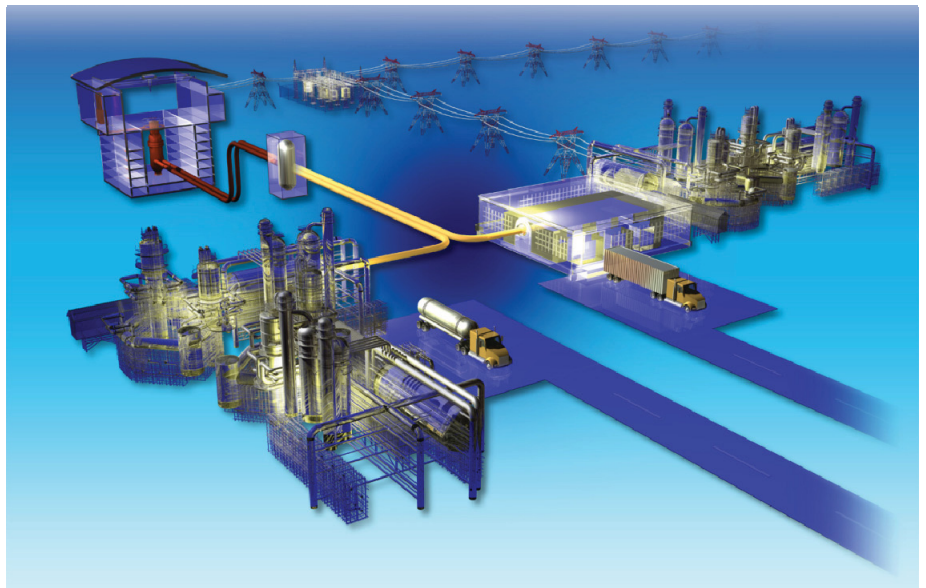


Technical Evaluation Study

Project No. 23843

HTGR-Integrated Hydrogen Production via Steam Methane Reforming (SMR) Process Analysis

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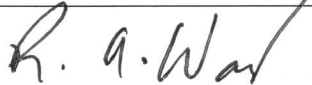


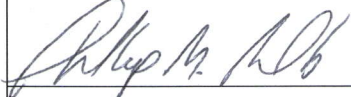
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NGNP Project	Technical Evaluation Study (TEV)	eCR Number: 582916
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EXECUTIVE SUMMARY

This technical evaluation (TEV) has been prepared as part of a study for the Next Generation Nuclear Plant Project to evaluate integration of High Temperature Gas-cooled Reactor (HTGR) technology with conventional chemical processes. This TEV addresses the integration of an HTGR for hydrogen production via steam methane reforming (SMR) of natural gas.

The HTGR can produce electricity and/or heat in the form of steam or high-temperature helium. In conventional chemical processes these products are generated by the combustion of fossil fuels such as coal and natural gas, resulting in significant greenhouse gas (GHG) emissions, including 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 and electricity could be integrated into conventional SMR processes for hydrogen production. For this study, the HTGR outlet temperature was set at 750°C (1,382°F) and a 40% electrical generation efficiency was assumed for the Rankine cycle.

Two baseline SMR cases were developed for this evaluation: one that includes carbon capture (CC) and one that does not. Similarly, two corresponding HTGR-integrated cases were developed. The hydrogen production capacity for all cases was held constant at 130 MMSCFD (approximately 1/100 of the current U.S. production capacity). Significant results from this evaluation are:

- Including CC in the conventional flowsheet results in an overall process efficiency penalty of 4.7%. However, this penalty is manifest as a reduction in export steam generation capability—hydrogen product yield is actually slightly better for the flowsheet that includes CC.
- For this application, the HTGR is used primarily to supply process heat. Only a small fraction of the energy from the HTGR is required as electrical power.
- Integration of nuclear heat can reduce the amount of natural gas required to produce a given quantity of hydrogen. If CC is not included in the design, these results indicate that natural gas consumption can be reduced by 15.3%. For a design that incorporates CC, the reduction is 11.6%. These results are reflected in the H₂ product yield, as shown in ES-1.
- CO₂ emissions are reduced as a result of integrating nuclear heat. If CC is not included in the design, nuclear integration results in a 15.4% reduction in CO₂ emissions. For a design that incorporates CC, the CO₂ emissions reduction is 38.6%. CO₂ disposition for each case is shown graphically in ES-2.
- The requirements to integrate HTGR technology with SMR for hydrogen production are quite modest— $\frac{1}{4}$ or less of a 600-MWt reactor would be sufficient to support a world-scale single-train hydrogen plant (130 MMSCFD). Large

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industrial processes may require multiple trains of this size to meet their hydrogen demand, therefore justifying construction of one or more HTGRs.

Based on the results of this study, several follow-on activities are recommended. The most significant of these recommendations are listed below.

- An economic analysis should be performed using the modeling results from this study as input. It is believed that such an analysis would further quantify the benefits of integrating HTGR heat with SMR technology for hydrogen production.
- It is likely that process results for the HTGR-integrated case could be improved if the HTGR temperature could be increased beyond 750°C (1,382°F). Hence, a study to quantify the performance improvement is recommended. Observations from this study indicate that CC should be included in the baseline configuration for the temperature sensitivity study.
- Additional work is warranted to scope out initial equipment design and further assess the feasibility of a HTGR-integrated SMR. As indicated in this report, the mode of heat transfer would shift from radiation in a conventional design to convection in a HTGR-integrated design.

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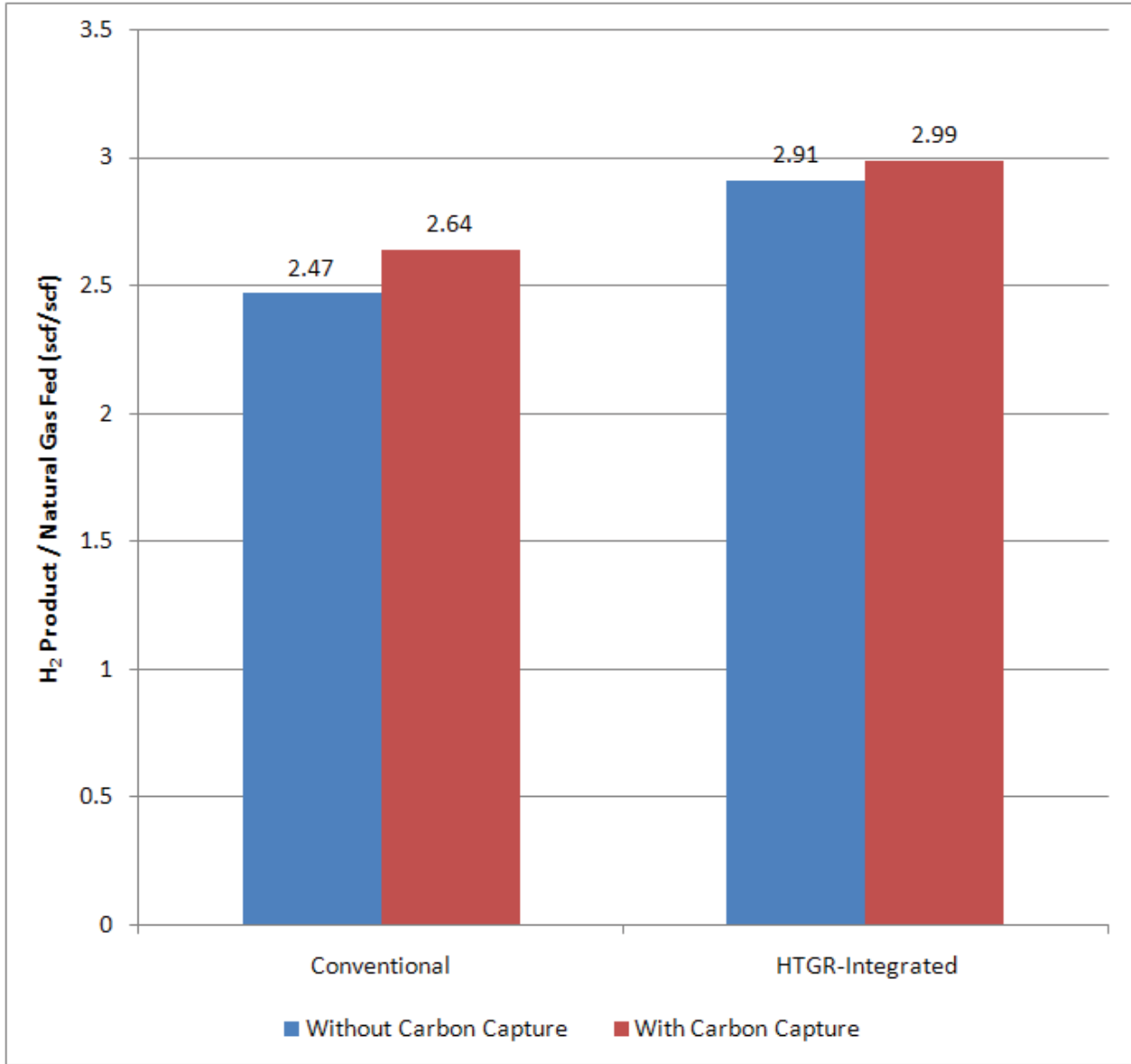


Figure ES-1. SMR H₂ product yield.

