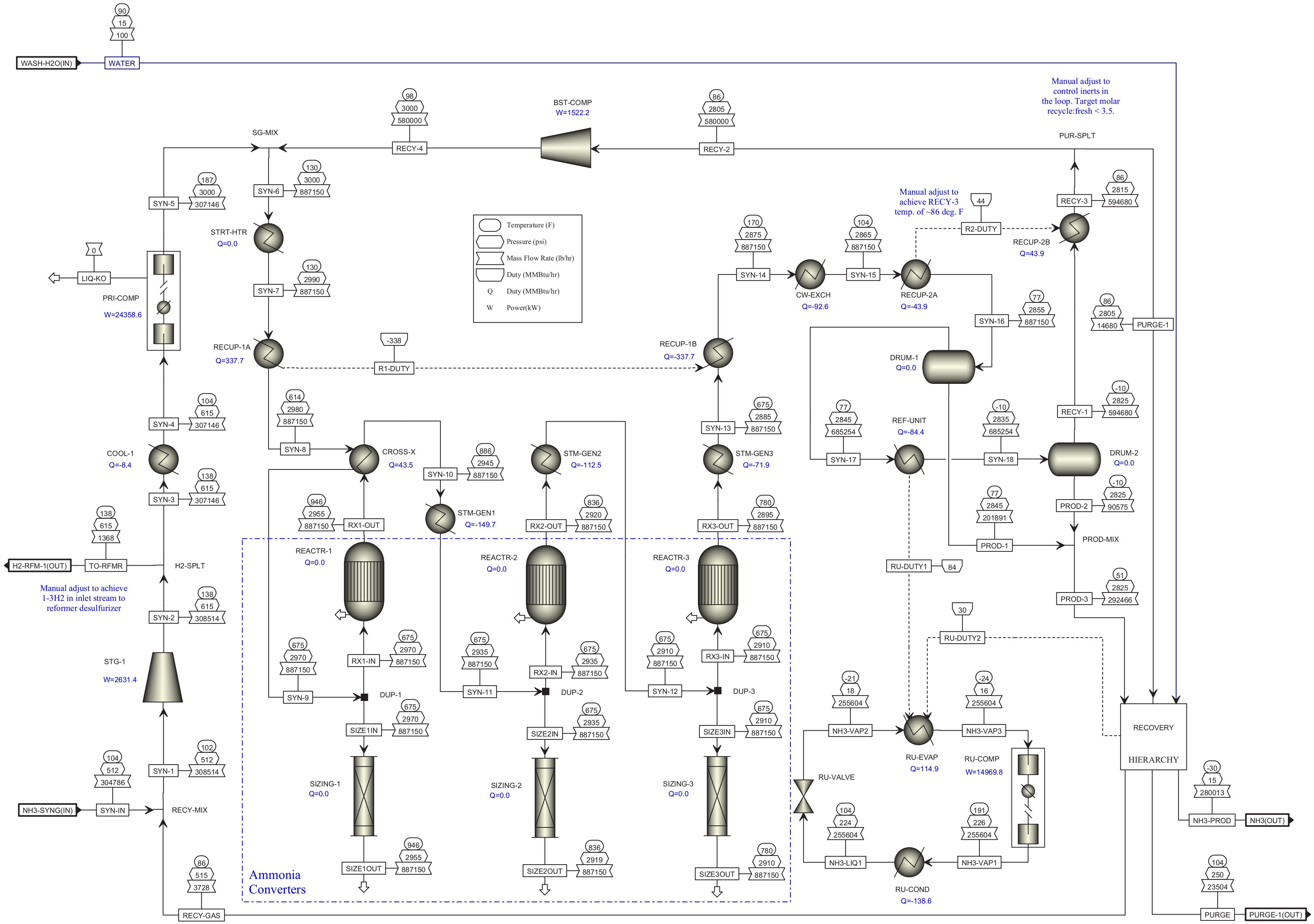
Syngas Cleaning & Conditioning



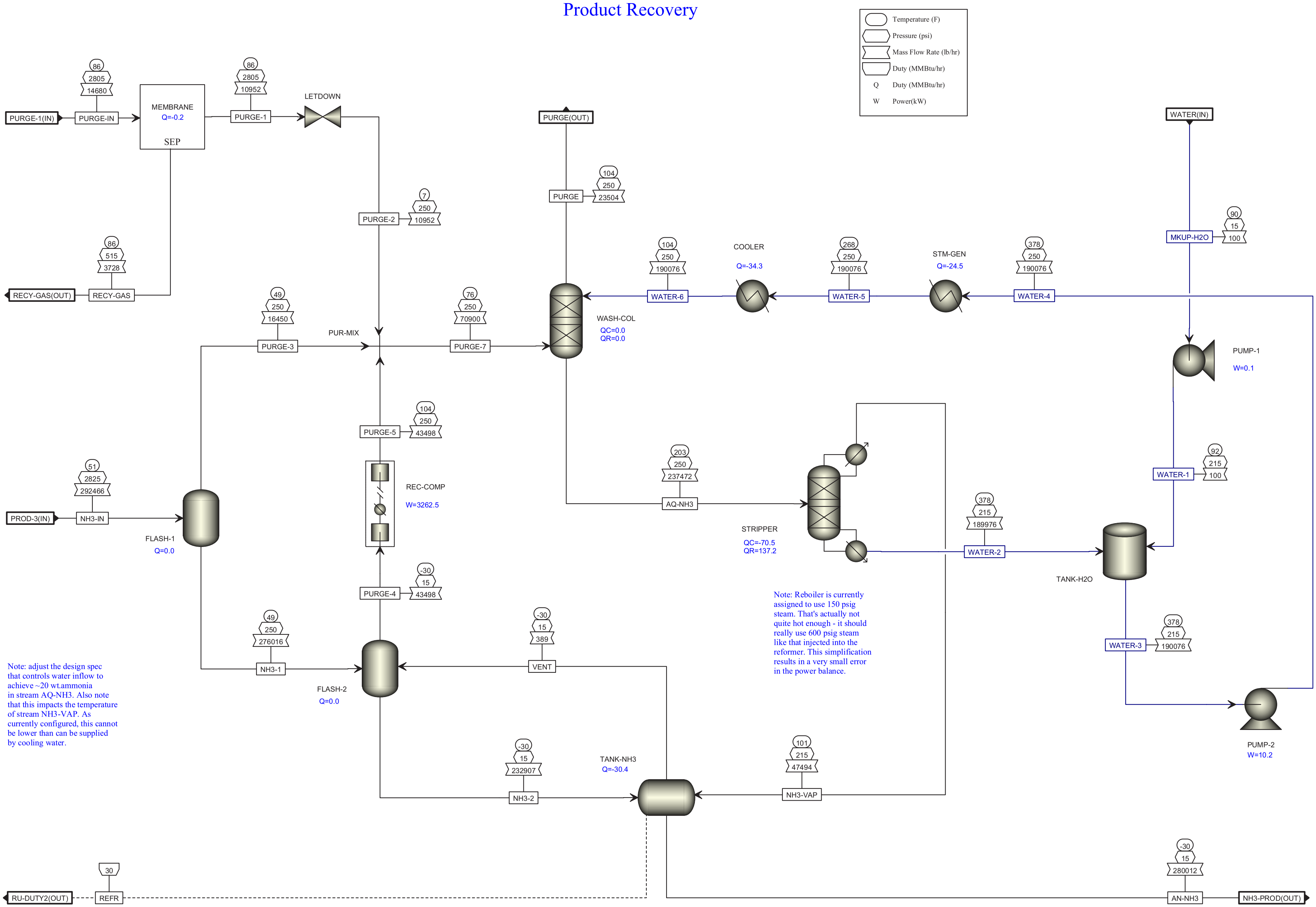
CO2 Compression



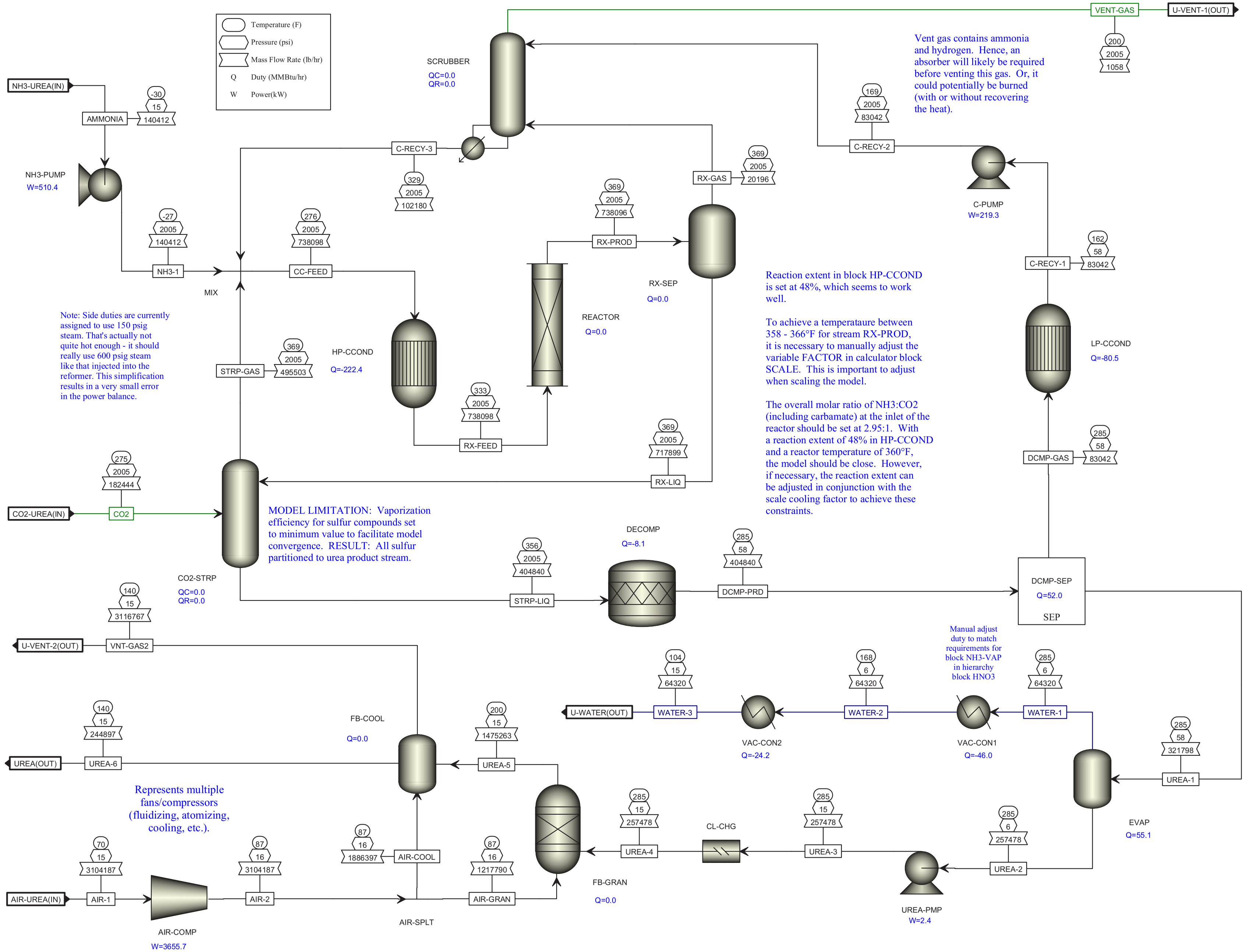
Ammonia Synthesis @ 3000 psi



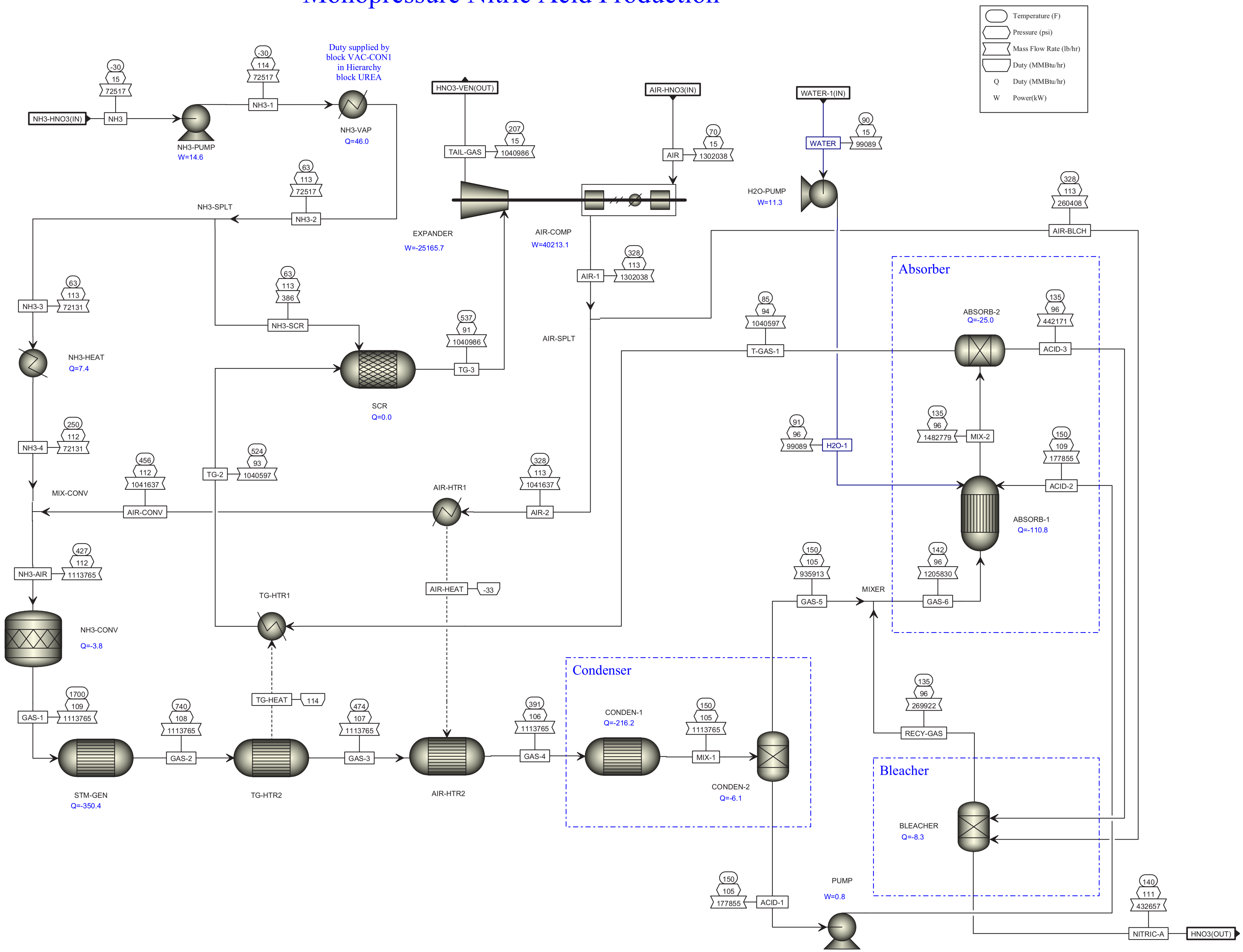
Product Recovery



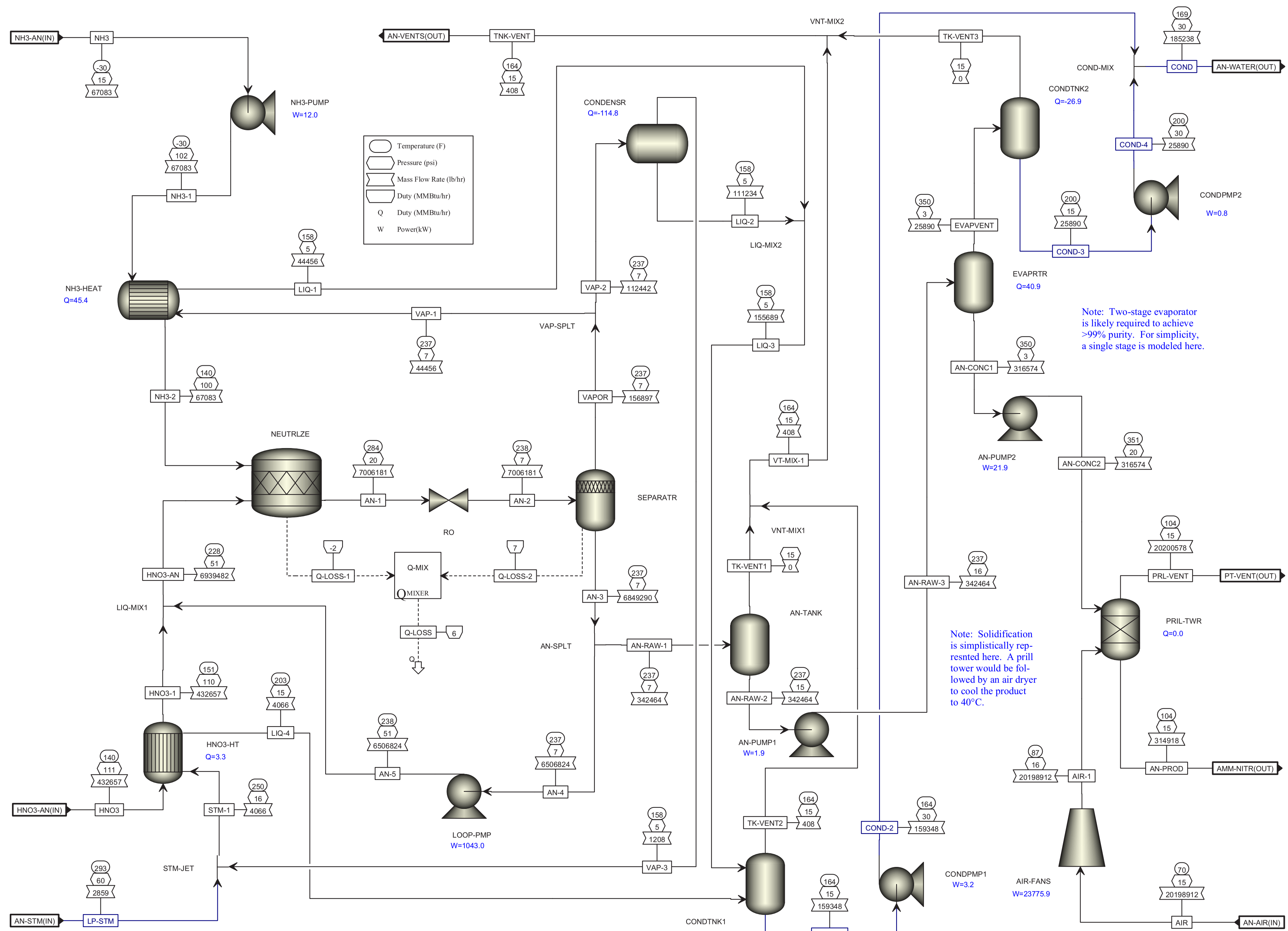
Urea Synthesis via the H.P. Stamicarbon CO2 Stripping Process



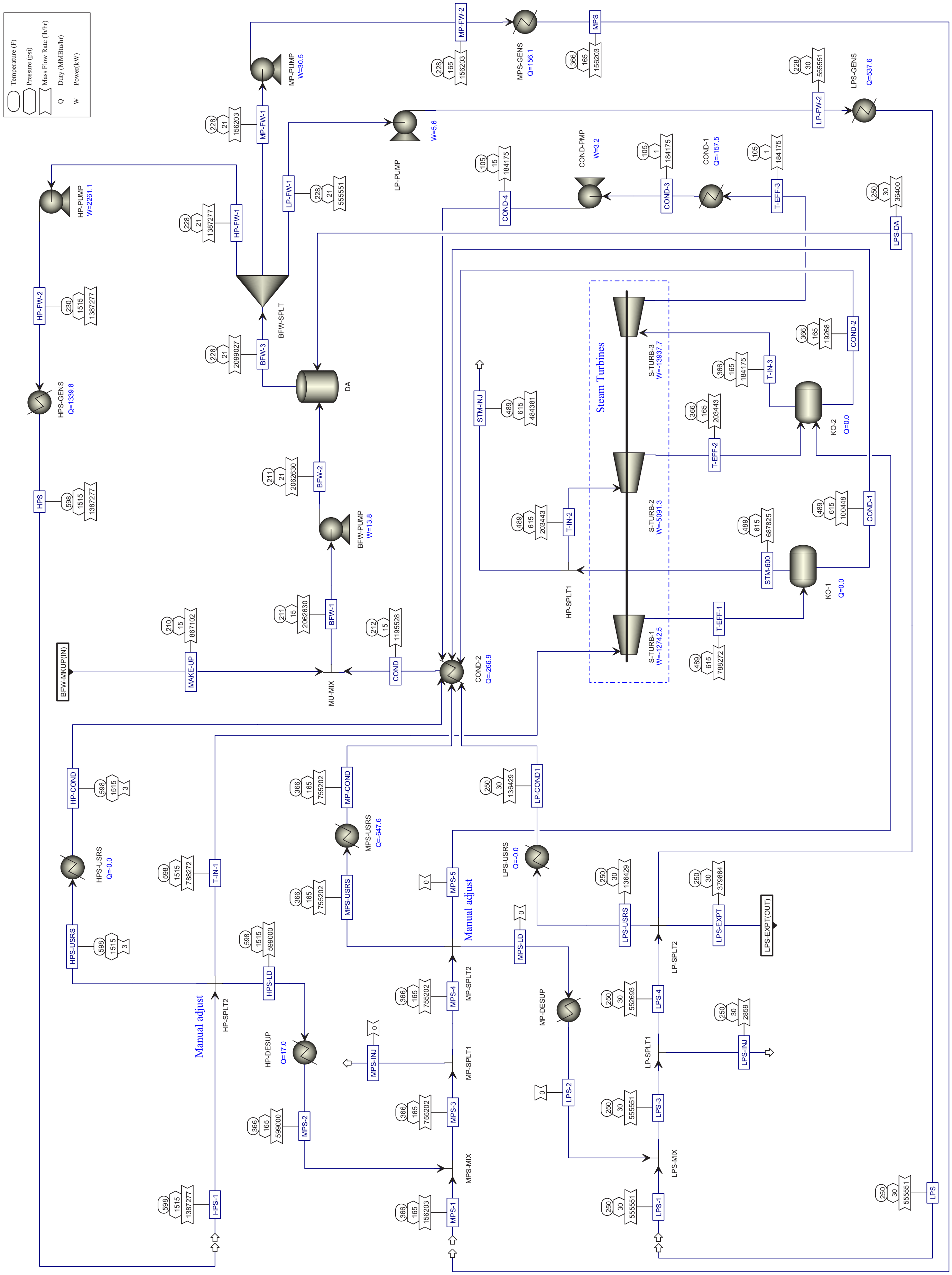
Monopressure Nitric Acid Production



Ammonium Nitrate Production (Uhde-I.G. Farbenindustrie Process)

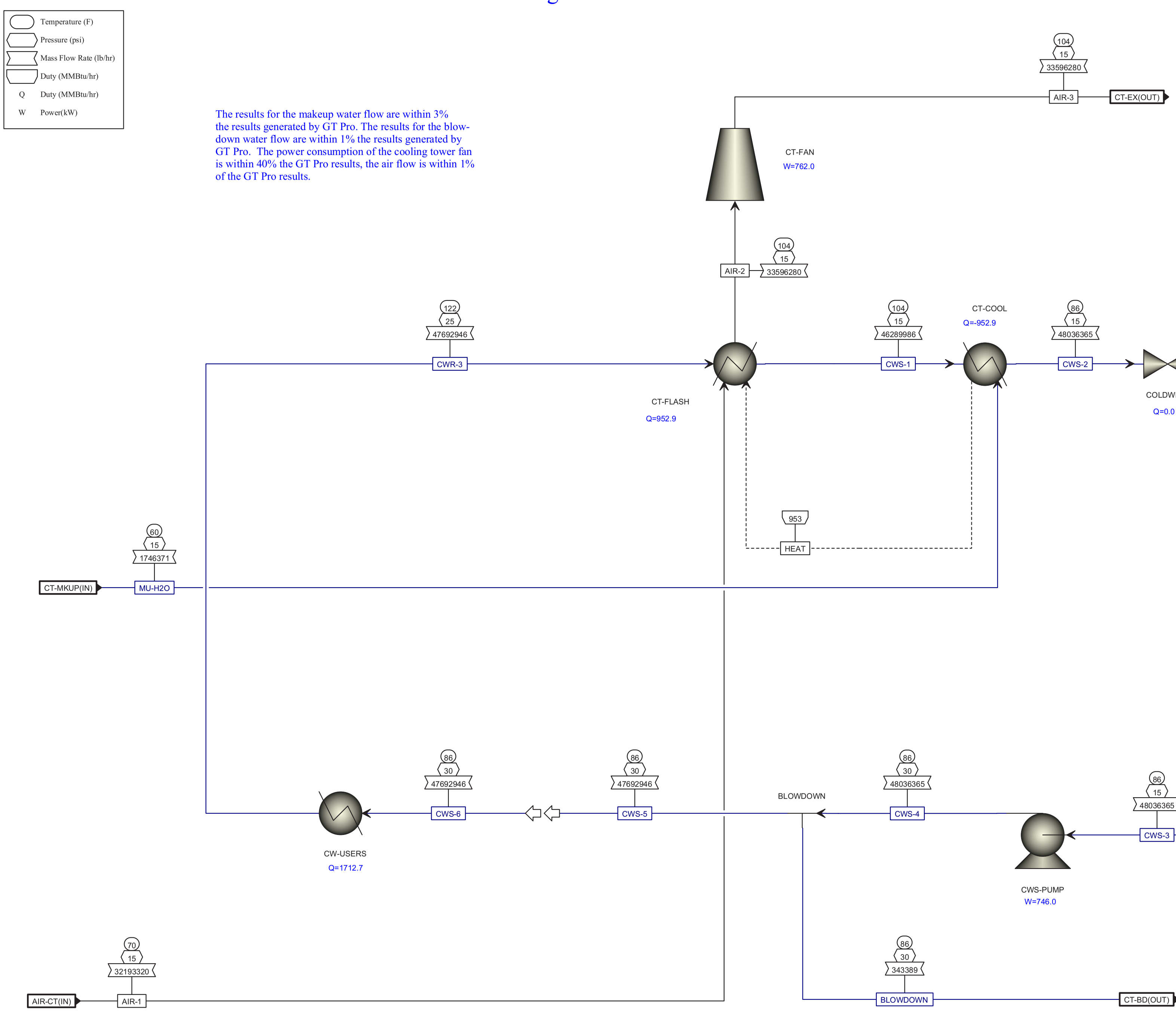


HRSG & Steam Turbines

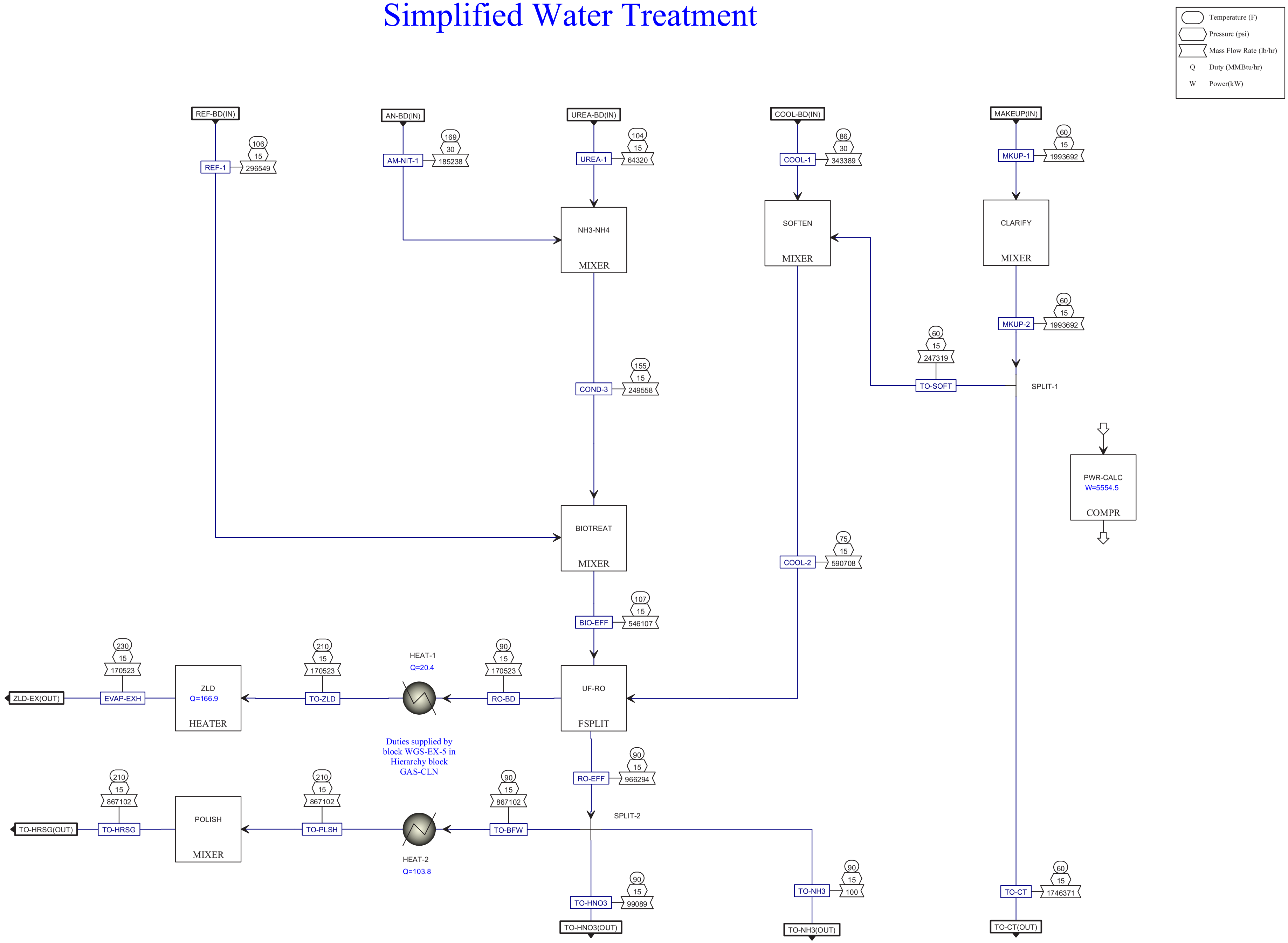


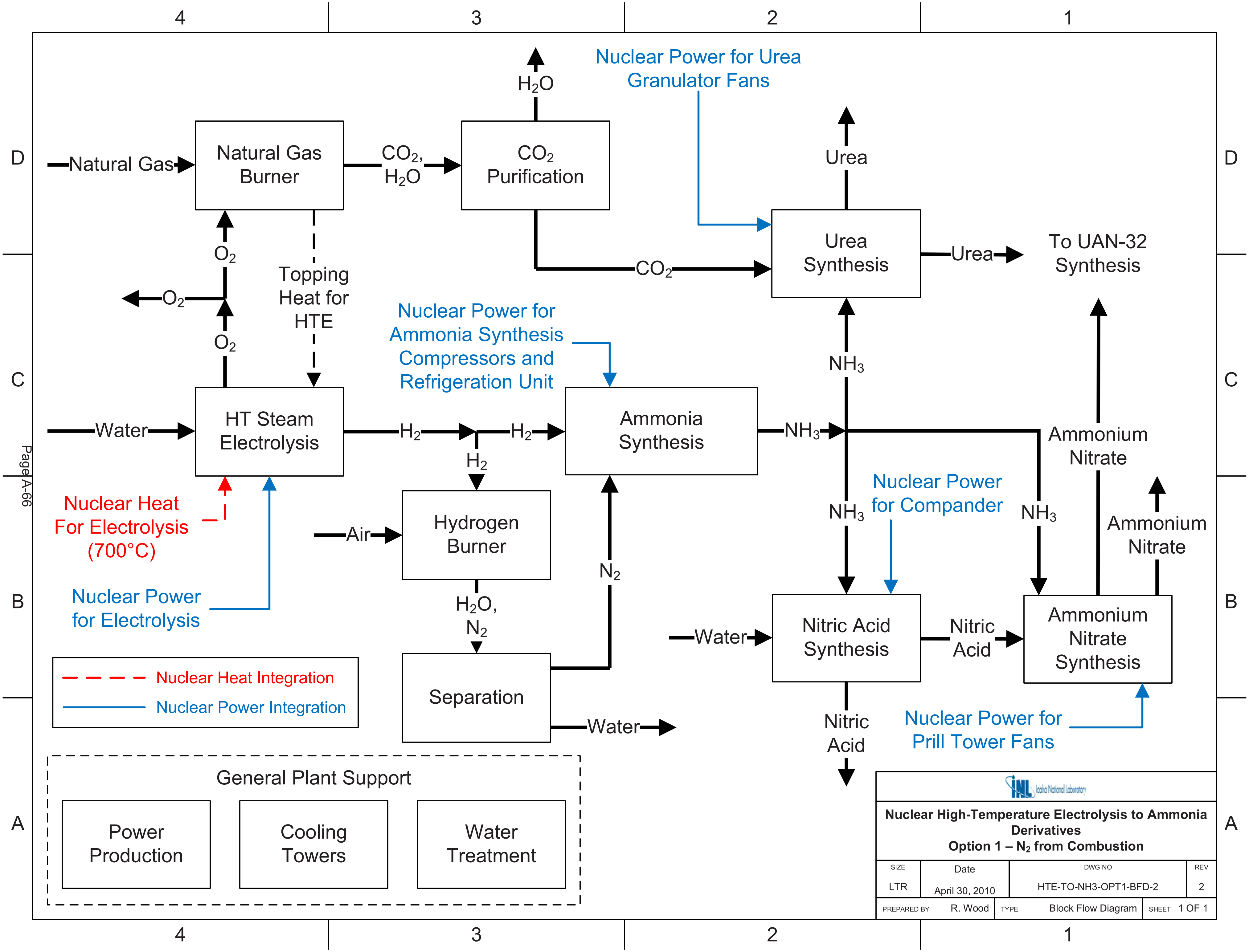
Cooling Tower

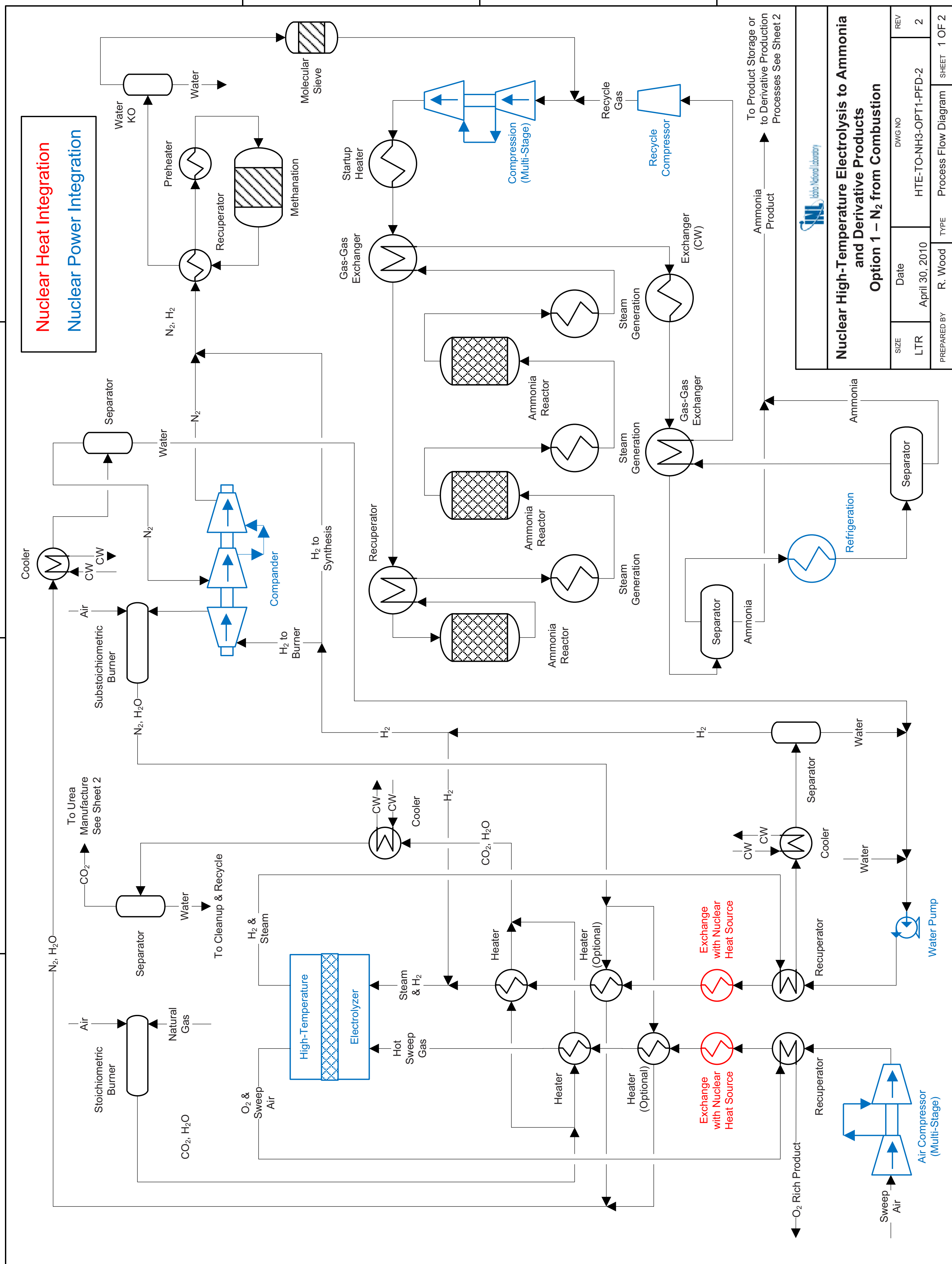
The results for the makeup water flow are within 3% the results generated by GT Pro. The results for the blow-down water flow are within 1% the results generated by GT Pro. The power consumption of the cooling tower fan is within 40% the GT Pro results, the air flow is within 1% of the GT Pro results.



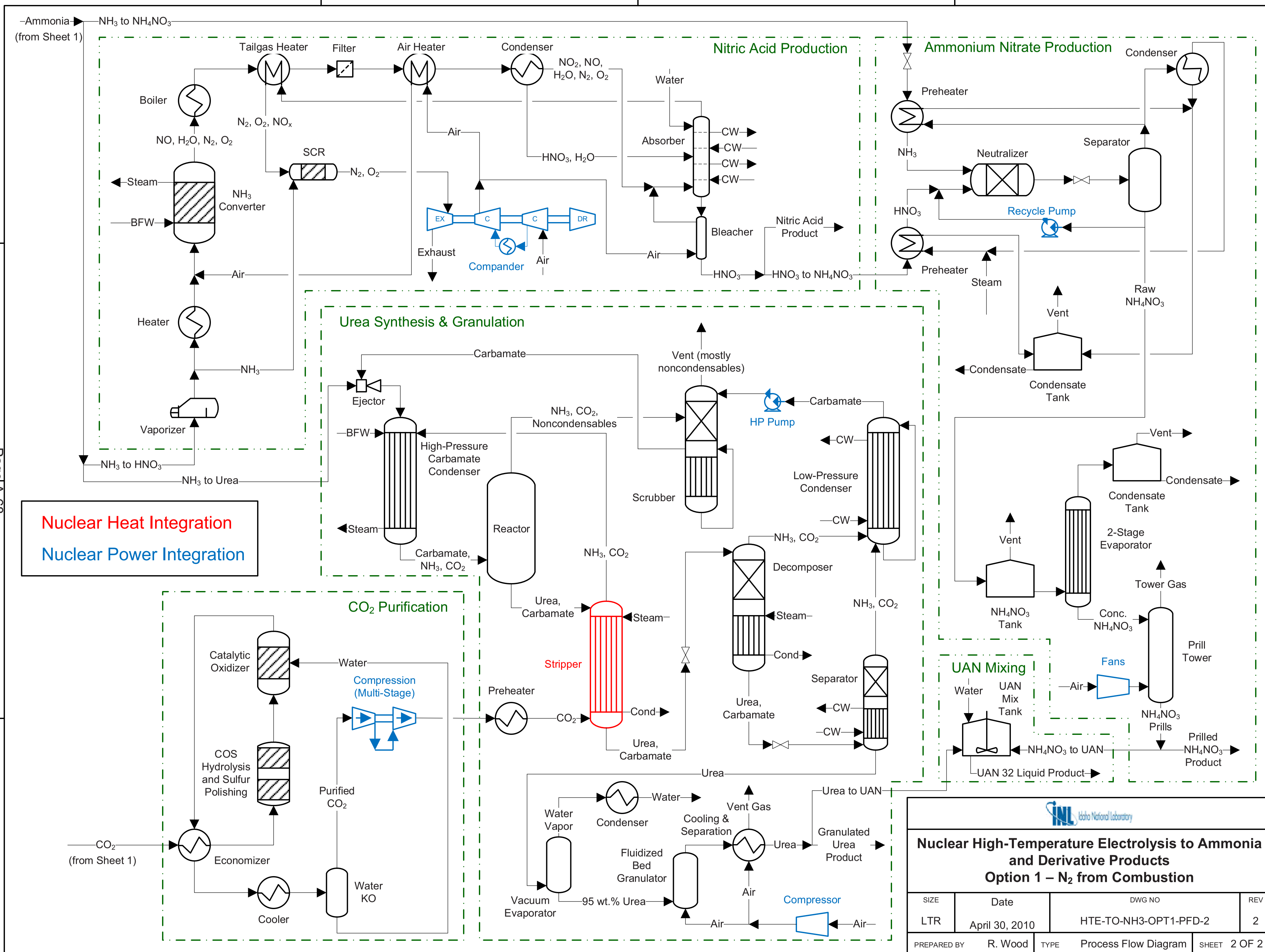
Simplified Water Treatment




Natural Gas
Natural Gas Burner
CO_2, H_2O
CO_2 Purification
H_2O
Nuclear Power for Urea Granulator Fans
Urea
Urea Synthesis
Urea
To UAN-32 Synthesis
O_2
O_2
O_2
Topping Heat for HTE
CO_2
Nuclear Power for Ammonia Synthesis Compressors and Refrigeration Unit
Water
HT Steam Electrolysis
H_2
H_2
H_2
Ammonia Synthesis
NH_3
NH_3
NH_3
NH_3
Ammonium Nitrate
Nuclear Power for Compander
Ammonium Nitrate
Nuclear Heat For Electrolysis (700°C)
Nuclear Power for Electrolysis
Air
Hydrogen Burner
H_2O, N_2
N_2
Water
Nitric Acid Synthesis
Nitric Acid
Ammonium Nitrate Synthesis
Nuclear Heat Integration
Nuclear Power Integration
Separation
Water
Nitric Acid
Nuclear Power for Prill Tower Fans
General Plant Support
Power Production
Cooling Towers
Water Treatment
Idaho National Laboratory
Nuclear High-Temperature Electrolysis to Ammonia Derivatives
Option 1 – N_2 from Combustion
SIZE LTR
Date April 30, 2010
DWG NO HTE-TO-NH3-OPT1-BFD-2
REV 2
PREPARED BY R. Wood
TYPE Block Flow Diagram
SHEET 1 OF 1



Nuclear Heat Integration
Nuclear Power Integration
Nuclear High-Temperature Electrolysis to Ammonia and Derivative Products
Option 1 – N_2 from Combustion
SIZE
LTR
Date
April 30, 2010
DWG NO
HTE-TO-NH3-OPT1-PFD-2
REV
2
PREPARED BY
R. Wood
TYPE
Process Flow Diagram
SHEET
1 OF 2
Idaho National Laboratory
Molecular Sieve
Water KO
Water
Preheater
Recuperator
Methanation
N_2, H_2
Startup Heater
Compression (Multi-Stage)
Recycle Gas
Recycle Compressor
To Product Storage or to Derivative Production Processes See Sheet 2
Ammonia Product
Gas-Gas Exchanger
Exchanger (CW)
Steam Generation
Ammonia Reactor
Separator
Cooler
CW
N_2
Compander
H_2 to Synthesis
H_2 to Burner
Air
Substoichiometric Burner
N_2, H_2O
Recuperator
Refrigeration
Ammonia
H_2
To Urea Manufacture See Sheet 2
CO_2
To Cleanup & Recycle
CO_2, H_2O
H_2 & Steam
Steam & H_2
High-Temperature Electrolyzer
Heater
Heater (Optional)
Exchange with Nuclear Heat Source
Water Pump
Hot Sweep Gas
O_2 & Sweep Air
Stoichiometric Burner
Natural Gas
N_2, H_2O
O_2 Rich Product
Sweep Air
Air Compressor (Multi-Stage)
A
B
C
D
1
2
3
4