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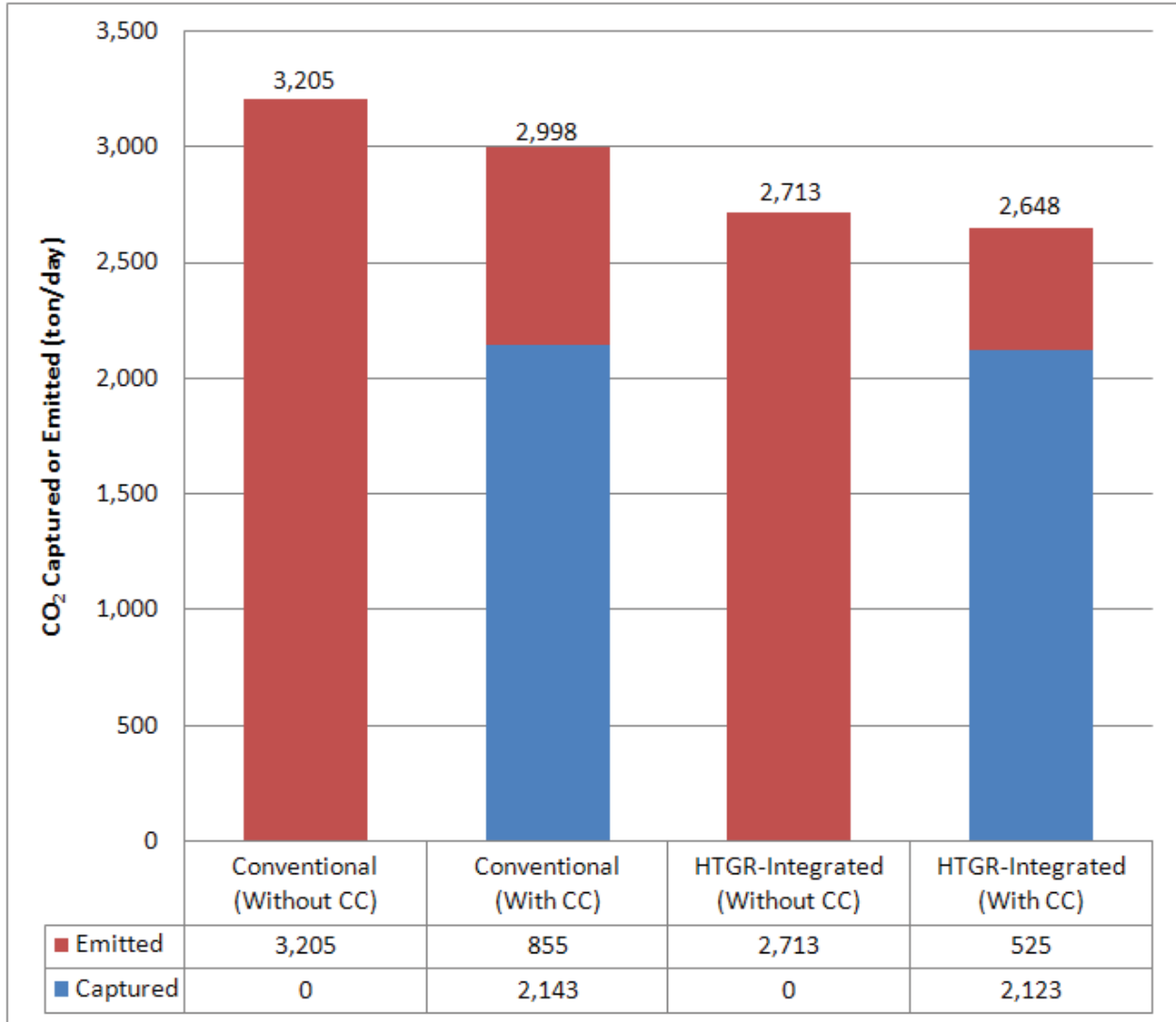


Figure ES-2. SMR CO₂ disposition.

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ACRONYMS AND NOMENCLATURE

CC	carbon capture
DOE	Department of Energy
GHG	greenhouse gases
HHV	higher heating value
HTGR	High Temperature Gas-cooled Reactor
HTSE	high-temperature steam electrolysis
INL	Idaho National Laboratory
MMBTU	1,000,000 British thermal units
MMSCF	1,000,000 standard cubic feet
NGNP	Next Generation Nuclear Plant
PSA	pressure swing absorption
ROT	reactor outlet temperature
SMR	steam methane reformer
TEV	technical evaluation

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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 *2005 Energy Policy Act* to promote reliance on safe, clean, economic nuclear energy and to establish a greenhouse-gas (GHG)-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 (1,382°F); this reflects the initial HTGR design and assumes a more conservative outlet temperature. Eventually temperatures of 950°C (1,722°F) are anticipated. Additionally, a 50°C (90°F) 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 (1,292°F). 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 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 HTGR-integrated process. This report describes how nuclear-generated heat or electricity could be integrated into conventional processes.

Table 1. Projected outputs of the NGNP.

HTGR Product	Product Description
Steam	540°C (1,004°F) and 17 MPa
High-Temperature Helium	Delivered at 700°C (1,292°F) and 9.1 MPa
Electricity	Generated by Rankine cycle with 40% thermal efficiency

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This TEV addresses potential integration opportunities for hydrogen production via steam methane reforming (SMR) of natural gas. More specifically, this TEV considers only integration of nuclear heat and electricity into the process—hydrogen production via high-temperature steam electrolysis (HTSE) is not addressed.^a The HTGR would produce electricity and heat, and would be located near the SMR facility. Details of the specific cases considered are described in Sections 2 and 3 of this report. A separate study should be conducted to assess the optimal siting of the HTGR with respect to the SMR 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 hydrogen production methods.

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 study.

This TEV first presents a general discussion of conventional steam methane reforming technologies used to produce hydrogen. Next, the specific processes selected for modeling are presented. A description of each process model is then provided. Finally, the results of the process modeling are discussed with emphasis placed on the impact of the HTGR integration.

2. PROCESS DESCRIPTION

2.1 Conventional Production of Hydrogen via SMR

A block flow diagram for a typical hydrogen production plant is shown in Figure 1. A majority of the natural gas fed to the plant is first desulfurized. The desulfurized natural gas is mixed with steam and reformed in catalyst-filled tubes via the following endothermic reactions to produce hydrogen:



a. HTGR-integrated hydrogen production using HTSE has previously been addressed in TEV-693, “Nuclear-Integrated Hydrogen Production Analysis.”

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Reformer temperatures between 800 and 870°C (1,472–1,598°F) are typical (Baade 2001), although lower temperatures (~760°C [1,400°F]) can be used to reduce the metallurgical requirements of the reforming tubes (Elshout 2010). Conversion of methane to hydrogen is improved by operating at higher temperatures. Many modern designs operate near the upper end of the temperature range, typically 860–880°C (1,580–1,616°F) to improve conversion (Uhde 2009).

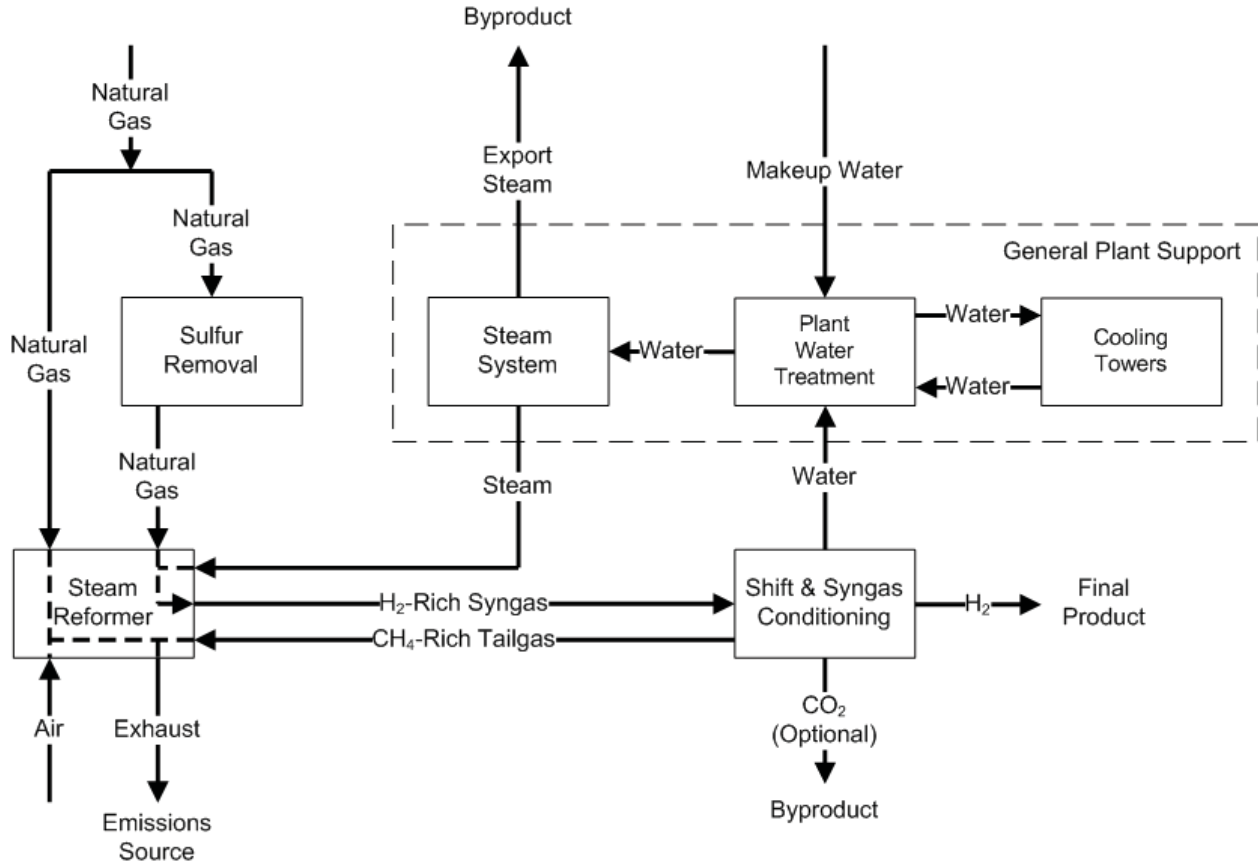


Figure 1. Typical block flow diagram for hydrogen production using SMR.

Adequate steam must also be fed to the reformer to prevent laydown of coke on the reforming catalyst. A steam-to-carbon molar ratio of three is typical for a natural gas feed consisting primarily of methane, but ratios as high as five have been used commercially to improve conversion in the reformer.