Nitric Acid Production
Ammonium Nitrate Production
Urea Synthesis & Granulation
CO$_2$ Purification
UAN Mixing
Nuclear Heat Integration
Nuclear Power Integration
Ammonia (from Sheet 1)
NH$_3$ to NH$_4$NO$_3$
NH$_3$ to HNO$_3$
NH$_3$ to Urea
Tailgas Heater
Filter
Air Heater
Condenser
NO$_2$, NO, H$_2$O, N$_2$, O$_2$
Water
Absorber
CW
Boiler
N$_2$, O$_2$, NO$_x$
NO, H$_2$O, N$_2$, O$_2$
SCR
N$_2$, O$_2$
Steam
BFW
NH$_3$ Converter
HNO$_3$, H$_2$O
Air
EX
C
DR
Exhaust
Compander
Bleacher
Nitric Acid Product
HNO$_3$
HNO$_3$ to NH$_4$NO$_3$
Heater
NH$_3$
Vaporizer
Preheater
Neutralizer
Separator
HNO$_3$
Recycle Pump
Raw NH$_4$NO$_3$
Vent
Condensate
Condensate Tank
Carbamate
Ejector
Vent (mostly noncondensables)
NH$_3$, CO$_2$, Noncondensables
HP Pump
High-Pressure Carbamate Condenser
Scrubber
Low-Pressure Condenser
Reactor
Carbamate, NH$_3$, CO$_2$
NH$_3$, CO$_2$
Decomposer
Urea, Carbamate
Stripper
Cond
Separator
2-Stage Evaporator
Tower Gas
NH$_4$NO$_3$ Tank
Conc. NH$_4$NO$_3$
Prill Tower
Fans
NH$_4$NO$_3$ Prills
Prilled NH$_4$NO$_3$ Product
UAN Mix Tank
NH$_4$NO$_3$ to UAN
UAN 32 Liquid Product
Catalytic Oxidizer
Compression (Multi-Stage)
COS Hydrolysis and Sulfur Polishing
Purified CO$_2$
CO$_2$ (from Sheet 1)
Economizer
Cooler
Water KO
CO$_2$
Urea
Water Vapor
Vacuum Evaporator
95 wt.% Urea
Fluidized Bed Granulator
Cooling & Separation
Vent Gas
Urea to UAN
Granulated Urea Product
Compressor
Nuclear High-Temperature Electrolysis to Ammonia and Derivative Products
Option 1 – N$_2$ from Combustion
SIZE LTR
Date April 30, 2010
DWG NO HTE-TO-NH3-OPT1-PFD-2
REV 2
PREPARED BY R. Wood
TYPE Process Flow Diagram
SHEET 2 OF 2

```
Calculator Block SUMMARY

   FEED SUMMARY:

     NATURAL GAS PROPERTIES:

       MASS FLOW =                            811. TON/DY
       VOLUME FLOW =                           36. MMSCFD @ 60°F
       HHV =                                23063. BTU/LB
       HHV =                                 1044. BTU/SCF @ 60°F
       ENERGY FLOW =                        37405. MMBTU/DY

       COMPOSITION:
          METHANE =                            93.571 MOL.%
          ETHANE =                              3.749 MOL.%
          PROPANE =                             0.920 MOL.%
          BUTANE =                              0.260 MOL.%
          PENTANE =                             0.040 MOL.%
          HEXANE =                              0.010 MOL.%
          NITROGEN =                            1.190 MOL.%
          OXYGEN =                              0.010 MOL.%
          CO2 =                                 0.250 MOL.%
          C4H10S =                              1. PPMV
          C2H6S =                               0. PPMV
          H2S =                                 0. PPMV

   INTERMEDIATE PRODUCT SUMMARY:

     RAW SYNGAS MASS FLOW =                287339. LB/HR
     RAW SYNGAS VOLUME FLOW =                 303. MMSCFD @ 60°F
     RAW SYNGAS COMPOSITION:
       H2                                     74.7 MOL.%
       CO                                      0.0 MOL.%
       CO2                                     0.0 MOL.%
       N2                                     24.9 MOL.%
       AR                                      0.3 MOL.%
       H2O                                     0.1 MOL.%
       CH4                                     0.0 MOL.%

     CLEANED SYNGAS MASS FLOW =            286569. LB/HR
     CLEANED SYNGAS VOLUME FLOW =             303. MMSCFD @ 60°F
     CLEANED SYNGAS COMPOSITION:
       H2                                     74.8 MOL.%
       N2                                     24.9 MOL.%
       AR                                      0.3 MOL.%
       CH4                                     0.0 MOL.%
       CO                                      0. PPMV
       CO2                                     0. PPMV
       H2O                                     0. PPMV

     AMMONIA MASS FLOW =                   280019. LB/HR
     AMMONIA MASS FLOW =                     3360. TON/DY
       AMMONIA PURITY =                        99.99 MOL.%

     NITRIC ACID MASS FLOW =               432644. LB/HR
     NITRIC ACID MASS FLOW =                 5192. TON/DY
       NITRIC ACID PURITY =                    57.93 WT.%
     HNO3 PRODUCED / NH3 FED =                  1.55 LB/LB

   FINAL PRODUCT SUMMARY:

     AMMONIA MASS FLOW =                       0. LB/HR
     AMMONIA MASS FLOW =                       0. TON/DY
       AMMONIA PURITY =                        99.99 MOL.%
```

```
  NITRIC ACID MASS FLOW =                        0. LB/HR
  NITRIC ACID MASS FLOW =                        0. TON/DY
    NITRIC ACID PURITY =                     57.93 WT.%

  UREA MASS FLOW =                      244883. LB/HR
  UREA MASS FLOW =                        2939. TON/DY
  UREA PRODUCED / AMMONIA FED =              1.74 LB/LB
  UREA PRODUCED / CO2 FED =                  1.33 LB/LB

  AMMONIUM NITRATE MASS FLOW =          314908. LB/HR
  AMMONIUM NITRATE MASS FLOW =            3779. TON/DY
  NH4NO3 PRODUCED / NITRIC ACID FED =        0.73 LB/LB
  NH4NO3 PRODUCED / AMMONIA FED =            4.69 LB/LB


POWER SUMMARY:

  ELECTRICAL GENERATORS:
    STEAM TURBINE POWER GENERATION =     159.1 MW
  GENERATOR SUBTOTAL =                   159.1 MW

  ELECTRICAL CONSUMERS:
    N2 GENERATION POWER CONSUMPTION =     18.8 MW
    HTE POWER CONSUMPTION =              918.1 MW
    POWER BLOCK POWER CONSUMPTION =        3.1 MW
    CO2 PROCESSING POWER CONSUMPTION =     5.6 MW
    AMMONIA SYNTH. POWER CONSUMPTION =    39.0 MW
    UREA SYNTHESIS POWER CONSUMPTION =     4.4 MW
    HNO3 SYNTHESIS POWER CONSUMPTION =    15.1 MW
    NH4NO3 SYNTH. POWER CONSUMPTION =     24.9 MW
    COOLING TOWER POWER CONSUMPTION =      2.5 MW
    WATER TREATMENT POWER CONSUMPTION =    6.6 MW
  CONSUMER SUBTOTAL =                   1037.9 MW

  NET PLANT POWER CONSUMPTION =          878.8 MW


WATER BALANCE:

  EVAPORATIVE LOSSES:
    COOLING TOWER EVAPORATION =         5595.4 GPM
    ZLD SYSTEM EVAPORATION =             383.1 GPM
  TOTAL EVAPORATIVE LOSSES =            5978.5 GPM

  WATER CONSUMED:
    BOILER FEED WATER MAKEUP =           915.0 GPM
    NITRIC ACID FEED WATER =             198.0 GPM
    COOLING TOWER MAKEUP =              5890.0 GPM
    HTE WATER REQUIREMENT =             1057.8 GPM
    AMMONIA WASH WATER REQUIREMENT =       0.1 GPM
  TOTAL WATER CONSUMED =                8060.9 GPM

  WATER GENERATED:
    CO2 COMPRESSION BLOWDOWN =           288.7 GPM
    UREA PRODUCTION BLOWDOWN =           128.9 GPM
    NH4NO3 PRODUCTION BLOWDOWN =         370.2 GPM
    COOLING TOWER BLOWDOWN =            1158.1 GPM
    H2 BURNER WATER BLOWDOWN =           164.1 GPM
  TOTAL WATER GENERATED =               2110.0 GPM

  PLANT WATER SUMMARY:
    NET MAKEUP WATER REQUIRED =         6334.0 GPM
```

```
   CO2 BALANCE:

     CO2 EMITTED (TOTAL) =                  11. TON/DY
     CO2 EMITTED (TOTAL) =                   0.2 MMSCFD @ 60˚F


   NUCLEAR INTEGRATION REQUIREMENTS:

     TOTAL ELECTRICITY DEMAND =            878.8 MW
       FROM HTE =                          918.1 MW
       FROM BALANCE OF PLANT =             -39.3 MW

     HYDROGEN DEMAND =                     715.1 TON/DY
       PURITY =                             99.9159 MOL.%

     TOTAL OXYGEN PRODUCTION =            5636.8 TON/DY
       FROM HTE =                         5636.8 TON/DY
         PURITY =                           99.8910 MOL.%
       PLANT O2 DEMAND =                  3131.1 TON/DY
       EXCESS O2 FOR SALE =               2505.7 TON/DY

     TOPPING HEAT SUPPLIED =                50.4 MMBTU/HR
       SUPPLY TEMPERATURE =               1874. ˚F
       RETURN TEMPERATURE =               1825. ˚F


Calculator Block AIRPROPS

   HUMIDITY DATA FOR STREAM PRI-AIR:
     HUMIDITY RATIO =                       43.5 GRAINS/LB

     RELATIVE HUMIDITY =                    39.9 %

Calculator Block REACTOR  Hierarchy: AMMONIA

   AMMONIA CATALYST BED SIZING CALCULATIONS
   (BASED ON IRON CATALYST KINETICS):

     REACTOR 1:
       DIAMETER =                10.84 FT
       LENGTH =                  54.19 FT
       VOLUME =                5000. FT3

     REACTOR 2:
       DIAMETER =                18.37 FT
       LENGTH =                  91.84 FT
       VOLUME =               24333. FT3

     REACTOR 3:
       DIAMETER =                22.98 FT
       LENGTH =                 114.87 FT
       VOLUME =               47630. FT3

     TOTAL CATALYST REQUIRED =  76963. FT3


Calculator Block PH       Hierarchy: NH4NO3

   TARGET NEUTRALIZER PH =   1.500
   ACTUAL NEUTRALIZER PH =   1.499


Calculator Block RATIO    Hierarchy: UREA
```

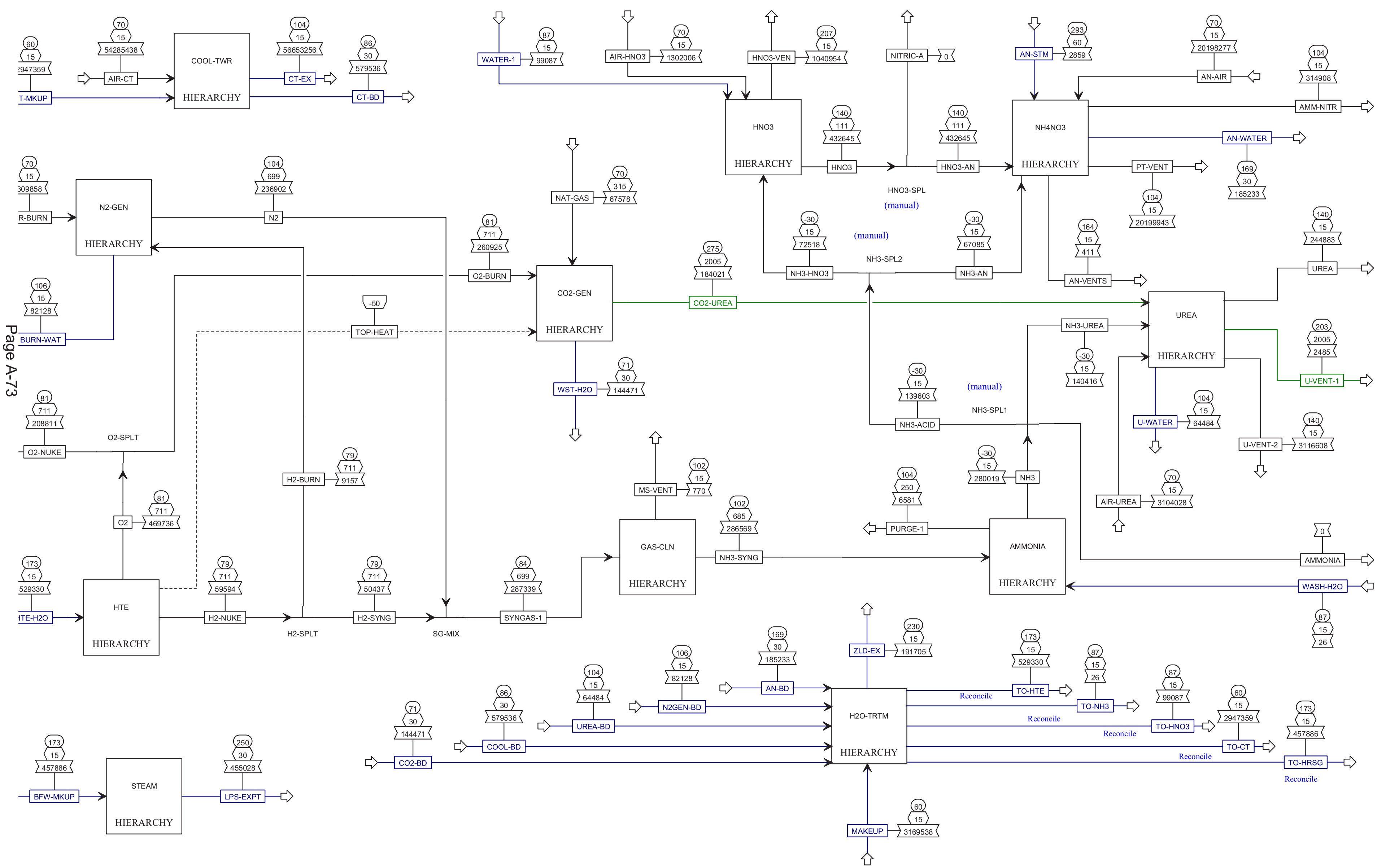
HTSE-to-Ammonia Results - Option 1

```
TARGET REACTOR NH3/CO2 RATIO =   2.950
ACTUAL REACTOR NH3/CO2 RATIO =   2.971
```

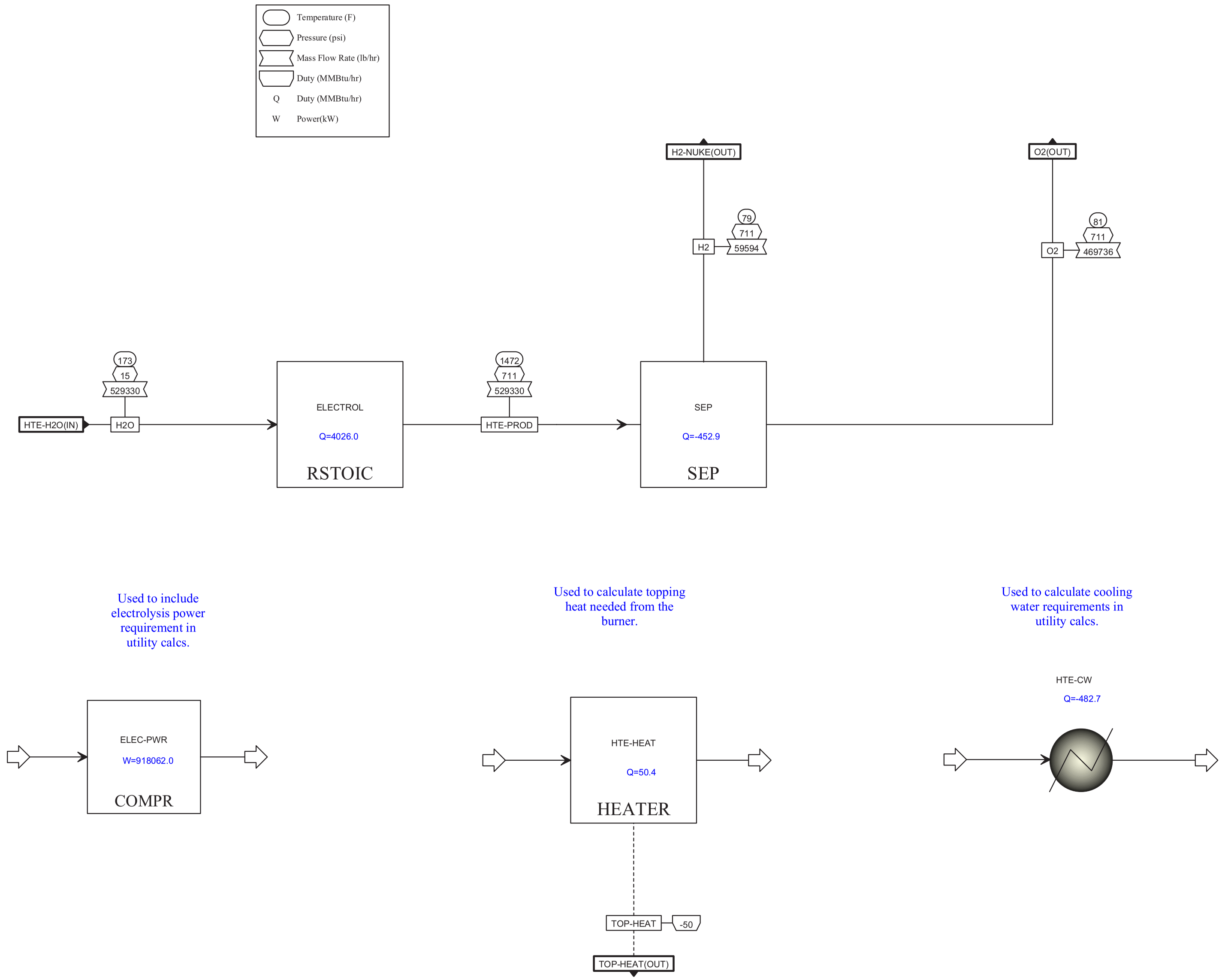
Temperature (F)
Pressure (psi)
Mass Flow Rate (lb/hr)
Duty (MMBtu/hr)

High Temperature Steam Electrolysis to Ammonia and Derivatives
N2 from H2 Combustion
(Derivatives produced in the correct ratio to make UAN32)

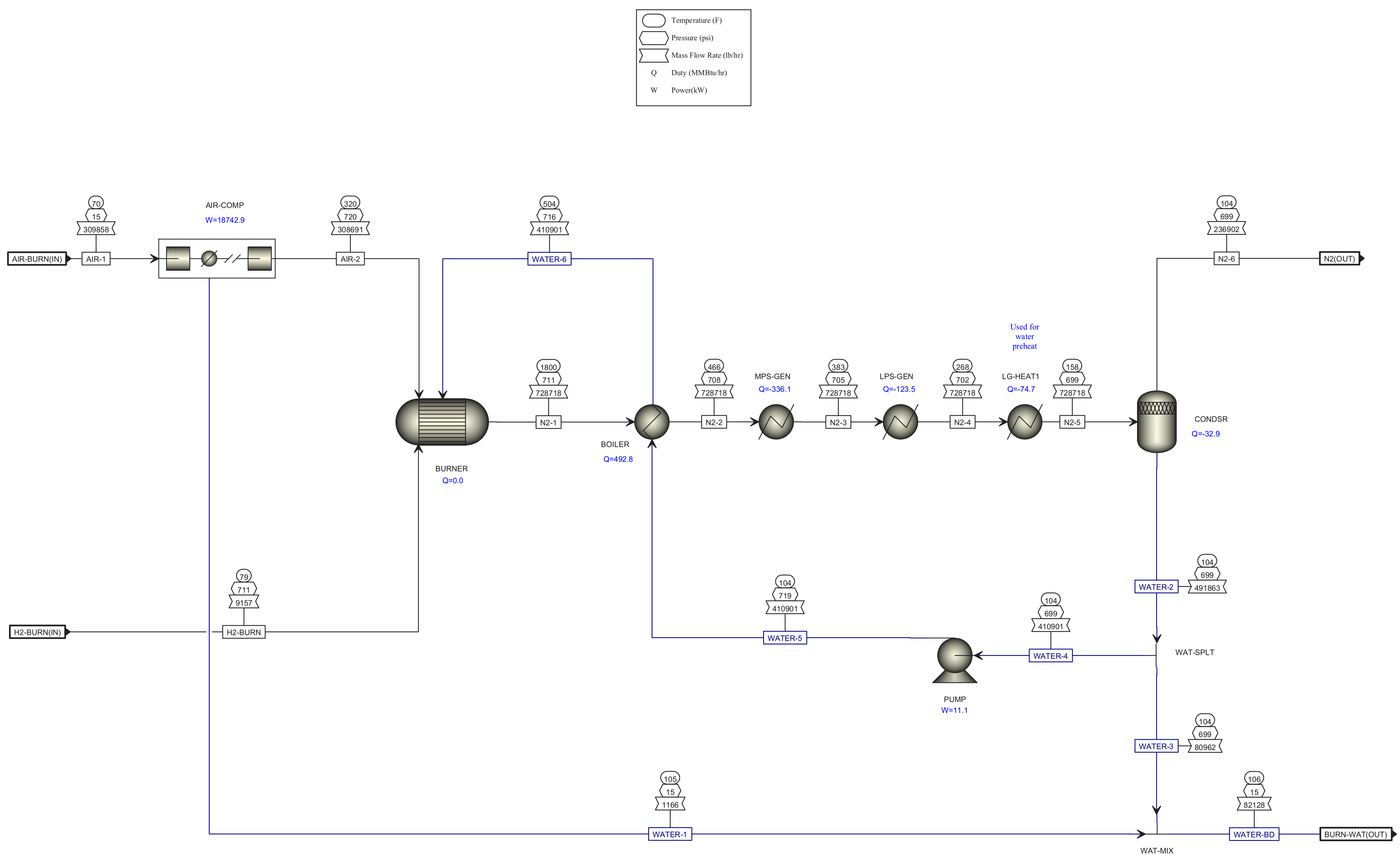
Color Legend
Water or Steam
CO2 Emissions Source



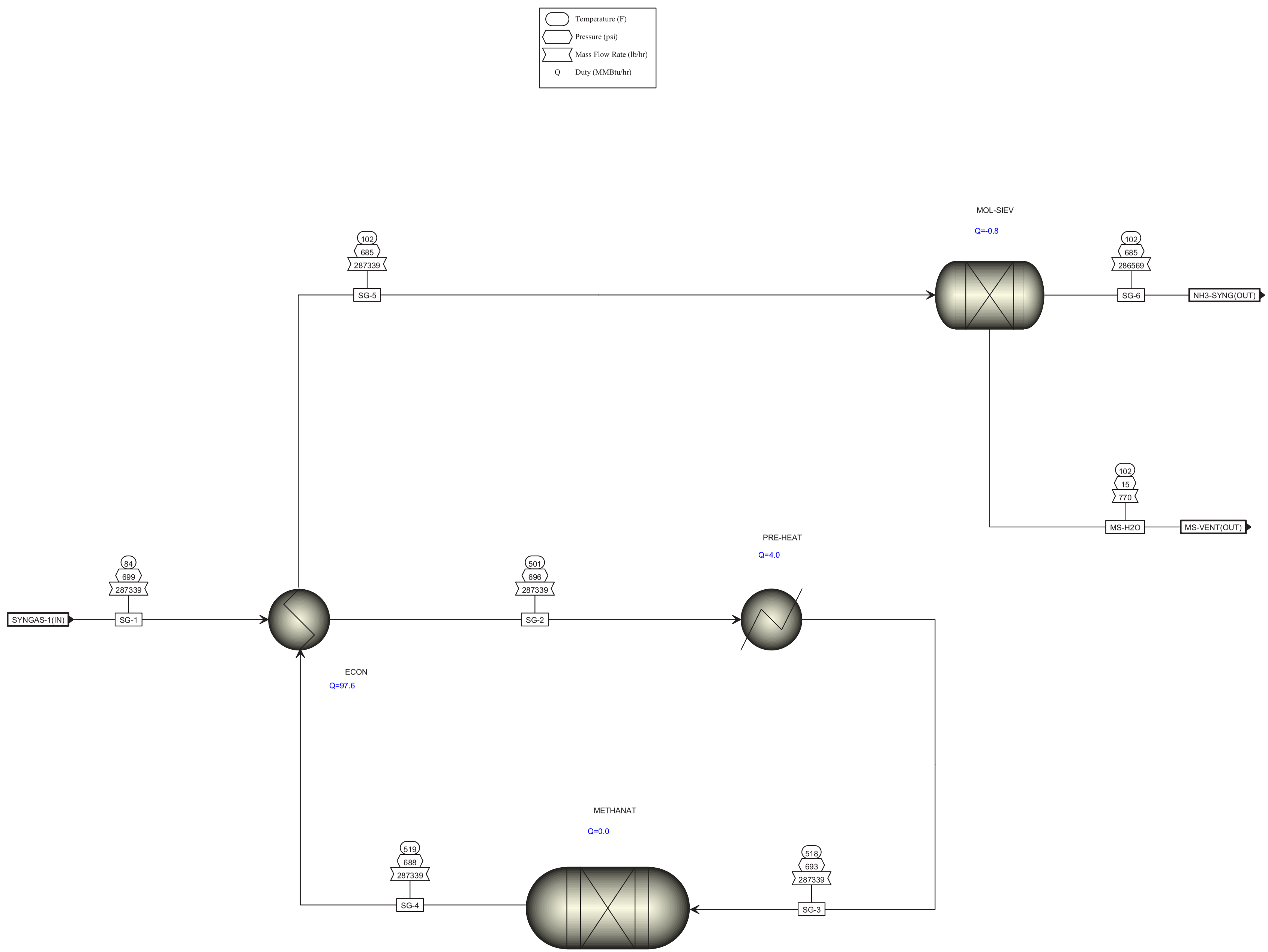
Simplified HTE Model



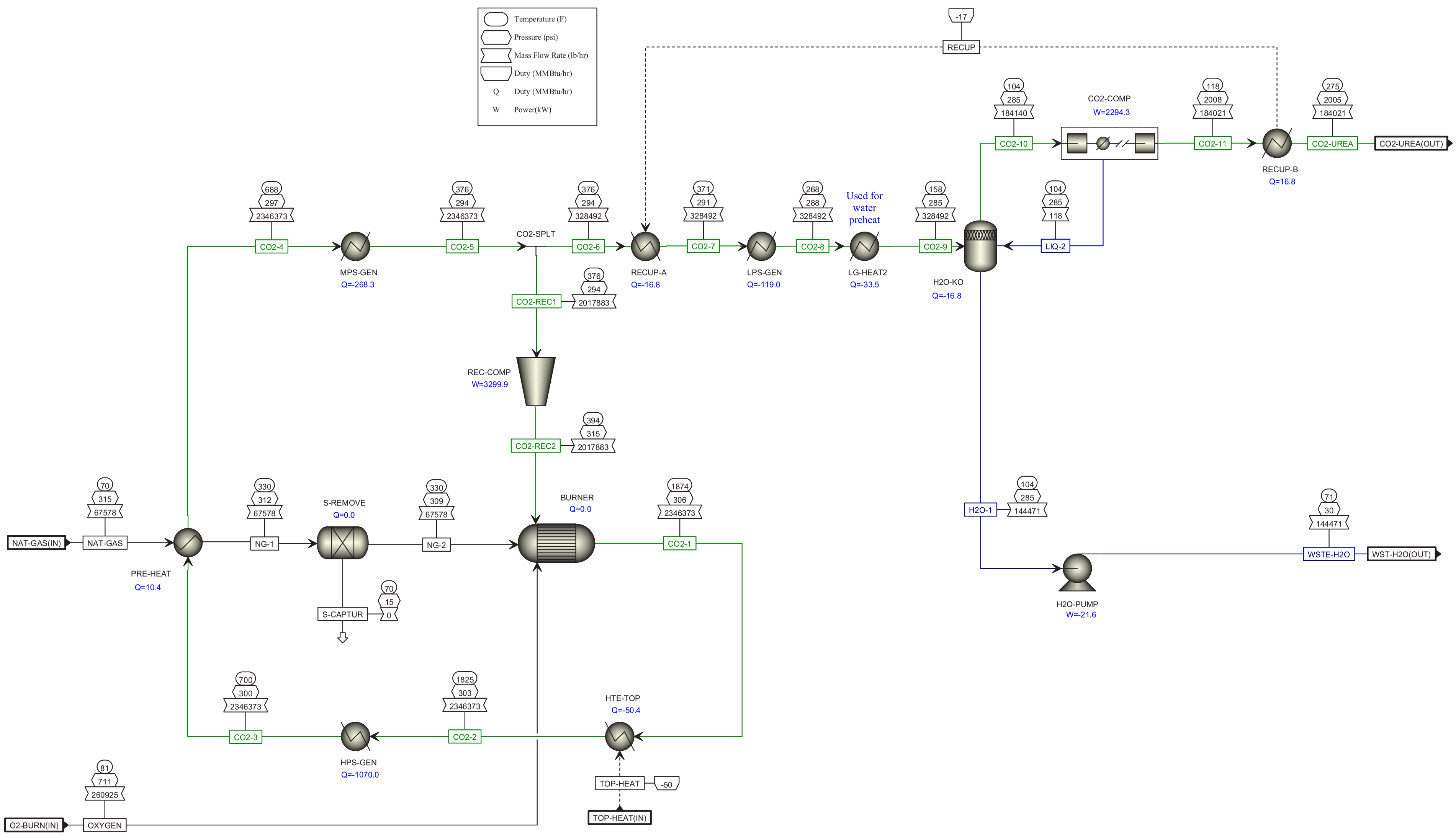
N2 Generation via H2-Air Combustion



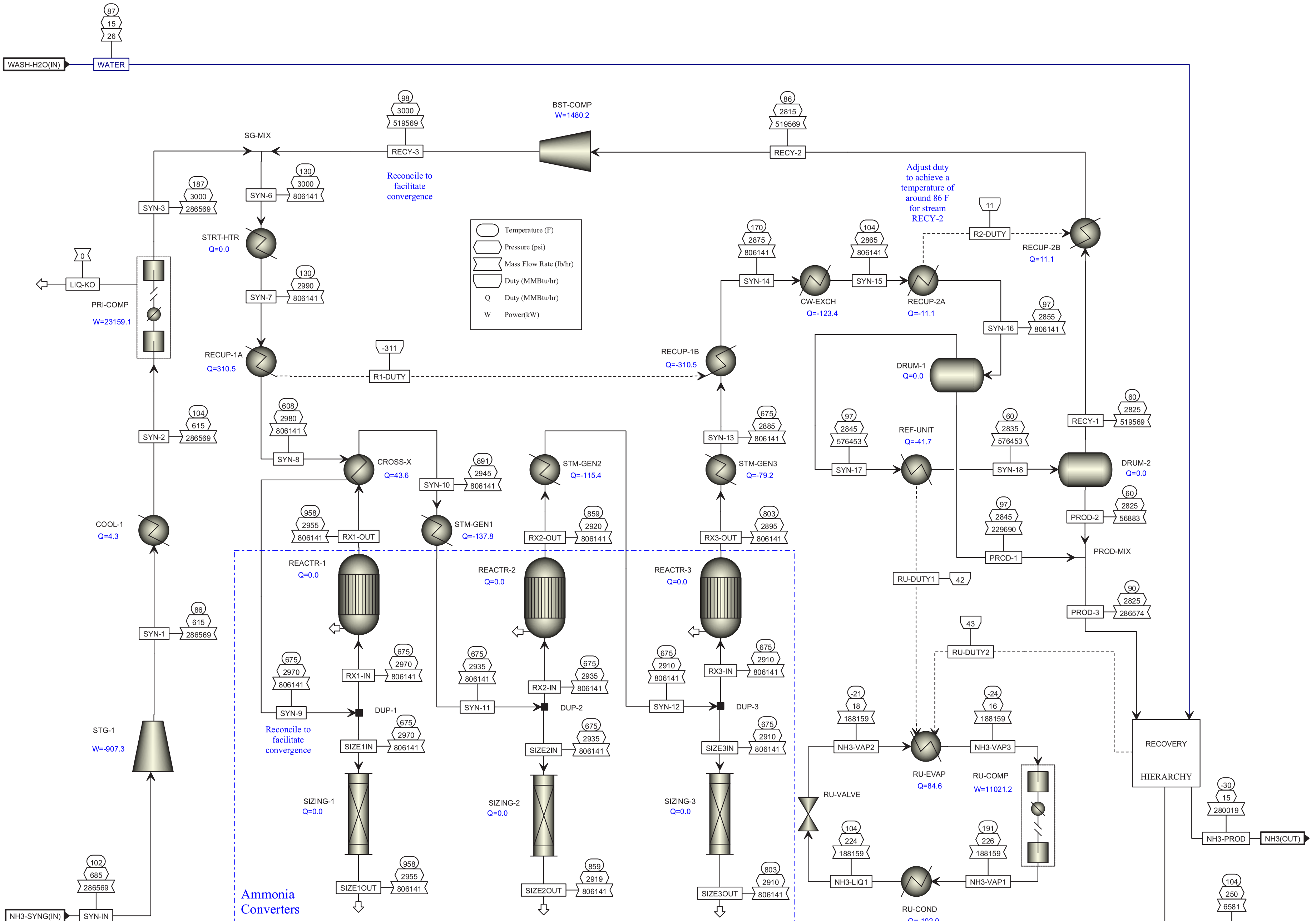
Syngas Cleaning & Conditioning



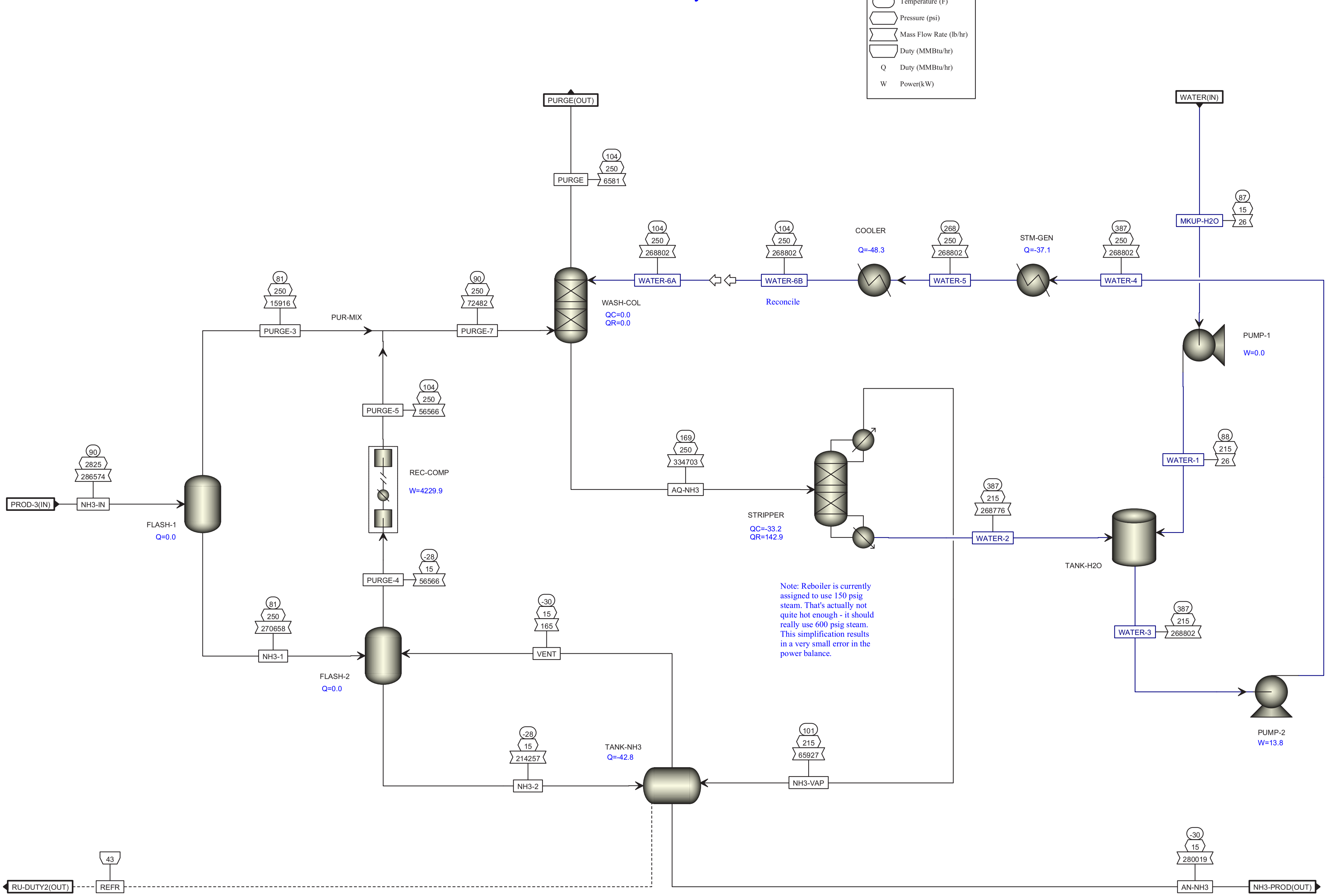
CO2 Generation



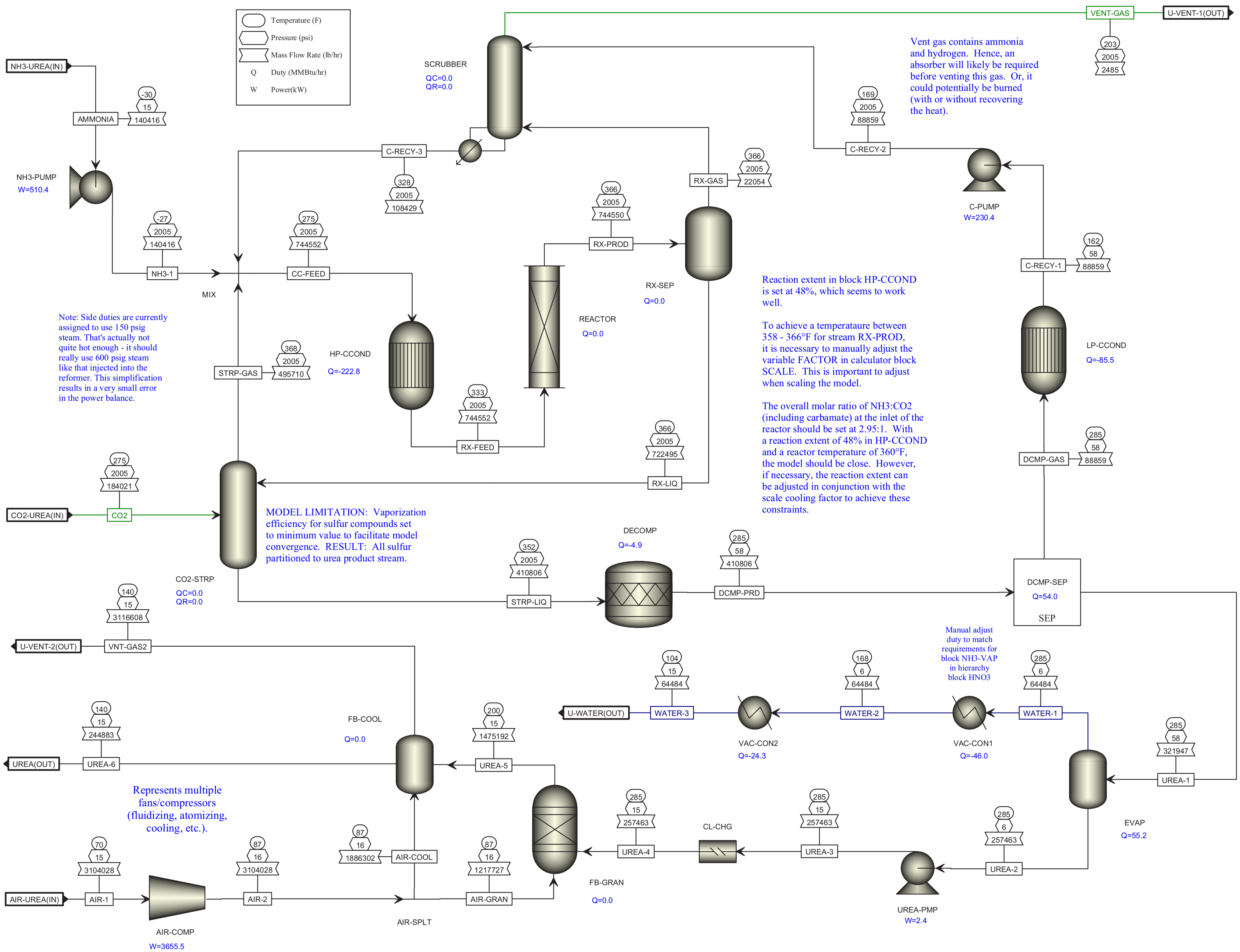
Ammonia Synthesis @ 3000 psi



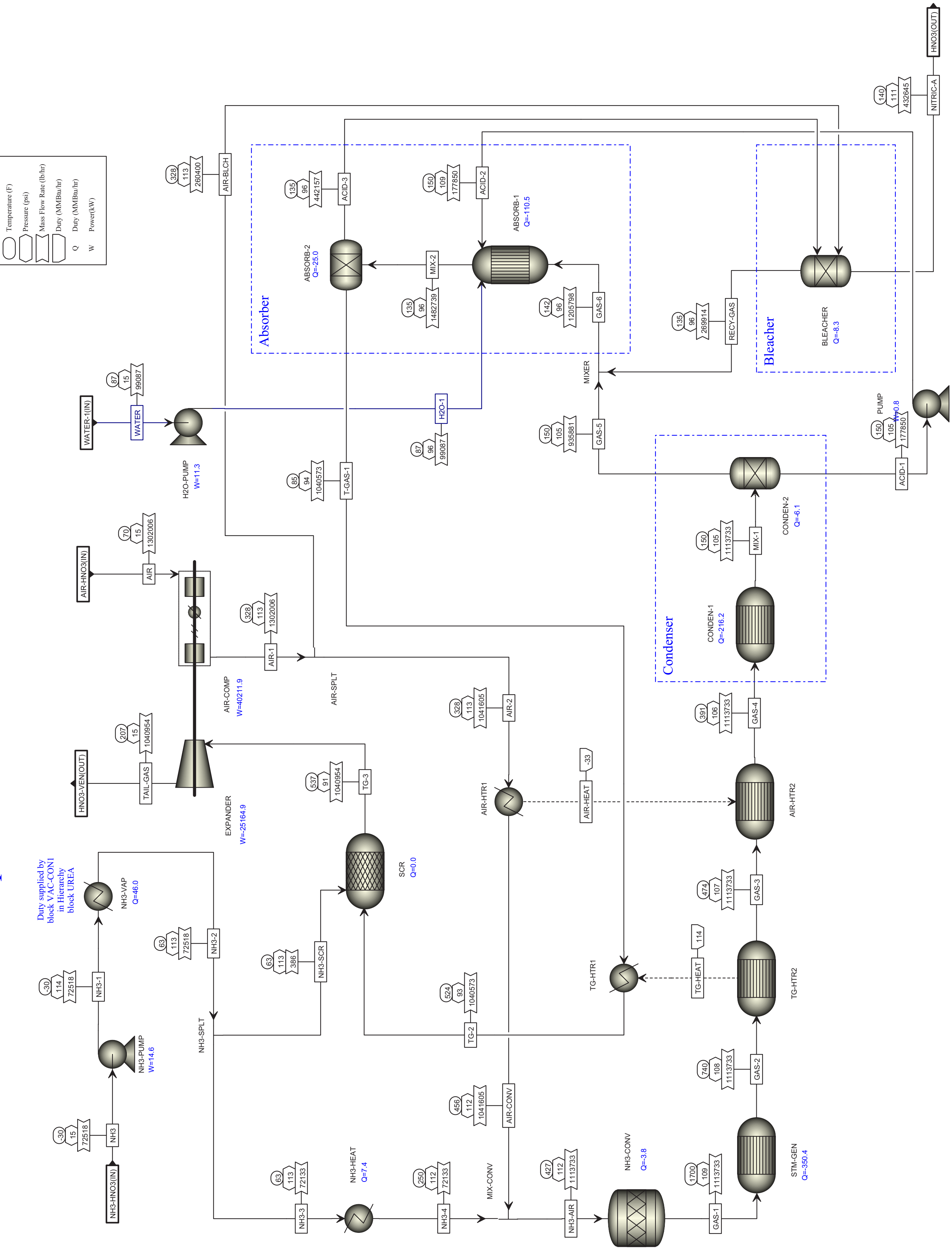
Product Recovery



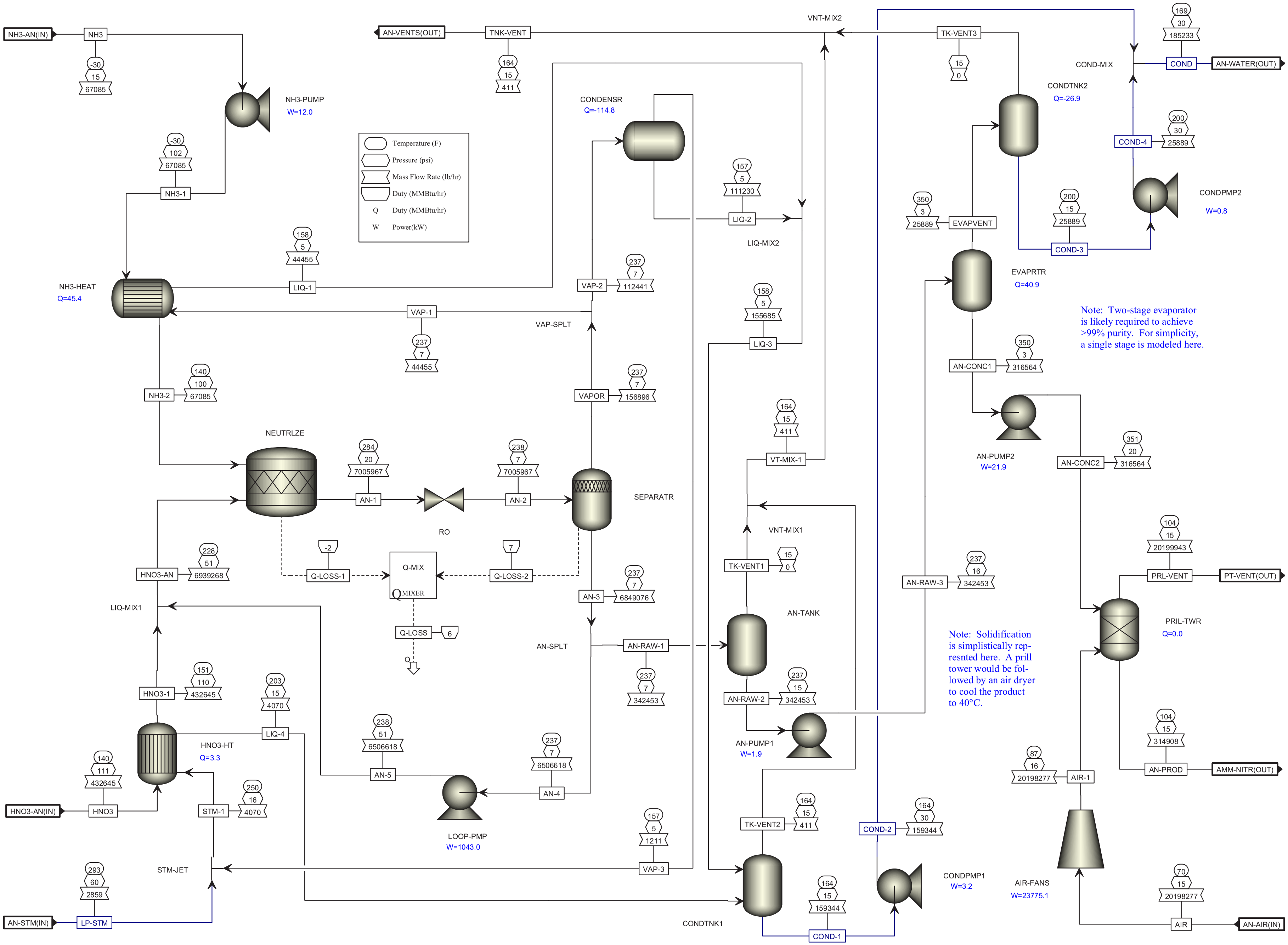
Urea Synthesis via the H.P. Stamicarbon CO2 Stripping Process