Due to the endothermic nature of the reforming reactions, heat must be supplied to the reformer. This is accomplished by firing natural gas to heat the reforming tubes. The amount of fresh natural gas that must be fired can be minimized by supplementing the burner with methane-rich tailgas that has been separated from

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the hydrogen product. Hence, there is a trade-off to be made in the design of a hydrogen plant:

1. Lower reformer temperature – A lower temperature reduces the metallurgical requirements for the reforming tubes. However, because a significant fraction of methane is unconverted through the reformer, more extensive separation is required and the overall size of the process equipment is somewhat increased.
2. Higher reformer temperature – A higher temperature increases the metallurgical requirements for the reforming tubes. However, methane conversion is higher resulting in a corresponding reduction in the size of the process equipment. Additionally, the separation requirements are somewhat reduced due to the lower methane concentration in the raw hydrogen product.

Another consideration in the design of a hydrogen plant is the desired amount of steam export. For a stand-alone hydrogen plant, it may be desirable to minimize steam export from the plant. However, if the hydrogen plant is located adjacent to another chemical process that requires process steam, it may be desirable to design the hydrogen plant for increased steam export.

As shown in Figure 1, the syngas produced in the reformer must be shifted and conditioned to produce a purified hydrogen product. The first step in purifying the reformed syngas involves conversion of carbon monoxide formed via Reaction 1 to hydrogen. This is accomplished using shift conversion and serves to maximize hydrogen production from the plant. Shift conversion is accomplished catalytically via the following exothermic reversible reaction:



This reaction proceeds faster at a higher temperature, but lower equilibrium CO concentration (and hence, maximum H₂ production) is favored at a lower temperature. Hence, a typical shift conversion arrangement consists of two catalytic beds in series with interstage heat exchange.^b The first bed, or high-temperature shift reactor, typically operates with an inlet temperature between 320 and 350°C (608–662°F). Due to the exothermic nature of the shift reaction, this temperature rises to approximately 420°C (788°F) at the outlet of the high-temperature shift reactor. The majority of the CO conversion is accomplished in the first stage where faster reaction kinetics are favorable due to the elevated operating temperature. This minimizes the size of the equipment.

^b See the process flow diagram presented in Appendix A for details of the two-bed shift conversion reactors and heat recuperation.

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Effluent from the first stage is cooled to approximately 200°C (392°F) prior to introduction to the second (low-temperature) shift converter. This stage converts a majority of the remaining CO to CO₂, thus maximizing hydrogen production by taking advantage of the favorable equilibrium conditions at the lower temperature.

Following shift conversion, there are several options for CO₂ removal and final purification of the hydrogen product. Two factors determine which option is selected: (1) the required purity of the hydrogen product, and (2) the need to sequester CO₂.

When low-purity hydrogen (<99%) is the desired product, two purifications steps are typically utilized: CO₂ removal and methanation. CO₂ is scrubbed out using any number of commercial solvents such as hot potassium carbonate, amines (such as MEA or MDEA), Selexol™, Rectisol™, or Sulfinol™. Once the bulk CO₂ has been absorbed from the hydrogen product, residual CO and CO₂ are removed in a methanation reactor, which converts carbon oxides to methane via the following reactions (Kohl 1997):



The primary impurity remaining in the hydrogen product is methane—a small amount as a result of methanation, and a larger amount as a result of incomplete methane conversion in the steam reformer. A typical hydrogen product is 98.2% H₂ and 1.8% CH₄, but may be as low as 92–95% H₂ (Baade 2001). Obviously the final purity is very closely tied to methane conversion in the reformer.

When high purity hydrogen (>99%) is the desired product, a single purification step is typically utilized: pressure swing adsorption (PSA). In years past, cryogenic separation was utilized, but this option has essentially been replaced by PSA in newer designs. The basic principle of PSA is that larger molecules and other light hydrocarbons can be selectively adsorbed on high surface area materials such as molecular sieves. Hydrogen has a weak affinity for adsorption. In a PSA system, a bed is operated under a pressurization–depressurization cycle. Multiple beds are typically used in order to provide a continuous flow of hydrogen even when some beds are in the depressurization (or desorption) cycle. Hydrogen recovery using a PSA system is typically 70–90%, with the ability to produce very pure hydrogen (99.999%). Note that in a typical PSA system, the desorbed reject gas used as fuel gas in the reformer typically contains a large fraction of CO₂ as well as combustible CO, CH₄, and H₂. Hence, the fuel gas can have a relatively low heating value. One advantage of the PSA system is that final product purity is decoupled from methane conversion in the reformer.

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When both a high-purity hydrogen product and carbon capture (CC) are required, two purification steps are typically implemented: CO₂ removal and final H₂ purification. As with the low hydrogen purity option described above, CO₂ can be absorbed using any number of commercial solvents. The hydrogen stream is further purified using PSA to remove methane and any traces of CO, CO₂, or other impurities. This results in a very pure hydrogen product stream, a sequestration-ready CO₂ stream, and a relatively high heating value fuel gas stream for use in the reformer.

Very few (if any) hydrogen plants currently sequester CO₂. The above option (CO₂ absorption followed by PSA for final H₂ purification) simply represents a combination of two well-proven technologies in the hydrogen manufacturing industry (i.e., low risk). A less-proven but potentially desirable option could be to use a specially tuned PSA system to simultaneously produce purified hydrogen, purified CO₂, and a high heating value fuel gas. Air Products has experimented with such a system at the pilot scale. Results indicated that the process can produce ultra-pure hydrogen (99.999%) at a recovery of 86–88% simultaneously with relatively pure carbon dioxide (99.4%) at a recovery in excess of 90% (Kohl 1997).

2.2 Nuclear-Integration Study Considerations

INL studies conducted during the last year have investigated hydrogen production via SMR for specific applications: Fischer-Tropsch synthesis for diesel fuel production (INL 2010a), methanol production with subsequent conversion to gasoline (INL 2010b), and ammonia production (INL 2010c). For each of these applications, reforming was tailored to produce the optimal synthesis gas. For this study, however, hydrogen is the final product rather than an intermediate product. Hence, to provide maximum flexibility for product use, a high-purity hydrogen plant configuration was selected for evaluation.

It is not standard practice to capture and sequester CO₂ as part of the H₂ manufacturing process. However, as environmental policy develops in regard to GHG emissions, it is conceivable that future H₂ plants may be required to reduce CO₂ emissions. Hence, for this study both CC and non-CC options were evaluated.

Another nuclear-integration issue that needed to be addressed in this study was the slight temperature mismatch between what can be supplied by a first-generation HTGR (700°C [1,292°F]) versus the optimal steam methane reformer operating temperature (760–880°C [1,400–1,616°F]). This discrepancy will be eliminated as future HTGR designs supply higher temperature heat (up to 900°C [1,652°F]). However, because this study assumes the HTGR can provide only 700°C (1,292°F) heat, some process modifications were required in order to

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utilize the available heat. The approach taken in this study was to separate the primary reformer into two zones or stages. In the first stage, nuclear heat replaces gas firing as the heat source. The partially reformed gas exiting the first stage is then routed to the second reformer stage. In this stage, additional heat is supplied by firing fuel gas to complete the reforming process.

3. PROCESS MODELING OVERVIEW

Plant models for hydrogen production via SMR 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 (model MacPro2,1), 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.

Four cases were originally identified for modeling, including flowsheets both with and without CC:

- Conventional SMR process without CC
- Conventional SMR process with CC
- HTGR-integrated SMR process without CC
- HTGR-integrated SMR process with CC.

For all cases, natural gas composition was taken from data published by Northwest Gas Association. Net capacity for the plant was set at 130 million scfd of hydrogen (29,000 lb/hr), which corresponds with the size of the largest single train hydrogen plant currently in operation in Europe.

For the Aspen models described in this analysis, a rigorous submodel of the nuclear power cycle has not yet been integrated. For this analysis, it is assumed that the reactor can provide hot helium for integration with the hydrogen plant as well as a Rankine cycle to generate electricity. A portion of the electricity generated by the HTGR plant is used to supply the needs of the hydrogen plant; the balance can be sold as a byproduct. Results for the nuclear power cycle were calculated separately using the UNISIM modeling

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package. 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.

A general model description for each case is presented below. Natural gas feed rate in each model was adjusted to achieve the desired net capacity for the plant of 130 million scfd of hydrogen, which matches the present capacity of a world-scale single-train hydrogen plant.

3.1 Conventional SMR Process (Without CC)

Figure 2 shows the block flow diagram for the conventional SMR case without CC. The proposed process includes unit operations for steam reforming, syngas shifting and conditioning, steam system, 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 (see Appendix B).