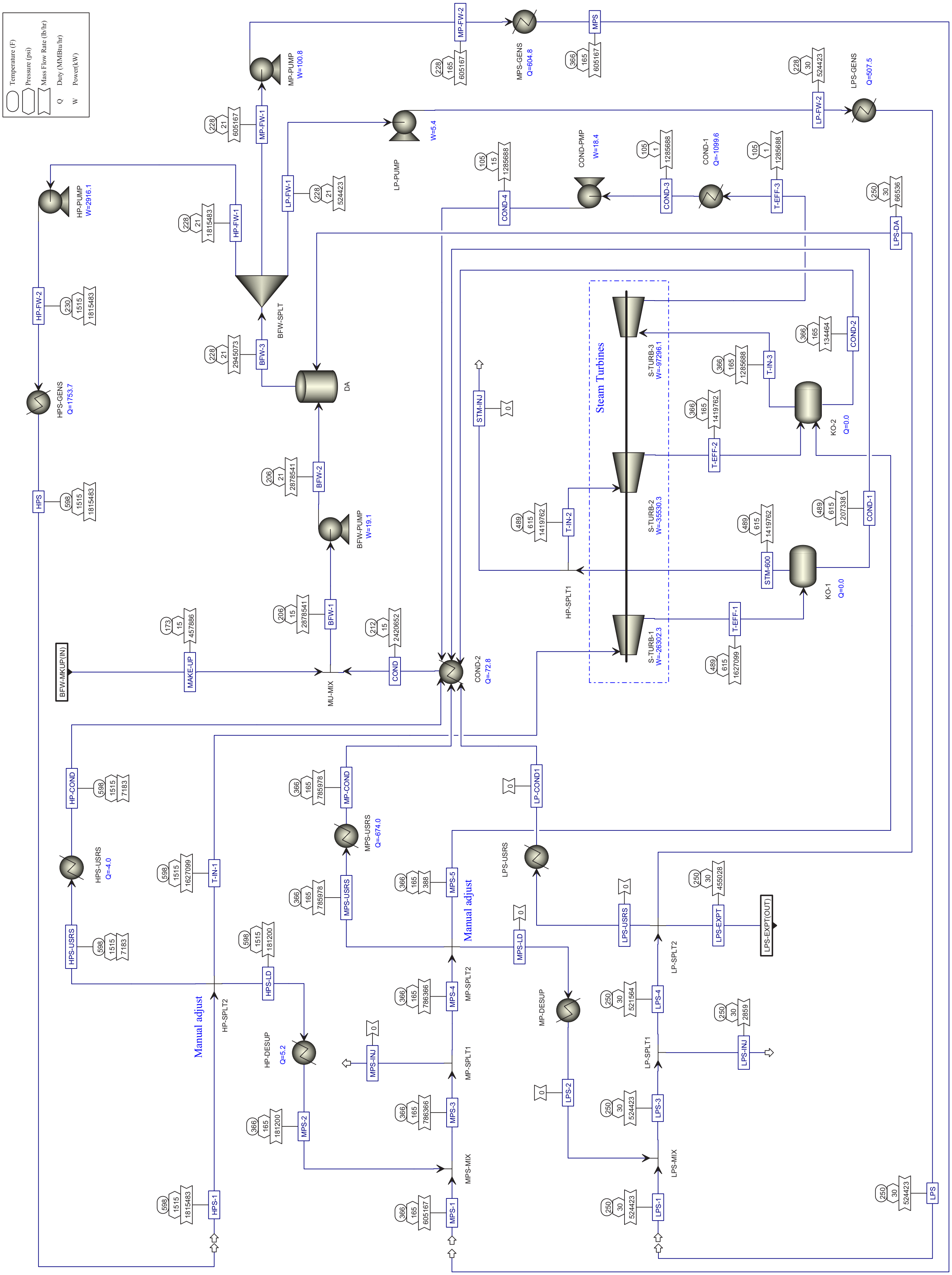
Monopressure Nitric Acid Production



Ammonium Nitrate Production (Uhde-I.G. Farbenindustrie Process)

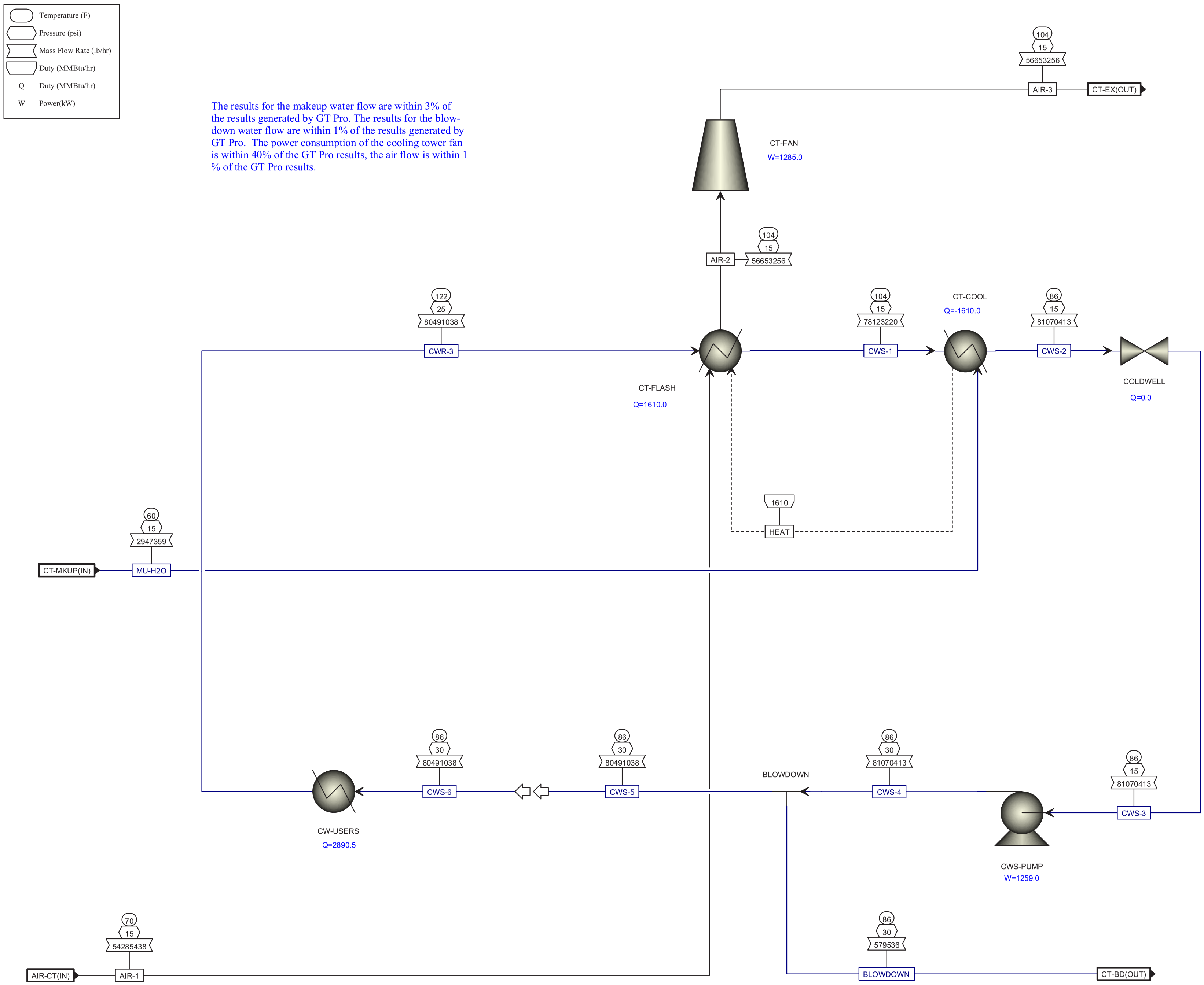


HRSG & Steam Turbines

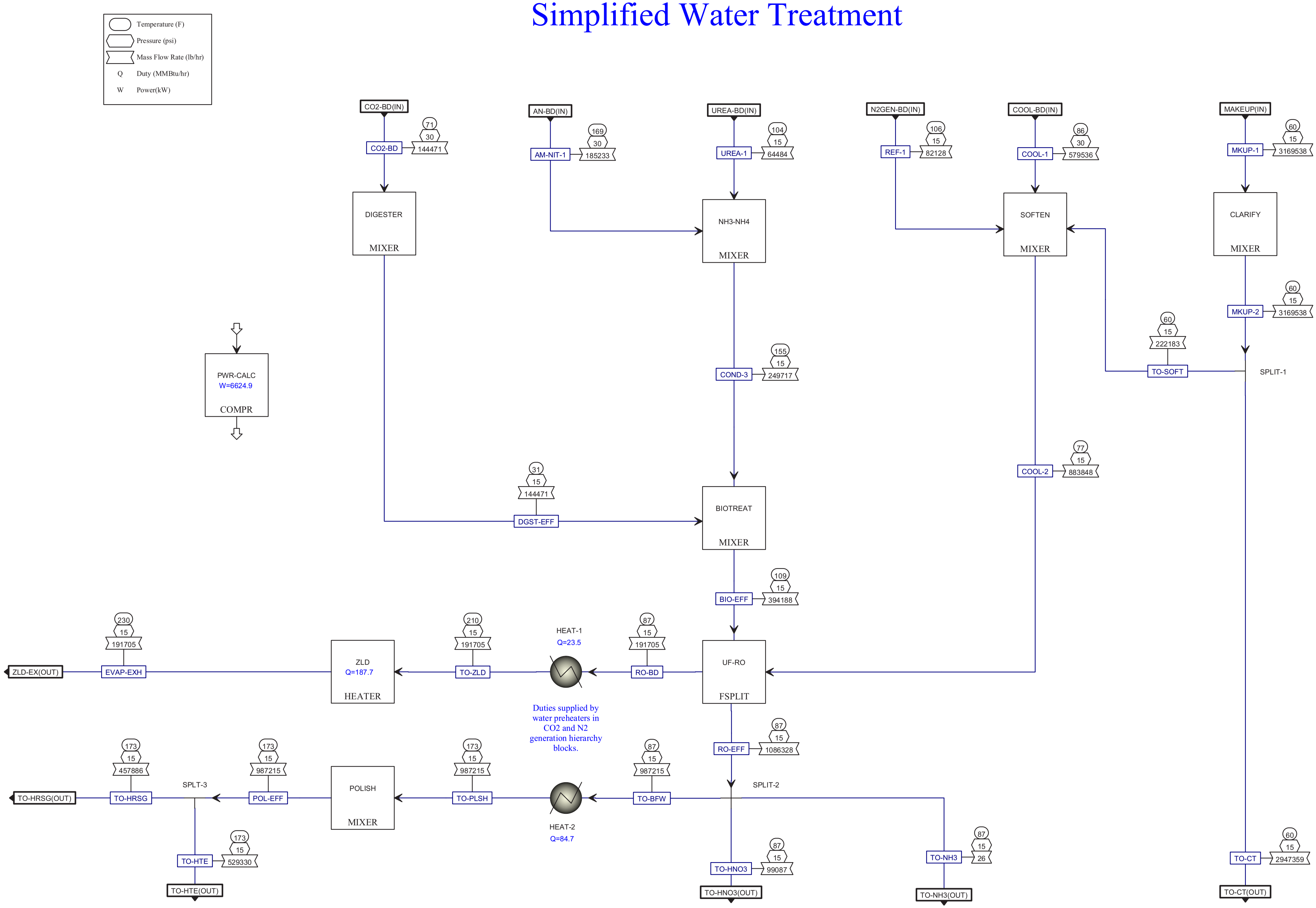


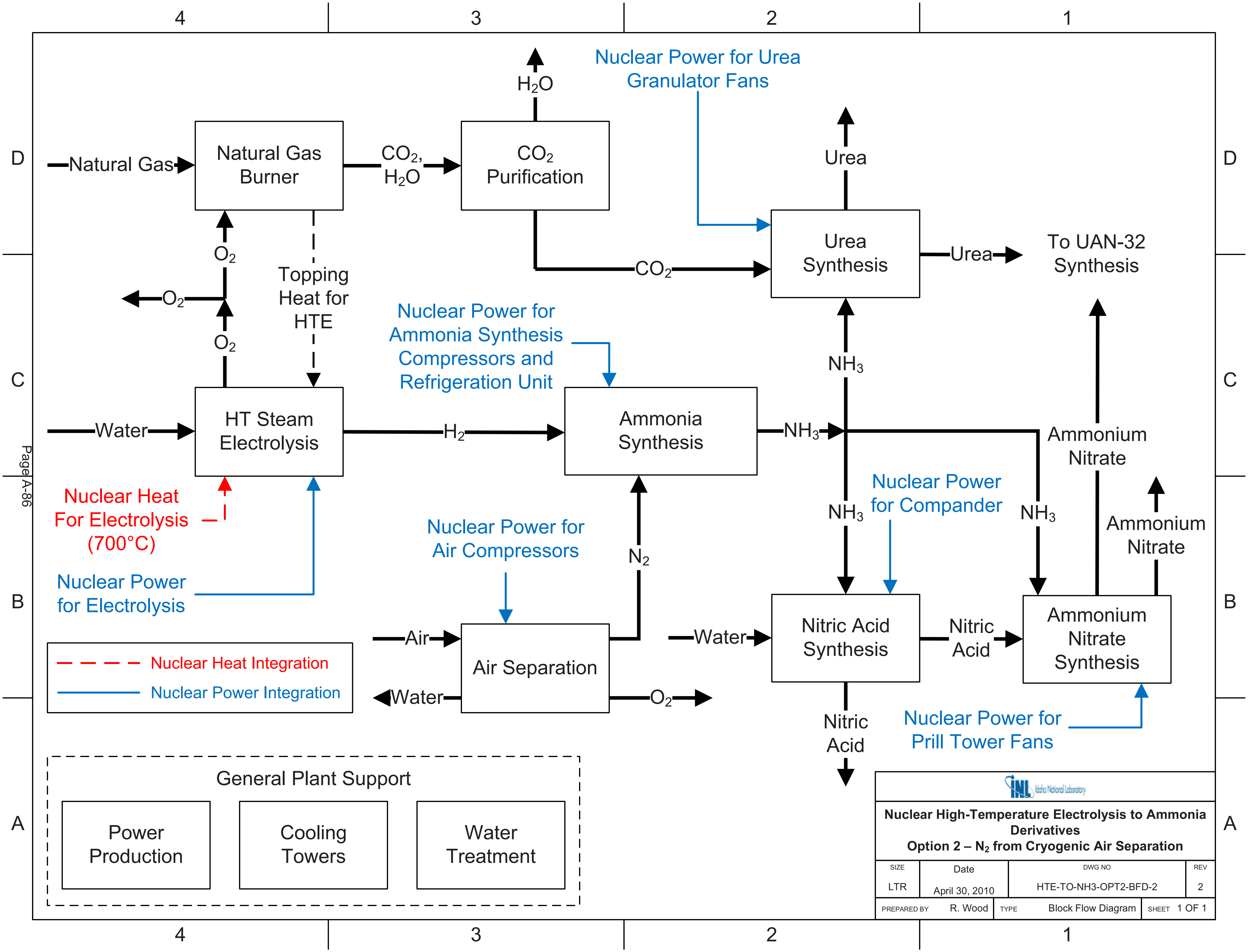
Cooling Tower

The results for the makeup water flow are within 3% of the results generated by GT Pro. The results for the blow-down water flow are within 1% of the results generated by GT Pro. The power consumption of the cooling tower fan is within 40% of the GT Pro results, the air flow is within 1 % of the GT Pro results.

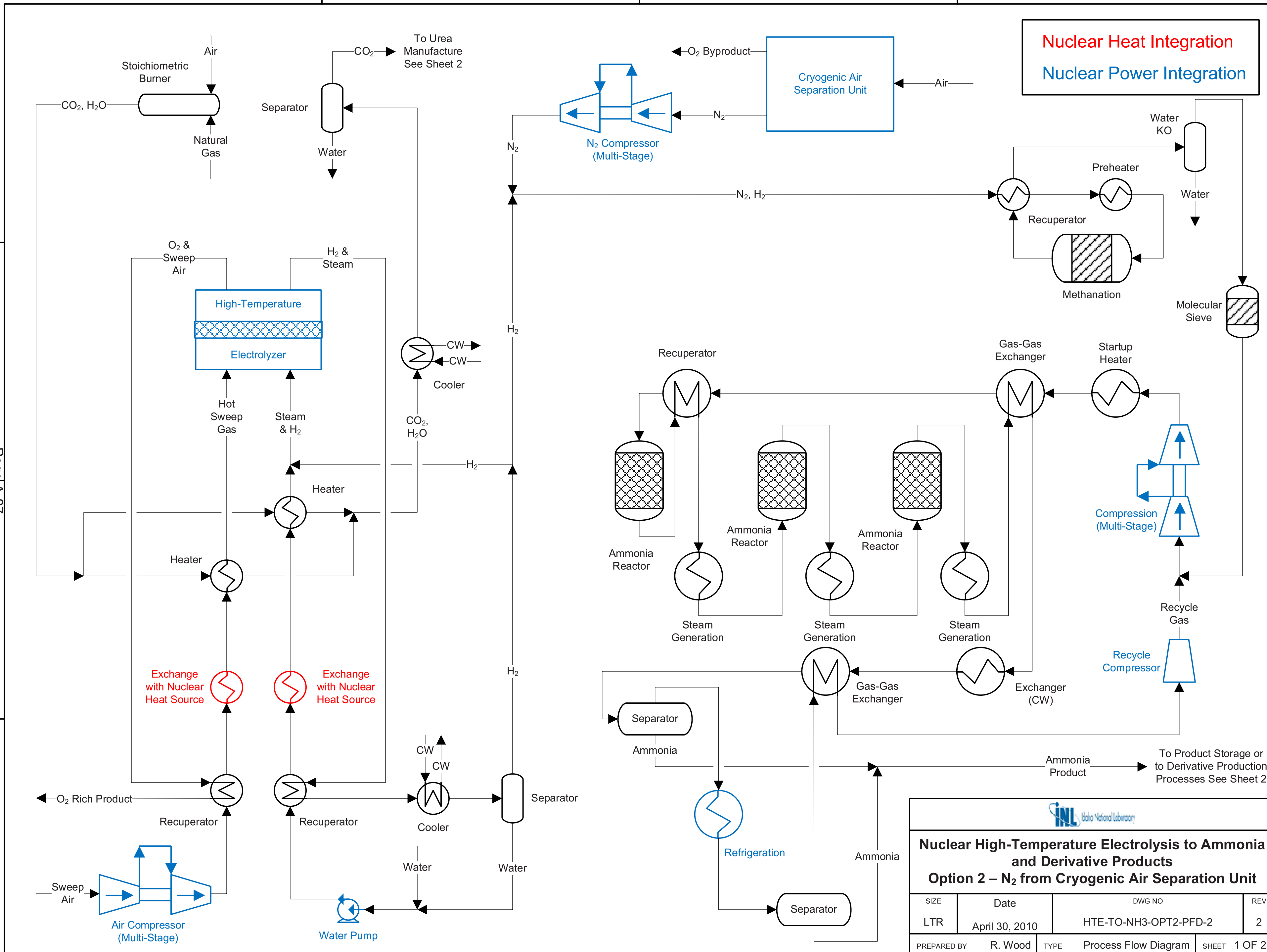


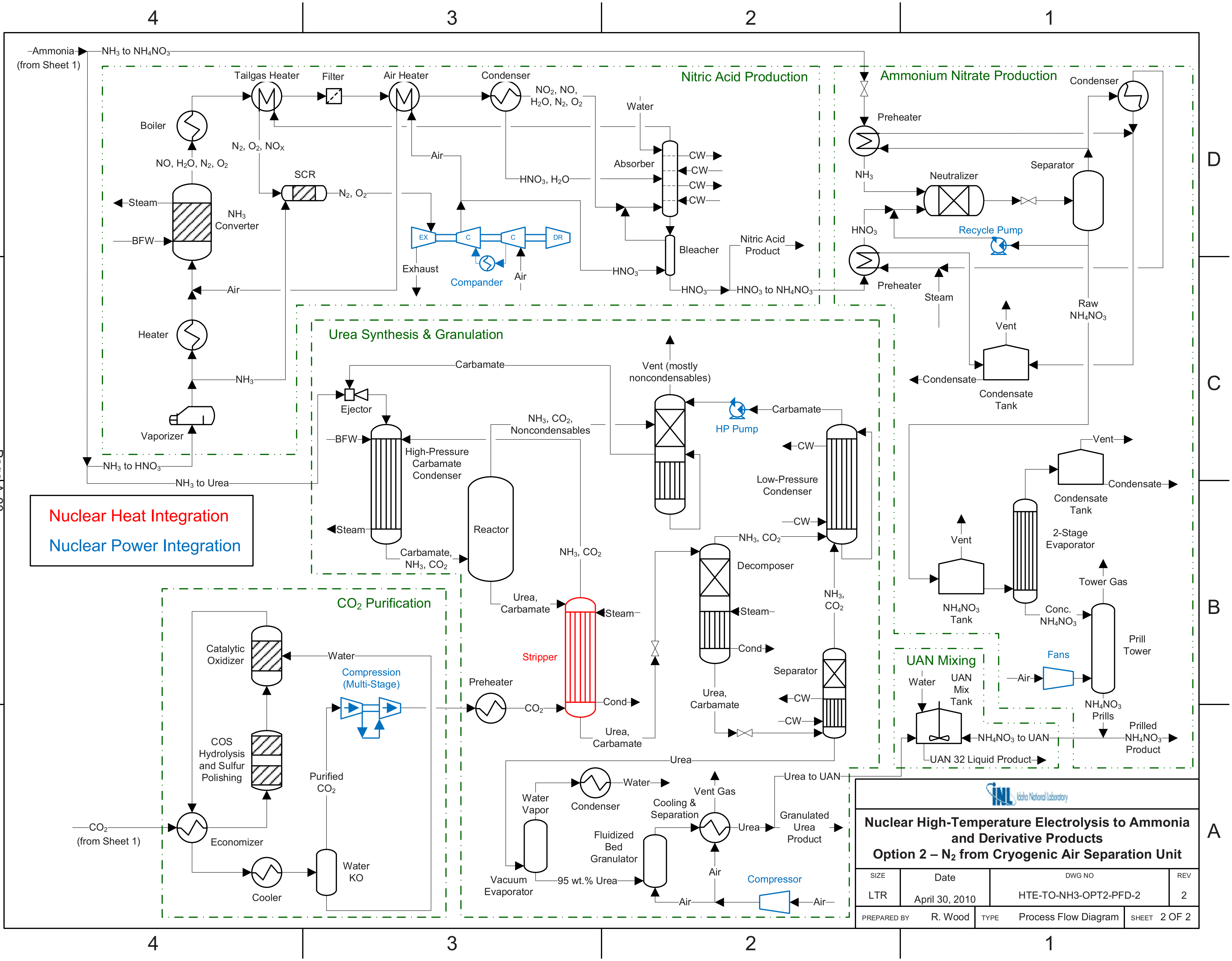
Simplified Water Treatment





Natural Gas
Natural Gas Burner
CO_2, H_2O
CO_2 Purification
H_2O
Nuclear Power for Urea Granulator Fans
Urea
Urea Synthesis
CO_2
Urea
To UAN-32 Synthesis
O_2
O_2
O_2
Topping Heat for HTE
Nuclear Power for Ammonia Synthesis Compressors and Refrigeration Unit
NH_3
Water
HT Steam Electrolysis
H_2
Ammonia Synthesis
NH_3
Ammonium Nitrate
Nuclear Heat For Electrolysis (700°C)
Nuclear Power for Compander
NH_3
NH_3
Ammonium Nitrate
Nuclear Power for Air Compressors
N_2
Nuclear Power for Electrolysis
Air
Air Separation
Water
O_2
Water
Nitric Acid Synthesis
Nitric Acid
Ammonium Nitrate Synthesis
Nuclear Heat Integration
Nuclear Power Integration
Nitric Acid
Nuclear Power for Prill Tower Fans
General Plant Support
Power Production
Cooling Towers
Water Treatment
Idaho National Laboratory
Nuclear High-Temperature Electrolysis to Ammonia Derivatives
Option 2 – N_2 from Cryogenic Air Separation
SIZE LTR
Date April 30, 2010
DWG NO HTE-TO-NH3-OPT2-BFD-2
REV 2
PREPARED BY R. Wood
TYPE Block Flow Diagram
SHEET 1 OF 1


Nuclear Heat Integration
Nuclear Power Integration
Stoichiometric Burner
Air
Natural Gas
CO2, H2O
Separator
Water
CO2
To Urea Manufacture See Sheet 2
O2 Byproduct
Cryogenic Air Separation Unit
Air
N2
N2 Compressor (Multi-Stage)
N2
N2, H2
Water KO
Preheater
Water
Recuperator
Methanation
Molecular Sieve
O2 & Sweep Air
H2 & Steam
High-Temperature
Electrolyzer
CW
CW
Cooler
CO2, H2O
Hot Sweep Gas
Steam & H2
H2
Heater
Heater
Exchange with Nuclear Heat Source
Exchange with Nuclear Heat Source
O2 Rich Product
Recuperator
Recuperator
Cooler
Separator
Sweep Air
Air Compressor (Multi-Stage)
Water Pump
Water
Water
Recuperator
Gas-Gas Exchanger
Startup Heater
Compression (Multi-Stage)
Ammonia Reactor
Ammonia Reactor
Ammonia Reactor
Steam Generation
Steam Generation
Steam Generation
Recycle Gas
Recycle Compressor
Gas-Gas Exchanger
Exchanger (CW)
Separator
Ammonia
Refrigeration
Separator
Ammonia
Ammonia Product
To Product Storage or to Derivative Production Processes See Sheet 2
Nuclear High-Temperature Electrolysis to Ammonia and Derivative Products
Option 2 – N2 from Cryogenic Air Separation Unit
SIZE
LTR
Date
April 30, 2010
DWG NO
HTE-TO-NH3-OPT2-PFD-2
REV
2
PREPARED BY
R. Wood
TYPE
Process Flow Diagram
SHEET
1 OF 2



Ammonia (from Sheet 1)
NH3 to NH4NO3
NH3 to HNO3
NH3 to Urea
Nitric Acid Production
Tailgas Heater
Filter
Air Heater
Condenser
NO2, NO, H2O, N2, O2
Water
Boiler
N2, O2, NOx
NO, H2O, N2, O2
SCR
N2, O2
Air
Absorber
CW
HNO3, H2O
Steam
BFW
NH3 Converter
EX
C
DR
Bleacher
Nitric Acid Product
HNO3
HNO3 to NH4NO3
Exhaust
Compander
Heater
Vaporizer
Ammonium Nitrate Production
Preheater
NH3
Neutralizer
Separator
Recycle Pump
Raw NH4NO3
Vent
Condensate
Condensate Tank
Urea Synthesis & Granulation
Carbamate
Ejector
Vent (mostly noncondensables)
HP Pump
NH3, CO2, Noncondensables
High-Pressure Carbamate Condenser
Low-Pressure Condenser
Reactor
Carbamate, NH3, CO2
NH3, CO2
Decomposer
Urea, Carbamate
Stripper
Cond
2-Stage Evaporator
NH4NO3 Tank
Conc. NH4NO3
Tower Gas
Prill Tower
Fans
NH4NO3 Prills
Prilled NH4NO3 Product
UAN Mixing
UAN Mix Tank
NH4NO3 to UAN
UAN 32 Liquid Product
CO2 Purification
Catalytic Oxidizer
Compression (Multi-Stage)
Preheater
CO2
COS Hydrolysis and Sulfur Polishing
Purified CO2
CO2 (from Sheet 1)
Economizer
Cooler
Water KO
Urea
Urea to UAN
Water Vapor
Condenser
Vent Gas
Cooling & Separation
Granulated Urea Product
Fluidized Bed Granulator
Vacuum Evaporator
95 wt.% Urea
Compressor
Nuclear Heat Integration
Nuclear Power Integration
Nuclear High-Temperature Electrolysis to Ammonia and Derivative Products
Option 2 – N2 from Cryogenic Air Separation Unit
SIZE LTR
Date April 30, 2010
DWG NO HTE-TO-NH3-OPT2-PFD-2
REV 2
PREPARED BY R. Wood
TYPE Process Flow Diagram
SHEET 2 OF 2

Calculator Block SUMMARY

FEED SUMMARY:

NATURAL GAS PROPERTIES:

MASS FLOW =	811. TON/DY
VOLUME FLOW =	36. MMSCFD @ 60°F
HHV =	23063. BTU/LB
HHV =	1044. BTU/SCF @ 60°F
ENERGY FLOW =	37404. MMBTU/DY

COMPOSITION:

METHANE =	93.571 MOL.%
ETHANE =	3.749 MOL.%
PROPANE =	0.920 MOL.%
BUTANE =	0.260 MOL.%
PENTANE =	0.040 MOL.%
HEXANE =	0.010 MOL.%
NITROGEN =	1.190 MOL.%
OXYGEN =	0.010 MOL.%
CO2 =	0.250 MOL.%
C4H10S =	1. PPMV
C2H6S =	0. PPMV
H2S =	0. PPMV

INTERMEDIATE PRODUCT SUMMARY:

RAW SYNGAS MASS FLOW =	283141. LB/HR
RAW SYNGAS VOLUME FLOW =	302. MMSCFD @ 60°F
RAW SYNGAS COMPOSITION:	
H2	74.9 MOL.%
CO	0.0 MOL.%
CO2	0.0 MOL.%
N2	25.0 MOL.%
AR	0.0 MOL.%
H2O	0.1 MOL.%
CH4	0.0 MOL.%

CLEANED SYNGAS MASS FLOW =	282763. LB/HR
CLEANED SYNGAS VOLUME FLOW =	302. MMSCFD @ 60°F
CLEANED SYNGAS COMPOSITION:	
H2	75.0 MOL.%
N2	25.0 MOL.%
AR	0.0 MOL.%
CH4	0.0 MOL.%
CO	0. PPMV
CO2	0. PPMV
H2O	0. PPMV

AMMONIA MASS FLOW =	279973. LB/HR
AMMONIA MASS FLOW =	3360. TON/DY
AMMONIA PURITY =	100.00 MOL.%

NITRIC ACID MASS FLOW =	432633. LB/HR
NITRIC ACID MASS FLOW =	5192. TON/DY
NITRIC ACID PURITY =	57.93 WT.%
HNO3 PRODUCED / NH3 FED =	1.55 LB/LB

FINAL PRODUCT SUMMARY:

AMMONIA MASS FLOW =	0. LB/HR
AMMONIA MASS FLOW =	0. TON/DY
AMMONIA PURITY =	100.00 MOL.%

```
  NITRIC ACID MASS FLOW =                      0. LB/HR
  NITRIC ACID MASS FLOW =                      0. TON/DY
    NITRIC ACID PURITY =                    57.93 WT.%

  UREA MASS FLOW =                        244876. LB/HR
  UREA MASS FLOW =                          2939. TON/DY
  UREA PRODUCED / AMMONIA FED =              1.74 LB/LB
  UREA PRODUCED / CO2 FED =                  1.33 LB/LB

  AMMONIUM NITRATE MASS FLOW =            314900. LB/HR
  AMMONIUM NITRATE MASS FLOW =              3779. TON/DY
  NH4NO3 PRODUCED / NITRIC ACID FED =        0.73 LB/LB
  NH4NO3 PRODUCED / AMMONIA FED =            4.69 LB/LB


POWER SUMMARY:

  ELECTRICAL GENERATORS:
    STEAM TURBINE POWER GENERATION =     131.0 MW
  GENERATOR SUBTOTAL =                   131.0 MW

  ELECTRICAL CONSUMERS:
    ASU POWER CONSUMPTION =               23.2 MW
    HTE POWER CONSUMPTION =              778.1 MW
    POWER BLOCK POWER CONSUMPTION =        3.0 MW
    CO2 PROCESSING POWER CONSUMPTION =     5.6 MW
    AMMONIA SYNTH. POWER CONSUMPTION =    38.5 MW
    UREA SYNTHESIS POWER CONSUMPTION =     4.4 MW
    HNO3 SYNTHESIS POWER CONSUMPTION =    15.1 MW
    NH4NO3 SYNTH. POWER CONSUMPTION =     24.9 MW
    COOLING TOWER POWER CONSUMPTION =      2.3 MW
    WATER TREATMENT POWER CONSUMPTION =    5.6 MW
  CONSUMER SUBTOTAL =                    900.6 MW

  NET PLANT POWER CONSUMPTION =          769.5 MW


WATER BALANCE:

  EVAPORATIVE LOSSES:
    COOLING TOWER EVAPORATION =         5079.4 GPM
    ZLD SYSTEM EVAPORATION =             307.0 GPM
  TOTAL EVAPORATIVE LOSSES =            5386.4 GPM

  WATER CONSUMED:
    BOILER FEED WATER MAKEUP =           645.3 GPM
    NITRIC ACID FEED WATER =             198.0 GPM
    COOLING TOWER MAKEUP =              5346.8 GPM
    HTE WATER REQUIREMENT =              896.5 GPM
    AMMONIA WASH WATER REQUIREMENT =       0.0 GPM
  TOTAL WATER CONSUMED =                7086.7 GPM

  WATER GENERATED:
    CO2 COMPRESSION BLOWDOWN =           288.7 GPM
    UREA PRODUCTION BLOWDOWN =           128.9 GPM
    NH4NO3 PRODUCTION BLOWDOWN =         370.1 GPM
    COOLING TOWER BLOWDOWN =            1051.3 GPM
  TOTAL WATER GENERATED =               1839.0 GPM

  PLANT WATER SUMMARY:
    NET MAKEUP WATER REQUIRED =         5554.7 GPM
```

```
CO2 BALANCE:

  CO2 EMITTED (TOTAL) =                    11. TON/DY
  CO2 EMITTED (TOTAL) =                     0.2 MMSCFD @ 60°F


NUCLEAR INTEGRATION REQUIREMENTS:

  TOTAL ELECTRICITY DEMAND =              769.5 MW
    FROM HTE =                            778.1 MW
    FROM BALANCE OF PLANT =                -8.5 MW

  HYDROGEN DEMAND =                       606.1 TON/DY
    PURITY =                               99.9159 MOL.%

  TOTAL OXYGEN PRODUCTION =              5566.0 TON/DY
    FROM ASU =                            788.6 TON/DY
      PURITY =                             99.4556 MOL.%
    FROM HTE =                           4777.5 TON/DY
      PURITY =                             99.8910 MOL.%
    PLANT O2 DEMAND =                    3131.0 TON/DY
    EXCESS O2 FOR SALE =                 2435.0 TON/DY

  TOPPING HEAT SUPPLIED =                  42.7 MMBTU/HR
    SUPPLY TEMPERATURE =                 1874. °F
    RETURN TEMPERATURE =                 1833. °F
```

Calculator Block AIRPROPS

```
HUMIDITY DATA FOR STREAM AIR-ASU:
  HUMIDITY RATIO =                         43.5 GRAINS/LB
  RELATIVE HUMIDITY =                      39.9 %
```

Calculator Block PH   Hierarchy: NH4NO3

```
TARGET NEUTRALIZER PH =   1.500
ACTUAL NEUTRALIZER PH =   1.499
```

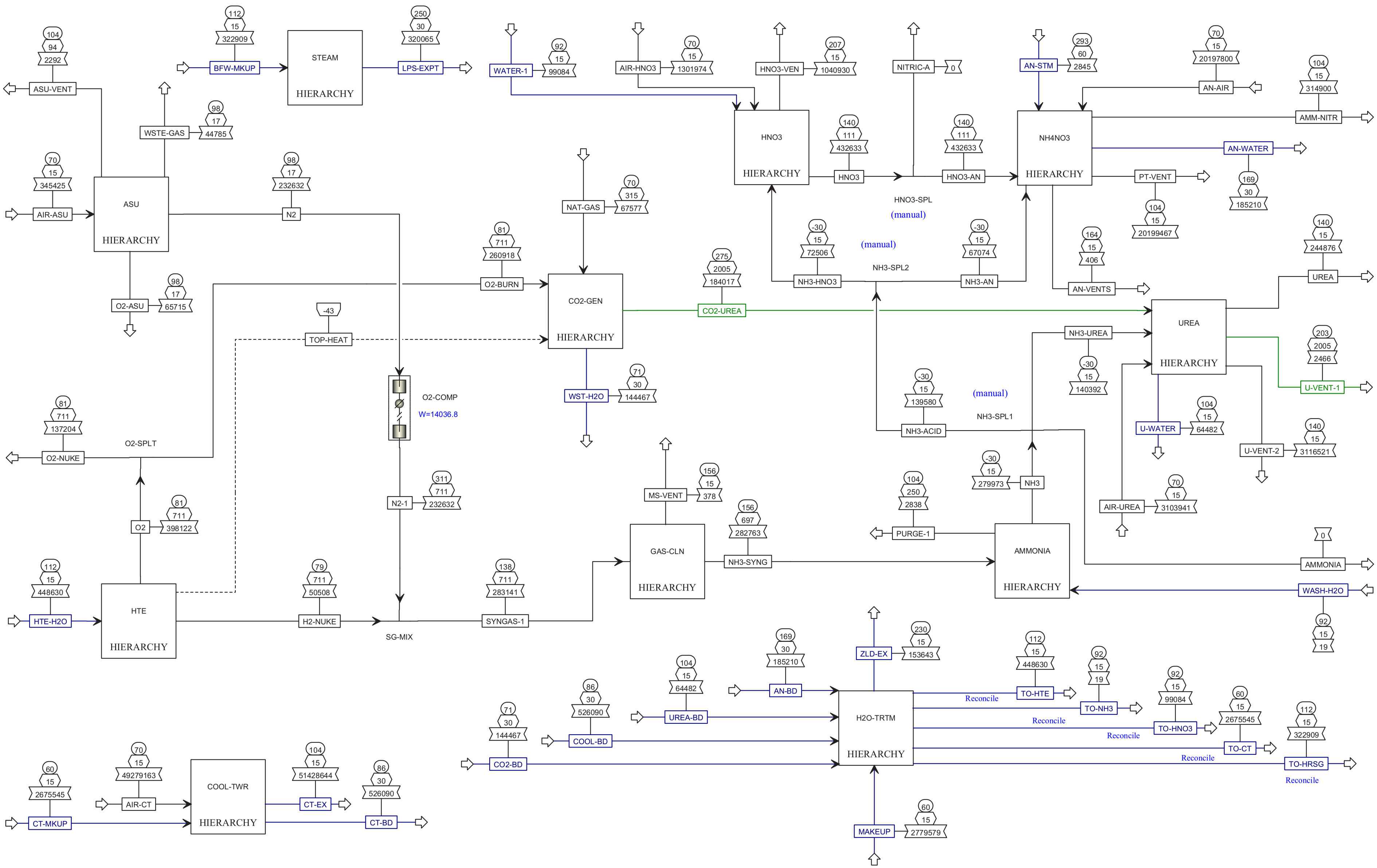
Calculator Block RATIO   Hierarchy: UREA

```
TARGET REACTOR NH3/CO2 RATIO =   2.950
ACTUAL REACTOR NH3/CO2 RATIO =   2.971
```

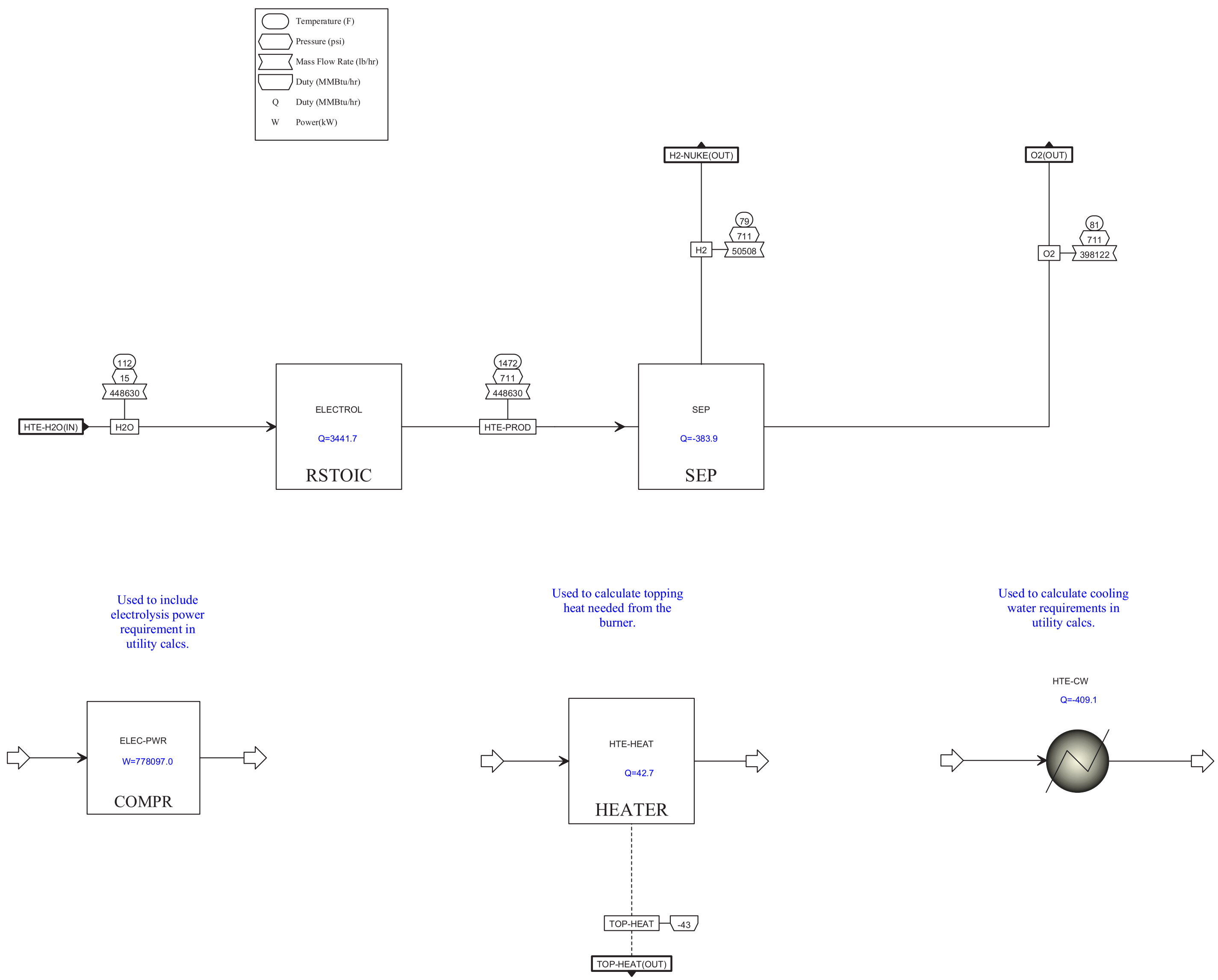
High Temperature Steam Electrolysis to Ammonia and Derivatives
N2 from Cryogenic Air Separation Unit (ASU)
(Derivatives produced in the correct ratio to make UAN32)

Temperature (F)
Pressure (psi)
Mass Flow Rate (lb/hr)
Duty (MMBtu/hr)
W Power(kW)

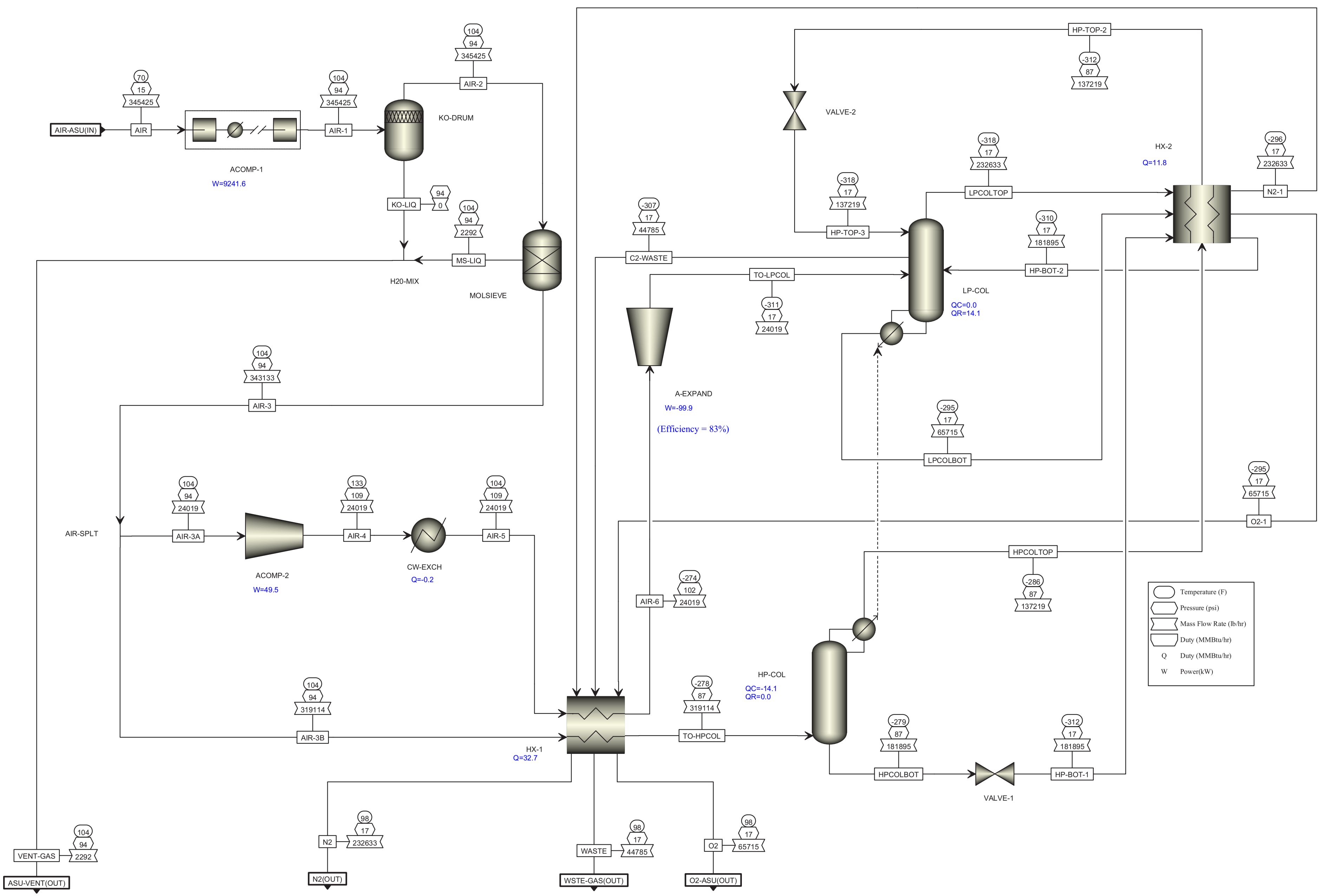
Color Legend
Water or Steam
CO2 Emissions Source



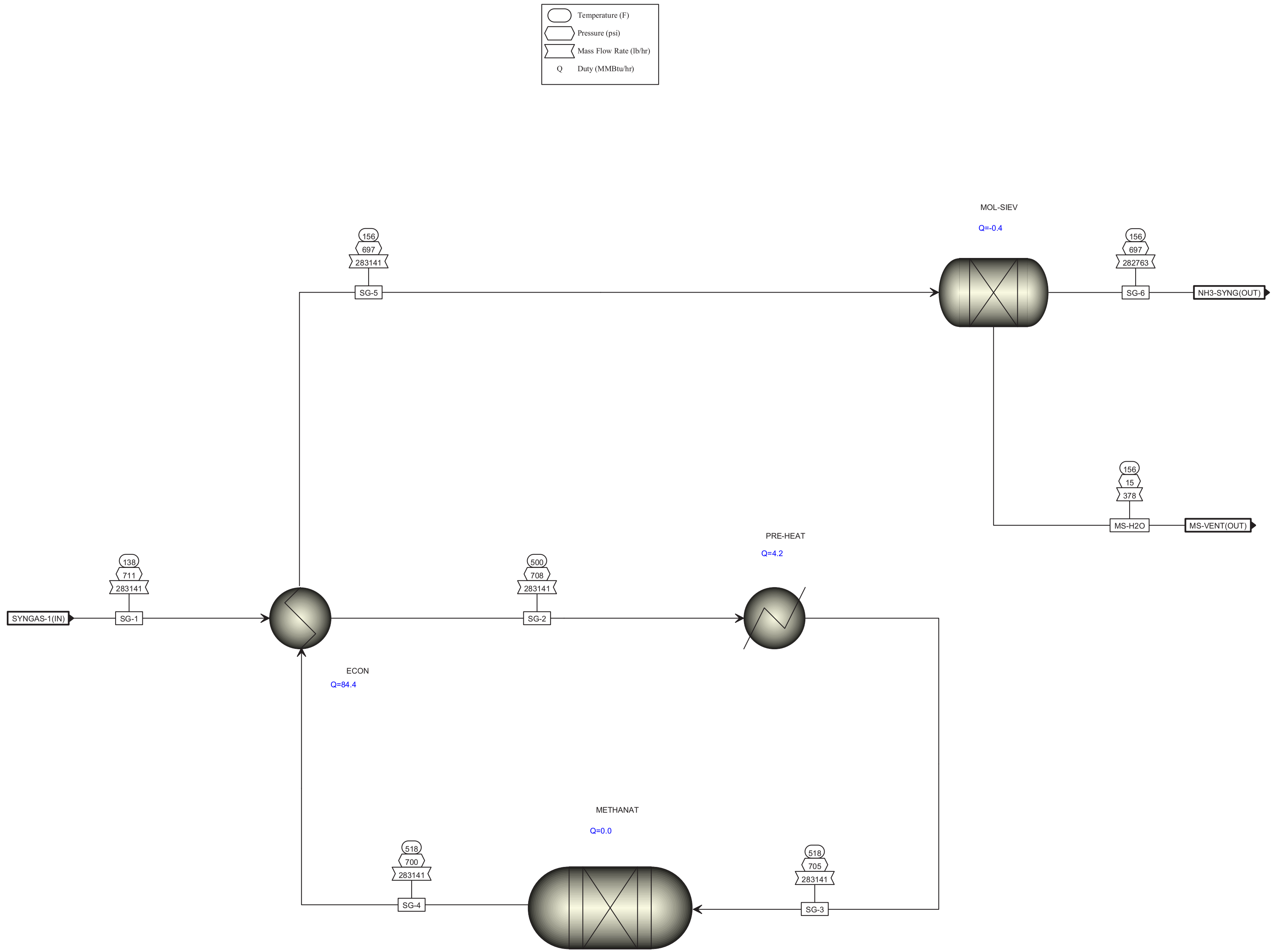
Simplified HTE Model



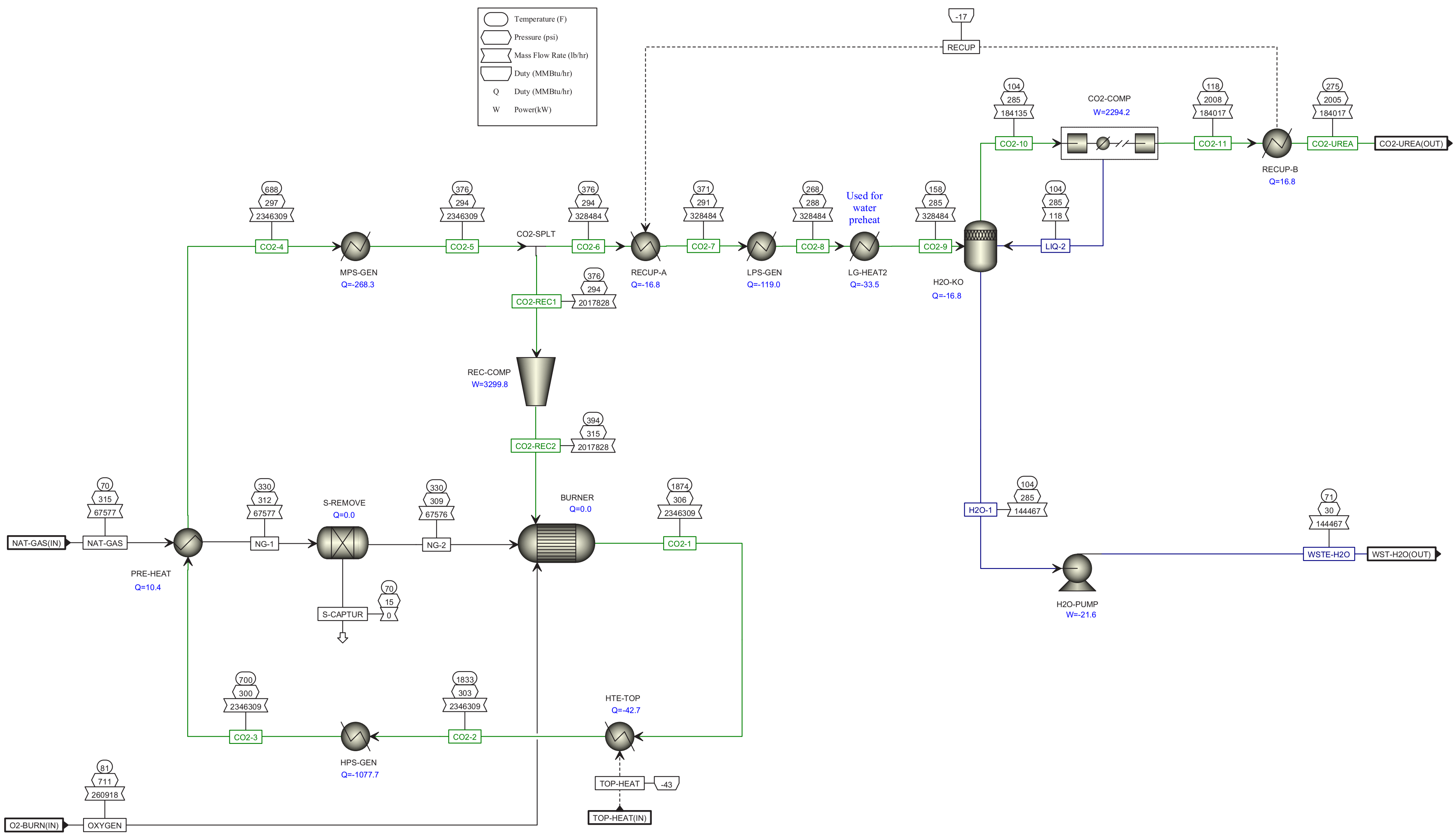
Air Separation Unit 99.5% O2 Purity



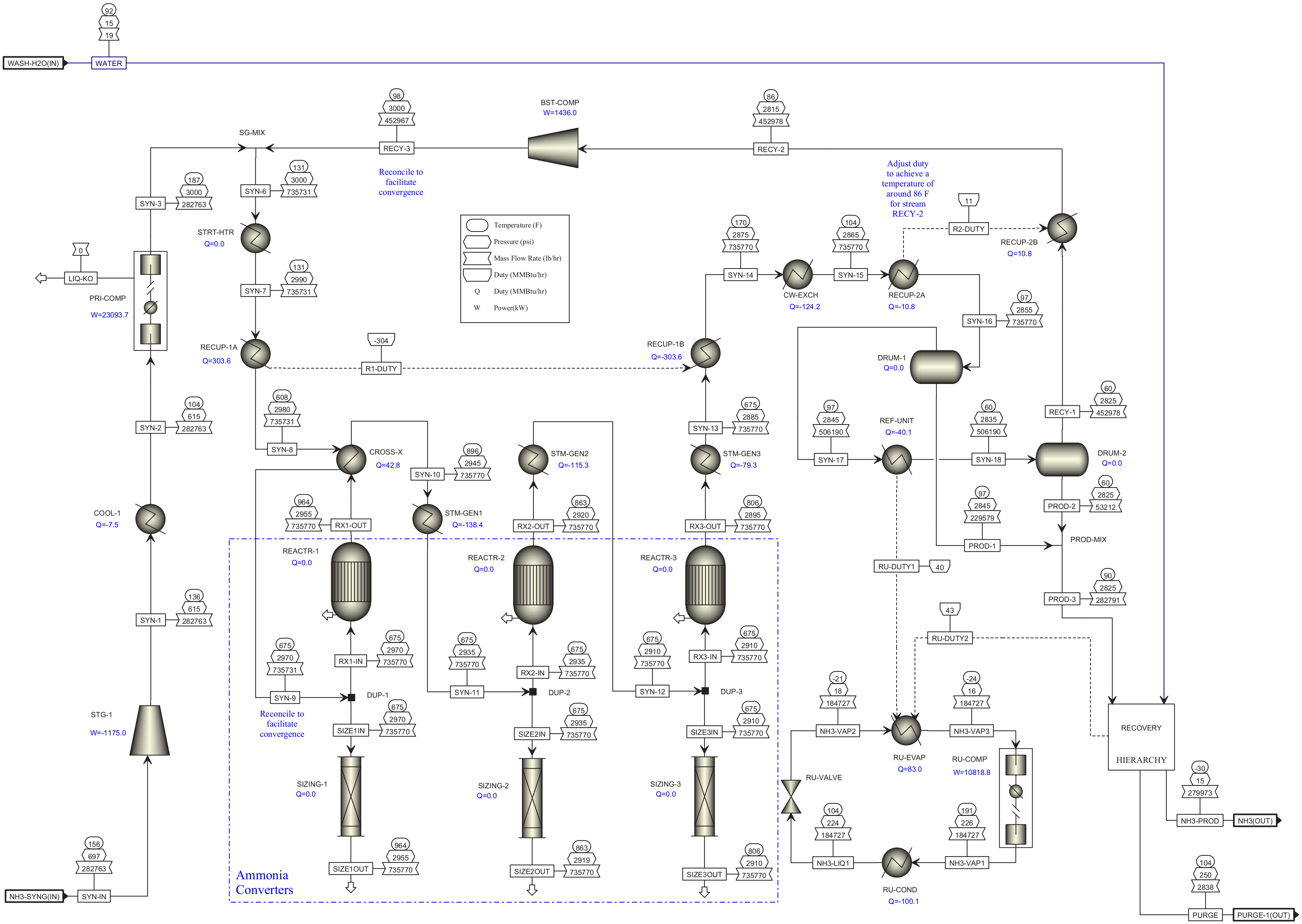
Syngas Cleaning & Conditioning



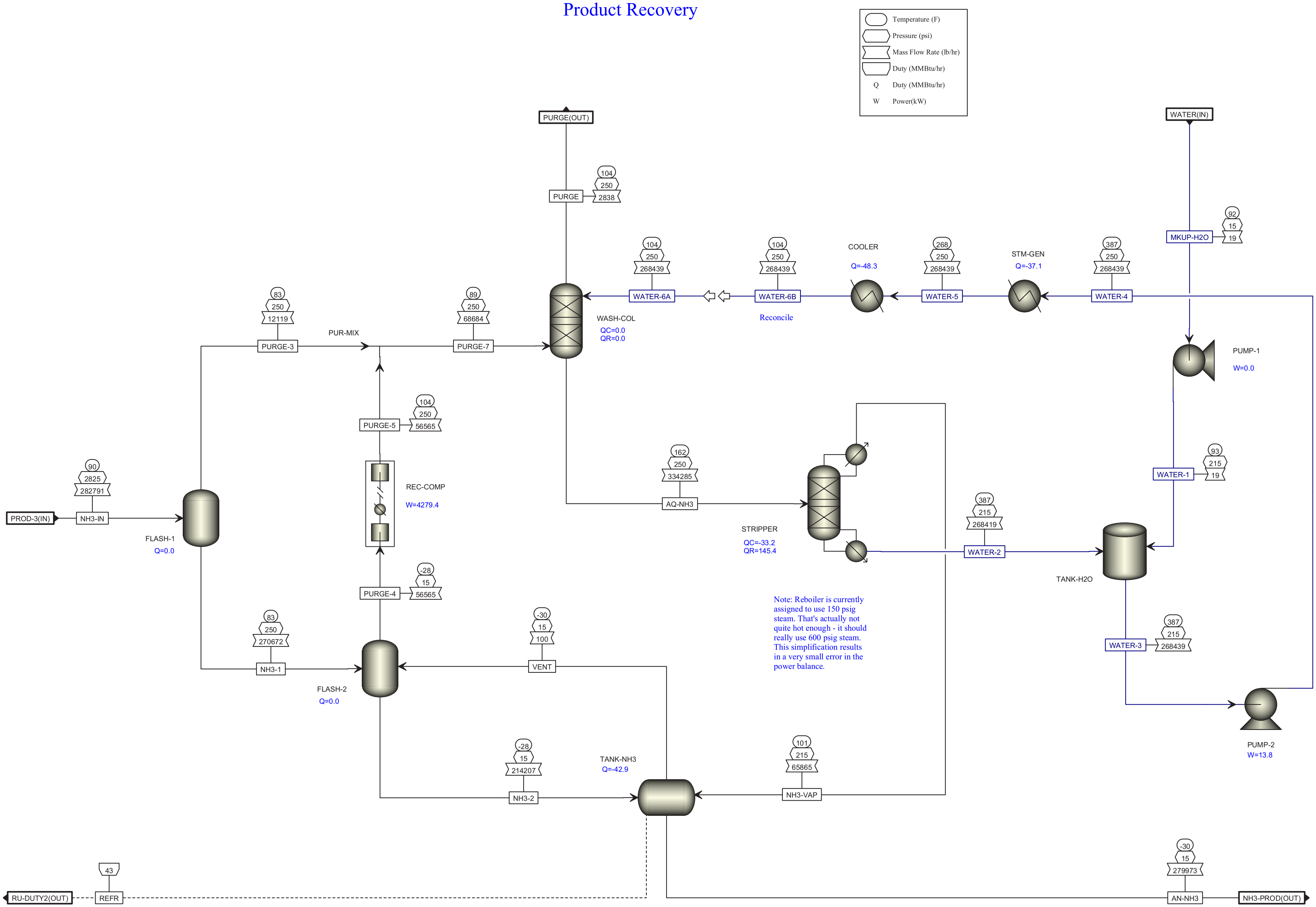
CO2 Generation



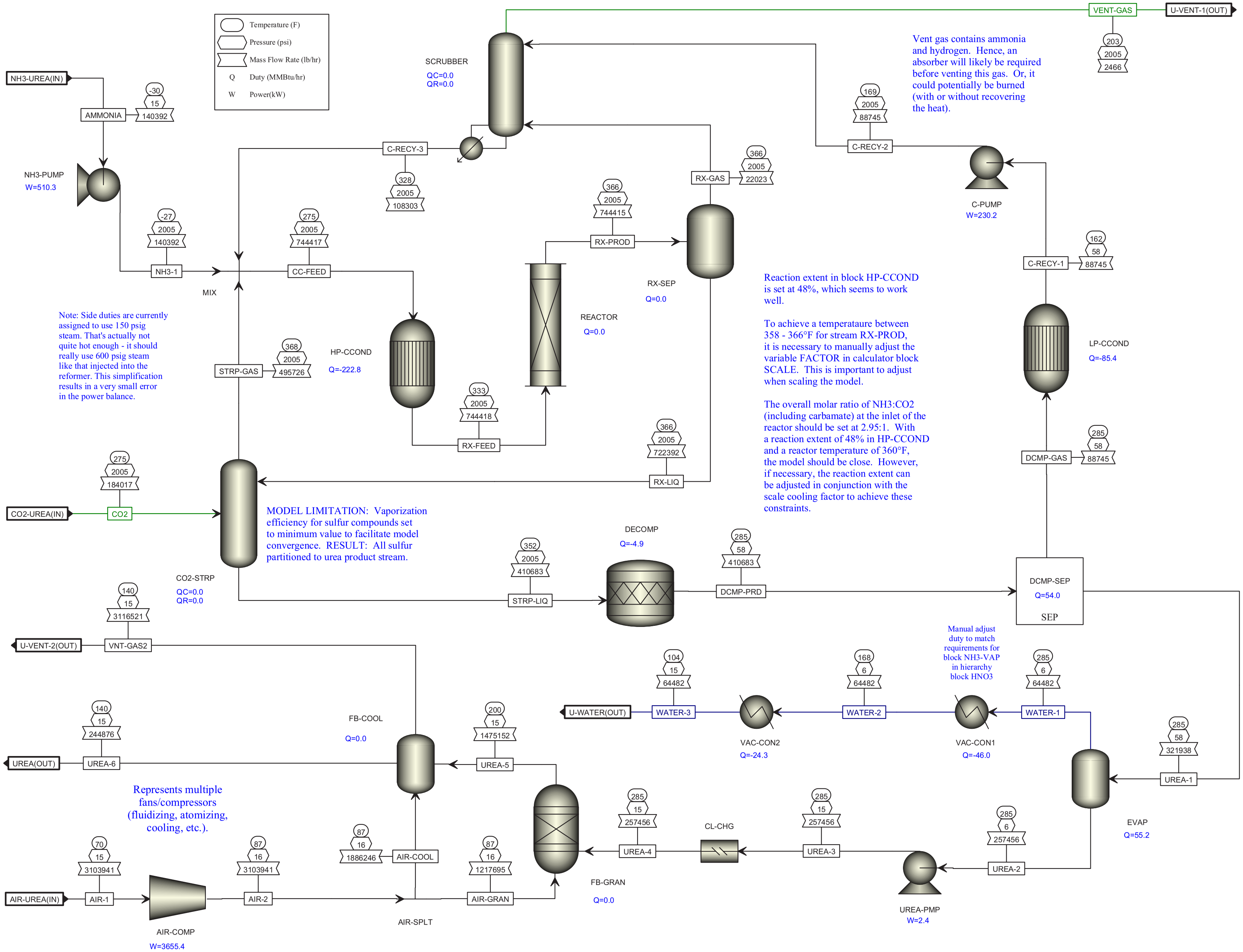
Ammonia Synthesis @ 3000 psi



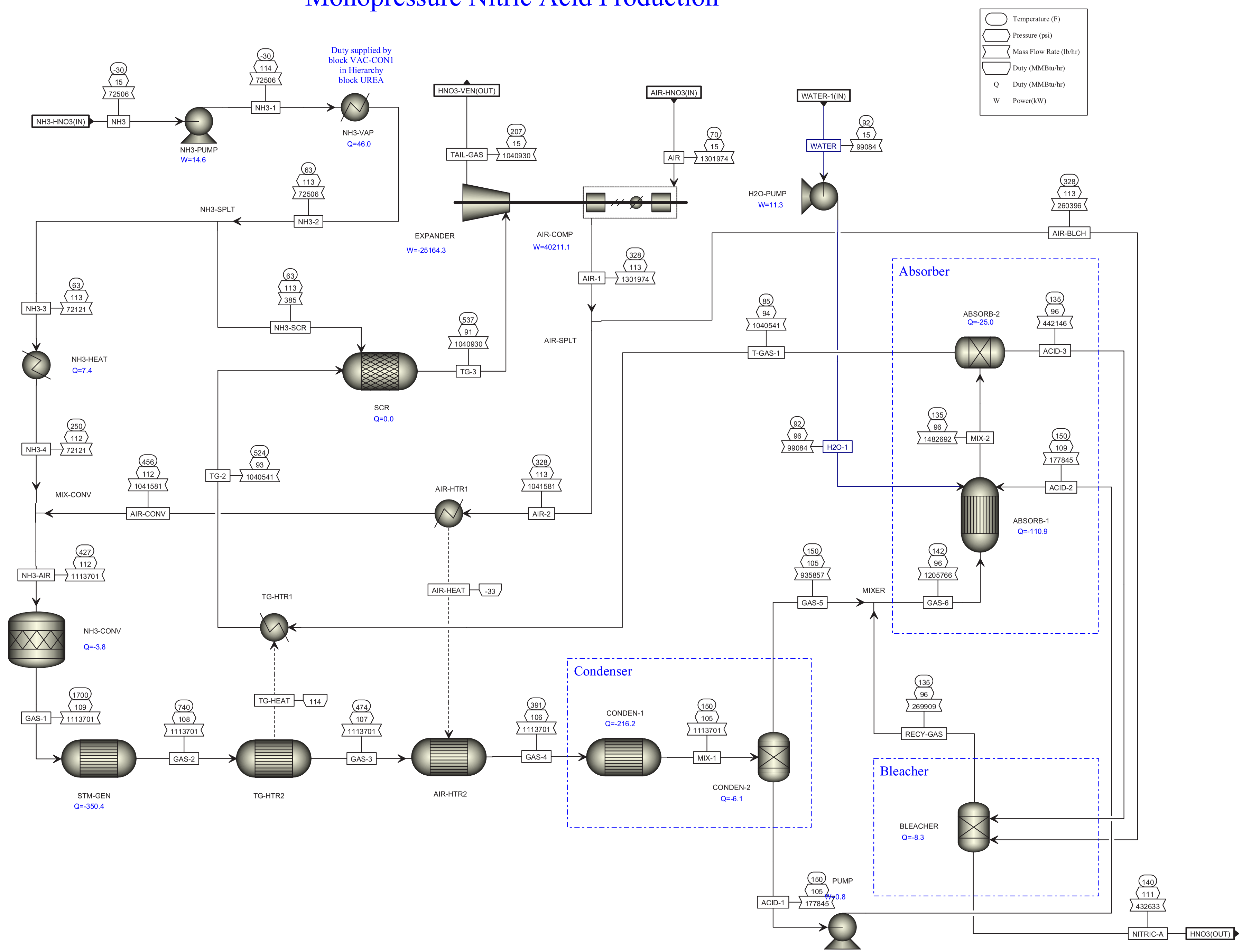
Product Recovery



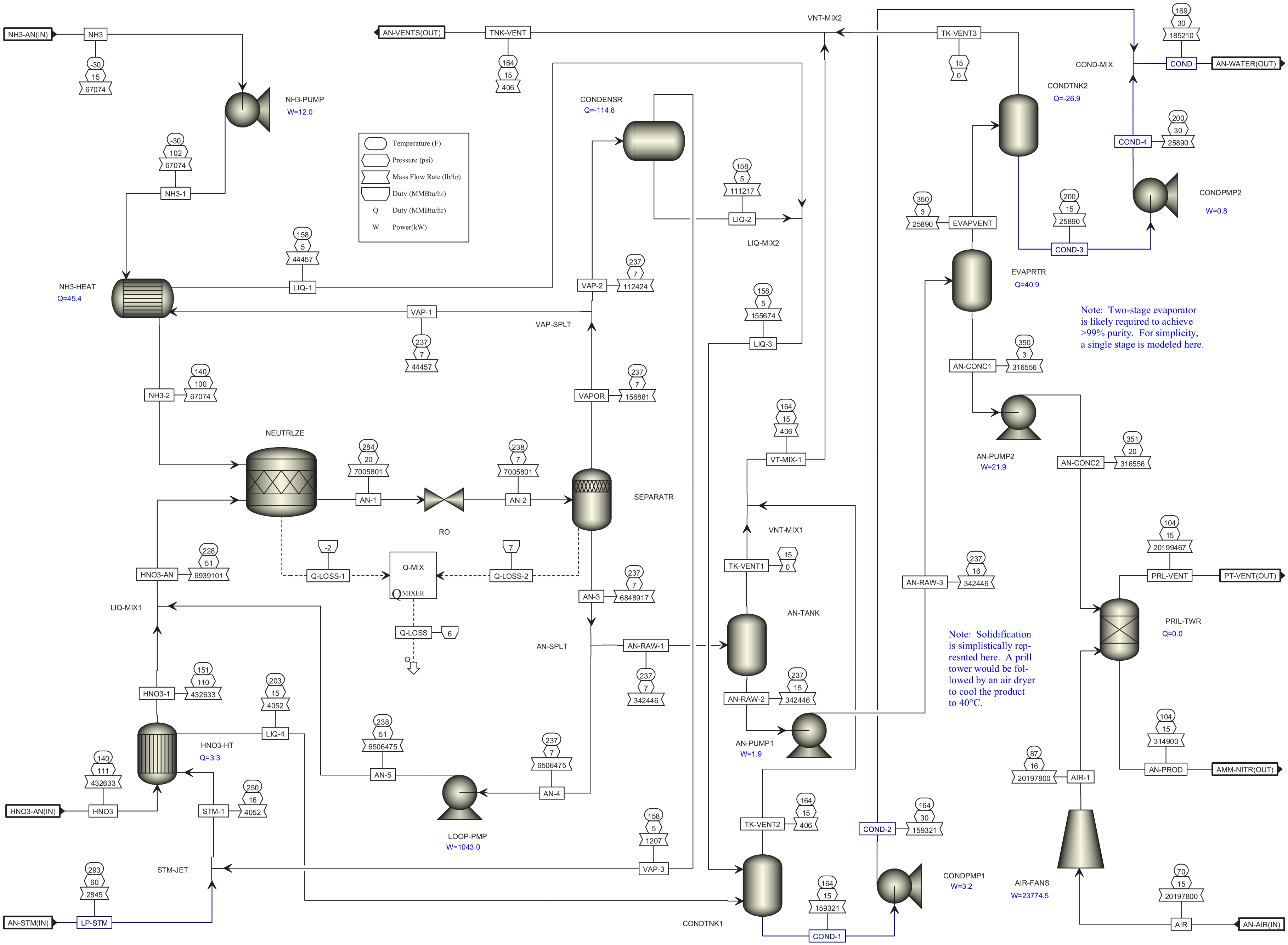
Urea Synthesis via the H.P. Stamicarbon CO2 Stripping Process



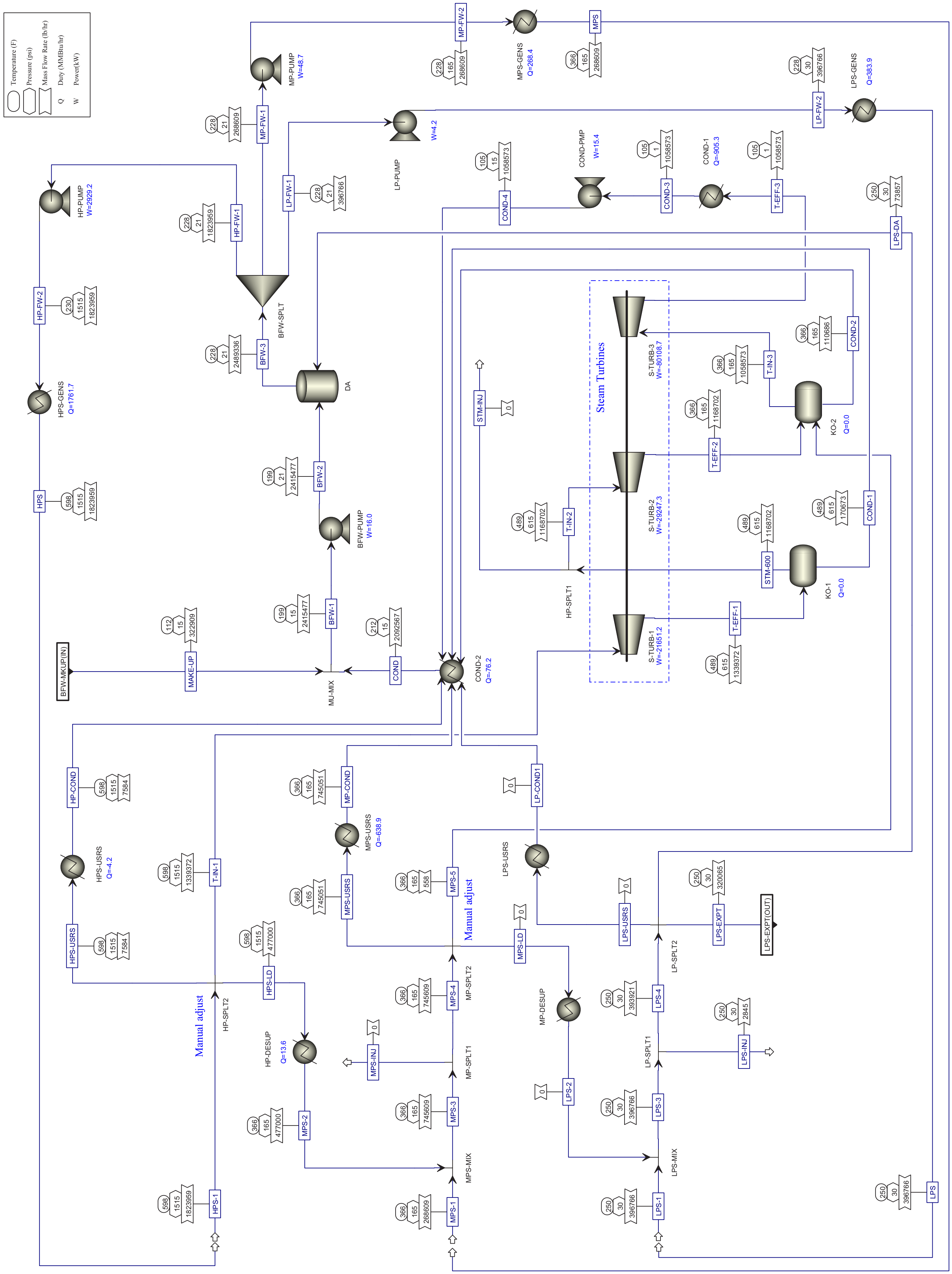
Monopressure Nitric Acid Production



Ammonium Nitrate Production (Uhde-I.G. Farbenindustrie Process)



HRSG & Steam Turbines



Cooling Tower

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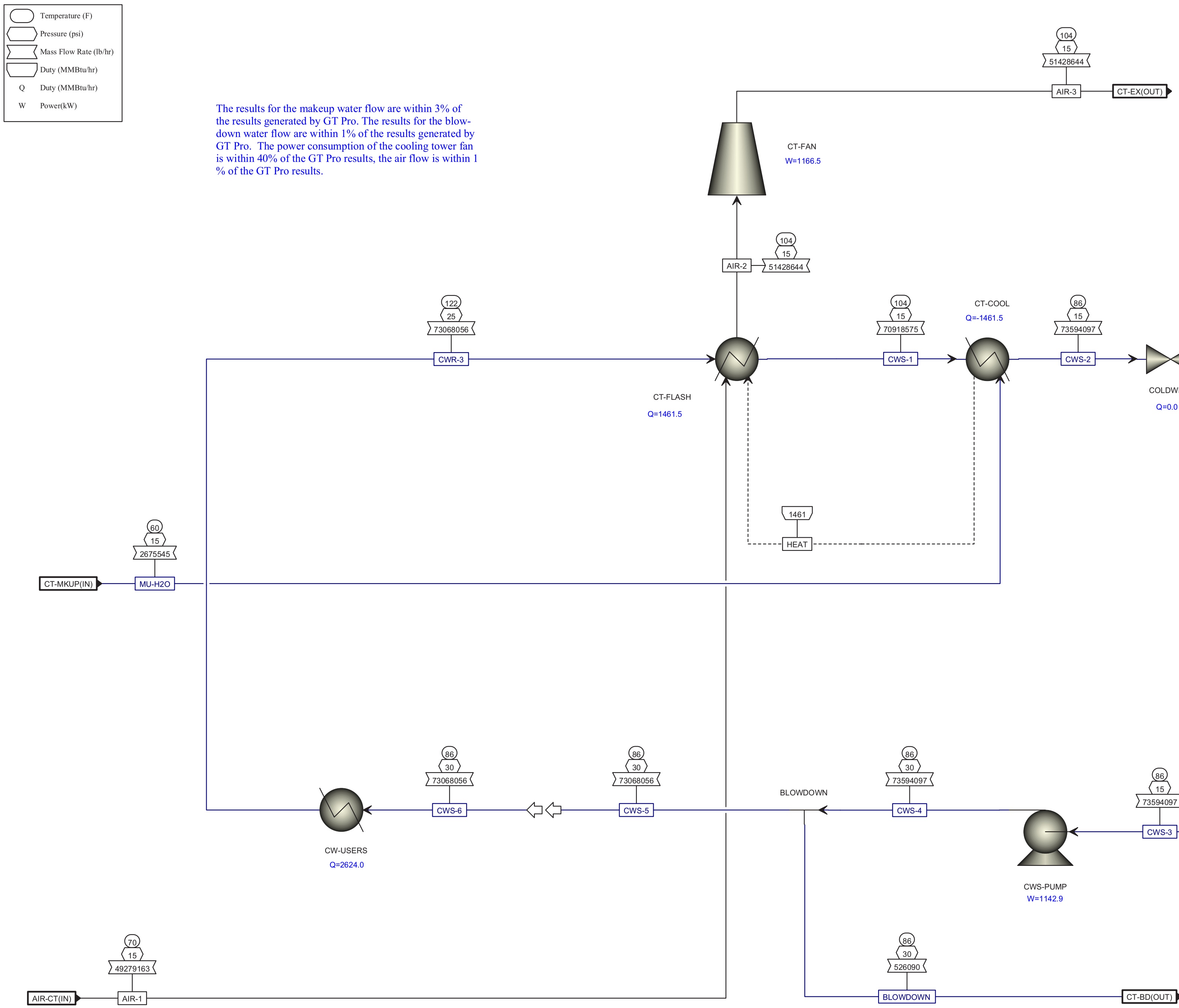
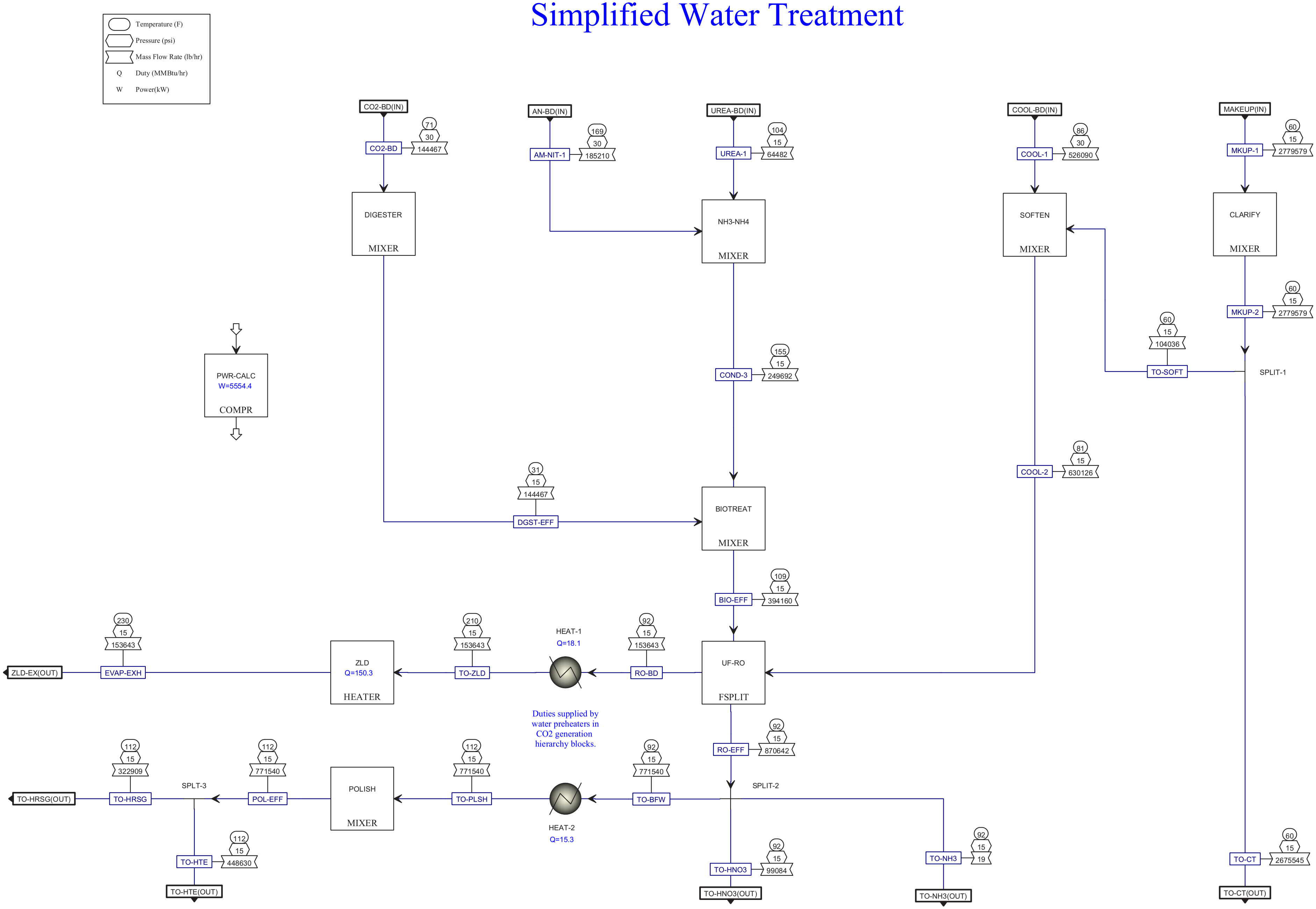


Simplified Water Treatment



Appendix B
Ammonia Capital Cost Estimates

INTEROFFICE MEMORANDUM

Idaho National Laboratory INL

Date: April 20, 2010

To: M. W. Patterson, Project Manager

From: B. W. Wallace, Cost Estimator BWW

Subject: NGNP Process Integration Estimates MA36-A through MA36-R

Per your request, Cost Estimating prepared cost estimates (Class 5) for the above-mentioned subject. The total estimated costs (TEC) rounded to the nearest $10 million, including contingency for these estimates are as follows:

MA36-A NGNP – Conventional Coal to Liquid $5,060,000,000
MA36-B NGNP – Nuclear Coal to Liquid $16,540,000,000
MA36-C NGNP – Conventional Gas to Liquid..............................$1,920,000,000
MA36-D NGNP – Nuclear Gas to Liquid $3,010,000,000
MA36-E NGNP – Conventional Coal to Synthetic Natural Gas......$1,920,000,000
MA36-F NGNP – Nuclear Coal to Synthetic Natural Gas $8,460,000,000
MA36-G NGNP – Conventional Coal to Methanol to Gas $5,940,000,000
MA36-H NGNP – Nuclear Coal to Methanol to Gas....................$18,470,000,000
MA36-I NGNP – Conventional Gas to Methanol to Gas $1,690,000,000
MA36-J NGNP – Nuclear Gas to Methanol to Gas.........................$3,030,000,000
MA36-L NGNP – Conventional Coal to Ammonia $2,150,000,000
MA36-M NGNP – Conventional Gas to Ammonia $1,660,000,000
MA36-N NGNP – Nuclear Gas to Ammonia $2,530,000,000
MA36-O NGNP – HTSE Ammonia w/o ASU $6,940,000,000
MA36-P NGNP – HTSE Ammonia w/ ASU..................................$6,300,000,000
MA36-Q NGNP – Conventional Steam Assisted Gravity Drained..$1,060,000,000
MA36-R NGNP – Nuclear Steam Assisted Gravity Drained..........$2,720,000,000

Please note the following:

A. These estimates are intended for use as an input to economic analysis models that will be used in a feasibility study. An allowance for contingency instead of management reserve has been included in the estimates to align more closely with commercial practices and terminology.
B. These estimates reflect the nth of a kind (NOAK) projects. Use of NOAK in these estimates represented a condition where construction of both the integrated capability and the high temperature gas reactor (HTGR) is common place and represents routine construction.
C. All costs represent 2009 values, and construction costs are considered to be overnight costs. Forward looking factors for escalation are not included in the estimates.
D. These cost estimates have been evaluated in the AACEI classification matrix as Class 5 estimates (*ref. DOE G. 430.1-1X, Appendix J*). The primary characteristic used in this guideline to define the classification category is the degree of project definition at this time.

The intent of this classification is to assist in the interpretation of the quality and value of the information available to prepare these cost estimates and the expected accuracy levels that can be produced. Per AACEI, a Class 5 indicates the lowest amount of project information, quality, and value with a graded approach to a Class 1, which indicates the highest amount of project information, quality, and value.

E. In addition to multiple internal reviews performed with the appropriate project team technical lead, an external review of the cost estimates has been performed by Washington Division United Research Services (URS) Corporation. A meeting with URS and the Next-Generation Nuclear Plant (NGNP) Process Integration team on April 7, 2010, allowed the project team and this estimator to discuss, in detail, the perceived scope, basis of estimates, assumptions, project risks, and the resources that make up this cost estimate. Comments from this review have been incorporated into these estimates to reflect a project team consensus of the documents.
F. It is intended that these estimates reflect commercial construction of the facilities addressed. It is not intended, nor are there provisions included in the estimates that would make them appropriate for construction at Idaho National Laboratory.

Refer to the cost estimating summary, detail, markup, and labor sheets with the cost breakdowns. Also included for your use are the cost estimate recapitulation sheets describing the basis and assumptions used in development of this estimate.

These estimates, the work, and the work breakdown structure are based on the information perceived by this estimator as to the scope of work to be completed. Any changes to the methodology used to prepare these estimates could have a significant effect on the cost estimates and/or schedules and should be reviewed by me. If you have any questions or comments, do not hesitate to contact me at 526-5896 or e-mail Bruce.Wallace@inl.gov.

Attachments

cc: (electronic only)
A. M. Gandrik
R. R. Honsinger
V. C. Maio
J. B. Martin
M. G. McKellar
L. O. Nelson
M. M. Plum
R. A. Wood
Estimate Files MA36-A through MA36-J and MA36-L through MA36-R

Uniform File Code: 8309
Disposition Authority: A16-1.5-b
Retention Schedule: Cut off at the end of each fiscal year. Destroy 10 years after cutoff.

NOTE: Original disposition authority, retention schedule, and Uniform Filing Code applied by the sender may not be appropriate for all recipients. Make adjustments as needed.