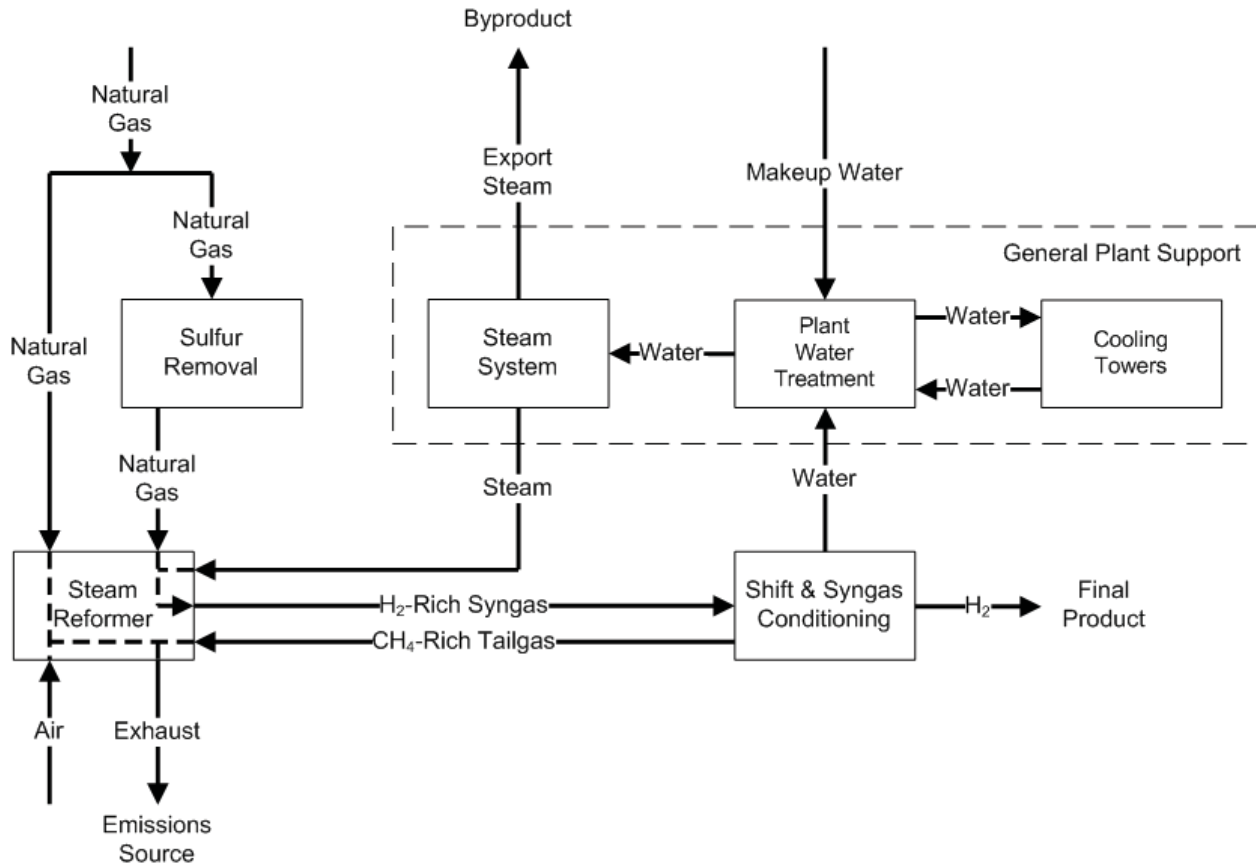


Figure 2. Block flow diagram for the conventional SMR case without CC.

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- **Sulfur Removal, Steam Reformer (REFORMER)** – For this study, a reformer temperature of 871°C (1,600°F) was selected to balance high methane conversion with reasonable reformer tube life. The steam-to-carbon inlet molar ratio was set at 3.0 in order to avoid carbon deposition on the reforming catalyst (Elshout 2010).

Natural gas is split into two streams. Of the total natural gas flow, 15.1% is burned to provide heat for the primary reformer. The remaining 84.9% of the natural gas flow is compressed to 460 psia and mixed with a small amount of hydrogen to achieve a concentration of 2 mol% H₂. The addition of a small amount of hydrogen is required to convert any sulfur in the natural gas to a chemical form that can be removed in the sulfur removal bed. The natural gas is then preheated to 221°C (430°F) and saturated with hot water. After saturation, the gas is further heated to 400°C (752°F) prior to hydrotreating to convert sulfur compounds to H₂S. Sulfur is then removed from the gas using a bed of zinc oxide. The desulfurized natural gas is then mixed with steam to achieve the desired steam-to-carbon molar ratio of 3.0. The resulting natural gas/steam mixture is then preheated to 700°C (1,292°F) prior to introduction to the reformer. A preformer is not included in this flowsheet, although it could be considered due to the relatively high preheat temperature.

The natural gas/steam mixture is fed to the primary reformer where methane is converted over a catalyst to CO, H₂, and CO₂. Using a Gibbs free energy minimization routine, methane conversion in the reformer is predicted to be 78.1% based on the specified conditions. 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 natural gas, syngas, water, and steam to provide preheat for these streams. In addition, the hot offgas from the reformer is used to preheat combustion air as well as raising 600 psig and 60 psig steam.

The effluent syngas from the primary reformer is cooled by exchange with boiler feed water to raise 600 psig steam. The resulting syngas has a H₂/CO ratio of 5.2 and contains 34.2 mol% H₂O, 5.3 mol% CO₂, and 3.5 mol% CH₄.

- **Shift & Syngas Conditioning (GAS-CLN)** – Figure 3 shows the shift and syngas conditioning scenario modeled for this case. Reformed gas is fed to the first shift converter at a temperature of 335°C (635°F) with a steam-to-dry-gas ratio of 0.52. Gas exits the first shift converter at a temperature of 409°C (769°F) due to the exothermic nature of the shift reaction. CO concentration is reduced to 3.2 mol%.

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Figure 3. Shift and syngas conditioning block flow diagram for cases without CC.

Steam is raised by exchange with the hot syngas exiting the first shift converter; the syngas is cooled to 191°C (375°F) prior to introduction to the second-stage shift converter. The steam-to-dry-gas ratio entering the second-stage shift converter is 0.38, which is above the minimum limit of 0.30 required to maintain good catalyst activity while inhibiting Fischer-Tropsch reactions. Gas exits the second shift converter at a temperature of 214°C (418°F) due to the exothermic nature of the shift reaction. CO concentration is reduced to 0.35 mol%. Additional steam is raised by exchange with the hot syngas exiting the second shift converter.

After recovering additional heat and condensing water from the shifted syngas, the PSA system is used to separate the purified H₂ product. Hydrogen recovery is specified at 88% and hydrogen purity is specified at 99.9% (although higher purities are achievable). A low-Btu fuel gas containing 56.5 mol% CO₂ is also produced by the PSA separation.

- Steam System (STEAM)** – Aspen Utility blocks are used to track all steam generation and use in the plant. This information is used as input to the steam system section of the model, allowing reconciliation of the entire plant steam balance. Excess steam produced by the process is available for export as a byproduct. For this case, 154 million Btu/hr of 600-psig steam and 35 million Btu/hr of 60-psig steam are available for export.
- 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.
- Plant Water Treatment (H2O-TRTM)** – Water treatment is simplistically modeled in Aspen Plus using a variety of separation blocks. INL is currently collaborating with water treatment vendors to develop the water treatment portion of the model. The existing water treatment

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scenario is a place holder and will be revised as information is received from the water treatment vendors. 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.

3.2 Conventional SMR Process (With CC)

Figure 4 shows the block flow diagram for the conventional SMR case with CC. The proposed process includes unit operations for steam reforming, syngas shifting and conditioning, steam system, 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 (see Appendix C). Because many of the unit operations remain unchanged from the conventional SMR case without CC, emphasis is placed on differences in configuration between the two cases.

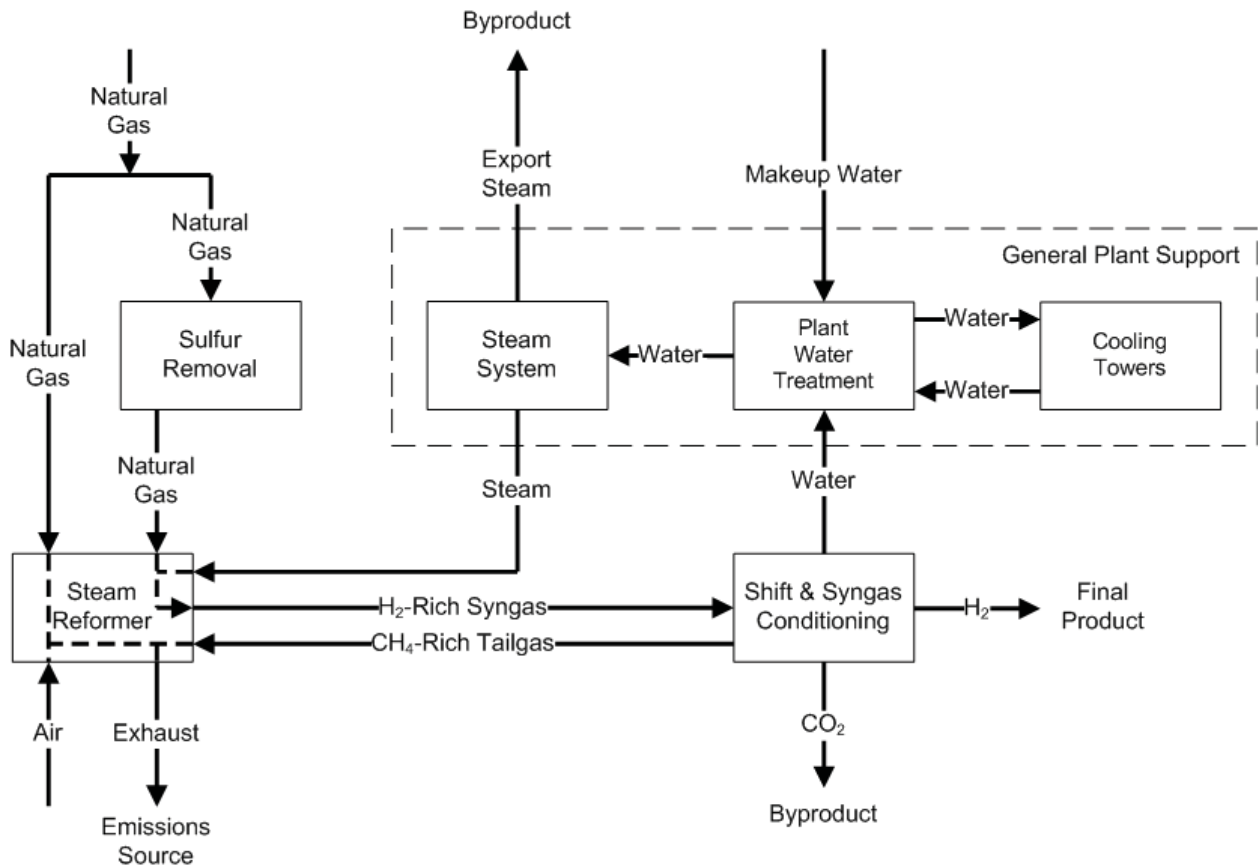


Figure 4. Block flow diagram for the conventional SMR case with CC.

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- Sulfur Removal, Steam Reformer (REFORMER)** – Steam reforming configuration and conditions in the SMR case with CC are nearly identical to those used in the conventional SMR case without CC. The primary difference is that the higher Btu content of the fuel gas in this case results in a lower demand for fresh natural gas to fire the reformer. The fraction of the total natural gas feed used for firing the reformer is reduced from 15.1% in the non-CC case to 9.0% for the CC case. Other key operating parameters are identical between the cases.
- Shift & Syngas Conditioning (GAS-CLN)** – Figure 5 shows the shift and syngas conditioning scenario modeled for this case. Shift conversion configuration and conditions are identical for the conventional SMR case with CC to the conventional SMR case without CC.