NGNP Conventional Coal to Ammonia

Project Name: NGNP Process Integration
Process: Conventional Coal to Ammonia
Estimate Number: MA36-L

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Process Component	Subtotal From Detail Sheets	Engineering %	Engineering	Contingency %	Contingency	Total Cost
Air Separation Unit (ASU)	$ 115,688,970	10%	$ 11,568,897	18%	$ 22,906,416	$ 150,164,283
Coal Preparation	$ 58,317,443	10%	$ 5,831,744	18%	$ 11,546,854	$ 75,696,041
Gasification	$ 187,583,875	10%	$ 18,758,388	18%	$ 37,141,607	$ 243,483,870
Water Gas Shift Reactor	$ 44,540,936	10%	$ 4,454,094	18%	$ 8,819,105	$ 57,814,135
Selexol	$ 51,726,242	10%	$ 5,172,624	18%	$ 10,241,796	$ 67,140,662
Pressure Swing Adsorption (PSA)	$ 56,185,670	10%	$ 5,618,567	18%	$ 11,124,763	$ 72,929,000
Methanation	$ 9,514,938	10%	$ 951,494	18%	$ 1,883,958	$ 12,350,390
Claus & SCOT	$ 47,476,883	10%	$ 4,747,688	18%	$ 9,400,423	$ 61,624,994
CO2 Compression	$ 8,864,886	10%	$ 886,489	18%	$ 1,755,247	$ 11,506,622
Ammonia Synthesis	$ 297,054,672	10%	$ 29,705,467	18%	$ 58,816,825	$ 385,576,965
Urea Synthesis	$ 288,347,019	10%	$ 28,834,702	18%	$ 57,092,710	$ 374,274,430
Nitric Acid Synthesis	$ 271,980,990	10%	$ 27,198,099	18%	$ 53,852,236	$ 353,031,325
Ammonium Nitrate Synthesis	$ 173,948,476	10%	$ 17,394,848	18%	$ 34,441,798	$ 225,785,122
Steam Turbines	$ 23,430,044	10%	$ 2,343,004	18%	$ 4,639,149	$ 30,412,197
Heat Recovery Steam Generator (HRSG)	$ 14,011,353	10%	$ 1,401,135	18%	$ 2,774,248	$ 18,186,737
Cooling Towers	$ 4,433,502	10%	$ 443,350	18%	$ 877,833	$ 5,754,685
Total Cost - Conventional Coal to Ammonia						$ 2,145,731,456
Total Cost Rounded to the Nearest $10M						$ 2,150,000,000

	Remarks
Checked By: SAW	
Approved By: John R	

COST ESTIMATE SUPPORT DATA RECAPITULATION

Project Title: NGNP Process Integration – Conventional Coal to Ammonia
Estimator: B. W. Wallace/CEP, R. R. Honsinger/CEP, J. B. Martin/CCT
Date: April 20, 2010
Estimate Type: Class 5
File: MA36-L
Approved By:

I. **PURPOSE:** *Brief description of the intent of how the estimate is to be used (i.e., for engineering study, comparative analysis, request for funding, proposal, etc.).*

It is expected that the capital costs identified in these estimates will be used in a model producing an economic analysis for each specific integrated application and subsequently will be considered in a related feasibility study.

II. **SCOPE OF WORK:** *Brief statement of the project's objective. Thorough overview and description of the proposed project. Identify work to be accomplished, as well as any specific work to be excluded.*

A. **Objective:**
Develop Class 5 estimates as defined by the Association for Advancement of Cost Engineering (AACEi) that will identify the current capital cost associated with a coal-to-ammonia process.

B. **Included:**
The scope of work required to achieve this objective includes the following:

1. Engineering
2. Construction of a new coal-to-ammonia refinery that consists of the following:
   a. Air separation unit
   b. Coal preparation
   c. Gasification process
   d. Water gas shift reactors
   e. Selexol process
   f. Pressure swing adsorption
   g. Methanation
   h. Claus and SCOT processes
   i. CO_2 compression
   j. Ammonia synthesis
   k. Urea synthesis
   l. Nitric acid synthesis
   m. Ammonium nitrate synthesis
   n. Steam turbines, internal to process
   o. Heat recovery steam generator, internal to process
   p. Cooling towers, internal to process
   q. Allowances for Balance of Plant (BOP)/offsite/outside of battery limits (OSBL), including the following:

(1.) Site development/improvements
(2.) Provisions for general and administrative buildings and structures
(3.) Provisions for OSBL piping
(4.) Provisions for OSBL instrumentation and control
(5.) Provisions for OSBL electrical
(6.) Provisions for facility supply and OSBL water systems
(7.) Provisions for site development/improvements
(8.) Project/construction management.

C. **Excluded**:
This scope of work specifically excludes the following elements:
1. Licensing and permitting costs
2. Operational costs
3. Land costs
4. Sales taxes
5. Royalties
6. Owner's fees and owner's costs.

III. **ESTIMATE METHODOLOGY**: *Overall methodology and rationale of how the estimate was developed (i.e., parametric, forced detail, bottoms up, etc.). Total dollars/hours and rough order magnitude (ROM) allocations of the methodologies used to develop the cost estimate.*

Consistent with the AACEi Class 5 estimates, the level of definition and engineering development available at the time they were prepared, their intended use in a feasibility study, and the time and resources available for their completion, the costs included in this estimate have been developed using parametric evaluations. These evaluations have used publicly available and published project costs to represent similar islands utilized in this project. Analysis and selection of the published costs used have been performed by the project technical lead and Cost Estimating. Suitability for use in this effort was determined considering the correctness and completeness of the data available, the manner in which total capital costs were represented, the age of the previously performed work, and the similarity to the capacity/trains required by this project. The specific sources, selected and used in this cost estimate, are identified in the capital cost estimate detail sheets. Adjustments have been made to these published costs using escalation factors identified in the Chemical Engineering Price Cost Index. Scaling of the published island costs has been accomplished using the six-tenths capacity factoring method. Any normalization to provide for geographic factors was considered using geographic factors available from RS Means Construction Cost Data references. Cost-estimating relationships have been used to identify allowances to complete the costs.

BOP/OSBL costs were determined by the project team, considering data provided by Shell Gasifier IGCC Base Case report NETL 2000, *Conceptual Cost Estimating Manual* Second Edition by John S. Page, and additional adjusted sources. Because the allowances identified did not show significant variability, the allowances identified in the NETL 2000 report were chosen for this effort in order to minimize the mixing of data sources.

IV. **BASIS OF THE ESTIMATE:** *Overall explanation of sources for resource pricing and schedules.*

A. **Quantification Basis:** *The source for the measurable quantities in the estimate that can be used in support of earned value management. Source documents may include drawings, design reports, engineers' notes, and other documentation upon which the estimate is originated.*

All islands and capacities have been provided to Cost Estimating by the respective project expert.

B. **Planning Basis:** *The source for the execution and strategies of the work that can be used to support the project execution plan, acquisition strategy, schedules, and market conditions and other documentation upon which the estimate is originated.*

1. All islands represent nth of a kind projects.
2. Projects will be constructed and operated by commercial entities.
3. All projects will be located in the U.S. Gulf Coast refinery region.
4. Costs are presented as overnight costs.
5. The cost estimate does not consider or address funding or labor resource restrictions. Sufficient funding and labor resources will be available in a manner that allows optimum usage of the funding and resources as estimated and scheduled.

C. **Cost Basis:** *The source for the costing on the estimate that can be used in support of earned value management, funding profiles, and schedule of values. Sources may include published costing references, judgment, actual costs, preliminary quotes or other documentation upon which the estimate is originated.*

1. All costs are represented as current value costs. Factors for forward-looking escalation and inflation factors are not included in this estimate.
2. Where required, published cost factors, as identified in the Chemical Engineering Plant Cost Index, will be applied to previous years' values to determine current year values.
3. Geographic location factors, as identified in RS Means Construction Cost Data reference manual, were considered for each source cost.
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8. Choi, 1996, Choi, Gerald N., et al, *Design/Economics of a Once-Through Natural Gas Fischer-Tropsch Plant with Power Co-Production*, Bechtel, 1996.
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12. Friedland, Robert J., et al., *Hydrogen Production Through Electrolysis*, NREL, June 2002.
13. Gray, 2004, Gray, David, et al, *Polygeneration of SNG, Hydrogen, Power, and Carbon Dioxide from Texas Lignite*, MTR-04, 2004-18, NETL, December 2004.
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16. Ivy, Johanna, *Summary of Electrolytic Hydrogen Production*, NREL September 2004.
17. Ibsen, Kelly, et al., *Equipment Design and Cost Estimation for Small Modular Biomass Systems, Synthesis Gas Cleanup, and Oxygen Separation Equipment*, NREL, May 2006.
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19. Kreutz, 2008, Kreutz, Thomas G., et al, "Fischer-Tropsch Fuels from Coal and Biomass," *25th Annual International Pittsburgh Coal Conference*, Pittsburgh, Princeton University, October 2008.
20. Loh, H. P., et al., *Process Equipment Cost Estimation*, DOE/NETL-2002/1169, NETL, 2002.
21. NETL, 2000, Shelton, W., et al., *Shell Gasifier IGCC Base Cases*, PED-IGCC-98-002, NETL, June 2000.
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24. NREL, 2005, Saur, Genevieve, *Wind-To-Hydrogen Project: Electrolyzer Capital Cost Study*, NREL, December 2008.
25. O'Brien, J. E., et al., *High-Temperature Electrolysis for Large-Scale Hydrogen and Syngas Production from Nuclear Energy – System Simulation and Economics*, INL, May 2009.
26. O'Brien, J. E., et al., *Parametric Study of Large-Scale Production of Syngas via High-Temperature Co-Electrolysis*, INL, January 2009.

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35. van der Ploeg, H. J., et al., *The Shell Coal Gasification Process for the US Industry*, Shell, October 2004.
36. WorleyParsons, 2002, Rameshni, Mahin, *Cost Effective Options to Expand SRU Capacity Using Oxygen*, WorleyParsons, May 2002.

V. **ESTIMATE QUALITY ASSURANCE:** *A listing of all estimate reviews that have taken place and the actions taken from those reviews.*

A review of the cost estimate was held on January 14, 2010, with the project team and the cost estimators. This review allowed for the project team to review and comment, in detail, on the perceived scope, basis of estimates, assumptions, project risks, and resources that make up this cost estimate. Comments from this review have been incorporated into this estimate to reflect a project team consensus of this document.

VI. **ASSUMPTIONS:** *Condition statements accepted or supposed true without proof of demonstration; statements adding clarification to scope. An assumption has a direct impact on total estimated cost.*

General Assumptions:

A. All costs are represented in 2009 values.

B. Costs that were included from sources representing years prior to 2009 have been normalized to 2009 values using the Chemical Engineering Plant Cost Index. This index was selected due to its widespread recognition and acceptance and its specific orientation toward work associated with chemical and refinery plants.

C. Capital costs are based on process islands. The majority of these islands are interchangeable, after factoring for the differing capacities, flowsheet-to-flowsheet.

D. All chemical processing and refinery processes will be located in the U.S. Gulf Coast region.

E. All costs considered to be balance of plant costs that can be specifically identified have been factored out of the reported source data and added into the estimate in a manner consistent with that identified in the NETL 2000 IGCC Base Cost report. Inclusion of the source costs in this manner normalizes all reported cost information to the bare-erected costs.

Coal to Ammonia

A. The air separation unit for this process requires an increase in oxygen output purity from 95 to 99.5%. A factor, based on INL simulations, of 1.36^0.6 was applied to the sources, which assumed 95% oxygen purity.
B. The NETL 2000 report lists the quench compressor separately from the gasification unit. The NETL 2007b report includes the cost of the quench compressor with the cost of the gasification unit. The costs were normalized to include both the quench compressor and gasification unit.
C. The WorleyParsons 2002 report includes engineering costs in the costs presented. Information from this report was factored by 0.9 to normalize the data by excluding the engineering allowance.

VII. CONTINGENCY GUIDELINE IMPLEMENTATION:

Contingency Methodologies: *Explanation of methodology used in determining overall contingency. Identify any specific drivers or items of concern.*

At a project risk review on December 9, 2009, the project team discussed risks to the project. An 18% allowance for capital construction contingency has been included at an island level based on the discussion and is included in the summary sheet. The contingency level that was included in the island cost source documents and additional threats and opportunities identified here were considered during this review. The contingency identified was considered by the project team and included in Cost and Performance Baseline for Fossil Energy Plants DOE/NETL-2007/1281 and similar reports. Typically, contingency allowance provided in these reports ranged from 15% to 20%. Since much of the data contained in this estimate has been derived from these reports, the project team has also chosen a level of contingency consistent with them.

A. **Threats**: *Uncertain events that are potentially negative or reduce the probability that the desired outcome will happen.*

1. The level of project definition/development that was available at the time the estimate was prepared represents a substantial risk to the project and is likely to occur. The high level at which elements were considered and included has the potential to include additional elements that are within the work scope but not sufficiently provided for or addressed at this level.
2. The estimate methodology employed is one of a stochastic parametrically evaluated process. This process used publicly available published costs that were related to the process required, costs were normalized using price indices,

and the cost was scaled to provide the required capacity. The cost-estimating relationships that were used represent typical costs for balance of plant allowances, but source cost data from which the initial island costs were derived were not completely descriptive of the elements included, not included, or simply referred to with different nomenclature or combined with other elements. While every effort has been made to correctly normalize and factor the costs for use in this effort, the risk exists that not all of these were correctly captured due to the varied information available.

3. This project is heavily dependent on metals, concrete, petroleum, and petroleum products. Competition for these commodities in today's environment due to global expansion, uncertainty, and product shortages affect the basic concepts of the supply and demand theories, thus increasing costs.
4. Impacts due to large quantities of materials, special alloy materials, fabrication capability, and labor availability could all represent conditions that may increase the total cost of the project.

B. **Opportunities:** *Uncertain events that could improve the results or improve the probability that the desired outcome will happen.*

1. Additional research and work performed with both vendors and potential owner/operators for a specific process or refinery may identify efficiencies and production means that have not been available for use in this analysis.
2. Recent historical data may identify and include technological advancements and efficiencies not included or reflected in the publicly available source data used in this effort.

Note: Contingency does not increase the overall accuracy of the estimate; it does, however, reduce the level of risk associated with the estimate. Contingency is intended to cover the inadequacies in the complete project scope definition, estimating methods, and estimating data. Contingency specifically excludes changes in project scope, unexpected work stoppages (e.g., strikes, disasters, and earthquakes) and excessive or unexpected inflation or currency fluctuations.

VIII. OTHER COMMENTS/CONCERNS SPECIFIC TO THE ESTIMATE:

None.

Detail Item Report - Air Separation Unit (ASU)

Project Name: NGNP Process Integration
Process: Conventional Coal to Ammonia
Estimate Number: MA36-L

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Shell IGCC Base Case (NETL 2000)	213,207	lb/hr	1	1999	$ 51,204,000	$ 51,204,000	$ 67,118,402	274,826	lb/hr	1	274,826	lb/hr	$ 78,161,957	$ 93,998,034
NETL Baseline Report (NETL 2007a)	1,728,789	lb/hr	2	2006	$ 287,187,000	$ 143,593,500	$ 147,157,470	274,826	lb/hr	1	274,826	lb/hr	$ 73,992,859	$ 88,984,251
Princeton Report (Kreutz 2008)	201,264	lb/hr	1	2007	$ 105,000,000	$ 105,000,000	$ 102,322,040	274,826	lb/hr	1	274,826	lb/hr	$ 123,351,439	$ 148,343,174
Hydrogen Report (Gray 2004)	296,583	lb/hr	1	2004	$ 76,000,000	$ 76,000,000	$ 87,600,180	274,826	lb/hr	1	274,826	lb/hr	$ 83,685,853	$ 83,685,853
Shell GTC Report (Shell 2004)	385,259	lb/hr	1	2004	$ 53,760,000	$ 53,760,000	$ 61,965,601	274,826	lb/hr	1	274,826	lb/hr	$ 50,598,038	$ 60,849,502
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	373,498	lb/hr	2	2006	$ 144,337,000	$ 72,168,500	$ 73,959,712	274,826	lb/hr	1	274,826	lb/hr	$ 93,255,602	$ 112,149,741

Source Selected:

Source	Reported Capacity	Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required	Trains Reqd.	Capacity per Train	Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Average of normalized and factored costs from NETL 2007a and Gray 2004										$ 78,839,356	$ 86,335,052

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 5,597,594	$ 6,129,789
Civil/Structural/Buildings	9.20%		$ 7,253,221	$ 7,942,825
Piping	7.10%		$ 5,597,594	$ 6,129,789
Control and Instrumentation	2.60%		$ 2,049,823	$ 2,244,711
Electrical Systems	8.00%		$ 6,307,148	$ 6,906,804
		Total Balance of Plant	$ 26,805,381	$ 29,353,918
		Total Balance of Plant Plus the Selected Source	$ 105,644,737	$ 115,688,970

Rationale for Selection:

NETL Baseline Report (NETL 2007a) and Hydrogen Report (Gray 2004) have been selected. An average cost of the two has been selected in order to not represent an overly agressive or conservative cost. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Note: The ASU cost was adjusted by a factor of "1.36^0.6" to account for the increase in oxygen output purity from 95% to 99.5%. The factor is based on INL simulations calculating the increase in capacity that would be needed to have the required purity output. The Gray 2004 report uses an oxygen purity of 99% and was not adjusted by the factor of "1.36^0.6."

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Coal to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-L**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity	Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required	Trains Reqd.	Capacity per Train	Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Shell IGCC Base Case (NETL 2000)	3,171 tpd	1	1999	$ 17,826,000	$ 17,826,000	$ 23,366,390	4,591 tpd	1	4,591 tpd	$ 29,174,734	$ 29,174,734
Hydrogen Report (Gray 2004)	7,787 tpd	1	2004	$ 47,000,000	$ 47,000,000	$ 54,173,796	4,591 tpd	1	4,591 tpd	$ 39,454,831	$ 39,454,831
Shell GTC Report (Shell 2004)	5,513 tpd	2	2004	$ 60,800,000	$ 30,400,000	$ 35,040,072	4,591 tpd	1	4,591 tpd	$ 47,586,129	$ 47,586,129
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	5,678 tpd	2	2006	$ 156,785,000	$ 78,392,500	$ 80,338,191	4,591 tpd	1	4,591 tpd	$ 107,190,386	$ 107,190,386

Source Selected:

Source	Reported Capacity	Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required	Trains Reqd.	Capacity per Train	Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Average of normalized and factored cost from Gray 2004 and Shell 2004 reports										$ 43,520,480	$ 43,520,480

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 3,089,954	$ 3,089,954
Civil/Structural/Buildings	9.20%		$ 4,003,884	$ 4,003,884
Piping	7.10%		$ 3,089,954	$ 3,089,954
Control and Instrumentation	2.60%		$ 1,131,532	$ 1,131,532
Electrical Systems	8.00%		$ 3,481,638	$ 3,481,638
		Total Balance of Plant	$ 14,796,963	$ 14,796,963
		Total Balance of Plant Plus the Selected Source	$ 58,317,443	$ 58,317,443

Rationale for Selection:

The Gray 2004 and the Shell 2004 reports identified recent actual costs that appear to be consistent with this project's needs. An average cost of the two has been selected in order to not represent an overly agressive or conservative cost. Cost factoring reflects the 6/10 rule. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Coal to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-L**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Gasifier														
Shell IGCC Base Case (NETL 2000)	2,977	tpd	1	1999	$ 87,802,000	$ 87,802,000	$ 115,091,203	4,215	tpd	1	4,215	tpd	$ 141,788,502	$ 141,788,502
Hydrogen Report (Gray 2004)	5,990	tpd	1	2004	$ 174,000,000	$ 174,000,000	$ 200,558,307	4,215	tpd	1	4,215	tpd	$ 162,424,203	$ 162,424,203
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	5,310	tpd	2	2006	$ 196,948,000	$ 98,474,000	$ 100,918,110	4,215	tpd	1	4,215	tpd	$ 133,168,072	$ 133,168,072
Shell GTC Report (Shell 2004)	5,201	tpd	2	2004	$ 202,240,000	$ 101,120,000	$ 116,554,345	4,215	tpd	1	4,215	tpd	$ 155,725,787	$ 155,725,787
Quench Compressor														
Shell IGCC Base Case (NETL 2000)	194,116	lb/hr	1	1999	$ 1,900,000	$ 1,900,000	$ 2,490,527	624,199	lb/hr	1	624,199	lb/hr	$ 5,019,360	$ 5,019,360

Source Selected:

Source	Reported Capacity	Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required	Trains Reqd.	Capacity per Train	Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Average of normalized and factored costs from NETL 2000 and NETL 2007b reports										$ 139,987,967	$ 139,987,967

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 9,939,146	$ 9,939,146
Civil/Structural/Buildings	9.20%		$ 12,878,893	$ 12,878,893
Piping	7.10%		$ 9,939,146	$ 9,939,146
Control and Instrumentation	2.60%		$ 3,639,687	$ 3,639,687
Electrical Systems	8.00%		$ 11,199,037	$ 11,199,037
		Total Balance of Plant	$ 47,595,909	$ 47,595,909
		Total Balance of Plant Plus the Selected Source	$ 187,583,875	$ 187,583,875

Rationale for Selection:

Shell IGCC Base Case (NETL 2000) and Shell IGCC Power Plant with CO2 Capture (NETL 2007b) are consistent in factored normalized cost per train, and in the size of trains required. An average cost of the two has been selected in order to not represent an overly agressive or conservative cost. Hydrogen Report (Gray 2004) was excluded as an unexplained and inconsistent outlier cost point. Cost factoring reflects the 6/10 rule. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Note: The reported cost of Gray 2004 is $87,000,000 for the gasification unit, and does not include a heat recovery unit. This cost has been doubled, based on information from an active vendor, UDHE, to account for the addition cost of the heat recovery unit. The quench compressor is listed as an independent line item in the NETL 2000 report. It is factored separately here to better fit the new process model. NETL 2007b includes quench compressor.

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Sources Considered:

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Princeton Report (Kreutz 2008)	815	MW	1	2007	$ 11,760,000	$ 11,760,000	$ 11,460,069	1,171	MW	1	1,171	MW	$ 14,245,269	$ 14,245,269
Hydrogen Report (Gray 2004)	48,243	lbmol/hr	1	2004	$ 23,000,000	$ 23,000,000	$ 26,510,581	70,333	lbmol/hr	1	70,333	lbmol/hr	$ 33,239,504	$ 33,239,504
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	1,714,460	lb/hr	4	2006	$ 12,367,000	$ 3,091,750	$ 3,168,487	1,360,918	lb/hr	1	1,360,918	lb/hr	$ 6,337,393	$ 6,337,393

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Hydrogen Report (Gray 2004)	48,243	lbmol/hr	1	2004	$ 23,000,000	$ 23,000,000	$ 26,510,581	70,333	lbmol/hr	1	70,333	lbmol/hr	$ 33,239,504	$ 33,239,504

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 2,360,005	$ 2,360,005
Civil/Structural/Buildings	9.20%		$ 3,058,034	$ 3,058,034
Piping	7.10%		$ 2,360,005	$ 2,360,005
Control and Instrumentation	2.60%		$ 864,227	$ 864,227
Electrical Systems	8.00%		$ 2,659,160	$ 2,659,160
		Total Balance of Plant	$ 11,301,432	$ 11,301,432
		Total Balance of Plant Plus the Selected Source	$ 44,540,936	$ 44,540,936

Rationale for Selection:

Princeton Report (Kreutz 2008) is the most recent cost available, the capacity per train most closely reflects this project's needs. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

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Sources Considered:

Source	Reported Capacity	Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required	Trains Reqd.	Capacity per Train	Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fluor/UOP Report (Fluor/UOP 2004)	31,226 lbmol/hr	1	2003	$ 22,862,000	$ 22,862,000	$ 29,117,771	49,957 lbmol/hr	1	49,957 lbmol/hr	$ 38,601,673	$ 38,601,673
Co-Production of Hydrogen Presentation (Chiesa 2003)	327 tph	1	2002	$ 32,800,000	$ 32,800,000	$ 42,450,961	96 tph	1	96 tph	$ 20,308,766	$ 20,308,766

Source Selected:

Source	Reported Capacity	Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required	Trains Reqd.	Capacity per Train	Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fluor/UOP Report (Fluor/UOP 2004)	31,226 lbmol/hr	1	2003	$ 22,862,000	$ 22,862,000	$ 29,117,771	49,957 lbmol/hr	1	49,957 lbmol/hr	$ 38,601,673	$ 38,601,673

Outside of Battery Limits - Including Balance of Plant and Offsites:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 2,740,719	$ 2,740,719
Civil/Structural/Buildings	9.20%		$ 3,551,354	$ 3,551,354
Piping	7.10%		$ 2,740,719	$ 2,740,719
Control and Instrumentation	2.60%		$ 1,003,643	$ 1,003,643
Electrical Systems	8.00%		$ 3,088,134	$ 3,088,134
		Total of Outside of Battery Limits - Including Balance of Plant and Offsites	$ 13,124,569	$ 13,124,569
		Total of Outside of Battery Limits - Including Balance of Plant and Offsites - Plus the Selected Sourc	$ 51,726,242	$ 51,726,242

Rationale for Selection:

Fluor/UOP 2004 was selected as the most recent cost point, because the reported capacity is similar the required capacity per train. The markups listed under 'Outside of Battery Limits' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Pressure Swing Adsorption (PSA)

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Coal to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-L**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fluor/UOP Report (Fluor/UOP 2004)	27,498	lbmol/hr	1	2003	$ 25,000,000	$ 25,000,000	$ 31,840,796	43,504	lbmol/hr	1	43,504	lbmol/hr	$ 41,929,605	$ 41,929,605

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fluor/UOP Report (Fluor/UOP 2004)	27,498	lbmol/hr	1	2003	$ 25,000,000	$ 25,000,000	$ 31,840,796	43,504	lbmol/hr	1	43,504	lbmol/hr	$ 41,929,605	$ 41,929,605

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 2,977,002	$ 2,977,002
Civil/Structural/Buildings	9.20%		$ 3,857,524	$ 3,857,524
Piping	7.10%		$ 2,977,002	$ 2,977,002
Control and Instrumentation	2.60%		$ 1,090,170	$ 1,090,170
Electrical Systems	8.00%		$ 3,354,368	$ 3,354,368
		Total Balance of Plant	$ 14,256,066	$ 14,256,066
		Total Balance of Plant Plus the Selected Source	$ 56,185,670	$ 56,185,670

Rationale for Selection:

Single source cost point. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Coal to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-L**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
DOE FE Report (DOE 1978)	1,000	tpd	1	1978	$ 1,467,000	$ 1,467,000	$ 3,432,834	3,358	tpd	1	3,358	tpd	$ 7,100,700	$ 7,100,700

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
DOE FE Report (DOE 1978)	1,000	tpd	1	1978	$ 1,467,000	$ 1,467,000	$ 3,432,834	3,358	tpd	1	3,358	tpd	$ 7,100,700	$ 7,100,700

Balance of Plant:

Description	Markup						Cost Per Train	Total Cost
Water Systems	7.10%						$ 504,150	$ 504,150
Civil/Structural/Buildings	9.20%						$ 653,264	$ 653,264
Piping	7.10%						$ 504,150	$ 504,150
Control and Instrumentation	2.60%						$ 184,618	$ 184,618
Electrical Systems	8.00%						$ 568,056	$ 568,056
						Total Balance of Plant	$ 2,414,238	$ 2,414,238
						Total Balance of Plant Plus the Selected Source	$ 9,514,938	$ 9,514,938

Rationale for Selection:

Single source island cost identified by the project technical lead. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Claus and SCOT

Project Name: NGNP Process Integration
Process: Conventional Coal to Ammonia
Estimate Number: MA36-L

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Claus and SCOT														
Princeton Report (Kreutz 2008)	12,586	lb/hr	1	2007	$ 33,800,000	$ 33,800,000	$ 32,937,952	12,762	lb/hr	1	12,762	lb/hr	$ 33,213,543	$ 33,213,543
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	142	tpd	1	2006	$ 22,794,000	$ 22,794,000	$ 24,926,373	153	tpd	1	153	tpd	$ 26,093,339	$ 26,093,339
Claus														
Shell IGCC Base Cases (NETL 2000)	6,496	lb/hr	1	1999	$ 9,964,000	$ 9,964,000	$ 13,060,850	12,762	lb/hr	1	12,762	lb/hr	$ 19,586,061	$ 19,586,061
Cost Effective Options to Expand SRU Capacity Using Oxygen (WorleyParsons 2002)	79	tpd	1	1999	$ 11,970,000	$ 11,970,000	$ 15,690,323	153	tpd	1	153	tpd	$ 23,276,222	$ 23,276,222
SCOT														
Shell IGCC Base Cases (NETL 2000)	6,496	lb/hr	1	1999	$ 4,214,000	$ 4,214,000	$ 5,523,728	12,762	lb/hr	1	12,762	lb/hr	$ 8,283,386	$ 8,283,386
Cost Effective Options to Expand SRU Capacity Using Oxygen (WorleyParsons 2002)	143	tpd	1	1999	$ 8,910,000	$ 8,910,000	$ 11,679,263	153	tpd	1	153	tpd	$ 12,154,288	$ 12,154,288

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
WorleyParsons 2002: Combined Claus and SCOT costs													$ 35,430,510	$ 35,430,510

Detail Item Report - Claus and SCOT

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Coal to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-L**	**Estimate Type:**	**Class 5**

Balance of Plant:

Description	Markup						Cost Per Train	Total Cost
Water Systems	7.10%						$ 2,515,566	$ 2,515,566
Civil/Structural/Buildings	9.20%						$ 3,259,607	$ 3,259,607
Piping	7.10%						$ 2,515,566	$ 2,515,566
Control and Instrumentation	2.60%						$ 921,193	$ 921,193
Electrical Systems	8.00%						$ 2,834,441	$ 2,834,441
						Total Balance of Plant	$ 12,046,373	$ 12,046,373
						Total Balance of Plant Plus the Selected Source	$ 47,476,883	$ 47,476,883

Rationale for Selection:

The WorleyParsons 2002 cost point was selected because of WorleyParsons' status as a working vendor in this industry. It is expected that this is the highest quality information available at this time. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Note: Costs from WorleyParsons 2002 have been multiplied by 0.9 to adjust for the included engineering costs. This factor was consistent with general process industry standards, and was selected with project team consensus.

Detail Item Report - CO2 Compression

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Coal to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-L**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Subcritical														
Princeton Report (Kreutz 2008)	10	MW	1	2007	$ 6,310,000	$ 6,310,000	$ 6,149,067	6	MW	1	6	MW	$ 4,730,392	$ 4,730,392
Supercritical														
Princeton Report (Kreutz 2008)	13	MW	1	2007	$ 9,520,000	$ 9,520,000	$ 9,277,198	1	MW	1	1	MW	$ 1,885,194	$ 1,885,194

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Kreutz 2008: Combined Subcritical and Supercritical Processes													$ 6,615,586	$ 6,615,586

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 469,707	$ 469,707
Civil/Structural/Buildings	9.20%		$ 608,634	$ 608,634
Piping	7.10%		$ 469,707	$ 469,707
Control and Instrumentation	2.60%		$ 172,005	$ 172,005
Electrical Systems	8.00%		$ 529,247	$ 529,247
		Total Balance of Plant	$ 2,249,299	$ 2,249,299
		Total Balance of Plant Plus the Selected Source	$ 8,864,886	$ 8,864,886

Rationale for Selection:

Single source cost point. Both subcritical and supercritical process costs were included under the CO2 Compression heading. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Project Name: NGNP Process Integration
Process: Conventional Coal to Ammonia
Estimate Number: MA36-L

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	360	tpd	1	2009	$ 44,000,000	$ 44,000,000	$ 44,000,000	3,358	tpd	2	1,679	tpd	$ 110,841,296	$ 221,682,591
Economics Ammonia Coal Gasification (Morel 1977)	1,000	tpd	1	1977	$ 50,748,000	$ 50,748,000	$ 80,243,904	3,358	tpd	2	1,679	tpd	$ 109,507,147	$ 219,014,295
Stamicarbon, Middle East Fertilizer Symposium, March 2009	4,297	tpd	1	2008	$ 800,000,000	$ 800,000,000	$ 779,596,498	3,358	tpd	2	1,679	tpd	$ 443,599,396	$ 887,198,792
Ammonia, Chem Systems, 1998	1,653	tpd	1	1998	$ 160,000,000	$ 160,000,000	$ 210,320,924	3,358	tpd	2	1,679	tpd	$ 212,264,936	$ 424,529,872

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	360	tpd	1	2009	$ 44,000,000	$ 44,000,000	$ 44,000,000	3,358	tpd	2	1,679	tpd	$ 110,841,296	$ 221,682,591

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 7,869,732	$ 15,739,464
Civil/Structural/Buildings	9.20%		$ 10,197,399	$ 20,394,798
Piping	7.10%		$ 7,869,732	$ 15,739,464
Control and Instrumentation	2.60%		$ 2,881,874	$ 5,763,747
Electrical Systems	8.00%		$ 8,867,304	$ 17,734,607
		Total Balance of Plant	$ 37,686,041	$ 75,372,081
		Total Balance of Plant Plus the Selected Source	$ 148,527,336	$ 297,054,672

Rationale for Selection:

The verbal cost was selected as recommended by the project team expert. The Stamicarbon information shows a roughly 350% increase in prices between 2003 and 2008 emphasizing the importance of using the most recent information available. The markups listed under 'Outside of Battery Limits' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Urea Synthesis

Project Name: NGNP Process Integration
Process: Conventional Coal to Ammonia
Estimate Number: MA36-L

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	625	tpd	1	2009	$ 85,000,000	$ 85,000,000	$ 85,000,000	2,939	tpd	1	2,939	tpd	$ 215,184,342	$ 215,184,342
Perry's Chemical Engineering Handbook, 7th Edition	200	tpd	1	1994	$ 8,800,000	$ 8,800,000	$ 12,240,152	2,939	tpd	1	2,939	tpd	$ 61,388,719	$ 61,388,719
PNNL	4681	tpd	1	2009	$ 182,000,000	$ 182,000,000	$ 182,000,000	2,939	tpd	1	2,939	tpd	$ 137,653,721	$ 137,653,721
Stamicarbon, Middle East Fertilizer Symposium, March 2009	3,582	tpd	1	2008	$ 550,000,000	$ 550,000,000	$ 535,972,592	2,939	tpd	1	2,939	tpd	$ 475,978,115	$ 475,978,115

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	625	tpd	1	2009	$ 85,000,000	$ 85,000,000	$ 85,000,000	2,939	tpd	1	2,939	tpd	$ 215,184,342	$ 215,184,342

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 15,278,088	$ 15,278,088
Civil/Structural/Buildings	9.20%		$ 19,796,959	$ 19,796,959
Piping	7.10%		$ 15,278,088	$ 15,278,088
Control and Instrumentation	2.60%		$ 5,594,793	$ 5,594,793
Electrical Systems	8.00%		$ 17,214,747	$ 17,214,747
		Total Balance of Plant	$ 73,162,676	$ 73,162,676
		Total Balance of Plant Plus the Selected Source	$ 288,347,019	$ 288,347,019

Rationale for Selection:

The verbal cost was selected as recommended by the project team expert. The Stamicarbon information shows a roughly 350% increase in prices between 2003 and 2008 emphasizing the importance of using the most recent information available. The markups listed under 'Outside of Battery Limits' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Project Name: NGNP Process Integration
Process: Conventional Coal to Ammonia
Estimate Number: MA36-L

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Perry's Chemical Engineering Handbook, 7th Edition	334	tpd	1	1994	$ 6,600,000	$ 6,600,000	$ 9,180,114	5,186	tpd	6	864	tpd	$ 16,239,973	$ 97,439,841
Fertilizer Manual	359	tpd	1	1998	$ 15,200,000	$ 15,200,000	$ 19,980,488	5,186	tpd	6	864	tpd	$ 33,828,481	$ 202,970,888

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fertilizer Manual	359	tpd	1	1998	$ 15,200,000	$ 15,200,000	$ 19,980,488	5,186	tpd	6	864	tpd	$ 33,828,481	$ 202,970,888

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 2,401,822	$ 14,410,933
Civil/Structural/Buildings	9.20%		$ 3,112,220	$ 18,673,322
Piping	7.10%		$ 2,401,822	$ 14,410,933
Control and Instrumentation	2.60%		$ 879,541	$ 5,277,243
Electrical Systems	8.00%		$ 2,706,279	$ 16,237,671
		Total Balance of Plant	$ 11,501,684	$ 69,010,102
		Total Balance of Plant Plus the Selected Source	$ 45,330,165	$ 271,980,990

Rationale for Selection:

The Fertilizer Manual was selected for being both the newest cost point and the most conservative. The data from Perry's Chemical Engineering Handbook is based on earlier data from Peters and Timmerhaus, making it even less desirable. The markups listed under 'Outside of Battery Limits' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Ammonium Nitrate Synthesis

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Coal to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-L**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Perry's Chemical Engineering Handbook, 7th Edition	334	tpd	1	1994	$ 6,000,000	$ 6,000,000	$ 8,345,558	3,779	tpd	3	1,260	tpd	$ 18,507,052	$ 55,521,157
Fertilizer Manual	1,395	tpd	1	1998	$ 35,000,000	$ 35,000,000	$ 46,007,702	3,779	tpd	3	1,260	tpd	$ 43,270,765	$ 129,812,295
EPA Report	1,200	tpd	1	1981	$ 9,480,000	$ 9,480,000	$ 13,434,154	3,779	tpd	3	1,260	tpd	$ 13,831,044	$ 41,493,133

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fertilizer Manual	1395.24	tpd	1.00	1998.00	35000000.00	35000000.00	46007702.18	3779.00	tpd	3.00	1259.67	tpd	43270765.12	129812295.36

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 3,072,224	$ 9,216,673
Civil/Structural/Buildings	9.20%		$ 3,980,910	$ 11,942,731
Piping	7.10%		$ 3,072,224	$ 9,216,673
Control and Instrumentation	2.60%		$ 1,125,040	$ 3,375,120
Electrical Systems	8.00%		$ 3,461,661	$ 10,384,984
		Total Balance of Plant	$ 14,712,060	$ 44,136,180
		Total Balance of Plant Plus the Selected Source	$ 57,982,825	$ 173,948,476

Rationale for Selection:

The Fertilizer Manual was selected for being both the newest cost point and the most conservative. The data from Perry's Chemical Engineering Handbook is based on earlier data from Peters and Timmerhaus, making it even less desirable. The markups listed under 'Outside of Battery Limits' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Coal to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-L**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Steam Turbine and HRSG														
Shell IGCC Base Cases (NETL 2000)	189	MW	1	1999	$ 50,671,000	$ 50,671,000	$ 66,419,744	47	MW	1	47	MW	$ 28,634,650	$ 28,634,650
Steam Turbine														
NETL Baseline Report (NETL 2007a)	401	MW	4	2006	$ 74,651,000	$ 18,662,750	$ 19,125,957	47	MW	1	47	MW	$ 12,062,694	$ 12,062,694
Princeton Report (Kreutz 2008)	275	MW	1	2007	$ 66,700,000	$ 66,700,000	$ 64,998,858	47	MW	1	47	MW	$ 22,375,764	$ 22,375,764
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	230	MW	1	2006	$ 44,515,000	$ 44,515,000	$ 45,619,856	47	MW	1	47	MW	$ 17,485,107	$ 17,485,107

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	230	MW	1	2006	$ 44,515,000	$ 44,515,000	$ 45,619,856	47	MW	1	47	MW	$ 17,485,107	$ 17,485,107

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 1,241,443	$ 1,241,443
Civil/Structural/Buildings	9.20%		$ 1,608,630	$ 1,608,630
Piping	7.10%		$ 1,241,443	$ 1,241,443
Control and Instrumentation	2.60%		$ 454,613	$ 454,613
Electrical Systems	8.00%		$ 1,398,809	$ 1,398,809
		Total Balance of Plant	$ 5,944,936	$ 5,944,936
		Total Balance of Plant Plus the Selected Source	$ 23,430,044	$ 23,430,044

Rationale for Selection:

Shell IGCC PowerPlant with CO2 Capture (NETL 2007b) is a recently reported cost point that closely reflects this project's requirements. The Princeton Report (Kreutz 2008) source for the steam turbine cost point is the NETL 2007b report. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Project Name: NGNP Process Integration
Process: Conventional Coal to Ammonia
Estimate Number: MA36-L

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Steam Turbine and HRSG														
Shell IGCC Base Cases (NETL 2000)	189	MW	1	1999	$ 50,671,000	$ 50,671,000	$ 66,419,744	47	MW	1	47	MW	$ 28,634,650	$ 28,634,650
HRSG														
NETL Baseline Report (NETL 2007a)	5,155,983	lb/hr	3	2006	$ 27,581,000	$ 9,193,667	$ 9,421,852	1,117,170	lb/hr	1	1,117,170	lb/hr	$ 7,276,017	$ 7,276,017
Princeton Report (Kreutz 2008)	355	MW	1	2007	$ 52,000,000	$ 52,000,000	$ 50,673,772	49	MW	1	49	MW	$ 15,411,146	$ 15,411,146
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	8,438,000	lb/hr	2	2006	$ 45,291,000	$ 22,645,500	$ 23,207,558	1,117,170	lb/hr	1	1,117,170	lb/hr	$ 10,456,234	$ 10,456,234

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	8,438,000	lb/hr	2	2006	$ 45,291,000	$ 22,645,500	$ 23,207,558	1,117,170	lb/hr	1	1,117,170	lb/hr	$ 10,456,234	$ 10,456,234

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 742,393	$ 742,393
Civil/Structural/Buildings	9.20%		$ 961,974	$ 961,974
Piping	7.10%		$ 742,393	$ 742,393
Control and Instrumentation	2.60%		$ 271,862	$ 271,862
Electrical Systems	8.00%		$ 836,499	$ 836,499
		Total Balance of Plant	$ 3,555,120	$ 3,555,120
		Total Balance of Plant Plus the Selected Source	$ 14,011,353	$ 14,011,353

Rationale for Selection:

Shell IGCC PowerPlant with CO2 Capture (NETL 2007b) is a recently reported cost point that closely reflects this project's requirements. The Princeton Report (Kreutz 2008) source for the steam turbine cost point is the NETL 2007b report. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Cooling Towers

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Coal to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-L**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Cooling Tower Depot	140,001	gpm	5	2009	$ 3,452,790	$ 690,558	$ 690,558	130,392	gpm	5	26,078	gpm	$ 661,717	$ 3,308,583

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Cooling Tower Depot	140,001	gpm	5	2009	$ 3,452,790	$ 690,558	$ 690,558	130,392	gpm	5	26,078	gpm	$ 661,717	$ 3,308,583

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 46,982	$ 234,909
Civil/Structural/Buildings	9.20%		$ 60,878	$ 304,390
Piping	7.10%		$ 46,982	$ 234,909
Control and Instrumentation	2.60%		$ 17,205	$ 86,023
Electrical Systems	8.00%		$ 52,937	$ 264,687
		Total Balance of Plant	$ 224,984	$ 1,124,918
		Total Balance of Plant Plus the Selected Source	$ 886,700	$ 4,433,502

Rationale for Selection:

Single source cost. Publically available current data. Calculated capital costs based on publically available cost data from a vendor regularly engaged in the building of cooling towers.

NGNP Conventional Gas to Ammonia Summary

Project Name: NGNP Process Integration
Process: Conventional Gas to Ammonia
Estimate Number: MA36-M

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Process Component	Subtotal From Detail Sheets	Engineering %	Engineering	Contingency %	Contingency	Total Cost
Steam Methane Reforming	$ 118,901,263	10%	$ 11,890,126	18%	$ 23,542,450	$ 154,333,839
Water Gas Shift Reactor	$ 39,287,822	10%	$ 3,928,782	18%	$ 7,778,989	$ 50,995,593
Selexol	$ 46,435,179	10%	$ 4,643,518	18%	$ 9,194,165	$ 60,272,862
Methanation	$ 9,520,038	10%	$ 952,004	18%	$ 1,884,967	$ 12,357,009
CO2 Compression	$ 12,478,201	10%	$ 1,247,820	18%	$ 2,470,684	$ 16,196,705
Ammonia Synthesis	$ 297,213,875	10%	$ 29,721,388	18%	$ 58,848,347	$ 385,783,610
Urea Synthesis	$ 288,347,019	10%	$ 28,834,702	18%	$ 57,092,710	$ 374,274,430
Nitric Acid Synthesis	$ 272,169,749	10%	$ 27,216,975	18%	$ 53,889,610	$ 353,276,334
Ammonium Nitrate Synthesis	$ 173,948,476	10%	$ 17,394,848	18%	$ 34,441,798	$ 225,785,122
Steam Turbines	$ 17,613,271	10%	$ 1,761,327	18%	$ 3,487,428	$ 22,862,026
Heat Recovery Steam Generator (HRSG)	$ -	10%	$ -	18%	$ -	$ -
Cooling Towers	$ 3,263,515	10%	$ 326,352	18%	$ 646,176	$ 4,236,043
Total Cost - Conventional Gas to Ammonia						$ 1,660,373,573
Total Cost Rounded to the Nearest $10M						$ 1,660,000,000

	Remarks
Checked By: SNW	
Approved By: John R	

COST ESTIMATE SUPPORT DATA RECAPITULATION

Project Title: NGNP Process Integration – Conventional Gas to Ammonia
Estimator: B. W. Wallace/CEP, R. R. Honsinger/CEP, J. B. Martin/CCT
Date: April 20, 2010
Estimate Type: Class 5
File: MA36-M
Approved By:

I. **PURPOSE:** *Brief description of the intent of how the estimate is to be used (i.e., for engineering study, comparative analysis, request for funding, proposal, etc.).*

It is expected that the capital costs identified in these estimates will be used in a model producing an economic analysis for each specific integrated application and subsequently will be considered in a related feasibility study.

II. **SCOPE OF WORK:** *Brief statement of the project's objective. Thorough overview and description of the proposed project. Identify work to be accomplished, as well as any specific work to be excluded.*

A. **Objective:**
Develop Class 5 estimates as defined by the Association for Advancement of Cost Engineering (AACEi) that will identify the current capital cost associated with a gas-to-ammonia process.