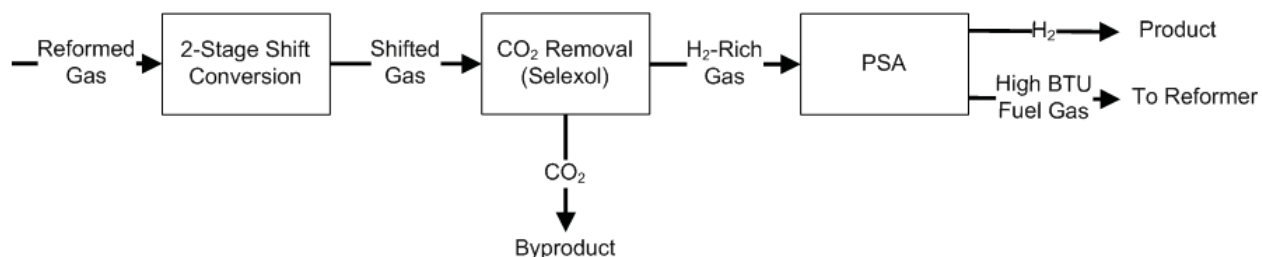


Figure 5. Shift and syngas conditioning block flow diagram for cases with CC.

Downstream of shift conversion, CO₂ removal is accomplished using the Selexol process. The inlet gas contains 18.9 mol% CO₂. The absorber and stripper columns in this system are not rigorously modeled; instead, a separation block is used to simplistically represent this process. CO₂ capture is specified at 99%, and CO₂ purity is specified at 99.1 mol%. Steam usage and electrical consumption of the Selexol process are estimated based on literature values (Burr 2008) and incorporated into the Aspen model results.

CO₂ recovered using the Selexol process is compressed to 2,000 psig using an eight-stage compressor. Power consumption for compression was tuned to match vendor information.

PSA configuration is identical for the conventional SMR case with CC to the conventional SMR case without CC. Hydrogen recovery is again specified at 88%, and hydrogen purity is specified at 99.9%. However, due to the difference in feed-gas composition, a high-Btu fuel gas containing only 1.3 mol% CO₂ is produced by the PSA separation.

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- Steam System (STEAM)** – The steam system in the conventional SMR case with CC is identical to that in the conventional SMR case without CC. For this case, 24 million Btu/hr of 600-psig steam is available for export.
- Cooling Towers (COOL-TWR)** – The cooling towers in the conventional SMR case with CC are identical to those in the conventional SMR case without CC.
- Plant Water Treatment (H2O-TRTM)** – The water treatment system in the conventional SMR case with CC is identical to that in the conventional SMR case without CC.

3.3 HTGR-integrated SMR Process (Without CC)

Figure 6 shows the block-flow diagram for the HTGR-integrated SMR process without CC. The proposed process includes the same unit operations as the conventional SMR process without CC: steam reforming, syngas shifting and conditioning, steam system, cooling towers, and water treatment. In addition, this configuration adds the HTGR system.

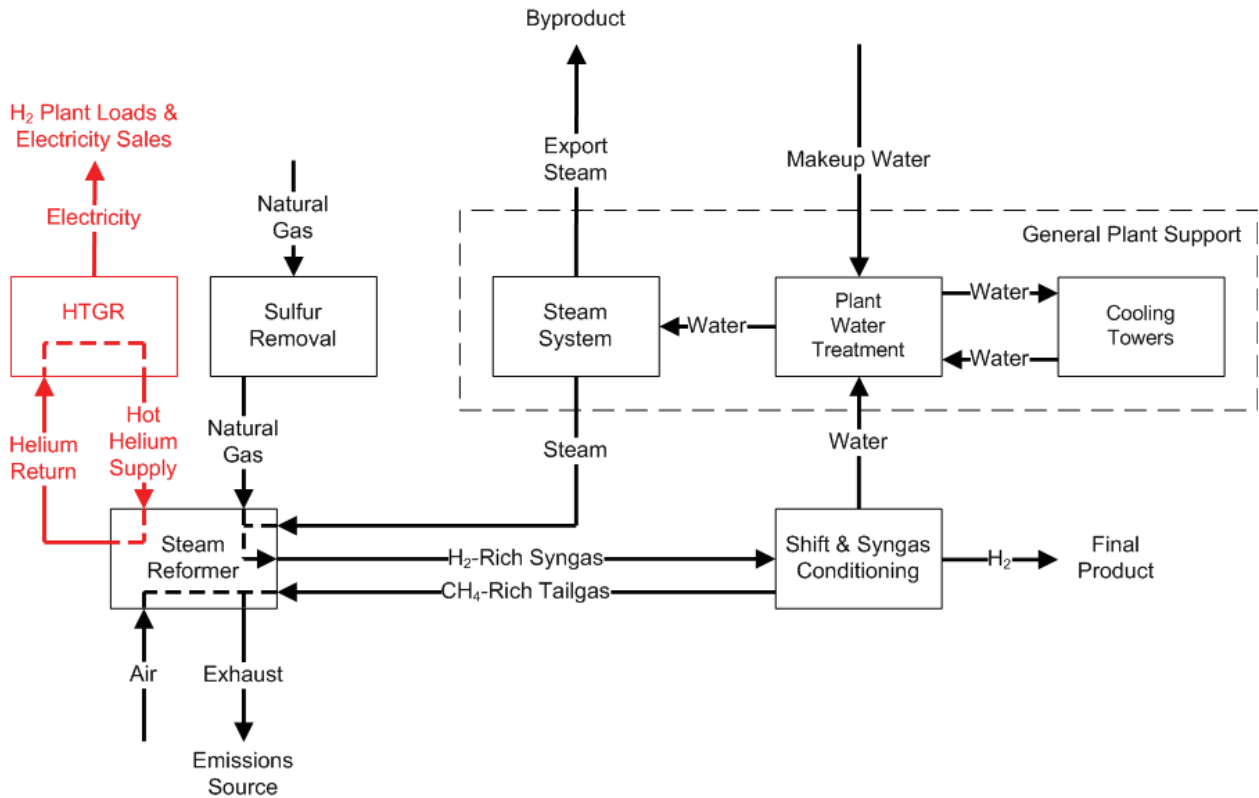


Figure 6. Block flow diagram for the HTGR-integrated SMR case without CC.

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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 (see Appendix D). Because many of the unit operations remain unchanged from the conventional SMR case without CC, emphasis is placed on differences in configuration between the two cases.

- Sulfur Removal, Steam Reformer (REFORMER)** – As previously discussed in Section 2.2 of this report, a slight temperature mismatch exists between what can be supplied by a first-generation HTGR (700°C [1,292°F]) versus the optimal steam methane reformer operating temperature (760–880°C [1,400–1,616°F]). Hence, some process modifications were required in order to utilize the available heat. The approach taken in this study was to separate the primary reformer into two zones or stages.

In the first stage, nuclear heat replaces gas firing as the heat source. As in the conventional SMR cases, the steam-to-carbon inlet molar ratio was set at 3.0. This stage operates with a reformer outlet temperature of 675°C (1247°F). Nuclear heat is also used to preheat natural gas, steam, and air. The required helium flow rate is set based on maintaining a 25°C (45°F) temperature difference between the helium inlet to the final preheater and the reformer feed-gas effluent. Methane conversion in this stage was 28.0%.

The partially reformed gas exiting the first stage is then routed to the second reformer stage. In this stage, additional heat is supplied by firing fuel gas to complete the reforming process. As in the conventional SMR cases, a final reformer temperature of 871°C (1,600°F) was selected to balance high methane conversion with reasonable reformer tube life. Methane conversion in this stage was 69.8%.

Other key operating parameters are identical to the conventional SMR cases, except that no natural gas is used for firing the reformer (i.e., fuel gas is used exclusively for this purpose in the second stage). This configuration results in a very tight (and perhaps unrealistic) temperature approach due to the low Btu content and amount of fuel gas available.

In addition, although reforming is split into two stages, overall methane conversion closely matches that of the conventional SMR cases (78.2%).

- Shift & Syngas Conditioning (GAS-CLN)** – The shift and syngas conditioning operations and configuration in the HTGR-integrated SMR case without CC is identical to that in the conventional SMR case without CC.

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- Steam System (STEAM)** – The steam system in the HTGR-integrated SMR case without CC is identical to that in the conventional SMR case without CC. For this case, 273 million Btu/hr of 600-psig steam and 0.4 million Btu/hr of 60-psig steam are available for export.
- Cooling Towers (COOL-TWR)** – The cooling towers in the HTGR-integrated SMR case without CC are identical to those in the conventional SMR case without CC.
- Plant Water Treatment (H2O-TRTM)** – The water treatment system in the HTGR-integrated SMR case without CC is identical to that in the conventional SMR case without CC.

3.4 HTGR-integrated SMR Process (With CC)

Figure 7 shows the block-flow diagram for the HTGR-integrated SMR process with CC. The proposed process includes the same unit operations as the conventional SMR process with CC: steam reforming, syngas shifting and conditioning, steam system, cooling towers, and water treatment. In addition, this configuration adds the HTGR system.

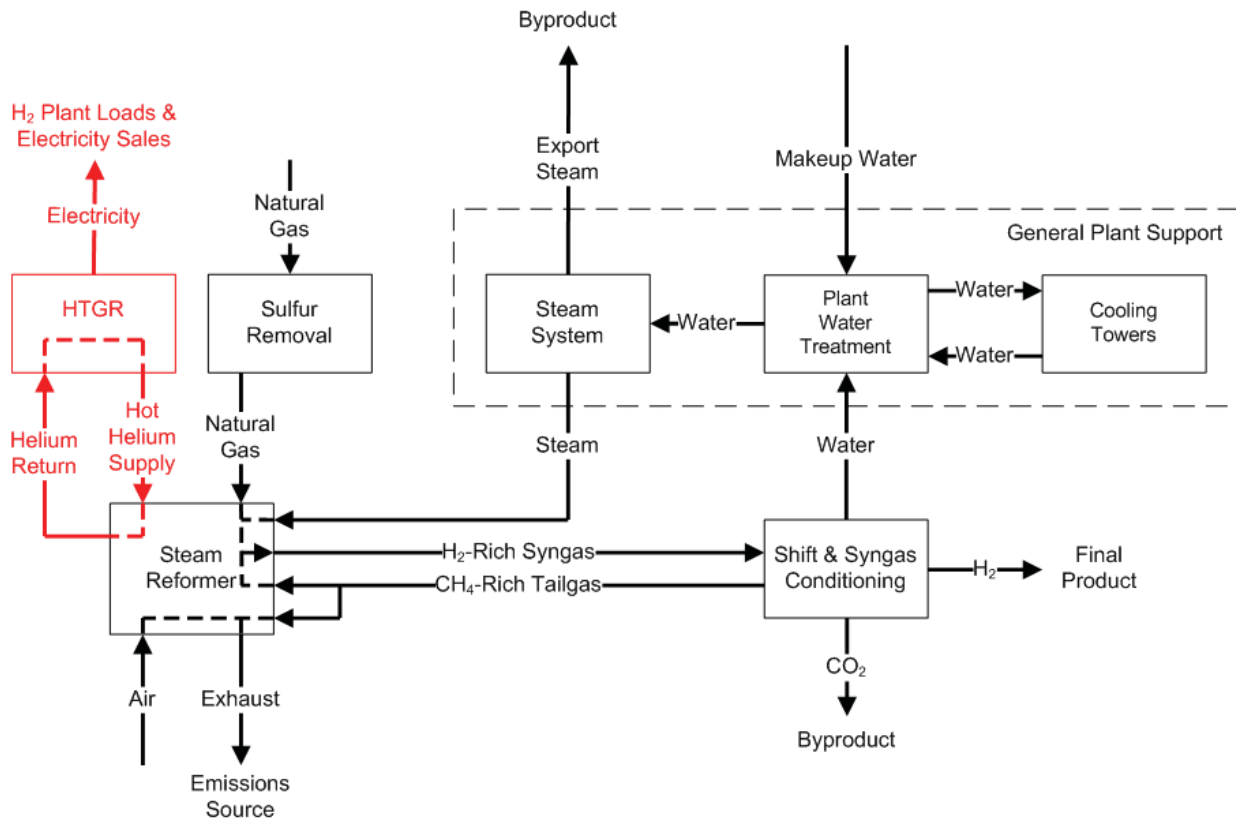


Figure 7. Block flow diagram for the HTGR-integrated SMR case with CC.

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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 (see Appendix E). Because many of the unit operations remain unchanged from the conventional SMR case with CC and the HTGR-integrated SMR case without CC, emphasis is placed on differences in configuration between the cases.

- **Sulfur Removal, Steam Reformer (REFORMER)** – Configuration of the reformer in the HTGR-integrated SMR case with CC is nearly identical to the configuration for the HTGR-integrated SMR case without CC. The principle difference is that for this case, a higher quality fuel gas is produced due to the removal of CO₂. This difference leads to the following process configuration changes:
 1. A temperature difference of 153°C (275°F) between the exhaust gas and the reformed syngas was specified. An adequate supply of fuel gas allows for a more realistic temperature approach in the second reforming stage.
 2. Excess fuel gas was recompressed and recycled with fresh natural gas feed to the reformer. This change was possible because a slight excess of fuel gas was available, and because this fuel gas contained primarily CH₄ and H₂. In this manner, hydrogen production can be maximized in relation to the amount of fresh natural gas feed.
 3. No H₂ product was used for hydrotreating in the reforming section of the plant. The hydrogen content in the recycled fuel gas was adequate to exceed a concentration of 2 mol% H₂ in the reformer feed stream (necessary for hydrotreating upstream of the sulfur removal bed). Hence, it was not necessary to divert a portion of the H₂ product for this purpose.
- **Shift & Syngas Conditioning (GAS-CLN)** – The shift and syngas conditioning operations and configuration in the HTGR-integrated SMR case with CC is identical to that in the conventional SMR case with CC.
- **Steam System (STEAM)** – The steam system in the HTGR-integrated SMR case with CC is identical to that in the conventional SMR case with CC. For this case, 176 million Btu/hr of 600-psig steam is available for export.

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- **Cooling Towers (COOL-TWR)** – The cooling towers in the HTGR-integrated SMR case with CC are identical to those in the conventional SMR case with CC.
- **Plant Water Treatment (H₂O-TRTM)** – The water treatment system in the HTGR-integrated SMR case with CC is identical to that in the conventional SMR case with CC.

4. PROCESS MODELING RESULTS AND OBSERVATIONS

Analysis of the SMR hydrogen production process indicates that it is beneficial to replace natural gas combustion in the reformer with heat supplied by an HTGR. Key results from the modeling are presented in Table 2, while high-level material and energy balances are presented in Figure 10. For the complete modeling results, see Appendix A. Aspen stream results for all cases are presented in electronic Appendixes B, C, D, and E.

From these results, including CC in the conventional flowsheet results in an overall process efficiency penalty of 4.7%. However, this penalty is manifest as a reduction in export steam generation capability—hydrogen product yield is actually slightly better for the flowsheet that includes CC.

Integration of nuclear heat can reduce the amount of natural gas required to produce a given quantity of hydrogen. If CC is not included in the design, these results indicate that natural gas consumption can be reduced by 15.3%. For a design that incorporates CC, the reduction is 11.6%. These results are reflected in the H₂ product yield, as shown in Figure 8.

Steam export is also increased as a result of integrating nuclear heat into the process. However, this is primarily due to the configuration of the reformer. In the design modeled here, nuclear heat is used to preheat many of the feed streams in the reforming island. This configuration makes a larger portion of heat in the reformer exhaust available for raising steam rather than preheating feed streams. An alternative could be considered to reduce both the export steam amount and the nuclear heat requirement. Reformer feed preheating (or a portion thereof) could be accomplished using heat from the exhaust stream rather than nuclear heat.

Power consumption in the conventional SMR process is quite modest. However, a small savings in power consumption is realized as a result of integrating nuclear heat into the process.

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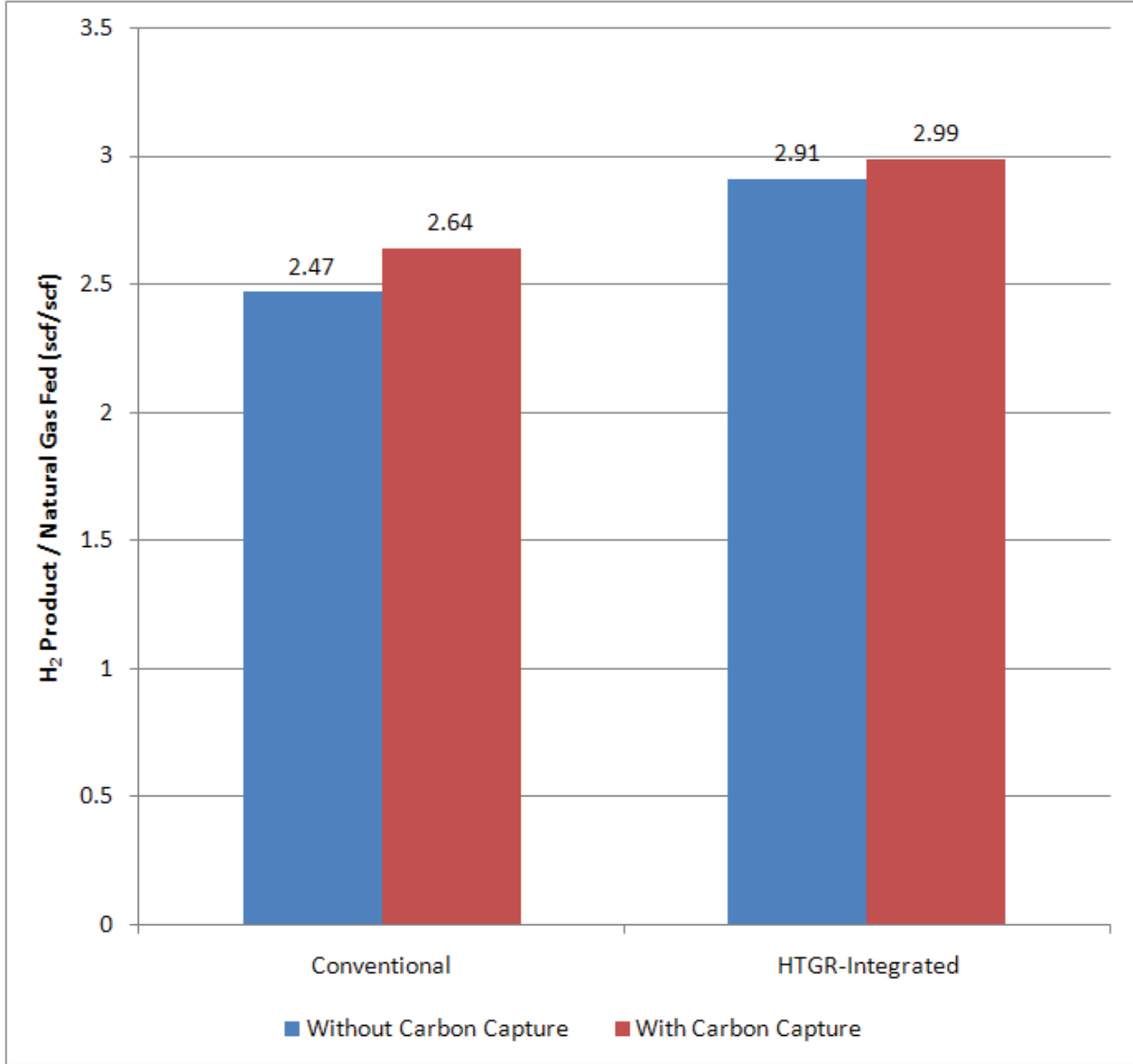


Figure 8. SMR H₂ product yield.

CO₂ emissions are also reduced as a result of integrating nuclear heat. If CC is not included in the design, nuclear integration results in a 15.4% reduction in CO₂ emissions. For a design that incorporates CC, the CO₂ emissions reduction is 38.6%. CO₂ disposition for each case is shown graphically in Figure 9.