B. **Included:**
The scope of work required to achieve this objective includes the following:
1. Engineering
2. Construction of a new gas-to-ammonia refinery that consists of the following:
   a. Steam methane reformer
   b. Water gas shift reactors
   c. Selexol process
   d. Methanation
   e. CO_2 compression
   f. Ammonia synthesis
   g. Urea synthesis
   h. Nitric acid synthesis
   i. Ammonium nitrate synthesis
   j. Steam turbines, internal to process
   k. Heat recovery steam generator, internal to process
   l. Cooling towers, internal to process
   m. Allowances for Balance of Plant (BOP)/offsite/outside of battery limits (OSBL), including the following:
      (1.) Site development/improvements
      (2.) Provisions for general and administrative buildings and structures
      (3.) Provisions for OSBL piping
      (4.) Provisions for OSBL instrumentation and control

(5.) Provisions for OSBL electrical
(6.) Provisions for facility supply and OSBL water systems
(7.) Provisions for site development/improvements
(8.) Project/construction management.

C. **Excluded:**
This scope of work specifically excludes the following elements:
1. Licensing and permitting costs
2. Operational costs
3. Land costs
4. Sales taxes
5. Royalties
6. Owner's fees and owner's costs.

III. **ESTIMATE METHODOLOGY:** *Overall methodology and rationale of how the estimate was developed (i.e., parametric, forced detail, bottoms up, etc.). Total dollars/hours and rough order magnitude (ROM) allocations of the methodologies used to develop the cost estimate.*

Consistent with the AACEi Class 5 estimates, the level of definition and engineering development available at the time they were prepared, their intended use in a feasibility study, and the time and resources available for their completion, the costs included in this estimate have been developed using parametric evaluations. These evaluations have used publicly available and published project costs to represent similar islands utilized in this project. Analysis and selection of the published costs used have been performed by the project technical lead and Cost Estimating. Suitability for use in this effort was determined considering the correctness and completeness of the data available, the manner in which total capital costs were represented, the age of the previously performed work, and the similarity to the capacity/trains required by this project. The specific sources, selected and used in this cost estimate, are identified in the capital cost estimate detail sheets. Adjustments have been made to these published costs using escalation factors identified in the Chemical Engineering Price Cost Index. Scaling of the published island costs has been accomplished using the six-tenths capacity factoring method. Any normalization to provide for geographic factors was considered using geographic factors available from RS Means Construction Cost Data references. Cost-estimating relationships have been used to identify allowances to complete the costs.

BOP/OSBL costs were determined by the project team, considering data provided by Shell Gasifier IGCC Base Case report NETL 2000, *Conceptual Cost Estimating Manual* Second Edition by John S. Page, and additional adjusted sources. Because the allowances identified did not show significant variability, the allowances identified in the NETL 2000 report were chosen for this effort in order to minimize the mixing of data sources.

IV. **BASIS OF THE ESTIMATE:** *Overall explanation of sources for resource pricing and schedules.*

A. **Quantification Basis:** *The source for the measurable quantities in the estimate that can be used in support of earned value management. Source documents may include drawings, design reports, engineers' notes, and other documentation upon which the estimate is originated.*

All islands and capacities have been provided to Cost Estimating by the respective project expert.

B. **Planning Basis:** *The source for the execution and strategies of the work that can be used to support the project execution plan, acquisition strategy, schedules, and market conditions and other documentation upon which the estimate is originated.*

1. All islands represent nth of a kind projects.
2. Projects will be constructed and operated by commercial entities.
3. All projects will be located in the U.S. Gulf Coast refinery region.
4. Costs are presented as overnight costs.
5. The cost estimate does not consider or address funding or labor resource restrictions. Sufficient funding and labor resources will be available in a manner that allows optimum usage of the funding and resources as estimated and scheduled.

C. **Cost Basis:** *The source for the costing on the estimate that can be used in support of earned value management, funding profiles, and schedule of values. Sources may include published costing references, judgment, actual costs, preliminary quotes or other documentation upon which the estimate is originated.*

1. All costs are represented as current value costs. Factors for forward-looking escalation and inflation factors are not included in this estimate.
2. Where required, published cost factors, as identified in the Chemical Engineering Plant Cost Index, will be applied to previous years' values to determine current year values.
3. Geographic location factors, as identified in RS Means Construction Cost Data reference manual, were considered for each source cost.
4. Apt, Jay, et al., *An Engineering-Economic Analysis of Syngas Storage*, NETL, July 2008.
5. AACEi, *Recommended Practices*, website, visited November 16, 2009, http://www.aacei.org/technical/rp.shtml.
6. Brown, L. C., et al., "Alternative Flowsheets for the Sulfur-Iodine Thermochemical Hydrogen Cycle," *General Atomics*, February 2003.
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8. Choi, 1996, Choi, Gerald N., et al, *Design/Economics of a Once-Through Natural Gas Fischer-Tropsch Plant with Power Co-Production*, Bechtel, 1996.

9. Dooley, J., et al, *Carbon Dioxide Capture and Geologic Storage*, Battelle, April 2006.
10. Douglas, Fred R., et al., *Conduction Technical and Economic Evaluations – as Applied for the Process and Utility Industries*, AACEi, April 1991.
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12. Friedland, Robert J., et al., *Hydrogen Production Through Electrolysis*, NREL, June 2002.
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24. NREL, 2005, Saur, Genevieve, *Wind-To-Hydrogen Project: Electrolyzer Capital Cost Study*, NREL, December 2008.
25. O'Brien, J. E., et al., *High-Temperature Electrolysis for Large-Scale Hydrogen and Syngas Production from Nuclear Energy – System Simulation and Economics*, INL, May 2009.
26. O'Brien, J. E., et al., *Parametric Study of Large-Scale Production of Syngas via High-Temperature Co-Electrolysis*, INL, January 2009.
27. Page, John S., *Conceptual Cost Estimating Manual – 2nd ed.,* Houston: Gulf Publishing Company, 1996.

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29. Ramsden, Todd, et al., *Current (2005) Hydrogen Production from Central Grid Electrolysis*, NREL, May 2008.
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35. van der Ploeg, H. J., et al., *The Shell Coal Gasification Process for the US Industry*, Shell, October 2004.
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V. **ESTIMATE QUALITY ASSURANCE:** *A listing of all estimate reviews that have taken place and the actions taken from those reviews.*

A review of the cost estimate was held on January 14, 2010, with the project team and the cost estimators. This review allowed for the project team to review and comment, in detail, on the perceived scope, basis of estimates, assumptions, project risks, and resources that make up this cost estimate. Comments from this review have been incorporated into this estimate to reflect a project team consensus of this document.

VI. **ASSUMPTIONS:** *Condition statements accepted or supposed true without proof of demonstration; statements adding clarification to scope. An assumption has a direct impact on total estimated cost.*

General Assumptions:

A. All costs are represented in 2009 values.
B. Costs that were included from sources representing years prior to 2009 have been normalized to 2009 values using the Chemical Engineering Plant Cost Index. This index was selected due to its widespread recognition and acceptance and its specific orientation toward work associated with chemical and refinery plants.
C. Capital costs are based on process islands. The majority of these islands are interchangeable, after factoring for the differing capacities, flowsheet-to-flowsheet.
D. All chemical processing and refinery processes will be located in the U.S. Gulf Coast region.
E. All costs considered to be balance of plant costs that can be specifically identified have been factored out of the reported source data and added into the estimate in a

manner consistent with that identified in the NETL 2000 IGCC Base Cost report. Inclusion of the source costs in this manner normalizes all reported cost information to the bare-erected costs.

Gas to Ammonia

The NREL 2006 report presented the steam methane reformer cost with the cost of the water gas shift reactors. This cost was normalized to exclude the cost of the water gas shift reactors.

VII. CONTINGENCY GUIDELINE IMPLEMENTATION:

Contingency Methodologies: *Explanation of methodology used in determining overall contingency. Identify any specific drivers or items of concern.*

At a project risk review on December 9, 2009, the project team discussed risks to the project. An 18% allowance for capital construction contingency has been included at an island level based on the discussion and is included in the summary sheet. The contingency level that was included in the island cost source documents and additional threats and opportunities identified here were considered during this review. The contingency identified was considered by the project team and included in Cost and Performance Baseline for Fossil Energy Plants DOE/NETL-2007/1281 and similar reports. Typically, contingency allowance provided in these reports ranged from 15% to 20%. Since much of the data contained in this estimate has been derived from these reports, the project team has also chosen a level of contingency consistent with them.

A. **Threats:** *Uncertain events that are potentially negative or reduce the probability that the desired outcome will happen.*

1. The level of project definition/development that was available at the time the estimate was prepared represents a substantial risk to the project and is likely to occur. The high level at which elements were considered and included has the potential to include additional elements that are within the work scope but not sufficiently provided for or addressed at this level.
2. The estimate methodology employed is one of a stochastic parametrically evaluated process. This process used publicly available published costs that were related to the process required, costs were normalized using price indices, and the cost was scaled to provide the required capacity. The cost-estimating relationships that were used represent typical costs for balance of plant allowances, but source cost data from which the initial island costs were derived were not completely descriptive of the elements included, not included, or simply referred to with different nomenclature or combined with other elements. While every effort has been made to correctly normalize and factor the costs for use in this effort, the risk exists that not all of these were correctly captured due to the varied information available.

3. This project is heavily dependent on metals, concrete, petroleum, and petroleum products. Competition for these commodities in today's environment due to global expansion, uncertainty, and product shortages affect the basic concepts of the supply and demand theories, thus increasing costs.
4. Impacts due to large quantities of materials, special alloy materials, fabrication capability, and labor availability could all represent conditions that may increase the total cost of the project.

B. **Opportunities**: *Uncertain events that could improve the results or improve the probability that the desired outcome will happen.*

1. Additional research and work performed with both vendors and potential owner/operators for a specific process or refinery may identify efficiencies and production means that have not been available for use in this analysis.
2. Recent historical data may identify and include technological advancements and efficiencies not included or reflected in the publicly available source data used in this effort.

Note: Contingency does not increase the overall accuracy of the estimate; it does, however, reduce the level of risk associated with the estimate. Contingency is intended to cover the inadequacies in the complete project scope definition, estimating methods, and estimating data. Contingency specifically excludes changes in project scope, unexpected work stoppages (e.g., strikes, disasters, and earthquakes) and excessive or unexpected inflation or currency fluctuations.

VIII. OTHER COMMENTS/CONCERNS SPECIFIC TO THE ESTIMATE:

None.

Detail Item Report - Methane Reforming

Project Name: NGNP Process Integration
Process: Conventional Gas to Ammonia
Estimate Number: MA36-M

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Natural Gas to Liquids Conversion Project (Raytheon 2000)	150	MMS CFD	1	1999	$ 79,000,000	$ 79,000,000	$ 103,553,507	88	MMS CFD	1	88	MMS CFD	$ 75,196,852	$ 88,732,286
NETL Natural Gas Report (Choi 1996)	100	MMS CFD	1	1996	$ 22,800,000	$ 22,800,000	$ 30,583,181	88	MMS CFD	1	88	MMS CFD	$ 28,325,152	$ 28,325,152
Gas Usage & Value The Technology and Economics of Natural Gas Use in the Process Industries (PennWell Corporation 2006)	4,700	tpd	2	1985	$ 168,150,000	$ 84,075,000	$ 132,328,312	3,227	tpd	1	3,227	tpd	$ 160,063,178	$ 160,063,178
NREL Report (NREL 2006)	183	MMS CFD	1	2002	$ 175,391,586	$ 175,391,586	$ 226,998,210	88	MMS CFD	1	88	MMS CFD	$ 146,395,026	$ 146,395,026

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Natural Gas to Liquids Conversion Project (Raytheon 2000)	150	MMS CFD	1	1999	$ 79,000,000	$ 79,000,000	$ 103,553,507	88	MMS CFD	1	88	MMS CFD	$ 75,196,852	$ 88,732,286

Detail Item Report - Methane Reforming

Project Name: NGNP Process Integration
Process: Conventional Gas to Ammonia
Estimate Number: MA36-M

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Balance of Plant:

Description	Markup						Cost Per Train	Total Cost
Water Systems	7.10%						$ 5,338,977	$ 6,299,992
Civil/Structural/Buildings	9.20%						$ 6,918,110	$ 8,163,370
Piping	7.10%						$ 5,338,977	$ 6,299,992
Control and Instrumentation	2.60%						$ 1,955,118	$ 2,307,039
Electrical Systems	8.00%						$ 6,015,748	$ 7,098,583
						Total Balance of Plant	$ 25,566,930	$ 30,168,977
						Total Balance of Plant Plus the Selected Source	$ 100,763,782	$ 118,901,263

Rationale for Selection:

Raytheon 2000 was selected as the most recent cost point, with technology most similar to the intended process. The adjusted PennWell Corporation 2006 cost point concurs with the Raytheon 2000 value. Costs presented in the Raytheon report have been increased by 18% to account for the difference between an autothermal process and traditional steam methane reforming process. The markups listed under 'Balance of Plant' are based on NETL 200 These markup values are comparable to additional published estimating guides, such as Page 1996.

Note: PennWell Corporation 2006 includes both the natural gas reformer and methanol synthesis units. The average cost for methanol synthesis, as calculated in this report, was subtracted from the total current cost for required trains cell so that this may be considered as a cost point for natural gas reforming. The NREL 2006 cost was scaled from the originally presented value to exclude the cost of the water gas shift reactors.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-M**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Hydrogen Report (Gray 2004)	48,243	lbmol/hr	1	2004	$ 23,000,000	$ 23,000,000	$ 26,510,581	57,059	lbmol/hr	1	57,059	lbmol/hr	$ 29,319,270	$ 29,319,270
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	1,714,460	lb/hr	4	2006	$ 12,367,000	$ 3,091,750	$ 3,168,487	893,700	lb/hr	1	903,998	lb/hr	$ 4,958,062	$ 4,958,062

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Hydrogen Report (Gray 2004)	48,243	lbmol/hr	1	2004	$ 23,000,000	$ 23,000,000	$ 26,510,581	57,059	lbmol/hr	1	57,059	lbmol/hr	$ 29,319,270	$ 29,319,270

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 2,081,668	$ 2,081,668
Civil/Structural/Buildings	9.20%		$ 2,697,373	$ 2,697,373
Piping	7.10%		$ 2,081,668	$ 2,081,668
Control and Instrumentation	2.60%		$ 762,301	$ 762,301
Electrical Systems	8.00%		$ 2,345,542	$ 2,345,542
		Total Balance of Plant	$ 9,968,552	$ 9,968,552
		Total Balance of Plant Plus the Selected Source	$ 39,287,822	$ 39,287,822

Rationale for Selection:

Gray 2004 is the most recent applicable cost available, the capacity per train most closely reflects this project's needs. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-M**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fluor/UOP Report (Fluor/UOP 2004)	31,226	lbmol/hr	1	2003	$ 22,862,000	$ 22,862,000	$ 29,117,771	41,734	lbmol/hr	1	41,734	lbmol/hr	$ 34,653,118	$ 34,653,118
Co-Production of Hydrogen Presentation (Chiesa 2003)	327	tph	1	2002	$ 32,800,000	$ 32,800,000	$ 42,450,961	149	tph	1	149	tph	$ 26,516,623	$ 26,516,623

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fluor/UOP Report (Fluor/UOP 2004)	31,226	lbmol/hr	1	2003	$ 22,862,000	$ 22,862,000	$ 29,117,771	41,734	lbmol/hr	1	41,734	lbmol/hr	$ 34,653,118	$ 34,653,118

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 2,460,371	$ 2,460,371
Civil/Structural/Buildings	9.20%		$ 3,188,087	$ 3,188,087
Piping	7.10%		$ 2,460,371	$ 2,460,371
Control and Instrumentation	2.60%		$ 900,981	$ 900,981
Electrical Systems	8.00%		$ 2,772,249	$ 2,772,249
		Total of Balance of Plant	$ 11,782,060	$ 11,782,060
		Total of Balance of Plant - Plus the Selected Source	$ 46,435,179	$ 46,435,179

Rationale for Selection:

Fluor/UOP 2004 was selected as the most recent cost point, because the reported capacity is similar the required capacity per train. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-M**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
DOE FE Report (DOE 1978)	1,000	tpd	1	1978	$ 1,467,000	$ 1,467,000	$ 3,432,834	3,361	tpd	1	3,361	tpd	$ 7,104,506	$ 7,104,506

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
DOE FE Report (DOE 1978)	1,000	tpd	1	1978	$ 1,467,000	$ 1,467,000	$ 3,432,834	3,361	tpd	1	3,361	tpd	$ 7,104,506	$ 7,104,506

Balance of Plant:

Description	Markup						Cost Per Train	Total Cost
Water Systems	7.10%						$ 504,420	$ 504,420
Civil/Structural/Buildings	9.20%						$ 653,615	$ 653,615
Piping	7.10%						$ 504,420	$ 504,420
Control and Instrumentation	2.60%						$ 184,717	$ 184,717
Electrical Systems	8.00%						$ 568,360	$ 568,360
						Total Balance of Plant	$ 2,415,532	$ 2,415,532
						Total Balance of Plant Plus the Selected Source	$ 9,520,038	$ 9,520,038

Rationale for Selection:

Single source island cost identified by the project technical lead. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996

Detail Item Report - CO_2 Compression

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-M**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Subcritical														
Princeton Report (Kreutz 2008)	10	MW	1	2007	$ 6,310,000	$ 6,310,000	$ 6,149,067	11	MW	1	11	MW	$ 6,645,555	$ 6,645,555
Supercritical														
Princeton Report (Kreutz 2008)	13	MW	1	2007	$ 9,520,000	$ 9,520,000	$ 9,277,198	2	MW	1	2	MW	$ 2,666,536	$ 2,666,536

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Kreutz 2008: Combined Subcritical and Supercritical Processes													$ 9,312,091	$ 9,312,091

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 661,158	$ 661,158
Civil/Structural/Buildings	9.20%		$ 856,712	$ 856,712
Piping	7.10%		$ 661,158	$ 661,158
Control and Instrumentation	2.60%		$ 242,114	$ 242,114
Electrical Systems	8.00%		$ 744,967	$ 744,967
		Total Balance of Plant	$ 3,166,111	$ 3,166,111
		Total Balance of Plant Plus the Selected Source	$ 12,478,201	$ 12,478,201

Rationale for Selection:

Single source cost point. Both subcritical and supercritical process costs were included under the CO2 Compression heading. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Project Name: NGNP Process Integration
Process: Conventional Gas to Ammonia
Estimate Number: MA36-M

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	360	tpd	1	2009	$ 44,000,000	$ 44,000,000	$ 44,000,000	3,361	tpd	2	1,681	tpd	$ 110,900,700	$ 221,801,399
Economics Ammonia Coal Gasification (Morel 1977)	1,000	tpd	1	1977	$ 50,748,000	$ 50,748,000	$ 80,243,904	3,361	tpd	2	1,681	tpd	$ 109,565,836	$ 219,131,673
Stamicarbon, Middle East Fertilizer Symposium, March 2009	4,297	tpd	1	2008	$ 800,000,000	$ 800,000,000	$ 779,596,498	3,361	tpd	2	1,681	tpd	$ 443,847,499	$ 887,694,999
Ammonia, Chem Systems, 1998	1,653	tpd	1	1998	$ 160,000,000	$ 160,000,000	$ 210,320,924	3,361	tpd	2	1,681	tpd	$ 212,413,385	$ 424,826,770

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	360	tpd	1	2009	$ 44,000,000	$ 44,000,000	$ 44,000,000	3,361	tpd	2	1,681	tpd	$ 110,900,700	$ 221,801,399

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 7,873,950	$ 15,747,899
Civil/Structural/Buildings	9.20%		$ 10,202,864	$ 20,405,729
Piping	7.10%		$ 7,873,950	$ 15,747,899
Control and Instrumentation	2.60%		$ 2,883,418	$ 5,766,836
Electrical Systems	8.00%		$ 8,872,056	$ 17,744,112
		Total Balance of Plant	$ 37,706,238	$ 75,412,476
		Total Balance of Plant Plus the Selected Sourc	$ 148,606,938	$ 297,213,875

Rationale for Selection:

The verbal cost was selected as recommended by the project team expert. The Stamicarbon information shows a roughly 350% increase in prices between 2003 and 2008 emphasizing the importance of using the most recent information available. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Urea Synthesis

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-M**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	625	tpd	1	2009	$ 85,000,000	$ 85,000,000	$ 85,000,000	2,939	tpd	1	2,939	tpd	$ 215,184,342	$ 215,184,342
Perry's Chemical Engineering Handbook, 7th Edition	200	tpd	1	1994	$ 8,800,000	$ 8,800,000	$ 12,240,152	2,939	tpd	1	2,939	tpd	$ 61,388,719	$ 61,388,719
PNNL	4681	tpd	1	2009	$ 182,000,000	$ 182,000,000	$ 182,000,000	2,939	tpd	1	2,939	tpd	$ 137,653,721	$ 137,653,721
Stamicarbon, Middle East Fertilizer Symposium, March 2009	3,582	tpd	1	2008	$ 550,000,000	$ 550,000,000	$ 535,972,592	2,939	tpd	1	2,939	tpd	$ 475,978,115	$ 475,978,115

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	625	tpd	1	2009	$ 85,000,000	$ 85,000,000	$ 85,000,000	2,939	tpd	1	2,939	tpd	$ 215,184,342	$ 215,184,342

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 15,278,088	$ 15,278,088
Civil/Structural/Buildings	9.20%		$ 19,796,959	$ 19,796,959
Piping	7.10%		$ 15,278,088	$ 15,278,088
Control and Instrumentation	2.60%		$ 5,594,793	$ 5,594,793
Electrical Systems	8.00%		$ 17,214,747	$ 17,214,747
		Total Balance of Plant	$ 73,162,676	$ 73,162,676
		Total Balance of Plant Plus the Selected Source	$ 288,347,019	$ 288,347,019

Rationale for Selection:

The verbal cost was selected as recommended by the project team expert. The Stamicarbon information shows a roughly 350% increase in prices between 2003 and 2008 emphasizing the importance of using the most recent information available. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-M**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Perry's Chemical Engineering Handbook, 7th Edition	334	tpd	1	1994	$ 6,600,000	$ 6,600,000	$ 9,180,114	5,192	tpd	6	865	tpd	$ 16,252,129	$ 97,512,774
Fertilizer Manual	359.38	tpd	1	1998	$ 15,200,000	$ 15,200,000	$ 19,980,488	5,192	tpd	6	865	tpd	$ 33,851,959	$ 203,111,753

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fertilizer Manual	359.38	tpd	1	1998	$ 15,200,000	$ 15,200,000	$ 19,980,488	5,192	tpd	6	865	tpd	$ 33,851,959	$ 203,111,753

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 2,403,489	$ 14,420,934
Civil/Structural/Buildings	9.20%		$ 3,114,380	$ 18,686,281
Piping	7.10%		$ 2,403,489	$ 14,420,934
Control and Instrumentation	2.60%		$ 880,151	$ 5,280,906
Electrical Systems	8.00%		$ 2,708,157	$ 16,248,940
		Total Balance of Plant	$ 11,509,666	$ 69,057,996
		Total Balance of Plant Plus the Selected Source	$ 45,361,625	$ 272,169,749

Rationale for Selection:

The Fertilizer Manual was selected for being both the newest cost point and the most conservative. The data from Perry's Chemical Engineering Handbook is based on earlier data from Peters and Timmerhaus, making it even less desirable. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Ammonium Nitrate Synthesis

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-M**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Perry's Chemical Engineering Handbook, 7th Edition	334.0303	tpd	1	1994	$ 6,000,000	$ 6,000,000	$ 8,345,558	3,779	tpd	3	1,260	tpd	$ 18,507,052	$ 55,521,157
Fertilizer Manual	1395.24	tpd	1	1998	$ 35,000,000	$ 35,000,000	$ 46,007,702	3,779	tpd	3	1,260	tpd	$ 43,270,765	$ 129,812,295
EPA Report	1200	tpd	1	1981	$ 9,480,000	$ 9,480,000	$ 13,434,154	3,779	tpd	3	1,260	tpd	$ 13,831,044	$ 41,493,133

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fertilizer Manual	1395.24	tpd	1	1998	$ 35,000,000	$ 35,000,000	$ 46,007,702	3,779	tpd	3	1,260	tpd	$ 43,270,765	$ 129,812,295

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 3,072,224	$ 9,216,673
Civil/Structural/Buildings	9.20%		$ 3,980,910	$ 11,942,731
Piping	7.10%		$ 3,072,224	$ 9,216,673
Control and Instrumentation	2.60%		$ 1,125,040	$ 3,375,120
Electrical Systems	8.00%		$ 3,461,661	$ 10,384,984
		Total Balance of Plant	$ 14,712,060	$ 44,136,180
		Total Balance of Plant Plus the Selected Source	$ 57,982,825	$ 173,948,476

Rationale for Selection:

The Fertilizer Manual was selected for being both the newest cost point and the most conservative. The data from Perry's Chemical Engineering Handbook is based on earlier data from Peters and Timmerhaus, making it even less desirable. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Project Name: NGNP Process Integration
Process: Conventional Gas to Ammonia
Estimate Number: MA36-M

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Steam Turbine and HRSG														
Shell IGCC Base Cases (NETL 2000)	189	MW	1	1999	$ 50,671,000	$ 50,671,000	$ 66,419,744	29	MW	1	29	MW	$ 21,539,455	$ 21,539,455
Steam Turbine														
NETL Baseline Report (NETL 2007a)	401	MW	4	2006	$ 74,651,000	$ 18,662,750	$ 19,125,957	29	MW	1	29	MW	$ 9,067,994	$ 9,067,994
Princeton Report (Kreutz 2008)	275	MW	1	2007	$ 66,700,000	$ 66,700,000	$ 64,998,858	29	MW	1	29	MW	$ 16,820,729	$ 16,820,729
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	230	MW	1	2006	$ 44,515,000	$ 44,515,000	$ 45,619,856	29	MW	1	29	MW	$ 13,144,232	$ 13,144,232

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	230	MW	1	2006	$ 44,515,000	$ 44,515,000	$ 45,619,856	29	MW	1	29	MW	$ 13,144,232	$ 13,144,232

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 933,241	$ 933,241
Civil/Structural/Buildings	9.20%		$ 1,209,269	$ 1,209,269
Piping	7.10%		$ 933,241	$ 933,241
Control and Instrumentation	2.60%		$ 341,750	$ 341,750
Electrical Systems	8.00%		$ 1,051,539	$ 1,051,539
		Total Balance of Plant	$ 4,469,039	$ 4,469,039
		Total Balance of Plant Plus the Selected Source	$ 17,613,271	$ 17,613,271

Rationale for Selection:

Shell IGCC PowerPlant with CO2 Capture (NETL 2007b) is a recently reported cost point that closely reflects this project's requirements. The Princeton Report (Kreutz 2008) source for the steam turbine cost point is the NETL 2007b report. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-M**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Steam Turbine and HRSG														
Shell IGCC Base Cases (NETL 2000)	189	MW	1	1999	$ 50,671,000	$ 50,671,000	$ 66,419,744	29	MW	1	29	MW	$ 21,525,776	$ 21,525,776
HRSG														
NETL Baseline Report (NETL 2007a)	5,155,983	lb/hr	3	2006	$ 27,581,000	$ 9,193,667	$ 9,421,852	-	lb/hr	1	-	lb/hr	$ -	$ -
Princeton Report (Kreutz 2008)	355	MW	1	2007	$ 52,000,000	$ 52,000,000	$ 50,673,772	-	MW	1	-	MW	$ -	$ -
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	8,438,000	lb/hr	2	2006	$ 45,291,000	$ 22,645,500	$ 23,207,558	-	lb/hr	1	-	lb/hr	$ -	$ -

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	8,438,000	lb/hr	2	2006	$ 45,291,000	$ 22,645,500	$ 23,207,558	-	lb/hr	1	-	lb/hr	$ -	$ -

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ -	$ -
Civil/Structural/Buildings	9.20%		$ -	$ -
Piping	7.10%		$ -	$ -
Control and Instrumentation	2.60%		$ -	$ -
Electrical Systems	8.00%		$ -	$ -
		Total Balance of Plant	$ -	$ -
		Total Balance of Plant Plus the Selected Source	$ -	$ -

Rationale for Selection:

Shell IGCC PowerPlant with CO2 Capture (NETL 2007b) is a recently reported cost point that closely reflects this project's requirements. The Princeton Report (Kreutz 2008) source for the steam turbine cost point is the NETL 2007b report. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Cooling Towers

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Conventional Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-M**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Cooling Tower Depot	123,824	gpm	5	2009	$ 3,141,680	$ 628,336	$ 628,336	93,996	gpm	4	23,499	gpm	$ 608,865	$ 2,435,459

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Cooling Tower Depot	123,824	gpm	5	2009	$ 3,141,680	$ 628,336	$ 628,336	93,996	gpm	4	23,499	gpm	$ 608,865	$ 2,435,459

Balance of Plant:

Description	Markup		Cost Per Train	Total Cost
Water Systems	7.10%		$ 43,229	$ 172,918
Civil/Structural/Buildings	9.20%		$ 56,016	$ 224,062
Piping	7.10%		$ 43,229	$ 172,918
Control and Instrumentation	2.60%		$ 15,830	$ 63,322
Electrical Systems	8.00%		$ 48,709	$ 194,837
		Total Balance of Plant	$ 207,014	$ 828,056
		Total Balance of Plant Plus the Selected Source	$ 815,879	$ 3,263,515

Rationale for Selection:

Single source cost. Publically available current data. Calculated capital costs based on publically available cost data from a vendor regularly engaged in the building of cooling towers. The markups listed under 'Balance of Plant' are based on NETL 2000. These markup values are comparable to additional published estimating guides, such as Page 1996.

NGNP Nuclear Gas to Ammonia Summary

Project Name: NGNP Process Integration
Process: Nuclear Gas to Ammonia
Estimate Number: MA36-N

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Process Component	Subtotal From Detail Sheets	Engineering %	Engineering	Contingency %	Contingency	Total Cost
High Temperature Gas Reactor (HTGR)	$ 768,613,333	0%	$ -	0%	$ -	$ 768,613,333
Rankine Power Cycle	$ 90,964,386	10%	$ 9,096,439	18%	$ 18,010,948	$ 118,071,773
Steam Methane Reforming	$ 105,413,531	10%	$ 10,541,353	18%	$ 20,871,879	$ 136,826,763
Water Gas Shift Reactor	$ 40,122,232	10%	$ 4,012,223	18%	$ 7,944,202	$ 52,078,657
Selexol	$ 46,833,260	10%	$ 4,683,326	18%	$ 9,272,985	$ 60,789,571
Methanation	$ 9,518,338	10%	$ 951,834	18%	$ 1,884,631	$ 12,354,803
CO2 Compression	$ 12,527,029	10%	$ 1,252,703	18%	$ 2,480,352	$ 16,260,083
Ammonia Synthesis	$ 297,160,814	10%	$ 29,716,081	18%	$ 58,837,841	$ 385,714,736
Urea Synthesis	$ 288,347,019	10%	$ 28,834,702	18%	$ 57,092,710	$ 374,274,430
Nitric Acid Synthesis	$ 272,169,749	10%	$ 27,216,975	18%	$ 53,889,610	$ 353,276,334
Ammonium Nitrate Synthesis	$ 173,948,476	10%	$ 17,394,848	18%	$ 34,441,798	$ 225,785,122
Steam Turbines	$ 18,653,382	10%	$ 1,865,338	18%	$ 3,693,370	$ 24,212,090
Heat Recovery Steam Generator (HRSG)	$ -	10%	$ -	18%	$ -	$ -
Cooling Towers	$ 3,610,480	10%	$ 361,048	18%	$ 714,875	$ 4,686,404
Total Cost - Nuclear Gas to Ammonia						$ 2,532,944,099
Total Cost Rounded to the Nearest $10M						$ 2,530,000,000

	Remarks
Checked By: SNW	
Approved By: John R	

4/20/2010

Rev. 03-04-10
Battelle Energy Alliance, LLC

COST ESTIMATE SUPPORT DATA RECAPITULATION

Project Title:	NGNP Process Integration – Nuclear Gas to Ammonia
Estimator:	B. W. Wallace/CEP, R. R. Honsinger/CEP, J. B. Martin/CCT
Date:	April 20, 2010
Estimate Type:	Class 5
File:	MA36-N
Approved By:	

I. **PURPOSE:** *Brief description of the intent of how the estimate is to be used (i.e., for engineering study, comparative analysis, request for funding, proposal, etc.).*

It is expected that the capital costs identified in these estimates will be used in a model producing an economic analysis for each specific integrated application and subsequently will be considered in a related feasibility study.

II. **SCOPE OF WORK:** *Brief statement of the project's objective. Thorough overview and description of the proposed project. Identify work to be accomplished, as well as any specific work to be excluded.*

A. **Objective:**
Develop Class 5 estimates as defined by the Association for Advancement of Cost Engineering (AACEi) that will identify the current capital cost associated with high-temperature gas reactors (HTGRs) integrated with a gas-to-ammonia process.

B. **Included:**
The scope of work required to achieve this objective includes the following:

1. Engineering
2. The allowance provided for the HTGR represents a complete and operable system. All elements required for construction of this nuclear reactor capability, including an initial steam generator, security systems, contingency, and owner's costs are included in the turn-key allowance. Owner's costs are included only in the case of the reactor capability. It is considered that the total value represents all inside of battery limits (ISBL) elements, outside of battery limits (OSBL) elements, site development, and all ancillary control and operational functions and capabilities.
3. Construction of a new integrated refinery capability to produce ammonia from natural gas that consists of the following:
   a. Overnight island-type costs for HTGRs
   b. Steam methane reformer
   c. Water gas shift reactors
   d. Selexol process
   e. Methanation
   f. CO_2 compression
   g. Ammonia synthesis
   h. Urea synthesis
   i. Nitric acid synthesis

j. Ammonium nitrate synthesis
k. Steam turbines, internal to process
l. Heat recovery steam generator, internal to process
m. Cooling towers, internal to process
n. Allowances for Balance of Plant (BOP)/offsite/OSBL, including the following:
   (1.) Site development/improvements
   (2.) Provisions for general and administrative buildings and structures
   (3.) Provisions for OSBL piping
   (4.) Provisions for OSBL instrumentation and control
   (5.) Provisions for OSBL electrical
   (6.) Provisions for facility supply and OSBL water systems
   (7.) Provisions for site development/improvements
   (8.) Project/construction management.

C. **Excluded:**
This scope of work specifically excludes the following elements:
1. Licensing and permitting costs
2. Operational costs
3. Land costs
4. Sales taxes
5. Royalties
6. Owner's fees and owner's costs, except those included for the HTGR
7. The allowance provided for the HTGR capability excludes all costs associated with materials development, or costs that would not be appropriately associated with an nth of a kind (NOAK) reactor/facility.

III. **ESTIMATE METHODOLOGY:** *Overall methodology and rationale of how the estimate was developed (i.e., parametric, forced detail, bottoms up, etc.). Total dollars/hours and rough order magnitude (ROM) allocations of the methodologies used to develop the cost estimate.*

Consistent with the AACEi Class 5 estimates, the level of definition and engineering development available at the time they were prepared, their intended use in a feasibility study, and the time and resources available for their completion, the costs included in this estimate have been developed using parametric evaluations. These evaluations have used publicly available and published project costs to represent similar islands utilized in this project. Analysis and selection of the published costs used have been performed by the project technical lead and Cost Estimating. Suitability for use in this effort was determined considering the correctness and completeness of the data available, the manner in which total capital costs were represented, the age of the previously performed work, and the similarity to the capacity/trains required by this project. The specific sources, selected and used in this cost estimate, are identified in the capital cost estimate detail sheets. Adjustments have been made to these published costs using escalation factors identified in the Chemical Engineering Price Cost Index. Scaling of the published island costs has been

accomplished using the six-tenths capacity factoring method. Costs included for the HTGR and power cycles have been identified and provided by the respective BEA subject matter experts. The total cost for each of these items has been linearly calculated from the respective base unit costs. Any normalization to provide for geographic factors was considered using geographic factors available from RS Means Construction Cost Data references. Cost-estimating relationships have been used to identify allowances to complete the costs.

It was identified to the Next Generation Nuclear Plant (NGNP) Process Integration team that the methodology employed by NGNP to develop the nuclear capability included constituents of parametric modeling, vendor quotes, actual costs, and proprietary costing databases. These preconceptual design estimates were reviewed by NGNP Project Engineering for credibility with regard to assumptions and bases of estimate and performed multiple studies to reconcile variations in the scope and assumptions within the three estimates.

BOP/OSBL costs were determined by the project team, considering data provided by Shell Gasifier IGCC Base Case report NETL 2000, *Conceptual Cost Estimating Manual* Second Edition by John S. Page, and additional adjusted sources. Because the allowances identified did not show significant variability, the allowances identified in the NETL 2000 report were chosen for this effort in order to minimize the mixing of data sources.

IV. **BASIS OF THE ESTIMATE:** *Overall explanation of sources for resource pricing and schedules.*

A. **Quantification Basis:** *The source for the measurable quantities in the estimate that can be used in support of earned value management. Source documents may include drawings, design reports, engineers' notes, and other documentation upon which the estimate is originated.*

All islands and capacities have been provided to Cost Estimating by the respective project expert.

B. **Planning Basis:** *The source for the execution and strategies of the work that can be used to support the project execution plan, acquisition strategy, schedules, and market conditions and other documentation upon which the estimate is originated.*

1. All islands and HTGRs represent NOAK projects.
2. Projects will be constructed and operated by commercial entities.
3. All projects, with the exception of the Steam-Assisted Gravity Drainage Project, will be located in the U.S. Gulf Coast refinery region.
4. Costs are presented as overnight costs.
5. The cost estimate does not consider or address funding or labor resource restrictions. Sufficient funding and labor resources will be available in a manner that allows optimum usage of the funding and resources as estimated and scheduled.

C. **Cost Basis:** *The source for the costing on the estimate that can be used in support of earned value management, funding profiles, and schedule of values. Sources may include published costing references, judgment, actual costs, preliminary quotes or other documentation upon which the estimate is originated.*

1. All costs are represented as current value costs. Factors for forward-looking escalation and inflation factors are not included in this estimate.
2. Where required, published cost factors, as identified in the Chemical Engineering Plant Cost Index, will be applied to previous years' values to determine current year values.
3. Geographic location factors, as identified in RS Means Construction Cost Data reference manual, were considered for each source cost.
4. The cost provided for the HTGR reflects internal BEA cost data that was developed for the HTGR and presented to the NGNP Process Integration team by L. Demmick. Considered in the cost is a pre-conceptual cost estimate prepared by three separate contractor teams. All contractor teams proposed 4-unit NOAK plants with thermal power levels between 2,000 MW_t and 2,400 MW_t at a cost of roughly $4B, including owner's cost. This equates to $1,667 to $2,000 per kW_t. For the purposes of this report, the nominal cost of an HTGR will be set at the upper end of this range, $2,000 per kW_t. This is a complete turnkey cost and includes engineering and construction of a NOAK HTGR, the power cycle, and contingency. The total HTGR cost for each process is calculated linearly as $1,708,333 per MWth of required capacity, excluding the cost of the power cycles.
5. The cost included for the power cycle was provided by the INL project team expert. The power cycle cost is based on the definition of a 240-MWe capacity and $618,176 per MWe. The total power cycle cost for each process is calculated linearly as $618,176 per MWe of required capacity. BOP, engineering, and contingency costs are added to the base cost.
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V. **ESTIMATE QUALITY ASSURANCE:** *A listing of all estimate reviews that have taken place and the actions taken from those reviews.*

A review of the cost estimate was held on January 14, 2010, with the project team and the cost estimators. This review allowed for the project team to review and comment, in detail, on the perceived scope, basis of estimates, assumptions, project risks, and resources that make up this cost estimate. Comments from this review have been incorporated into this estimate to reflect a project team consensus of this document.

VI. **ASSUMPTIONS:** *Condition statements accepted or supposed true without proof of demonstration; statements adding clarification to scope. An assumption has a direct impact on total estimated cost.*

General Assumptions:

A. All costs are represented in 2009 values.

B. Costs that were included from sources representing years prior to 2009 have been normalized to 2009 values using the Chemical Engineering Plant Cost Index. This index was selected due to its widespread recognition and acceptance and its specific orientation toward work associated with chemical and refinery plants.

C. Capital costs are based on process islands. The majority of these islands are interchangeable, after factoring for the differing capacities, flowsheet-to-flowsheet.

D. All chemical processing and refinery processes will be located in the U.S. Gulf Coast region.

E. All costs considered to be BOP costs that can be specifically identified have been factored out of the reported source data and added into the estimate in a manner consistent with that identified in the NETL 2000 IGCC Base Cost report. Inclusion of the source costs in this manner normalizes all reported cost information to the bare-erected costs.

HTGR:

A. The linearly scalable cost included for an HTGR reflects an NOAK reactor with a 750°C-operating temperature.

B. HTGR is considered to be linearly scalable, by required capacity, per the direction of the project team. This allows the process integration feasibility studies to showcase the financial analysis of the process without the added burden of integer quantity 600-MWth HTGRs.

C. The allowance represents a turnkey condition for the reactor and its supporting infrastructure.

D. A high-temperature, high-pressure steam generator is included in the cost represented for HTGR.

E. A contingency allowance is included in the HTGR cost, but is not identified as a separate line item in this estimate. This allowance was identified and included by the NGNP HTGR project team.

F. Total cost range, including contingency, for HTGR is -50%, +100%.

G. Cost included for the power cycle reflects NOAK research and manufacturing developments to allow for assumed high pressures and temperatures.

H. The power cycle is considered to be linearly scalable, by required capacity, per the direction of the project team. This allows the process integration feasibility studies to showcase the financial analysis of the process.

Gas to Ammonia

The NREL 2006 report presented the steam methane reformer cost with the cost of the water gas shift reactors. This cost was normalized to exclude the cost of the water gas shift reactors.

VII. **CONTINGENCY GUIDELINE IMPLEMENTATION:**

Contingency Methodologies: *Explanation of methodology used in determining overall contingency. Identify any specific drivers or items of concern.*

At a project risk review on December 9, 2009, the project team discussed risks to the project. An 18% allowance for capital construction contingency has been included at an island level based on the discussion and is included in the summary sheet. The contingency level that was included in the island cost source documents and additional threats and opportunities identified here were considered during this review. The contingency identified was considered by the project team and included in Cost and Performance Baseline for Fossil Energy Plants DOE/NETL-2007/1281 and similar reports. Typically, contingency allowance provided in these reports ranged from 15% to 20%. Since much of the data contained in this estimate has been derived from these reports, the project team has also chosen a level of contingency consistent with them.

While the level of contingency provided for the HTGR capability is not identified as a line item, the cost data provided to the NGNP Process Integration team was identified as including an appropriate allocation for contingency. No additional contingency has been added to this element.

A. **Threats:** *Uncertain events that are potentially negative or reduce the probability that the desired outcome will happen.*

1. The singularly largest threat to this estimate surrounds the lump sum cost included for the HTGR reactor(s). While the overriding assumption is that the HTGR will be NOAK, currently, a complete HTGR has not been commissioned.
2. The level of project definition/development that was available at the time the estimate was prepared represents a substantial risk to the project and is likely to occur. The high level at which elements were considered and included has the potential to include additional elements that are within the work scope but not sufficiently provided for or addressed at this level.
3. The estimate methodology employed is one of a stochastic parametrically evaluated process. This process used publicly available published costs that were related to the process required, costs were normalized using price indices, and the cost was scaled to provide the required capacity. The cost-estimating relationships that were used represent typical costs for BOP allowances, but source cost data from which the initial island costs were derived were not completely descriptive of the elements included, not included, or simply referred to with different nomenclature or combined with other elements. While every effort has been made to correctly normalize and factor the costs for use in this effort, the risk exists that not all of these were correctly captured due to the varied information available.
4. This project is heavily dependent on metals, concrete, petroleum, and petroleum products. Competition for these commodities in today's environment due to global expansion, uncertainty, and product shortages affect the basic concepts of the supply and demand theories, thus increasing costs.
5. Impacts due to large quantities of materials, special alloy materials, fabrication capability, and labor availability could all represent conditions that may increase the total cost of the project.

B. **Opportunities:** *Uncertain events that could improve the results or improve the probability that the desired outcome will happen.*

1. Additional research and work performed with both vendors and potential owner/operators for a specific process or refinery may identify efficiencies and production means that have not been available for use in this analysis.
2. Recent historical data may identify and include technological advancements and efficiencies not included or reflected in the publicly available source data used in this effort.

Note: Contingency does not increase the overall accuracy of the estimate; it does, however, reduce the level of risk associated with the estimate. Contingency is intended to cover the inadequacies in the complete project scope definition, estimating methods, and estimating data. Contingency specifically excludes changes in project scope, unexpected work stoppages (e.g., strikes, disasters, and earthquakes) and excessive or unexpected inflation or currency fluctuations.

VIII. OTHER COMMENTS/CONCERNS SPECIFIC TO THE ESTIMATE:

None.

Detail Item Report - High Temperature Gas Reactor (HTGR)

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Nuclear Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-N**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Internal Cost Data (INL 2009)	1	MWth		2009	$ 1,708,333	$ 1,708,333	$ 1,708,333	450	MWth				$ 768,613,333	$ 768,613,333

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Internal Cost Data (INL 2009)	1	MWth		2009	$ 1,708,333	$ 1,708,333	$ 1,708,333	450	MWth				$ 768,613,333	$ 768,613,333

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	0.00%		$ -	$ -
Civil/Structural/Buildings	0.00%		$ -	$ -
Piping	0.00%		$ -	$ -
Control and Instrumentation	0.00%		$ -	$ -
Electrical Systems	0.00%		$ -	$ -
		Total Balance of Plant	$ -	$ -
		Total Balance of Plant Plus the Selected Source	$ 768,613,333	$ 768,613,333

Basis of Estimate Notes:

Single source cost point. This cost has been provided by the subcontracted subject matter expert L. Demick to the INL NGNP Process Integration team. This cost represents a complete turnkey cost. The cost of an HTGR reactor, as provided by L. Demick, is $2,000,000 per MWth required. This cost used has been reduced to $1,708,333 per MWth to exclude the cost of power cycles.

Detail Item Report - Rankine Cycle - Case 11, Supercritical PC Case

Project Name: NGNP Process Integration
Process: Nuclear Gas to Ammonia
Estimate Number: MA36-N

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Internal Cost Data (INL 2009)	240	MWe	1	2009	$ 148,362,255	$ 148,362,255	$ 148,362,255	106	MWe	1	106	MWe	$ 90,964,386	$ 90,964,386

Summary:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Internal Cost Data (INL 2009)	240	MWe	1	2009	$ 148,362,255	$ 148,362,255	$ 148,362,255	106	MWe	1	106	MWe	$ 90,964,386	$ 90,964,386

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	0.00%		$ -	$ -
Civil/Structural/Buildings	0.00%		$ -	$ -
Piping	0.00%		$ -	$ -
Control and Instrumentation	0.00%		$ -	$ -
Electrical Systems	0.00%		$ -	$ -
		Total Balance of Plant	$ -	$ -
		Total Balance of Plant Plus the Selected Source	$ 90,964,386	$ 90,964,386

Basis of Estimate Notes:

Single source cost. The reported costs are from the INL project team expert. The reported cost represents a Rankine power cycle, excluding the steam generator. The cost is based on information found in NETL 2007b, which has been adjusted and customized for this project by the INL project team expert. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996. The allowances have been adjusted and customized for this project based on estimator judgment. The reduced civil/structural/buildings allowance accounts for the buildings that are included in the Rankine power cycle cost. Water and electrical systems BOP allowances are included in the reported cost for the Rankine power cycle.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Nuclear Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-N**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Natural Gas to Liquids Conversion Project (Raytheon 2000)	150	MMS CFD	1	1999	$ 79,000,000	$ 79,000,000	$ 103,553,507	72	MMS CFD	1	72	MMS CFD	$ 66,666,792	$ 78,666,814
NETL Natural Gas Report (Choi 1996)	100	MMS CFD	1	1996	$ 22,800,000	$ 22,800,000	$ 30,583,181	72	MMS CFD	1	72	MMS CFD	$ 25,112,049	$ 25,112,049
Gas Usage & Value The Technology and Economics of Natural Gas Use in the Process Industries (PennWell Corporation 2006)	4,700	tpd	2	1985	$ 168,150,000	$ 84,075,000	$ 132,328,312	2,538	tpd	2	1,269	tpd	$ 91,430,065	$ 182,860,130
NREL Report (NREL 2006)	183	MMS CFD	1	2002	$ 175,391,586	$ 175,391,586	$ 226,998,210	72	MMS CFD	2	36	MMS CFD	$ 85,572,315	$ 171,144,630

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Natural Gas to Liquids Conversion Project (Raytheon 2000)	150	MMS CFD	1	1999	$ 79,000,000	$ 79,000,000	$ 103,553,507	72	MMS CFD	1	72	MMS CFD	$ 66,666,792	$ 78,666,814

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 4,733,342	$ 5,585,344
Civil/Structural/Buildings	9.20%		$ 6,133,345	$ 7,237,347
Piping	7.10%		$ 4,733,342	$ 5,585,344
Control and Instrumentation	2.60%		$ 1,733,337	$ 2,045,337
Electrical Systems	8.00%		$ 5,333,343	$ 6,293,345
		Total Balance of Plant	$ 22,666,709	$ 26,746,717
		Total Balance of Plant Plus the Selected Source	$ 89,333,501	$ 105,413,531

Rationale for Selection:

Raytheon 2000 was selected as the most recent cost point, with technology most similar to the intended process. The adjusted PennWell Corporation 2006 cost point concurs with the Raytheon 2000 value. Costs presented in th Raytheon report have been increased by 18% to account for the difference between an autothermal process and traditional steam methane reforming process. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Note: PennWell Corporation 2006 includes both the natural gas reformer and methanol synthesis units. The average cost for methanol synthesis, as calculated in this report, was subtracted from the total current cost for required trains cell so that this may be considered as a cost point for natural gas reforming. The NREL 2006 cost was scaled from the originally presented value to exclude the cost of the water gas shift reactors.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Nuclear Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-N**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Princeton Report (Kreutz 2008)	66,742	MMBTU/day	1	2007	$ 11,760,000	$ 11,760,000	$ 11,460,069	-	MMBTU/day	1	-	MMBTU/day	$ -	$ -
Hydrogen Report (Gray 2004)	48,243	lbmol/hr	1	2004	$ 23,000,000	$ 23,000,000	$ 26,510,581	59,093	lbmol/hr	1	59,093	lbmol/hr	$ 29,941,964	$ 29,941,964
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	1,714,460	lb/hr	4	2006	$ 12,367,000	$ 3,091,750	$ 3,168,487	941,756	lb/hr	1	941,756	lbmol/hr	$ 5,081,296	$ 5,081,296

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Hydrogen Report (Gray 2004)	48,243	lbmol/hr	1	2004	$ 23,000,000	$ 23,000,000	$ 26,510,581	59,093	lbmol/hr	1	59,093	lbmol/hr	$ 29,941,964	$ 29,941,964

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 2,125,879	$ 2,125,879
Civil/Structural/Buildings	9.20%		$ 2,754,661	$ 2,754,661
Piping	7.10%		$ 2,125,879	$ 2,125,879
Control and Instrumentation	2.60%		$ 778,491	$ 778,491
Electrical Systems	8.00%		$ 2,395,357	$ 2,395,357
		Total Balance of Plant	$ 10,180,268	$ 10,180,268
		Total Balance of Plant Plus the Selected Source	$ 40,122,232	$ 40,122,232

Rationale for Selection:

Hydrogen Report (Gray 2004) has been selected for use based on the advise of the technical lead. It is considered to be the most similar in design and function to the water gas shift required for this project. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Nuclear Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-N**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fluor/UOP Report (Fluor/UOP 2004)	31,226	lbmol/hr	1	2003	$ 22,862,000	$ 22,862,000	$ 29,117,771	42,332	lbmol/hr	1	42,332	lbmol/hr	$ 34,950,194	$ 34,950,194
Co-Production of Hydrogen Presentation (Chiesa 2003)	327	tph	1	2002	$ 32,800,000	$ 32,800,000	$ 42,450,961	150	tph	1	150	tph	$ 26,615,433	$ 26,615,433

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fluor/UOP Report (Fluor/UOP 2004)	31,226	lbmol/hr	1	2003	$ 22,862,000	$ 22,862,000	$ 29,117,771	42,332	lbmol/hr	1	42,332	lbmol/hr	$ 34,950,194	$ 34,950,194

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 2,481,464	$ 2,481,464
Civil/Structural/Buildings	9.20%		$ 3,215,418	$ 3,215,418
Piping	7.10%		$ 2,481,464	$ 2,481,464
Control and Instrumentation	2.60%		$ 908,705	$ 908,705
Electrical Systems	8.00%		$ 2,796,015	$ 2,796,015
		Total Balance of Plant	$ 11,883,066	$ 11,883,066
		Total Balance of Plant Plus the Selected Source	$ 46,833,260	$ 46,833,260

Rationale for Selection:

Fluor/UOP 2004 was selected as the most recent cost point, because the reported capacity is similar the required capacity per train. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Methanation

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Nuclear Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-N**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
DOE FE Report (DOE 1978)	1,000	tpd	1	1978	$ 1,467,000	$ 1,467,000	$ 3,432,834	3,360	tpd	1	3,360	tpd	$ 7,103,237	$ 7,103,237

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
DOE FE Report (DOE 1978)	1,000	tpd	1	1978	$ 1,467,000	$ 1,467,000	$ 3,432,834	3,360	tpd	1	3,360	tpd	$ 7,103,237	$ 7,103,237

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 504,330	$ 504,330
Civil/Structural/Buildings	9.20%		$ 653,498	$ 653,498
Piping	7.10%		$ 504,330	$ 504,330
Control and Instrumentation	2.60%		$ 184,684	$ 184,684
Electrical Systems	8.00%		$ 568,259	$ 568,259
		Total Balance of Plant	$ 2,415,101	$ 2,415,101
		Total Balance of Plant Plus the Selected Source	$ 9,518,338	$ 9,518,338

Rationale for Selection:

Single source island cost identified by the project technical lead. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such
Page 1996.

Detail Item Report - CO2 Compression

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Nuclear Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-N**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Subcritical														
Princeton Report (Kreutz 2008)	10	MW	1	2007	$ 6,310,000	$ 6,310,000	$ 6,149,067	11	MW	1	11	MW	$ 6,671,585	$ 6,671,585
Supercritical														
Princeton Report (Kreutz 2008)	13	MW	1	2007	$ 9,520,000	$ 9,520,000	$ 9,277,198	2	MW	1	2	MW	$ 2,676,943	$ 2,676,943

Source Selected:

Source	Reported Capacity	Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required	Trains Reqd.	Capacity per Train	Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Kreutz 2008: Combined Subcritical and Supercritical Processes										$ 9,348,529	$ 9,348,529

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 663,746	$ 663,746
Civil/Structural/Buildings	9.20%		$ 860,065	$ 860,065
Piping	7.10%		$ 663,746	$ 663,746
Control and Instrumentation	2.60%		$ 243,062	$ 243,062
Electrical Systems	8.00%		$ 747,882	$ 747,882
		Total Balance of Plant	$ 3,178,500	$ 3,178,500
		Total Balance of Plant Plus the Selected Source	$ 12,527,029	$ 12,527,029

Rationale for Selection:

Single source cost point. Both subcritical and supercritical process costs were included under the CO2 Compression heading. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Ammonia Synthesis

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Nuclear Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-N**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	360	tpd	1	2009	$ 44,000,000	$ 44,000,000	$ 44,000,000	3,360	tpd	2	1,680	tpd	$ 110,880,901	$ 221,761,801
Economics Ammonia Coal Gasification (Morel 1977)	1,000	tpd	1	1977	$ 50,748,000	$ 50,748,000	$ 80,243,904	3,360	tpd	2	1,680	tpd	$ 109,546,276	$ 219,092,551
Stamicarbon, Middle East Fertilizer Symposium, March 2009	4,297	tpd	1	2008	$ 800,000,000	$ 800,000,000	$ 779,596,498	3,360	tpd	2	1,680	tpd	$ 443,768,260	$ 887,536,519
Ammonia, Chem Systems, 1998	1,653	tpd	1	1998	$ 160,000,000	$ 160,000,000	$ 210,320,924	3,360	tpd	2	1,680	tpd	$ 212,375,463	$ 424,750,926

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	360	tpd	1	2009	$ 44,000,000	$ 44,000,000	$ 44,000,000	3,360	tpd	2	1,680	tpd	$ 110,880,901	$ 221,761,801

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 7,872,544	$ 15,745,088
Civil/Structural/Buildings	9.20%		$ 10,201,043	$ 20,402,086
Piping	7.10%		$ 7,872,544	$ 15,745,088
Control and Instrumentation	2.60%		$ 2,882,903	$ 5,765,807
Electrical Systems	8.00%		$ 8,870,472	$ 17,740,944
		Total Balance of Plant	$ 37,699,506	$ 75,399,012
		Total Balance of Plant Plus the Selected Source	$ 148,580,407	$ 297,160,814

Rationale for Selection:

The verbal cost was selected as recommended by the project team expert. The Stamicarbon information shows a roughly 350% increase in prices between 2003 and 2008 emphasizing the importance of using the most recent information available. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Nuclear Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-N**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	625	tpd	1	2009	$ 85,000,000	$ 85,000,000	$ 85,000,000	2,939	tpd	1	2,939	tpd	$ 215,184,342	$ 215,184,342
Perry's Chemical Engineering Handbook, 7th Edition	200	tpd	1	1994	$ 8,800,000	$ 8,800,000	$ 12,240,152	2,939	tpd	1	2,939	tpd	$ 61,388,719	$ 61,388,719
PNNL	4681	tpd	1	2009	$ 182,000,000	$ 182,000,000	$ 182,000,000	2,939	tpd	1	2,939	tpd	$ 137,653,721	$ 137,653,721
Stamicarbon, Middle East Fertilizer Symposium, March 2009	3,582	tpd	1	2008	$ 550,000,000	$ 550,000,000	$ 535,972,592	2,939	tpd	1	2,939	tpd	$ 475,978,115	$ 475,978,115

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	625	tpd	1	2009	$ 85,000,000	$ 85,000,000	$ 85,000,000	2,939	tpd	1	2,939	tpd	$ 215,184,342	$ 215,184,342

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 15,278,088	$ 15,278,088
Civil/Structural/Buildings	9.20%		$ 19,796,959	$ 19,796,959
Piping	7.10%		$ 15,278,088	$ 15,278,088
Control and Instrumentation	2.60%		$ 5,594,793	$ 5,594,793
Electrical Systems	8.00%		$ 17,214,747	$ 17,214,747
		Total Balance of Plant	$ 73,162,676	$ 73,162,676
		Total Balance of Plant Plus the Selected Source	$ 288,347,019	$ 288,347,019

Rationale for Selection:

The verbal cost was selected as recommended by the project team expert. The Stamicarbon information shows a roughly 350% increase in prices between 2003 and 2008 emphasizing the importance of using the most recent information available. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Nitric Acid Synthesis

Project Name: NGNP Process Integration
Process: Nuclear Gas to Ammonia
Estimate Number: MA36-N

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Perry's Chemical Engineering Handbook, 7th Edition	334	tpd	1	1994	$ 6,600,000	$ 6,600,000	$ 9,180,114	5,192	tpd	6	865	tpd	$ 16,251,244	$ 97,507,466
Fertilizer Manual	359	tpd	1	1998	$ 15,200,000	$ 15,200,000	$ 19,980,488	5,192	tpd	6	865	tpd	$ 33,851,959	$ 203,111,753

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fertilizer Manual	359	tpd	1	1998	$ 15,200,000	$ 15,200,000	$ 19,980,488	5,192	tpd	6	865	tpd	$ 33,851,959	$ 203,111,753

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 2,403,489	$ 14,420,934
Civil/Structural/Buildings	9.20%		$ 3,114,380	$ 18,686,281
Piping	7.10%		$ 2,403,489	$ 14,420,934
Control and Instrumentation	2.60%		$ 880,151	$ 5,280,906
Electrical Systems	8.00%		$ 2,708,157	$ 16,248,940
		Total Balance of Plant	$ 11,509,666	$ 69,057,996
		Total Balance of Plant Plus the Selected Source	$ 45,361,625	$ 272,169,749

Rationale for Selection:

The Fertilizer Manual was selected for being both the newest cost point and the most conservative. The data from Perry's Chemical Engineering Handbook is based on earlier data from Peters and Timmerhaus, making it even less desirable. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Ammonium Nitrate Synthesis

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Nuclear Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-N**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Perry's Chemical Engineering Handbook, 7th Edition	334	tpd	1	1994	$ 6,000,000	$ 6,000,000	$ 8,345,558	3,779	tpd	3	1,260	tpd	$ 18,507,052	$ 55,521,157
Fertilizer Manual	1,395	tpd	1	1998	$ 35,000,000	$ 35,000,000	$ 46,007,702	3,779	tpd	3	1,260	tpd	$ 43,270,765	$ 129,812,295
EPA Report	1,200	tpd	1	1981	$ 9,480,000	$ 9,480,000	$ 13,434,154	3,779	tpd	3	1,260	tpd	$ 13,831,044	$ 41,493,133

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fertilizer Manual	1,395	tpd	1	1998	$ 35,000,000	$ 35,000,000	$ 46,007,702	3,779	tpd	3	1,260	tpd	$ 43,270,765	$ 129,812,295

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 3,072,224	$ 9,216,673
Civil/Structural/Buildings	9.20%		$ 3,980,910	$ 11,942,731
Piping	7.10%		$ 3,072,224	$ 9,216,673
Control and Instrumentation	2.60%		$ 1,125,040	$ 3,375,120
Electrical Systems	8.00%		$ 3,461,661	$ 10,384,984
		Total Balance of Plant	$ 14,712,060	$ 44,136,180
		Total Balance of Plant Plus the Selected Source	$ 57,982,825	$ 173,948,476

Rationale for Selection:

The Fertilizer Manual was selected for being both the newest cost point and the most conservative. The data from Perry's Chemical Engineering Handbook is based on earlier data from Peters and Timmerhaus, making it even less desirable. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Nuclear Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-N**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Steam Turbine and HRSG														
Shell IGCC Base Cases (NETL 2000)	189	MW	1	1999	$ 50,671,000	$ 50,671,000	$ 66,419,744	32	MW	1	32	MW	$ 22,796,931	$ 22,796,931
Steam Turbine														
NETL Baseline Report (NETL 2007a)	401	MW	4	2006	$ 74,651,000	$ 18,662,750	$ 19,125,957	32	MW	1	32	MW	$ 9,603,484	$ 9,603,484
Princeton Report (Kreutz 2008)	275	MW	1	2007	$ 66,700,000	$ 66,700,000	$ 64,998,858	32	MW	1	32	MW	$ 17,814,038	$ 17,814,038
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	230	MW	1	2006	$ 44,515,000	$ 44,515,000	$ 45,619,856	32	MW	1	32	MW	$ 13,920,435	$ 13,920,435

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	230	MW	1	2006	$ 44,515,000	$ 44,515,000	$ 45,619,856	32	MW	1	32	MW	$ 13,920,435	$ 13,920,435

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 988,351	$ 988,351
Civil/Structural/Buildings	9.20%		$ 1,280,680	$ 1,280,680
Piping	7.10%		$ 988,351	$ 988,351
Control and Instrumentation	2.60%		$ 361,931	$ 361,931
Electrical Systems	8.00%		$ 1,113,635	$ 1,113,635
		Total Balance of Plant	$ 4,732,948	$ 4,732,948
		Total Balance of Plant Plus the Selected Source	$ 18,653,382	$ 18,653,382

Rationale for Selection:

Shell IGCC PowerPlant with CO2 Capture (NETL 2007b) is a recently reported cost point that closely reflects this project's requirements. The Princeton Report (Kreutz 2008) source for the steam turbine cost point is the NETL 2007b report. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - HRSG

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Nuclear Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-N**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Steam Turbine and HRSG														
Shell IGCC Base Cases (NETL 2000)	189	MW	1	1999	$ 50,671,000	$ 50,671,000	$ 66,419,744	-	MW	1	-	MW	$ -	$ -
HRSG														
NETL Baseline Report (NETL 2007a)	5,155,983	lb/hr	3	2006	$ 27,581,000	$ 9,193,667	$ 9,421,852	-	lb/hr	1	-	lb/hr	$ -	$ -
Princeton Report (Kreutz 2008)	355	MW	1	2007	$ 52,000,000	$ 52,000,000	$ 50,673,772	-	MW	1	-	MW	$ -	$ -
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	8,438,000	lb/hr	2	2006	$ 45,291,000	$ 22,645,500	$ 23,207,558	-	lb/hr	1	-	lb/hr	$ -	$ -

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	8,438,000	lb/hr	2	2006	$ 45,291,000	$ 22,645,500	$ 23,207,558	-	lb/hr	1	-	lb/hr	$ -	$ -

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ -	$ -
Civil/Structural/Buildings	9.20%		$ -	$ -
Piping	7.10%		$ -	$ -
Control and Instrumentation	2.60%		$ -	$ -
Electrical Systems	8.00%		$ -	$ -
		Total Balance of Plant	$ -	$ -
		Total Balance of Plant Plus the Selected Source	$ -	$ -

Rationale for Selection:

Shell IGCC PowerPlant with CO2 Capture (NETL 2007b) is a recently reported cost point that closely reflects this project's requirements. The Princeton Report (Kreutz 2008) source for the steam turbine cost point is the NETL 2007b report. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**Nuclear Gas to Ammonia**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-N**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Cooling Tower Depot	123,006	gpm	3	2009	$ 3,140,030	$ 1,046,677	$ 1,046,677	95,310	gpm	3	31,770	gpm	$ 898,129	$ 2,694,388

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Cooling Tower Depot	123,006	gpm	3	2009	$ 3,140,030	$ 1,046,677	$ 1,046,677	95,310	gpm	3	31,770	gpm	$ 898,129	$ 2,694,388

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 63,767	$ 191,302
Civil/Structural/Buildings	9.20%		$ 82,628	$ 247,884
Piping	7.10%		$ 63,767	$ 191,302
Control and Instrumentation	2.60%		$ 23,351	$ 70,054
Electrical Systems	8.00%		$ 71,850	$ 215,551
		Total Balance of Plant	$ 305,364	$ 916,092
		Total Balance of Plant Plus the Selected Source	$ 1,203,493	$ 3,610,480

Rationale for Selection:

Single source cost. Publically available current data. Calculated capital costs based on publically available cost data from a vendor regularly engaged in the building of cooling towers. The allowances listed under 'Balance of Pla are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

NGNP HTSE Ammonia w/o ASU Summary

Project Name: NGNP Process Integration
Process: HTSE Ammonia w/o ASU
Estimate Number: MA36-O

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Process Component	Subtotal From Detail Sheets	Engineering %	Engineering	Contingency %	Contingency	Total Cost
High Temperature Gas Reactor (HTGR)	$ 4,201,101,415	0%	$ -	0%	$ -	$ 4,201,101,415
Rankine Power Cycle	$ 615,345,051	10%	$ 61,534,505	18%	$ 121,838,320	$ 798,717,876
High Temperature Steam Electrolysis (HTSE)	$ 363,429,475	10%	$ 36,342,947	18%	$ 71,959,036	$ 471,731,458
N2 Generation	$ 17,287,060	10%	$ 1,728,706	18%	$ 3,422,838	$ 22,438,603
CO2 Generation	$ 15,022,364	10%	$ 1,502,236	18%	$ 2,974,428	$ 19,499,029
Methanation	$ 9,518,338	10%	$ 951,834	18%	$ 1,884,631	$ 12,354,803
Ammonia Synthesis	$ 297,160,814	10%	$ 29,716,081	18%	$ 58,837,841	$ 385,714,736
Urea Synthesis	$ 288,347,019	10%	$ 28,834,702	18%	$ 57,092,710	$ 374,274,430
Nitric Acid Synthesis	$ 272,169,749	10%	$ 27,216,975	18%	$ 53,889,610	$ 353,276,334
Ammonium Nitrate Synthesis	$ 173,948,476	10%	$ 17,394,848	18%	$ 34,441,798	$ 225,785,122
Steam Turbines	$ 49,012,114	10%	$ 4,901,211	18%	$ 9,704,398	$ 63,617,723
Heat Recovery Steam Generator (HRSG)	$ -	10%	$ -	18%	$ -	$ -
Cooling Towers	$ 5,735,762	10%	$ 573,576	18%	$ 1,135,681	$ 7,445,019
Total Cost - HTSE Ammonia w/o ASU						$ 6,935,956,549
Total Cost Rounded to the Nearest $10M						$ 6,940,000,000

	Remarks
Checked By: SNW	
Approved By: John R	

COST ESTIMATE SUPPORT DATA RECAPITULATION

Project Title: NGNP Process Integration – HTSE Ammonia without ASU
Estimator: B. W. Wallace/CEP, R. R. Honsinger/CEP, J. B. Martin/CCT
Date: April 20, 2010
Estimate Type: Class 5
File: MA36-O
Approved By:

I. **PURPOSE:** *Brief description of the intent of how the estimate is to be used (i.e., for engineering study, comparative analysis, request for funding, proposal, etc.).*

It is expected that the capital costs identified in these estimates will be used in a model producing an economic analysis for each specific integrated application and subsequently will be considered in a related feasibility study.

II. **SCOPE OF WORK:** *Brief statement of the project's objective. Thorough overview and description of the proposed project. Identify work to be accomplished, as well as any specific work to be excluded.*

A. **Objective:**
Develop Class 5 estimates as defined by the Association for Advancement of Cost Engineering (AACEi) that will identify the current capital cost associated with high-temperature gas reactors (HTGRs) integrated with a nuclear ammonia without an air separation unit process.

B. **Included:**
The scope of work required to achieve this objective includes the following:

1. Engineering
2. The allowance provided for the HTGR represents a complete and operable system. All elements required for construction of this nuclear reactor capability, including an initial steam generator, security systems, contingency, and owner's costs are included in the turn-key allowance. Owner's costs are included only in the case of the reactor capability. It is considered that the total value represents all inside of battery limits (ISBL) elements, outside of battery limits (OSBL) elements, site development, and all ancillary control and operational functions and capabilities.
3. Construction of a new integrated refinery capability to produce ammonia that consists of the following:
   a. Overnight island-type costs for HTGRs
   b. High-temperature steam electrolysis (HTSE) hydrogen production unit
   c. H_2 combustor (N_2 generation)
   d. Natural gas combustor (CO_2 generation)
   e. Methanation
   f. Ammonia synthesis
   g. Urea synthesis
   h. Nitric acid synthesis

i. Ammonium nitrate synthesis
j. Steam turbines, internal to process
k. Heat recovery steam generator, internal to process
l. Cooling towers, internal to process
m. Allowances for Balance of Plant (BOP)/offsite/OSBL, including the following:
   (1.) Site development/improvements
   (2.) Provisions for general and administrative buildings and structures
   (3.) Provisions for OSBL piping
   (4.) Provisions for OSBL instrumentation and control
   (5.) Provisions for OSBL electrical
   (6.) Provisions for facility supply and OSBL water systems
   (7.) Provisions for site development/improvements
   (8.) Project/construction management.

C. **Excluded:**

This scope of work specifically excludes the following elements:

1. Licensing and permitting costs
2. Operational costs
3. Land costs
4. Sales taxes
5. Royalties
6. Owner's fees and owner's costs, except those included for the HTGR
7. The allowance provided for the HTGR capability excludes all costs associated with materials development, or costs that would not be appropriately associated with an nth of a kind (NOAK) reactor/facility.

III. **ESTIMATE METHODOLOGY:** *Overall methodology and rationale of how the estimate was developed (i.e., parametric, forced detail, bottoms up, etc.). Total dollars/hours and rough order magnitude (ROM) allocations of the methodologies used to develop the cost estimate.*

Consistent with the AACEi Class 5 estimates, the level of definition and engineering development available at the time they were prepared, their intended use in a feasibility study, and the time and resources available for their completion, the costs included in this estimate have been developed using parametric evaluations. These evaluations have used publicly available and published project costs to represent similar islands utilized in this project. Analysis and selection of the published costs used have been performed by the project technical lead and Cost Estimating. Suitability for use in this effort was determined considering the correctness and completeness of the data available, the manner in which total capital costs were represented, the age of the previously performed work, and the similarity to the capacity/trains required by this project. The specific sources, selected and used in this cost estimate, are identified in the capital cost estimate detail sheets. Adjustments have been made to these published costs using escalation factors identified in the Chemical Engineering Price Cost Index. Scaling of the published island costs has been

accomplished using the six-tenths capacity factoring method. Costs included for the HTGR, power cycles, and HTSE, have been identified and provided by the respective BEA subject matter experts. The total cost for each of these items has been linearly calculated from the respective base unit costs. Any normalization to provide for geographic factors was considered using geographic factors available from RS Means Construction Cost Data references. Cost-estimating relationships have been used to identify allowances to complete the costs.

It was identified to the Next Generation Nuclear Plant (NGNP) Process Integration team that the methodology employed by NGNP to develop the nuclear capability included constituents of parametric modeling, vendor quotes, actual costs, and proprietary costing databases. These preconceptual design estimates were reviewed by NGNP Project Engineering for credibility with regard to assumptions and bases of estimate and performed multiple studies to reconcile variations in the scope and assumptions within the three estimates.

BOP/OSBL costs were determined by the project team, considering data provided by Shell Gasifier IGCC Base Case report NETL 2000, *Conceptual Cost Estimating Manual* Second Edition by John S. Page, and additional adjusted sources. Because the allowances identified did not show significant variability, the allowances identified in the NETL 2000 report were chosen for this effort in order to minimize the mixing of data sources.

IV. **BASIS OF THE ESTIMATE**: *Overall explanation of sources for resource pricing and schedules.*

A. **Quantification Basis**: *The source for the measurable quantities in the estimate that can be used in support of earned value management. Source documents may include drawings, design reports, engineers' notes, and other documentation upon which the estimate is originated.*

All islands and capacities have been provided to Cost Estimating by the project respective expert.

B. **Planning Basis**: *The source for the execution and strategies of the work that can be used to support the project execution plan, acquisition strategy, schedules, and market conditions and other documentation upon which the estimate is originated.*

1. All islands and HTGRs represent NOAK projects.
2. Projects will be constructed and operated by commercial entities.
3. All projects, with the exception of the Steam-Assisted Gravity Drainage Project, will be located in the U.S. Gulf Coast refinery region.
4. Costs are presented as overnight costs.
5. The cost estimate does not consider or address funding or labor resource restrictions. Sufficient funding and labor resources will be available in a manner that allows optimum usage of the funding and resources as estimated and scheduled.

C. **Cost Basis:** *The source for the costing on the estimate that can be used in support of earned value management, funding profiles, and schedule of values. Sources may include published costing references, judgment, actual costs, preliminary quotes or other documentation upon which the estimate is originated.*

1. All costs are represented as current value costs. Factors for forward-looking escalation and inflation factors are not included in this estimate.
2. Where required, published cost factors, as identified in the Chemical Engineering Plant Cost Index, will be applied to previous years' values to determine current year values.
3. Geographic location factors, as identified in RS Means Construction Cost Data reference manual, were considered for each source cost.
4. The cost provided for the HTGR reflects internal BEA cost data that was developed for the HTGR and presented to the NGNP Process Integration team by L. Demmick. Considered in the cost is a pre-conceptual cost estimate prepared by three separate contractor teams. All contractor teams proposed 4-unit NOAK plants with thermal power levels between 2,000 MW_t and 2,400 MW_t at a cost of roughly $4B, including owner's cost. This equates to $1,667 to $2,000 per kW_t. For the purposes of this report, the nominal cost of an HTGR will be set at the upper end of this range, $2,000 per kW_t. This is a complete turnkey cost and includes engineering and construction of a NOAK HTGR, the power cycle, and contingency. The total HTGR cost for each process is calculated linearly as $1,708,333 per MWth of required capacity, excluding the cost of the power cycles.
5. The cost included for the power cycle was provided by the INL project team expert. The power cycle cost is based on the definition of a 240-MWe capacity and $618,176 per MWe. The total power cycle cost for each process is calculated linearly as $618,176 per MWe of required capacity. BOP, engineering, and contingency costs are added to the base cost.
6. The cost included for HTSE was provided by the INL project team expert. The total HTSE cost for each process is calculated linearly as $36,120,156 per kg/s of required capacity. BOP, engineering, and contingency costs are added to the base cost.
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V. **ESTIMATE QUALITY ASSURANCE:** *A listing of all estimate reviews that have taken place and the actions taken from those reviews.*

A review of the cost estimate was held on January 14, 2010, with the project team and the cost estimators. This review allowed for the project team to review and comment, in detail, on the perceived scope, basis of estimates, assumptions, project risks, and resources that make up this cost estimate. Comments from this review have been incorporated into this estimate to reflect a project team consensus of this document.

VI. **ASSUMPTIONS:** *Condition statements accepted or supposed true without proof of demonstration; statements adding clarification to scope. An assumption has a direct impact on total estimated cost.*

General Assumptions:

A. All costs are represented in 2009 values.

B. Costs that were included from sources representing years prior to 2009 have been normalized to 2009 values using the Chemical Engineering Plant Cost Index. This index was selected due to its widespread recognition and acceptance and its specific orientation toward work associated with chemical and refinery plants.

C. Capital costs are based on process islands. The majority of these islands are interchangeable, after factoring for the differing capacities, flowsheet-to-flowsheet.

D. All chemical processing and refinery processes will be located in the U.S. Gulf Coast region.

E. All costs considered to be BOP costs that can be specifically identified have been factored out of the reported source data and added into the estimate in a manner consistent with that identified in the NETL 2000 IGCC Base Cost report. Inclusion of the source costs in this manner normalizes all reported cost information to the bare-erected costs.

HTGR:

A. The linearly scalable cost included for an HTGR reflects an NOAK reactor with a 750°C-operating temperature.
B. HTGR is considered to be linearly scalable, by required capacity, per the direction of the project team. This allows the process integration feasibility studies to showcase the financial analysis of the process without the added burden of integer quantity 600-MWth HTGRs.
C. The allowance represents a turnkey condition for the reactor and its supporting infrastructure.
D. A high-temperature, high-pressure steam generator is included in the cost represented for HTGR.
E. A contingency allowance is included in the HTGR cost, but is not identified as a separate line item in this estimate. This allowance was identified and included by the NGNP HTGR project team.
F. Total cost range, including contingency, for HTGR is -50%, +100%.
G. Cost included for the power cycle reflects NOAK research and manufacturing developments to allow for assumed high pressures and temperatures.
H. The power cycle is considered to be linearly scalable, by required capacity, per the direction of the project team. This allows the process integration feasibility studies to showcase the financial analysis of the process.
I. The cost included for HTSE reflects NOAK research and manufacturing developments, which will increase the expected lifespan of the electrolysis cells.
J. The HTSE is considered to be linearly scalable, by required capacity, per the direction of the project team. This allows the process integration feasibility studies to showcase the financial analysis of the process.

HTSE Ammonia without Air Separation Unit

Some estimated island costs are based on figures from a verbal conversation with Casale, a leading world vendor of process industry services. These costs were used in cases where other acceptable costs were not available.

VII. CONTINGENCY GUIDELINE IMPLEMENTATION:

Contingency Methodologies: *Explanation of methodology used in determining overall contingency. Identify any specific drivers or items of concern.*

At a project risk review on December 9, 2009, the project team discussed risks to the project. An 18% allowance for capital construction contingency has been included at an island level based on the discussion and is included in the summary sheet. The contingency level that was included in the island cost source documents and additional threats and opportunities identified here were considered during this review. The contingency identified was considered by the project team and included in Cost and Performance Baseline for Fossil Energy Plants DOE/NETL-2007/1281 and similar reports. Typically, contingency allowance provided in these reports ranged from 15% to 20%. Since much of the data contained in this estimate has been derived from these reports, the project team has also chosen a level of contingency consistent with them.

While the level of contingency provided for the HTGR capability is not identified as a line item, the cost data provided to the NGNP Process Integration team was identified as including an appropriate allocation for contingency. No additional contingency has been added to this element.

A. **Threats**: *Uncertain events that are potentially negative or reduce the probability that the desired outcome will happen.*

1. The singularly largest threat to this estimate surrounds the lump sum cost included for the HTGR reactor(s). This is followed by the HTSE process, where applicable. While the overriding assumption is that these elements will be NOAK, currently, a complete HTGR has not been commissioned and the HTSE has been successfully developed in an integrated laboratory-scale model, but has not been completed in either pilot plant or production scales.
2. The level of project definition/development that was available at the time the estimate was prepared represents a substantial risk to the project and is likely to occur. The high level at which elements were considered and included has the potential to include additional elements that are within the work scope but not sufficiently provided for or addressed at this level.
3. The estimate methodology employed is one of a stochastic parametrically evaluated process. This process used publicly available published costs that were related to the process required, costs were normalized using price indices, and the cost was scaled to provide the required capacity. The cost-estimating relationships that were used represent typical costs for BOP allowances, but source cost data from which the initial island costs were derived were not completely descriptive of the elements included, not included, or simply referred to with different nomenclature or combined with other elements. While every effort has been made to correctly normalize and factor the costs for use in this effort, the risk exists that not all of these were correctly captured due to the varied information available.
4. This project is heavily dependent on metals, concrete, petroleum, and petroleum products. Competition for these commodities in today’s environment due to global expansion, uncertainty, and product shortages affects the basic concepts of the supply and demand theories, thus increasing costs.
5. Impacts due to large quantities of materials, special alloy materials, fabrication capability, and labor availability could all represent conditions that may increase the total cost of the project.

B. **Opportunities:** *Uncertain events that could improve the results or improve the probability that the desired outcome will happen.*

1. Additional research and work performed with both vendors and potential owner/operators for a specific process or refinery may identify efficiencies and production means that have not been available for use in this analysis.
2. Recent historical data may identify and include technological advancements and efficiencies not included or reflected in the publicly available source data used in this effort.

Note: Contingency does not increase the overall accuracy of the estimate; it does, however, reduce the level of risk associated with the estimate. Contingency is intended to cover the inadequacies in the complete project scope definition, estimating methods, and estimating data. Contingency specifically excludes changes in project scope, unexpected work stoppages (e.g., strikes, disasters, and earthquakes) and excessive or unexpected inflation or currency fluctuations.

VIII. **OTHER COMMENTS/CONCERNS SPECIFIC TO THE ESTIMATE:**

None.

Detail Item Report - High Temperature Gas Reactor (HTGR)

Project Name: NGNP Process Integration
Process: HTSE Ammonia w/o ASU
Estimate Number: MA36-O

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Internal Cost Data (INL 2009)	1	MWth		2009	$ 1,708,333	$ 1,708,333	$ 1,708,333	2,459	MWth				$ 4,201,101,415	$ 4,201,101,415

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Internal Cost Data (INL 2009)	1	MWth		2009	$ 1,708,333	$ 1,708,333	$ 1,708,333	2,459	MWth				$ 4,201,101,415	$ 4,201,101,415

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	0.00%		$ -	$ -
Civil/Structural/Buildings	0.00%		$ -	$ -
Piping	0.00%		$ -	$ -
Control and Instrumentation	0.00%		$ -	$ -
Electrical Systems	0.00%		$ -	$ -
		Total Balance of Plant	$ -	$ -
		Total Balance of Plant Plus the Selected Source	$ 4,201,101,415	$ 4,201,101,415

Basis of Estimate Notes:

Single source cost point. This cost has been provided by the subcontracted subject matter expert L. Demick to the INL NGNP Process Integration team. This cost represents a complete turnkey cost. The cost of an HTGR reactor, as provided by L. Demick, is $2,000,000 per MWth required. This cost used has been reduced to $1,708,333 per MWth to exclude the cost of power cycles.

Detail Item Report - Rankine Cycle - Case 11, Supercritical PC Case

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia w/o ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-O**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Internal Cost Data (INL 2009)	240	MWe	1	2009	$ 148,362,255	$ 148,362,255	$ 148,362,255	879	MWe	5	176	MWe	$ 123,069,010	$ 615,345,051

Summary:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Internal Cost Data (INL 2009)	240	MWe	1	2009	$ 148,362,255	$ 148,362,255	$ 148,362,255	879	MWe	5	176	MWe	$ 123,069,010	$ 615,345,051

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	0.00%		$ -	$ -
Civil/Structural/Buildings	0.00%		$ -	$ -
Piping	0.00%		$ -	$ -
Control and Instrumentation	0.00%		$ -	$ -
Electrical Systems	0.00%		$ -	$ -
		Total Balance of Plant	$ -	$ -
		Total Balance of Plant Plus the Selected Source	$ 123,069,010	$ 615,345,051

Basis of Estimate Notes:

Single source cost. The reported costs are from the INL project team expert. The reported cost represents a Rankine power cycle, excluding the steam generator. The cost is based on information found in NETL 2007b, which has been adjusted and customized for this project by the INL project team expert. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996. The allowances have been adjusted and customized for this project based on estimator judgment. The reduced civil/structural/buildings allowance accounts for the buildings that are included in the Rankine power cycle cost. Water and electrical systems BOP allowances are included in the reported cost for the Rankine power cycle.

Detail Item Report - High Temperature Steam Electrolysis (HTSE)

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia w/o ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-O**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Feasibility Study (INL 2009)	1.00	kg/s	1	2009	$ 36,120,156	$ 36,120,156	$ 36,120,156	7.51	kg/s				$ 271,216,026	$ 271,216,026

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Feasibility Study (INL 2009)	1.00	kg/s	1	2009	$ 36,120,156	$ 36,120,156	$ 36,120,156	7.51	kg/s				$ 271,216,026	$ 271,216,026

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 19,256,338	$ 19,256,338
Civil/Structural/Buildings	9.20%		$ 24,951,874	$ 24,951,874
Piping	7.10%		$ 19,256,338	$ 19,256,338
Control and Instrumentation	2.60%		$ 7,051,617	$ 7,051,617
Electrical Systems	8.00%		$ 21,697,282	$ 21,697,282
		Total Balance of Plant	$ 92,213,449	$ 92,213,449
		Total Balance of Plant Plus the Selected Source	$ 363,429,475	$ 363,429,475

Rationale for Selection:

Single source cost. The reported costs are from the INL project team expert. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Project Name: NGNP Process Integration
Process: HTSE Ammonia w/o ASU
Estimate Number: MA36-O

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
N2 Generator Cost (Wood 2009)	239,265	lb/hr	1	2007	$13,317,500	$ 13,317,500	$ 12,977,845	236,902	lb/hr	1	236,902	lb/hr	$ 12,900,791	$ 12,900,791

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
N2 Generator Cost (Wood 2009)	239,265	lb/hr	1	2007	$13,317,500	$ 13,317,500	$ 12,977,845	236,902	lb/hr	1	236,902	lb/hr	$ 12,900,791	$ 12,900,791

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 915,956	$ 915,956
Civil/Structural/Buildings	9.20%		$ 1,186,873	$ 1,186,873
Piping	7.10%		$ 915,956	$ 915,956
Control and Instrumentation	2.60%		$ 335,421	$ 335,421
Electrical Systems	8.00%		$ 1,032,063	$ 1,032,063
		Total Balance of Plant	$ 4,386,269	$ 4,386,269
		Total Balance of Plant Plus the Selected Source	$ 17,287,060	$ 17,287,060

Rationale for Selection:

Single source cost point. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - CO2 Generation

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia w/o ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-O**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
CO2 Compression - Subcritical														
Princeton Report (Kreutz 2008)	10	MW	1	2007	$ 6,310,000	$ 6,310,000	$ 6,149,067	2	MW	1	2	MW	$ 2,329,179	$ 2,329,179
CO2 Compression - Supercritical														
Princeton Report (Kreutz 2008)	13	MW	1	2007	$ 9,520,000	$ 9,520,000	$ 9,277,198	0.3	MW	1	0.3	MW	$ 987,887	$ 987,887
CO2 Generation														
CO2 Generator (Wood 2009)	184,095	lb/hr	1	2007	$ 8,102,200	$ 8,102,200	$ 7,895,558	184,021	lb/hr	1	184,021	lb/hr	$ 7,893,654	$ 7,893,654

Source Selected:

Source	Reported Capacity	Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required	Trains Reqd.	Capacity per Train	Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Kreutz 2008: Combined Subcritical and Supercritical Processes										$ 11,210,720	$ 11,210,720

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 795,961	$ 795,961
Civil/Structural/Buildings	9.20%		$ 1,031,386	$ 1,031,386
Piping	7.10%		$ 795,961	$ 795,961
Control and Instrumentation	2.60%		$ 291,479	$ 291,479
Electrical Systems	8.00%		$ 896,858	$ 896,858
		Total Balance of Plant	$ 3,811,645	$ 3,811,645
		Total Balance of Plant Plus the Selected Source	$ 15,022,364	$ 15,022,364

Rationale for Selection:

Single source cost point. The only CO2 generation source considered was Wood 2009. This cost was supplemented with CO2 compression costs from Kreutz 2008 to represent a full island cost. Both subcritical and supercritical process costs were included under the CO2 Compression heading. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, su as Page 1996.

Detail Item Report - Methanation

Project Name: NGNP Process Integration
Process: HTSE Ammonia w/o ASU
Estimate Number: MA36-O

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
DOE FE Report (DOE 1978)	1,000	tpd	1	1978	$ 1,467,000	$ 1,467,000	$ 3,432,834	3,360	tpd	1	3,360	tpd	$ 7,103,237	$ 7,103,237

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
DOE FE Report (DOE 1978)	1,000	tpd	1	1978	$ 1,467,000	$ 1,467,000	$ 3,432,834	3,360	tpd	1	3,360	tpd	$ 7,103,237	$ 7,103,237

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 504,330	$ 504,330
Civil/Structural/Buildings	9.20%		$ 653,498	$ 653,498
Piping	7.10%		$ 504,330	$ 504,330
Control and Instrumentation	2.60%		$ 184,684	$ 184,684
Electrical Systems	8.00%		$ 568,259	$ 568,259
		Total Balance of Plant	$ 2,415,101	$ 2,415,101
		Total Balance of Plant Plus the Selected Source	$ 9,518,338	$ 9,518,338

Rationale for Selection:

Single source island cost identified by the project technical lead. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such
Page 1996.