As can be seen from these results, the nuclear requirements are quite modest— $\frac{1}{4}$ or less of a 600-MW_t reactor would be sufficient to support a world-scale single-train hydrogen plant (130 MMSCFD). A majority of the energy from the HTGR would be required as process heat, with only a small fraction required as electrical power.

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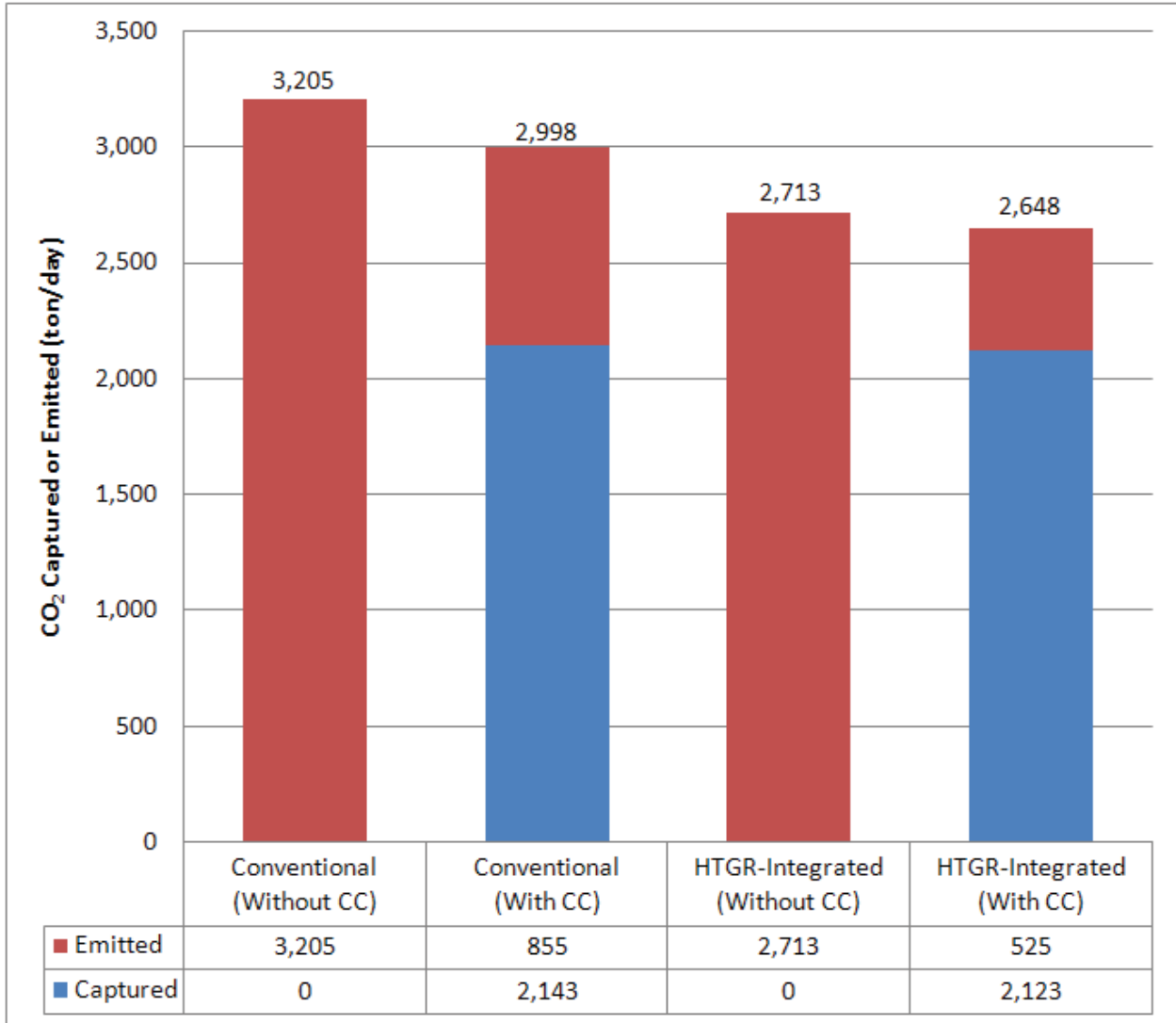


Figure 9. SMR CO₂ disposition.

Figure 10 shows that integration of an HTGR increases water consumption of the plant. However, this difference is primarily attributed to the increase in steam export for the HTGR-integrated processes. The water calculations do not currently assume that condensate will be returned from exported steam.

In addition to the results summarized above and in Table 2 and Figure 10, the following observations can be made:

- Heat transfer in a conventional reformer is largely accomplished via radiation due to the high temperatures resulting from combustion. In a HTGR-integrated scenario with a lower temperature heat source, a complete redesign of the reformer would be required to facilitate convective heat transfer rather than

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radiative heat transfer. One possible option would be to design the reformer similar to a shell and tube heat exchanger with catalyst packed in the tubes. A second option would be to design the reformer similar to a compact heat exchanger, again with catalyst packed in the tubes or coated on the tube surfaces.

2. Comparing a case without CC to a case with CC indicates that carbon removal from the fuel gas is beneficial (aside from the additional electrical power consumption required for CO₂ compression). In the conventional cases, CC produced a higher quality fuel gas. This in turn ensured that the fuel gas was capable of providing the heat required for reforming (i.e., it may be necessary to supplement the fuel gas with a small amount of natural gas in the conventional case without CC). In addition, carbon removal increases the amount of hydrogen that can be produced from a given amount of natural gas, albeit at the expense of producing additional steam for export. In the HTGR-integrated cases, a similar observation can be made—carbon removal increases hydrogen production slightly at the expense of producing additional export steam.

If the PSA system alone could be used to produce both pure hydrogen and CO₂, the benefits of CC could be realized without the need for additional equipment (i.e., an absorber system).

3. For the HTGR-integrated case without CC, the fuel gas contains a significant fraction of CO₂. The viable options for this stream are to burn it in the reformer to provide the required heat for reformation, or to burn it to raise steam or generate power.

For the HTGR-integrated case with CC, however, the fuel gas is relatively free of CO₂. Hence, a new option is available for this fuel gas stream—excess fuel gas (above and beyond what is required to supply reformation heat) can be recycled with the fresh natural gas to the reformer tubes. This option will maximize hydrogen production.

These options seem to indicate that the CC case could benefit substantially from raising the HTGR reactor outlet temperature (ROT). The increased temperature would shift more of the reforming duty from fuel gas combustion to the HTGR; hence, more fuel gas would be available for recycle to the reformer tubes to directly increase the hydrogen production rate. In contrast, for the case without CC, an increase in HTGR ROT would merely increase the amount of steam available for export.

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Table 2. SMR hydrogen production modeling case study results.

	Conventional (Without CC)	Conventional (With CC)	HTGR-Integrated (Without CC)	HTGR-Integrated (With CC)
Inputs				
Natural gas feed rate (MMSCFD) ¹	52.5	49.2	44.5	43.5
# 600-MWt HTGRs required ²	n/a	n/a	0.22	0.25
Outputs				
Hydrogen (MMSCFD) ¹	130	130	130	130
Steam export (MMBtu/hr)	189.2	23.6	273.2	176.2
600-psig steam	153.9	23.6	272.8	176.2
60-psig steam	35.3	0	0.4	0
Performance Metrics				
H ₂ product/natural gas feed (scf/scf)	2.47	2.64	2.91	2.99
Steam export/H ₂ product (lb/lb)	6.40	0.80	9.20	5.94
Process thermal efficiency, HHV basis (%) ²	82.5	77.8	85.0	80.3
Utility Usage				
Total Power (MW)	6.3	14.4	4.6	13.7
Natural gas reforming	3.8	3.3	1.7	2.0
Syngas purification	0.0	2.3	0.0	2.3
Steam system	0.3	0.2	0.4	0.4
CO ₂ compression	0.0	7.1	0.0	7.0
Cooling towers	0.1	0.1	0.1	0.1
Water treatment	2.1	1.4	2.4	2.0
Total Water (gpm) ³	943	650	1,114	958
Evaporation rate (gpm)	296	341	300	348
CO₂ Emissions				
Captured (ton/day CO ₂)	0	2,143	0	2,123
Emitted (ton/day CO ₂)	3,205	855	2,713	525
Nuclear Integration Summary				
Electricity demand (MW _E)	n/a	n/a	4.6	13.7
Process heat demand (MW _T)	n/a	n/a	119	117
Helium flow rate (ton/hr)	n/a	n/a	335	332
Helium supply temperature	n/a	n/a	700°C (1,292°F)	700°C (1,292°F)
Helium return temperature	n/a	n/a	428°C (803°F)	431°C (807°F)