Detail Item Report - Ammonia Synthesis

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia w/o ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-O**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	360	tpd	1	2009	$ 44,000,000	$ 44,000,000	$ 44,000,000	3,360	tpd	2	1,680	tpd	$ 110,880,901	$ 221,761,801
Economics Ammonia Coal Gasification (Morel 1977)	1,000	tpd	1	1977	$ 50,748,000	$ 50,748,000	$ 80,243,904	3,360	tpd	2	1,680	tpd	$ 109,546,276	$ 219,092,551
Stamicarbon, Middle East Fertilizer Symposium, March 2009	4,297	tpd	1	2008	$ 800,000,000	$ 800,000,000	$ 779,596,498	3,360	tpd	2	1,680	tpd	$ 443,768,260	$ 887,536,519
Ammonia, Chem Systems, 1998	1,653	tpd	1	1998	$ 160,000,000	$ 160,000,000	$ 210,320,924	3,360	tpd	2	1,680	tpd	$ 212,375,463	$ 424,750,926

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	360	tpd	1	2009	$ 44,000,000	$ 44,000,000	$ 44,000,000	3,360	tpd	2	1,680	tpd	$ 110,880,901	$ 221,761,801

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 7,872,544	$ 15,745,088
Civil/Structural/Buildings	9.20%		$ 10,201,043	$ 20,402,086
Piping	7.10%		$ 7,872,544	$ 15,745,088
Control and Instrumentation	2.60%		$ 2,882,903	$ 5,765,807
Electrical Systems	8.00%		$ 8,870,472	$ 17,740,944
		Total Balance of Plant	$ 37,699,506	$ 75,399,012
		Total Balance of Plant Plus the Selected Source	$ 148,580,407	$ 297,160,814

Rationale for Selection:

The verbal cost was selected as recommended by the project team expert. The Stamicarbon information shows a roughly 350% increase in prices between 2003 and 2008 emphasizing the importance of using the most recent information available. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia w/o ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-O**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	625	tpd	1	2009	$ 85,000,000	$ 85,000,000	$ 85,000,000	2,939	tpd	1	2,939	tpd	$ 215,184,342	$ 215,184,342
Perry's Chemical Engineering Handbook, 7th Edition	200	tpd	1	1994	$ 8,800,000	$ 8,800,000	$ 12,240,152	2,939	tpd	1	2,939	tpd	$ 61,388,719	$ 61,388,719
PNNL	4681	tpd	1	2009	$ 182,000,000	$ 182,000,000	$ 182,000,000	2,939	tpd	1	2,939	tpd	$ 137,653,721	$ 137,653,721
Stamicarbon, Middle East Fertilizer Symposium, March 2009	3,582	tpd	1	2008	$ 550,000,000	$ 550,000,000	$ 535,972,592	2,939	tpd	1	2,939	tpd	$ 475,978,115	$ 475,978,115

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	625	tpd	1	2009	$ 85,000,000	$ 85,000,000	$ 85,000,000	2,939	tpd	1	2,939	tpd	$ 215,184,342	$ 215,184,342

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 15,278,088	$ 15,278,088
Civil/Structural/Buildings	9.20%		$ 19,796,959	$ 19,796,959
Piping	7.10%		$ 15,278,088	$ 15,278,088
Control and Instrumentation	2.60%		$ 5,594,793	$ 5,594,793
Electrical Systems	8.00%		$ 17,214,747	$ 17,214,747
		Total Balance of Plant	$ 73,162,676	$ 73,162,676
		Total Balance of Plant Plus the Selected Source	$ 288,347,019	$ 288,347,019

Rationale for Selection:

The verbal cost was selected as recommended by the project team expert. The Stamicarbon information shows a roughly 350% increase in prices between 2003 and 2008 emphasizing the importance of using the most recent information available. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia w/o ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-O**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Perry's Chemical Engineering Handbook, 7th Edition	334	tpd	1	1994	$ 6,600,000	$ 6,600,000	$ 9,180,114	5,192	tpd	6	865	tpd	$ 16,252,129	$ 97,512,774
Fertilizer Manual	359	tpd	1	1998	$ 15,200,000	$ 15,200,000	$ 19,980,488	5,192	tpd	6	865	tpd	$ 33,851,959	$ 203,111,753

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fertilizer Manual	359	tpd	1	1998	$ 15,200,000	$ 15,200,000	$ 19,980,488	5,192	tpd	6	865	tpd	$ 33,851,959	$ 203,111,753

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 2,403,489	$ 14,420,934
Civil/Structural/Buildings	9.20%		$ 3,114,380	$ 18,686,281
Piping	7.10%		$ 2,403,489	$ 14,420,934
Control and Instrumentation	2.60%		$ 880,151	$ 5,280,906
Electrical Systems	8.00%		$ 2,708,157	$ 16,248,940
		Total Balance of Plant	$ 11,509,666	$ 69,057,996
		Total Balance of Plant Plus the Selected Source	$ 45,361,625	$ 272,169,749

Rationale for Selection:

The Fertilizer Manual was selected for being both the newest cost point and the most conservative. The data from Perry's Chemical Engineering Handbook is based on earlier data from Peters and Timmerhaus, making it even less desirable. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Ammonium Nitrate Synthesis

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia w/o ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-O**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity	Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required	Trains Reqd.	Capacity per Train	Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Perry's Chemical Engineering Handbook, 7th Edition	334 tpd	1	1994	$ 6,000,000	$ 6,000,000	$ 8,345,558	3,779 tpd	3	1,260 tpd	$ 18,507,052	$ 55,521,157
Fertilizer Manual	1,395 tpd	1	1998	$ 35,000,000	$ 35,000,000	$ 46,007,702	3,779 tpd	3	1,260 tpd	$ 43,270,765	$ 129,812,295
EPA Report	1,200 tpd	1	1981	$ 9,480,000	$ 9,480,000	$ 13,434,154	3,779 tpd	3	1,260 tpd	$ 13,831,044	$ 41,493,133

Source Selected:

Source	Reported Capacity	Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required	Trains Reqd.	Capacity per Train	Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fertilizer Manual	1,395 tpd	1	1998	$ 35,000,000	$ 35,000,000	$ 46,007,702	3,779 tpd	3	1,260 tpd	$ 43,270,765	$ 129,812,295

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 3,072,224	$ 9,216,673
Civil/Structural/Buildings	9.20%		$ 3,980,910	$ 11,942,731
Piping	7.10%		$ 3,072,224	$ 9,216,673
Control and Instrumentation	2.60%		$ 1,125,040	$ 3,375,120
Electrical Systems	8.00%		$ 3,461,661	$ 10,384,984
		Total Balance of Plant	$ 14,712,060	$ 44,136,180
		Total Balance of Plant Plus the Selected Source	$ 57,982,825	$ 173,948,476

Rationale for Selection:

The Fertilizer Manual was selected for being both the newest cost point and the most conservative. The data from Perry's Chemical Engineering Handbook is based on earlier data from Peters and Timmerhaus, making it even less desirable. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia w/o ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-O**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Steam Turbine and HRSG														
Shell IGCC Base Cases (NETL 2000)	189	MW	1	1999	$ 50,671,000	$ 50,671,000	$ 66,419,744	159	MW	1	159	MW	$ 59,899,366	$ 59,899,366
Steam Turbine														
NETL Baseline Report (NETL 2007a)	401	MW	4	2006	$ 74,651,000	$ 18,662,750	$ 19,125,957	159	MW	1	159	MW	$ 25,233,335	$ 25,233,335
Princeton Report (Kreutz 2008)	275	MW	1	2007	$ 66,700,000	$ 66,700,000	$ 64,998,858	159	MW	1	159	MW	$ 46,806,721	$ 46,806,721
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	230	MW	1	2006	$ 44,515,000	$ 44,515,000	$ 45,619,856	159	MW	1	159	MW	$ 36,576,204	$ 36,576,204

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	230	MW	1	2006	$ 44,515,000	$ 44,515,000	$ 45,619,856	159	MW	1	159	MW	$ 36,576,204	$ 36,576,204

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 2,596,910	$ 2,596,910
Civil/Structural/Buildings	9.20%		$ 3,365,011	$ 3,365,011
Piping	7.10%		$ 2,596,910	$ 2,596,910
Control and Instrumentation	2.60%		$ 950,981	$ 950,981
Electrical Systems	8.00%		$ 2,926,096	$ 2,926,096
		Total Balance of Plant	$ 12,435,909	$ 12,435,909
		Total Balance of Plant Plus the Selected Source	$ 49,012,114	$ 49,012,114

Rationale for Selection:

Shell IGCC PowerPlant with CO2 Capture (NETL 2007b) is a recently reported cost point that closely reflects this project's requirements. The Princeton Report (Kreutz 2008) source for the steam turbine cost point is the NETL 2007b report. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia w/o ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-O**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Steam Turbine and HRSG														
Shell IGCC Base Cases (NETL 2000)	189	MW	1	1999	$ 50,671,000	$ 50,671,000	$ 66,419,744	-	MW	1	-	MW	$ -	$ -
HRSG														
NETL Baseline Report (NETL 2007a)	5,155,983	lb/hr	3	2006	$ 27,581,000	$ 9,193,667	$ 9,421,852	-	lb/hr	1	-	lb/hr	$ -	$ -
Princeton Report (Kreutz 2008)	355	MW	1	2007	$ 52,000,000	$ 52,000,000	$ 50,673,772	-	MW	1	-	MW	$ -	$ -
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	8,438,000	lb/hr	2	2006	$ 45,291,000	$ 22,645,500	$ 23,207,558	-	lb/hr	1	-	lb/hr	$ -	$ -

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	8,438,000	lb/hr	2	2006	$ 45,291,000	$ 22,645,500	$ 23,207,558	-	lb/hr	1	-	lb/hr	$ -	$ -

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ -	$ -
Civil/Structural/Buildings	9.20%		$ -	$ -
Piping	7.10%		$ -	$ -
Control and Instrumentation	2.60%		$ -	$ -
Electrical Systems	8.00%		$ -	$ -
		Total Balance of Plant	$ -	$ -
		Total Balance of Plant Plus the Selected Source	$ -	$ -

Rationale for Selection:

Shell IGCC PowerPlant with CO2 Capture (NETL 2007b) is a recently reported cost point that closely reflects this project's requirements. The Princeton Report (Kreutz 2008) source for the steam turbine cost point is the NETL 2007b report. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia w/o ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-O**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Cooling Tower Depot	182,142	gpm	5	2009	$ 4,611,840	$ 922,368	$ 922,368	160,853	gpm	5	32,171	gpm	$ 856,084	$ 4,280,419

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Cooling Tower Depot	182,142	gpm	5	2009	$ 4,611,840	$ 922,368	$ 922,368	160,853	gpm	5	32,171	gpm	$ 856,084	$ 4,280,419

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 60,782	$ 303,910
Civil/Structural/Buildings	9.20%		$ 78,760	$ 393,799
Piping	7.10%		$ 60,782	$ 303,910
Control and Instrumentation	2.60%		$ 22,258	$ 111,291
Electrical Systems	8.00%		$ 68,487	$ 342,434
		Total Balance of Plant	$ 291,069	$ 1,455,343
		Total Balance of Plant Plus the Selected Source	$ 1,147,152	$ 5,735,762

Rationale for Selection:

Single source cost. Publically available current data. Calculated capital costs based on publically available cost data from a vendor regularly engaged in the building of cooling towers. The allowances listed under 'Balance of Pla
are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

NGNP HTSE Ammonia with ASU Summary

Project Name: NGNP Process Integration
Process: HTSE Ammonia with ASU
Estimate Number: MA36-P

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Process Component	Subtotal From Detail Sheets	Engineering %	Engineering	Contingency %	Contingency	Total Cost
High Temperature Gas Reactor (HTGR)	$ 3,666,008,904	0%	$ -	0%	$ -	$ 3,666,008,904
Rankine Power Cycle	$ 568,211,288	10%	$ 56,821,129	18%	$ 112,505,835	$ 737,538,252
High Temperature Steam Electrolysis (HTSE)	$ 308,019,178	10%	$ 30,801,918	18%	$ 60,987,797	$ 399,808,893
Air Separation Unit (ASU)	$ 49,029,331	10%	$ 4,902,933	18%	$ 9,707,808	$ 63,640,072
CO2 Generation	$ 15,022,226	10%	$ 1,502,223	18%	$ 2,974,401	$ 19,498,850
Methanation	$ 9,518,338	10%	$ 951,834	18%	$ 1,884,631	$ 12,354,803
Ammonia Synthesis	$ 297,160,814	10%	$ 29,716,081	18%	$ 58,837,841	$ 385,714,736
Urea Synthesis	$ 288,347,019	10%	$ 28,834,702	18%	$ 57,092,710	$ 374,274,430
Nitric Acid Synthesis	$ 272,169,749	10%	$ 27,216,975	18%	$ 53,889,610	$ 353,276,334
Ammonium Nitrate Synthesis	$ 173,948,476	10%	$ 17,394,848	18%	$ 34,441,798	$ 225,785,122
Steam Turbines	$ 43,617,835	10%	$ 4,361,783	18%	$ 8,636,331	$ 56,615,949
Cooling Towers	$ 5,927,569	10%	$ 592,757	18%	$ 1,173,659	$ 7,693,985
Total Cost - HTSE Ammonia with ASU						$ 6,302,210,330
Total Cost Rounded to the Nearest $10M						$ 6,300,000,000

	Remarks
Checked By: SAW	
Approved By: John P...	

4/20/2010

Battelle Energy Alliance, LLC

COST ESTIMATE SUPPORT DATA RECAPITULATION

Project Title: NGNP Process Integration – HTSE Ammonia with ASU
Estimator: B. W. Wallace/CEP, R. R. Honsinger/CEP, J. B. Martin/CCT
Date: April 20, 2010
Estimate Type: Class 5
File: MA36-P
Approved By:

I. **PURPOSE:** *Brief description of the intent of how the estimate is to be used (i.e., for engineering study, comparative analysis, request for funding, proposal, etc.).*

It is expected that the capital costs identified in these estimates will be used in a model producing an economic analysis for each specific integrated application and subsequently will be considered in a related feasibility study.

II. **SCOPE OF WORK:** *Brief statement of the project's objective. Thorough overview and description of the proposed project. Identify work to be accomplished, as well as any specific work to be excluded.*

A. **Objective:**
Develop Class 5 estimates as defined by the Association for Advancement of Cost Engineering (AACEi) that will identify the current capital cost associated with high-temperature gas reactors (HTGRs) integrated with a nuclear ammonia with the air separation unit process.

B. **Included:**
The scope of work required to achieve this objective includes the following:

1. Engineering
2. The allowance provided for the HTGR represents a complete and operable system. All elements required for construction of this nuclear reactor capability, including an initial steam generator, security systems, contingency, and owner's costs are included in the turn-key allowance. Owner's costs are included only in the case of the reactor capability. It is considered that the total value represents all inside of battery limits (ISBL) elements, outside of battery limits (OSBL) elements, site development, and all ancillary control and operational functions and capabilities.
3. Construction of a new integrated refinery capability to produce ammonia that consists of the following:
   a. Overnight island-type costs for HTGRs
   b. High-temperature steam electrolysis (HTSE) hydrogen production unit
   c. Air separation unit
   d. Natural gas combustor (CO_2 generation)
   e. Methanation
   f. Ammonia synthesis
   g. Urea synthesis
   h. Nitric acid synthesis

i. Ammonium nitrate synthesis
j. Steam turbines, internal to process
k. Heat recovery steam generator, internal to process
l. Cooling towers, internal to process
m. Allowances for Balance of Plant (BOP)/offsite/OSBL, including the following:
   (1.) Site development/improvements
   (2.) Provisions for general and administrative buildings and structures
   (3.) Provisions for OSBL piping
   (4.) Provisions for OSBL instrumentation and control
   (5.) Provisions for OSBL electrical
   (6.) Provisions for facility supply and OSBL water systems
   (7.) Provisions for site development/improvements
   (8.) Project/construction management.

C. **Excluded:**
This scope of work specifically excludes the following elements:
1. Licensing and permitting costs
2. Operational costs
3. Land costs
4. Sales taxes
5. Royalties
6. Owner's fees and owner's costs, except those included for the HTGR
7. The allowance provided for the HTGR capability excludes all costs associated with materials development, or costs that would not be appropriately associated with an nth of a kind (NOAK) reactor/facility.

III. **ESTIMATE METHODOLOGY:** *Overall methodology and rationale of how the estimate was developed (i.e., parametric, forced detail, bottoms up, etc.). Total dollars/hours and rough order magnitude (ROM) allocations of the methodologies used to develop the cost estimate.*

Consistent with the AACEi Class 5 estimates, the level of definition and engineering development available at the time they were prepared, their intended use in a feasibility study, and the time and resources available for their completion, the costs included in this estimate have been developed using parametric evaluations. These evaluations have used publicly available and published project costs to represent similar islands utilized in this project. Analysis and selection of the published costs used have been performed by the project technical lead and Cost Estimating. Suitability for use in this effort was determined considering the correctness and completeness of the data available, the manner in which total capital costs were represented, the age of the previously performed work, and the similarity to the capacity/trains required by this project. The specific sources, selected and used in this cost estimate, are identified in the capital cost estimate detail sheets. Adjustments have been made to these published costs using escalation factors identified in the Chemical Engineering Price Cost Index. Scaling of the published island costs has been

accomplished using the six-tenths capacity factoring method. Costs included for the HTGR, power cycles, and HTSE, have been identified and provided by the respective BEA subject matter experts. The total cost for each of these items has been linearly calculated from the respective base unit costs. Any normalization to provide for geographic factors was considered using geographic factors available from RS Means Construction Cost Data references. Cost-estimating relationships have been used to identify allowances to complete the costs.

It was identified to the Next Generation Nuclear Plant (NGNP) Process Integration team that the methodology employed by NGNP to develop the nuclear capability included constituents of parametric modeling, vendor quotes, actual costs, and proprietary costing databases. These preconceptual design estimates were reviewed by NGNP Project Engineering for credibility with regard to assumptions and bases of estimate and performed multiple studies to reconcile variations in the scope and assumptions within the three estimates.

BOP/OSBL costs were determined by the project team, considering data provided by Shell Gasifier IGCC Base Case report NETL 2000, *Conceptual Cost Estimating Manual* Second Edition by John S. Page, and additional adjusted sources. Because the allowances identified did not show significant variability, the allowances identified in the NETL 2000 report were chosen for this effort in order to minimize the mixing of data sources.

IV. **BASIS OF THE ESTIMATE**: *Overall explanation of sources for resource pricing and schedules.*

A. **Quantification Basis**: *The source for the measurable quantities in the estimate that can be used in support of earned value management. Source documents may include drawings, design reports, engineers' notes, and other documentation upon which the estimate is originated.*

All islands and capacities have been provided to Cost Estimating by the respective project expert.

B. **Planning Basis**: *The source for the execution and strategies of the work that can be used to support the project execution plan, acquisition strategy, schedules, and market conditions and other documentation upon which the estimate is originated.*

1. All islands and HTGRs represent NOAK projects.
2. Projects will be constructed and operated by commercial entities.
3. All projects, with the exception of the Steam-Assisted Gravity Drainage Project, will be located in the U.S. Gulf Coast refinery region.
4. Costs are presented as overnight costs.
5. The cost estimate does not consider or address funding or labor resource restrictions. Sufficient funding and labor resources will be available in a manner that allows optimum usage of the funding and resources as estimated and scheduled.

C. **Cost Basis:** *The source for the costing on the estimate that can be used in support of earned value management, funding profiles, and schedule of values. Sources may include published costing references, judgment, actual costs, preliminary quotes or other documentation upon which the estimate is originated.*

1. All costs are represented as current value costs. Factors for forward-looking escalation and inflation factors are not included in this estimate.
2. Where required, published cost factors, as identified in the Chemical Engineering Plant Cost Index, will be applied to previous years' values to determine current year values.
3. Geographic location factors, as identified in RS Means Construction Cost Data reference manual, were considered for each source cost.
4. The cost provided for the HTGR reflects internal BEA cost data that was developed for the HTGR and presented to the NGNP Process Integration team by L. Demmick. Considered in the cost is a pre-conceptual cost estimate prepared by three separate contractor teams. All contractor teams proposed 4-unit NOAK plants with thermal power levels between 2,000 MW_t and 2,400 MW_t at a cost of roughly \$4B, including owner's cost. This equates to \$1,667 to \$2,000 per kW_t. For the purposes of this report, the nominal cost of an HTGR will be set at the upper end of this range, \$2,000 per kW_t. This is a complete turnkey cost and includes engineering and construction of a NOAK HTGR, the power cycle, and contingency. The total HTGR cost for each process is calculated linearly as \$1,708,333 per MWth of required capacity, excluding the cost of the power cycles.
5. The cost included for the power cycle was provided by the INL project team expert. The power cycle cost is based on the definition of a 240-MWe capacity and \$618,176 per MWe. The total power cycle cost for each process is calculated linearly as \$618,176 per MWe of required capacity. BOP, engineering, and contingency costs are added to the base cost.
6. The cost included for HTSE was provided by the INL project team expert. The total HTSE cost for each process is calculated linearly as \$36,120,156 per kg/s of required capacity. BOP, engineering, and contingency costs are added to the base cost.
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V. **ESTIMATE QUALITY ASSURANCE:** *A listing of all estimate reviews that have taken place and the actions taken from those reviews.*

A review of the cost estimate was held on January 14, 2010, with the project team and the cost estimators. This review allowed for the project team to review and comment, in detail, on the perceived scope, basis of estimates, assumptions, project risks, and resources that make up this cost estimate. Comments from this review have been incorporated into this estimate to reflect a project team consensus of this document.

VI. **ASSUMPTIONS:** *Condition statements accepted or supposed true without proof of demonstration; statements adding clarification to scope. An assumption has a direct impact on total estimated cost.*

General Assumptions:

A. All costs are represented in 2009 values.

B. Costs that were included from sources representing years prior to 2009 have been normalized to 2009 values using the Chemical Engineering Plant Cost Index. This index was selected due to its widespread recognition and acceptance and its specific orientation toward work associated with chemical and refinery plants.

C. Capital costs are based on process islands. The majority of these islands are interchangeable, after factoring for the differing capacities, flowsheet-to-flowsheet.

D. All chemical processing and refinery processes will be located in the U.S. Gulf Coast region.

E. All costs considered to be BOP costs that can be specifically identified have been factored out of the reported source data and added into the estimate in a manner consistent with that identified in the NETL 2000 IGCC Base Cost report. Inclusion of the source costs in this manner normalizes all reported cost information to the bare-erected costs.

HTGR:

A. The linearly scalable cost included for an HTGR reflects an NOAK reactor with a 750°C-operating temperature.
B. HTGR is considered to be linearly scalable, by required capacity, per the direction of the project team. This allows the process integration feasibility studies to showcase the financial analysis of the process without the added burden of integer quantity 600-MWth HTGRs.
C. The allowance represents a turnkey condition for the reactor and its supporting infrastructure.
D. A high-temperature, high-pressure steam generator is included in the cost represented for HTGR.
E. A contingency allowance is included in the HTGR cost, but is not identified as a separate line item in this estimate. This allowance was identified and included by the NGNP HTGR project team.
F. Total cost range, including contingency, for HTGR is -50%, +100%.
G. Cost included for the power cycle reflects NOAK research and manufacturing developments to allow for assumed high pressures and temperatures.
H. The power cycle is considered to be linearly scalable, by required capacity, per the direction of the project team. This allows the process integration feasibility studies to showcase the financial analysis of the process.
I. The cost included for HTSE reflects NOAK research and manufacturing developments, which will increase the expected lifespan of the electrolysis cells.
J. The HTSE is considered to be linearly scalable, by required capacity, per the direction of the project team. This allows the process integration feasibility studies to showcase the financial analysis of the process.

HTSE Ammonia with Air Separation Unit

Some estimated island costs are based on figures from a verbal conversation with Casale, a leading world vendor of process industry services. These costs were used in cases where other acceptable costs were not available.

VII. CONTINGENCY GUIDELINE IMPLEMENTATION:

Contingency Methodologies: *Explanation of methodology used in determining overall contingency. Identify any specific drivers or items of concern.*

At a project risk review on December 9, 2009, the project team discussed risks to the project. An 18% allowance for capital construction contingency has been included at an island level based on the discussion and is included in the summary sheet. The contingency level that was included in the island cost source documents and additional threats and opportunities identified here were considered during this review. The contingency identified was considered by the project team and included in Cost and Performance Baseline for Fossil Energy Plants DOE/NETL-2007/1281 and similar reports. Typically, contingency allowance provided in these reports ranged from 15% to 20%. Since much of the data contained in this estimate has been derived from these reports, the project team has also chosen a level of contingency consistent with them.

While the level of contingency provided for the HTGR capability is not identified as a line item, the cost data provided to the NGNP Process Integration team was identified as including an appropriate allocation for contingency. No additional contingency has been added to this element.

A. **Threats**: *Uncertain events that are potentially negative or reduce the probability that the desired outcome will happen.*

1. The singularly largest threat to this estimate surrounds the lump sum cost included for the HTGR reactor(s). This is followed by the HTSE process, where applicable. While the overriding assumption is that these elements will be NOAK, currently, a complete HTGR has not been commissioned and the HTSE has been successfully developed in an integrated laboratory-scale model, but has not been completed in either pilot plant or production scales.
2. The level of project definition/development that was available at the time the estimate was prepared represents a substantial risk to the project and is likely to occur. The high level at which elements were considered and included has the potential to include additional elements that are within the work scope but not sufficiently provided for or addressed at this level.
3. The estimate methodology employed is one of a stochastic parametrically evaluated process. This process used publicly available published costs that were related to the process required, costs were normalized using price indices, and the cost was scaled to provide the required capacity. The cost-estimating relationships that were used represent typical costs for BOP allowances, but source cost data from which the initial island costs were derived were not completely descriptive of the elements included, not included, or simply referred to with different nomenclature or combined with other elements. While every effort has been made to correctly normalize and factor the costs for use in this effort, the risk exists that not all of these were correctly captured due to the varied information available.
4. This project is heavily dependent on metals, concrete, petroleum, and petroleum products. Competition for these commodities in today's environment due to global expansion, uncertainty, and product shortages affects the basic concepts of the supply and demand theories, thus increasing costs.
5. Impacts due to large quantities of materials, special alloy materials, fabrication capability, and labor availability could all represent conditions that may increase the total cost of the project.

B. **Opportunities:** *Uncertain events that could improve the results or improve the probability that the desired outcome will happen.*

1. Additional research and work performed with both vendors and potential owner/operators for a specific process or refinery may identify efficiencies and production means that have not been available for use in this analysis.
2. Recent historical data may identify and include technological advancements and efficiencies not included or reflected in the publicly available source data used in this effort.

Note: Contingency does not increase the overall accuracy of the estimate; it does, however, reduce the level of risk associated with the estimate. Contingency is intended to cover the inadequacies in the complete project scope definition, estimating methods, and estimating data. Contingency specifically excludes changes in project scope, unexpected work stoppages (e.g., strikes, disasters, and earthquakes) and excessive or unexpected inflation or currency fluctuations.

VIII. OTHER COMMENTS/CONCERNS SPECIFIC TO THE ESTIMATE:

None.

Detail Item Report - High Temperature Gas Reactor (HTGR)

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia with ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-P**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Internal Cost Data (INL 2009)	1	MWth		2009	$ 1,708,333	$ 1,708,333	$ 1,708,333	2,146	MWth				$ 3,666,008,904	$ 3,666,008,904

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Internal Cost Data (INL 2009)	1	MWth		2009	$ 1,708,333	$ 1,708,333	$ 1,708,333	2,146	MWth				$ 3,666,008,904	$ 3,666,008,904

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	0.00%		$ -	$ -
Civil/Structural/Buildings	0.00%		$ -	$ -
Piping	0.00%		$ -	$ -
Control and Instrumentation	0.00%		$ -	$ -
Electrical Systems	0.00%		$ -	$ -
		Total Balance of Plant	$ -	$ -
		Total Balance of Plant Plus the Selected Source	$ 3,666,008,904	$ 3,666,008,904

Basis of Estimate Notes:

Single source cost point. This cost has been provided by the subcontracted subject matter expert L. Demick to the INL NGNP Process Integration team. This cost represents a complete turnkey cost. The cost of an HTGR reactor, as provided by L. Demick, is $2,000,000 per MWth required. This cost used has been reduced to $1,708,333 per MWth to exclude the cost of power cycles.

Detail Item Report - Rankine Cycle - Case 11, Supercritical PC Case

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia with ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-P**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Internal Cost Data (INL 2009)	240	MWe	1	2009	$ 148,362,255	$ 148,362,255	$ 148,362,255	770	MWe	5	154	MWe	$ 113,642,258	$ 568,211,288

Summary:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Internal Cost Data (INL 2009)	240	MWe	1	2009	$ 148,362,255	$ 148,362,255	$ 148,362,255	770	MWe	5	154	MWe	$ 113,642,258	$ 568,211,288

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	0.00%		$ -	$ -
Civil/Structural/Buildings	0.00%		$ -	$ -
Piping	0.00%		$ -	$ -
Control and Instrumentation	0.00%		$ -	$ -
Electrical Systems	0.00%		$ -	$ -
		Total Balance of Plant	$ -	$ -
		Total Balance of Plant Plus the Selected Source	$ 113,642,258	$ 568,211,288

Basis of Estimate Notes:

Single source cost. The reported costs are from the INL project team expert. The reported cost represents a Rankine power cycle, excluding the steam generator. The cost is based on information found in NETL 2007b, which has been adjusted and customized for this project by the INL project team expert. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996. The allowances have been adjusted and customized for this project based on estimator judgment. The reduced civil/structural/buildings allowance accounts for the buildings that are included in the Rankine power cycle cost. Water and electrical systems BOP allowances are included in the reported cost for the Rankine power cycle.

Detail Item Report - High Temperature Steam Electrolysis (HTSE)

Project Name: NGNP Process Integration
Process: HTSE Ammonia with ASU
Estimate Number: MA36-P

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Internal Cost Data (INL 2009)	1.00	kg/s		2009	$ 36,120,156	$ 36,120,156	$ 36,120,156	6.36	kg/s				$ 229,865,058	$ 229,865,058

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
INL Internal Cost Data (INL 2009)	1.00	kg/s		2009	$ 36,120,156	$ 36,120,156	$ 36,120,156	6.36	kg/s				$ 229,865,058	$ 229,865,058

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 16,320,419	$ 16,320,419
Civil/Structural/Buildings	9.20%		$ 21,147,585	$ 21,147,585
Piping	7.10%		$ 16,320,419	$ 16,320,419
Control and Instrumentation	2.60%		$ 5,976,492	$ 5,976,492
Electrical Systems	8.00%		$ 18,389,205	$ 18,389,205
		Total Balance of Plant	$ 78,154,120	$ 78,154,120
		Total Balance of Plant Plus the Selected Source	$ 308,019,178	$ 308,019,178

Basis of Estimate Notes:

Single source cost. The reported costs are from the INL project team expert. The cost is based on information from Harvego 2008, Solid State Energy Conversion Alliance, and discussions between INL engineers and Ceramatec and Proton Energy. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Air Separation Unit (ASU)

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia with ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-P**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Shell IGCC Base Case (NETL 2000)	213,207	lb/hr	1	1999	$ 51,204,000	$ 51,204,000	$ 67,118,402	65,715	lb/hr	1	65,715	lb/hr	$ 33,125,271	$ 39,836,648
NETL Baseline Report (NETL 2007a)	1,728,789	lb/hr	2	2006	$ 287,187,000	$ 143,593,500	$ 147,157,470	65,715	lb/hr	1	65,715	lb/hr	$ 31,358,395	$ 37,711,792
Princeton Report (Kreutz 2008)	201,264	lb/hr	1	2007	$ 105,000,000	$ 105,000,000	$ 102,322,040	65,715	lb/hr	1	65,715	lb/hr	$ 52,276,709	$ 62,868,281
Hydrogen Report (Gray 2004)	296,583	lb/hr	1	2004	$ 76,000,000	$ 76,000,000	$ 87,600,180	65,715	lb/hr	1	65,715	lb/hr	$ 35,466,315	$ 35,466,315
Shell GTC Report (Shell 2004)	385,259	lb/hr	1	2004	$ 53,760,000	$ 53,760,000	$ 61,965,601	65,715	lb/hr	1	65,715	lb/hr	$ 21,443,600	$ 25,788,201
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	373,498	lb/hr	2	2006	$ 144,337,000	$ 72,168,500	$ 73,959,712	65,715	lb/hr	1	65,715	lb/hr	$ 39,522,003	$ 47,529,396

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Average of normalized and factored costs from NETL 2007a and Gray 2004													$ 33,412,355	$ 36,589,053

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 2,372,277	$ 2,597,823
Civil/Structural/Buildings	9.20%		$ 3,073,937	$ 3,366,193
Piping	7.10%		$ 2,372,277	$ 2,597,823
Control and Instrumentation	2.60%		$ 868,721	$ 951,315
Electrical Systems	8.00%		$ 2,672,988	$ 2,927,124
		Total Balance of Plant	$ 11,360,201	$ 12,440,278
		Total Balance of Plant Plus the Selected Source	$ 44,772,555	$ 49,029,331

Rationale for Selection:

NETL Baseline Report (NETL 2007a) and Hydrogen Report (Gray 2004) have been selected. An average cost of the two has been selected in order to not represent an overly agressive or conservative cost. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Note: The base ASU cost was multiplied by "1.36^0.6" to account for the increase in oxygen output purity from 95% to 99.5%. The adjustment is based on INL simulations calculating the increase in capacity that would be needed have the required purity output. The Gray 2004 report uses an oxygen purity of 99% and was not adjusted by the "1.36^0.6."

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia with ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-P**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
CO2 Compression - Subcritical														
Princeton Report (Kreutz 2008)	10	MW	1	2007	$ 6,310,000	$ 6,310,000	$ 6,149,067	2	MW	1	2	MW	$ 2,329,179	$ 2,329,179
CO2 Compression - Supercritical														
Princeton Report (Kreutz 2008)	13	MW	1	2007	$ 9,520,000	$ 9,520,000	$ 9,277,198	0.3	MW	1	0.3	MW	$ 987,887	$ 987,887
CO2 Generation														
CO2 Generator (Wood 2009)	184,095	lb/hr	1	2007	$ 8,102,200	$ 8,102,200	$ 7,895,558	184,017	lb/hr	1	184,017	lb/hr	$ 7,893,551	$ 7,893,551

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Kreutz 2008: Combined Subcritical and Supercritical Processes													$ 11,210,617	$ 11,210,617

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 795,954	$ 795,954
Civil/Structural/Buildings	9.20%		$ 1,031,377	$ 1,031,377
Piping	7.10%		$ 795,954	$ 795,954
Control and Instrumentation	2.60%		$ 291,476	$ 291,476
Electrical Systems	8.00%		$ 896,849	$ 896,849
		Total Balance of Plant	$ 3,811,610	$ 3,811,610
		Total Balance of Plant Plus the Selected Source	$ 15,022,226	$ 15,022,226

Rationale for Selection:

Single source cost point. The only CO2 generation source considered was Wood 2009. This cost was supplemented with CO2 compression costs from Kreutz 2008 to represent a full island cost. Both subcritical and supercritical process costs were included under the CO2 Compression heading. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, su as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia with ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-P**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
DOE FE Report (DOE 1978)	1,000	tpd	1	1978	$ 1,467,000	$ 1,467,000	$ 3,432,834	3,360	tpd	1	3,360	tpd	$ 7,103,237	$ 7,103,237

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
DOE FE Report (DOE 1978)	1,000	tpd	1	1978	$ 1,467,000	$ 1,467,000	$ 3,432,834	3,360	tpd	1	3,360	tpd	$ 7,103,237	$ 7,103,237

Balance of Plant:

Description	% of Total Cost						Cost Per Train	Total Cost
Water Systems	7.10%						$ 504,330	$ 504,330
Civil/Structural/Buildings	9.20%						$ 653,498	$ 653,498
Piping	7.10%						$ 504,330	$ 504,330
Control and Instrumentation	2.60%						$ 184,684	$ 184,684
Electrical Systems	8.00%						$ 568,259	$ 568,259
						Total Balance of Plant	$ 2,415,101	$ 2,415,101
						Total Balance of Plant Plus the Selected Source	$ 9,518,338	$ 9,518,338

Rationale for Selection:

Single source island cost identified by the project technical lead. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such
Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia with ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-P**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	360	tpd	1	2009	$ 44,000,000	$ 44,000,000	$ 44,000,000	3,360	tpd	2	1,680	tpd	$ 110,880,901	$ 221,761,801
Economics Ammonia Coal Gasification (Morel 1977)	1,000	tpd	1	1977	$ 50,748,000	$ 50,748,000	$ 80,243,904	3,360	tpd	2	1,680	tpd	$ 109,546,276	$ 219,092,551
Stamicarbon, Middle East Fertilizer Symposium, March 2009	4,297	tpd	1	2008	$ 800,000,000	$ 800,000,000	$ 779,596,498	3,360	tpd	2	1,680	tpd	$ 443,768,260	$ 887,536,519
Ammonia, Chem Systems, 1998	1,653	tpd	1	1998	$ 160,000,000	$ 160,000,000	$ 210,320,924	3,360	tpd	2	1,680	tpd	$ 212,375,463	$ 424,750,926

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	360	tpd	1	2009	$ 44,000,000	$ 44,000,000	$ 44,000,000	3,360	tpd	2	1,680	tpd	$ 110,880,901	$ 221,761,801

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 7,872,544	$ 15,745,088
Civil/Structural/Buildings	9.20%		$ 10,201,043	$ 20,402,086
Piping	7.10%		$ 7,872,544	$ 15,745,088
Control and Instrumentation	2.60%		$ 2,882,903	$ 5,765,807
Electrical Systems	8.00%		$ 8,870,472	$ 17,740,944
		Total Balance of Plant	$ 37,699,506	$ 75,399,012
		Total Balance of Plant Plus the Selected Source	$ 148,580,407	$ 297,160,814

Rationale for Selection:

The verbal cost was selected as recommended by the project team expert. The Stamicarbon information shows a roughly 350% increase in prices between 2003 and 2008 emphasizing the importance of using the most recent information available. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia with ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-P**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	625	tpd	1	2009	$ 85,000,000	$ 85,000,000	$ 85,000,000	2,939	tpd	1	2,939	tpd	$ 215,184,342	$ 215,184,342
Perry's Chemical Engineering Handbook, 7th Edition	200	tpd	1	1994	$ 8,800,000	$ 8,800,000	$ 12,240,152	2,939	tpd	1	2,939	tpd	$ 61,388,719	$ 61,388,719
PNNL	4681	tpd	1	2009	$ 182,000,000	$ 182,000,000	$ 182,000,000	2,939	tpd	1	2,939	tpd	$ 137,653,721	$ 137,653,721
Stamicarbon, Middle East Fertilizer Symposium, March 2009	3,582	tpd	1	2008	$ 550,000,000	$ 550,000,000	$ 535,972,592	2,939	tpd	1	2,939	tpd	$ 475,978,115	$ 475,978,115

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Vendor - Verbal 2009	625	tpd	1	2009	$ 85,000,000	$ 85,000,000	$ 85,000,000	2,939	tpd	1	2,939	tpd	$ 215,184,342	$ 215,184,342

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 15,278,088	$ 15,278,088
Civil/Structural/Buildings	9.20%		$ 19,796,959	$ 19,796,959
Piping	7.10%		$ 15,278,088	$ 15,278,088
Control and Instrumentation	2.60%		$ 5,594,793	$ 5,594,793
Electrical Systems	8.00%		$ 17,214,747	$ 17,214,747
		Total Balance of Plant	$ 73,162,676	$ 73,162,676
		Total Balance of Plant Plus the Selected Source	$ 288,347,019	$ 288,347,019

Rationale for Selection:

The verbal cost was selected as recommended by the project team expert. The Stamicarbon information shows a roughly 350% increase in prices between 2003 and 2008 emphasizing the importance of using the most recent information available. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Nitric Acid Synthesis

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia with ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-P**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Perry's Chemical Engineering Handbook, 7th Edition	334	tpd	1	1994	$ 6,600,000	$ 6,600,000	$ 9,180,114	5,192	tpd	6	865	tpd	$ 16,251,244	$ 97,507,466
Fertilizer Manual	359	tpd	1	1998	$ 15,200,000	$ 15,200,000	$ 19,980,488	5,192	tpd	6	865	tpd	$ 33,851,959	$ 203,111,753

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fertilizer Manual	359	tpd	1	1998	$ 15,200,000	$ 15,200,000	$ 19,980,488	5,192	tpd	6	865	tpd	$ 33,851,959	$ 203,111,753

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 2,403,489	$ 14,420,934
Civil/Structural/Buildings	9.20%		$ 3,114,380	$ 18,686,281
Piping	7.10%		$ 2,403,489	$ 14,420,934
Control and Instrumentation	2.60%		$ 880,151	$ 5,280,906
Electrical Systems	8.00%		$ 2,708,157	$ 16,248,940
		Total Balance of Plant	$ 11,509,666	$ 69,057,996
		Total Balance of Plant Plus the Selected Source	$ 45,361,625	$ 272,169,749

Rationale for Selection:

The Fertilizer Manual was selected for being both the newest cost point and the most conservative. The data from Perry's Chemical Engineering Handbook is based on earlier data from Peters and Timmerhaus, making it even less desirable. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Ammonium Nitrate Synthesis

Project Name: NGNP Process Integration
Process: HTSE Ammonia with ASU
Estimate Number: MA36-P

Client: M. Patterson
Prepared By: B. Wallace, R. Honsinger, J. Martin
Estimate Type: Class 5

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Perry's Chemical Engineering Handbook, 7th Edition	334	tpd	1	1994	$ 6,000,000	$ 6,000,000	$ 8,345,558	3,779	tpd	3	1,260	tpd	$ 18,507,052	$ 55,521,157
Fertilizer Manual	1,395	tpd	1	1998	$ 35,000,000	$ 35,000,000	$ 46,007,702	3,779	tpd	3	1,260	tpd	$ 43,270,765	$ 129,812,295
EPA Report	1,200	tpd	1	1981	$ 9,480,000	$ 9,480,000	$ 13,434,154	3,779	tpd	3	1,260	tpd	$ 13,831,044	$ 41,493,133

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Fertilizer Manual	1,395	tpd	1	1998	$ 35,000,000	$ 35,000,000	$ 46,007,702	3,779	tpd	3	1,260	tpd	$ 43,270,765	$ 129,812,295

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 3,072,224	$ 9,216,673
Civil/Structural/Buildings	9.20%		$ 3,980,910	$ 11,942,731
Piping	7.10%		$ 3,072,224	$ 9,216,673
Control and Instrumentation	2.60%		$ 1,125,040	$ 3,375,120
Electrical Systems	8.00%		$ 3,461,661	$ 10,384,984
		Total Balance of Plant	$ 14,712,060	$ 44,136,180
		Total Balance of Plant Plus the Selected Source	$ 57,982,825	$ 173,948,476

Rationale for Selection:

The Fertilizer Manual was selected for being both the newest cost point and the most conservative. The data from Perry's Chemical Engineering Handbook is based on earlier data from Peters and Timmerhaus, making it even less desirable. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia with ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-P**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Steam Turbine and HRSG														
Shell IGCC Base Cases (NETL 2000)	189	MW	1	1999	$ 50,671,000	$ 50,671,000	$ 66,419,744	131	MW	1	131	MW	$ 53,340,709	$ 53,340,709
Steam Turbine														
NETL Baseline Report (NETL 2007a)	401	MW	4	2006	$ 74,651,000	$ 18,662,750	$ 19,125,957	131	MW	1	131	MW	$ 22,456,151	$ 22,456,151
Princeton Report (Kreutz 2008)	275	MW	1	2007	$ 66,700,000	$ 66,700,000	$ 64,998,858	131	MW	1	131	MW	$ 41,655,168	$ 41,655,168
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	230	MW	1	2006	$ 44,515,000	$ 44,515,000	$ 45,619,856	131	MW	1	131	MW	$ 32,550,623	$ 32,550,623

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Shell IGCC Power Plant with CO2 Capture (NETL 2007b)	230	MW	1	2006	$ 44,515,000	$ 44,515,000	$ 45,619,856	131	MW	1	131	MW	$ 32,550,623	$ 32,550,623

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 2,311,094	$ 2,311,094
Civil/Structural/Buildings	9.20%		$ 2,994,657	$ 2,994,657
Piping	7.10%		$ 2,311,094	$ 2,311,094
Control and Instrumentation	2.60%		$ 846,316	$ 846,316
Electrical Systems	8.00%		$ 2,604,050	$ 2,604,050
		Total Balance of Plant	$ 11,067,212	$ 11,067,212
		Total Balance of Plant Plus the Selected Source	$ 43,617,835	$ 43,617,835

Rationale for Selection:

Shell IGCC PowerPlant with CO2 Capture (NETL 2007b) is a recently reported cost point that closely reflects this project's requirements. The Princeton Report (Kreutz 2008) source for the steam turbine cost point is the NETL 2007b report. The allowances listed under 'Balance of Plant' are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.

Detail Item Report - Cooling Towers

Project Name:	**NGNP Process Integration**	**Client:**	**M. Patterson**
Process:	**HTSE Ammonia with ASU**	**Prepared By:**	**B. Wallace, R. Honsinger, J. Martin**
Estimate Number:	**MA36-P**	**Estimate Type:**	**Class 5**

Sources Considered:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Cooling Tower Depot	156,524	gpm	5	2009	$ 4,611,840	$ 922,368	$ 922,368	146,019	gpm	5	29,204	gpm	$ 884,712	$ 4,423,559

Source Selected:

Source	Reported Capacity		Reported Trains	Report Cost Year	Reported Cost	Reporting Year Cost Per Train	Normalized Cost Per Train using CEPCI Index	Capacity Required		Trains Reqd.	Capacity per Train		Factored Cost per Train from Normalized Cost	Total Current Cost for Required Trains
Cooling Tower Depot	156,524	gpm	5	2009	$ 4,611,840	$ 922,368	$ 922,368	146,019	gpm	5	29,204	gpm	$ 884,712	$ 4,423,559

Balance of Plant:

Description	% of Total Cost		Cost Per Train	Total Cost
Water Systems	7.10%		$ 62,815	$ 314,073
Civil/Structural/Buildings	9.20%		$ 81,393	$ 406,967
Piping	7.10%		$ 62,815	$ 314,073
Control and Instrumentation	2.60%		$ 23,003	$ 115,013
Electrical Systems	8.00%		$ 70,777	$ 353,885
		Total Balance of Plant	$ 300,802	$ 1,504,010
		Total Balance of Plant Plus the Selected Source	$ 1,185,514	$ 5,927,569

Rationale for Selection:

Single source cost. Publically available current data. Calculated capital costs based on publically available cost data from a vendor regularly engaged in the building of cooling towers. The allowances listed under 'Balance of Pla are based on NETL 2000. These allowance values are comparable to additional published estimating guides, such as Page 1996.