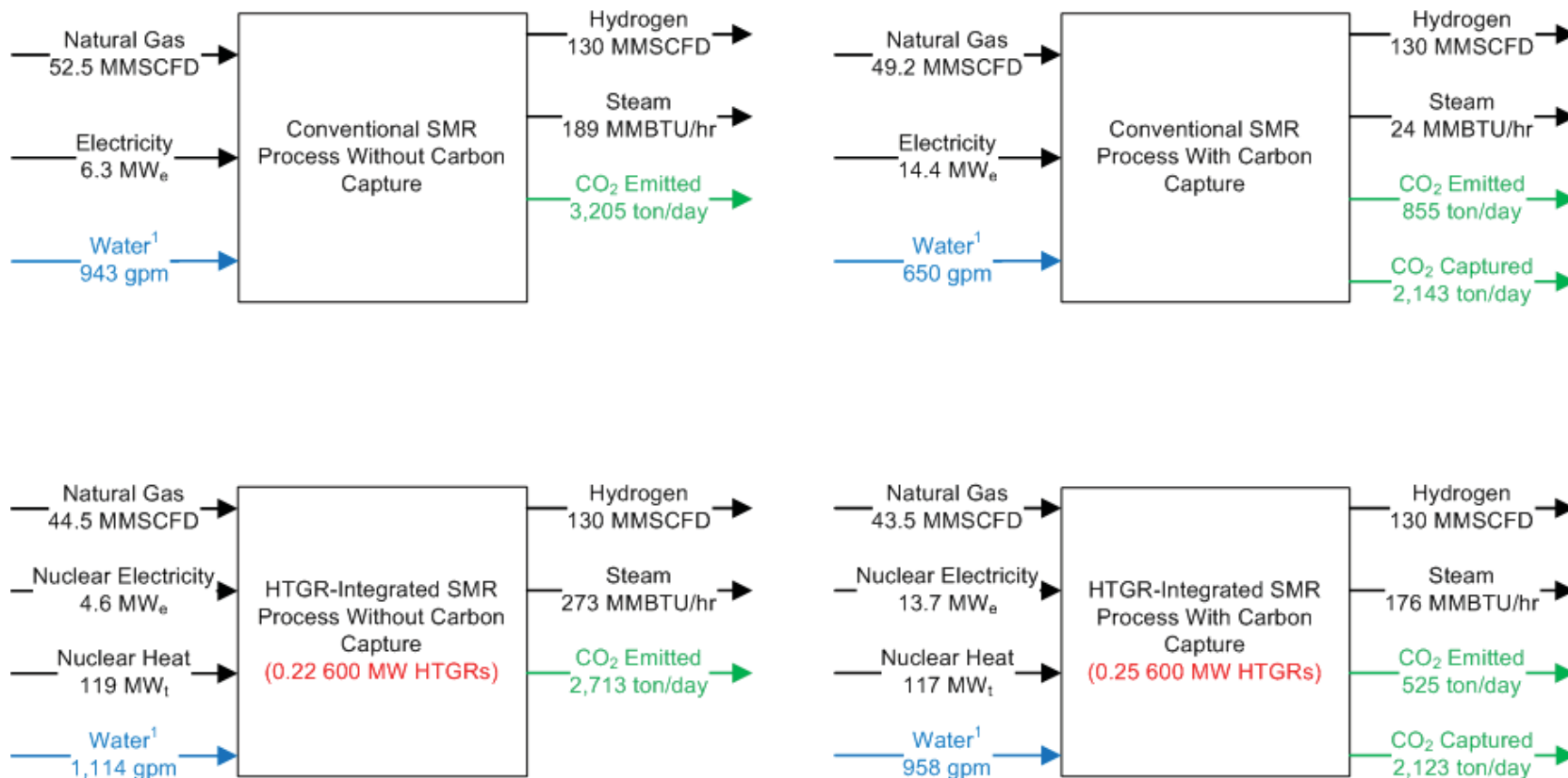
¹Standard temperature of 15.56°C (60°F).

²Assumes a power generation efficiency of 33% for fossil-based power production and 40% for HTGR-based power production.

³Does not include heat rejection for the nuclear plant.

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¹Does not include heat rejection requirement for the nuclear plant.

Figure 10. SMR hydrogen production modeling case material balance summary.

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5. FUTURE WORK AND RECOMMENDATIONS

The following items should be performed in the future to refine and build upon the work summarized in this report:

- A separate study should be conducted to assess the optimal siting of the HTGR with respect to the SMR facilities, balancing safety concerns associated with separation distance and heat losses associated with transporting high-temperature heat long distances.
- A rigorous Aspen Plus submodel of the HTGR 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.
- An economic analysis should be performed using the modeling results from this study as input. It is believed that such an analysis would further quantify the benefits of integrating HTGR heat with SMR technology for hydrogen production.
- It is likely that process results for the HTGR-integrated case could be improved if the HTGR temperature could be increased beyond 750°C (1,382°F). Hence, a study to quantify the performance improvement is recommended. Observations from this study indicate that CC should be included in the baseline configuration for the temperature sensitivity study.
- Additional work is warranted to scope out initial equipment design and further assess the feasibility of a HTGR-integrated SMR. As indicated in this report, the mode of heat transfer would shift from radiation in a conventional design to convection in a HTGR-integrated design.

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7. APPENDIXES

Appendix A, Detailed Modeling Results and Flowsheets

Appendix B, [Electronic] Conventional SMR Without CC Stream Results.xlsx

Appendix C, [Electronic] Conventional SMR With CC Stream Results.xlsx

Appendix D, [Electronic] HTGR-integrated SMR Without CC Stream Results.xlsx

Appendix E, [Electronic] HTGR-integrated SMR With CC Stream Results.xlsx

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Appendix A
Detailed Modeling Results and Flowsheets

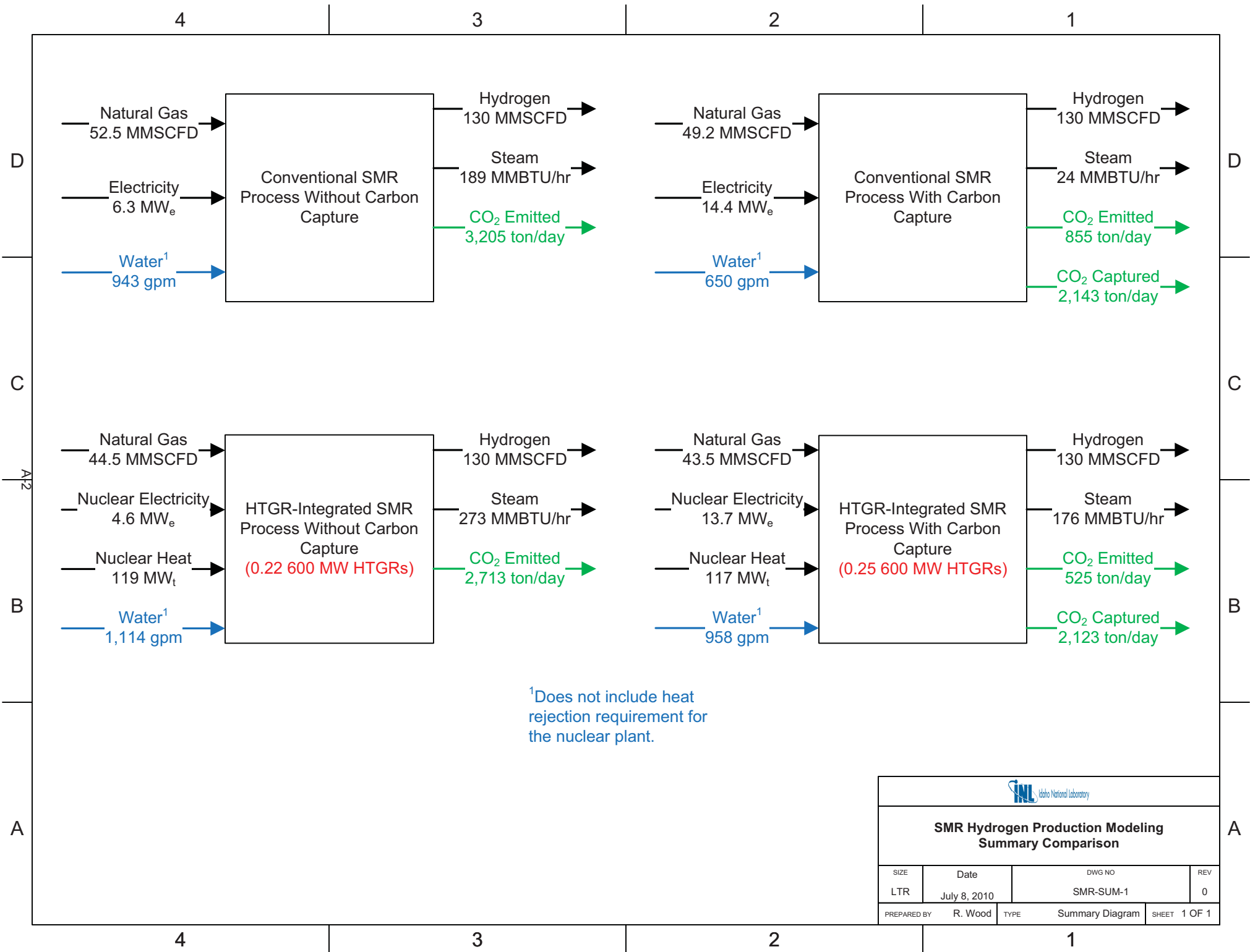
SMR Case Summary

	Conventional (Without CC)	Conventional (With CC)	Nuclear-Assisted (Without CC)	Nuclear-Assisted (With CC)
Inputs				
Natural Gas Feed Rate (MMSCFD) ¹	52.5	49.2	44.5	43.5
# 600 MWt HTGRs Required ²	n/a	n/a	0.22	0.25
Outputs				
Hydrogen (MMSCFD) ¹	130	130	130	130
Steam Export (MMBtu/hr)	189.2	23.6	273.2	176.2
600 psig Steam	153.9	23.6	272.8	176.2
60 psig Steam	35.3	0	0.4	0
Performance Metrics				
H ₂ Product / Natural Gas Fed (scf/scf)	2.47	2.64	2.91	2.99
Steam Export / H ₂ Product (lb/lb)	6.40	0.80	9.20	5.94
Process Thermal Efficiency, HHV Basis (%) ²	82.5	77.8	85.0	80.3
Utility Usage				
Total Power (MW)	6.3	14.4	4.6	13.7
Natural Gas Reforming	3.8	3.3	1.7	2.0
Syngas Purification	0.0	2.3	0.0	2.3
Steam System	0.3	0.2	0.4	0.4
CO ₂ Compression	0.0	7.1	0.0	7.0
Cooling Towers	0.1	0.1	0.1	0.1
Water Treatment	2.1	1.4	2.4	2.0
Total Water (gpm) ³	943	650	1,114	958
Evaporation Rate (gpm)	296	341	300	348
CO₂ Emissions				
Captured (ton/day CO ₂)	0	2,143	0	2,123
Emitted (ton/day CO ₂)	3,205	855	2,713	525
Nuclear Integration Summary				
Electricity Demand (MW _E)	n/a	n/a	4.6	13.7
Process Heat Demand (MW _T)	n/a	n/a	119	117
Helium Flow Rate (ton/hr)	n/a	n/a	335	332
Helium Supply Temperature (deg. F)	n/a	n/a	1292	1292
Helium Return Temperature (deg. F)	n/a	n/a	803	807

¹Standard temperature of 60 degrees F.

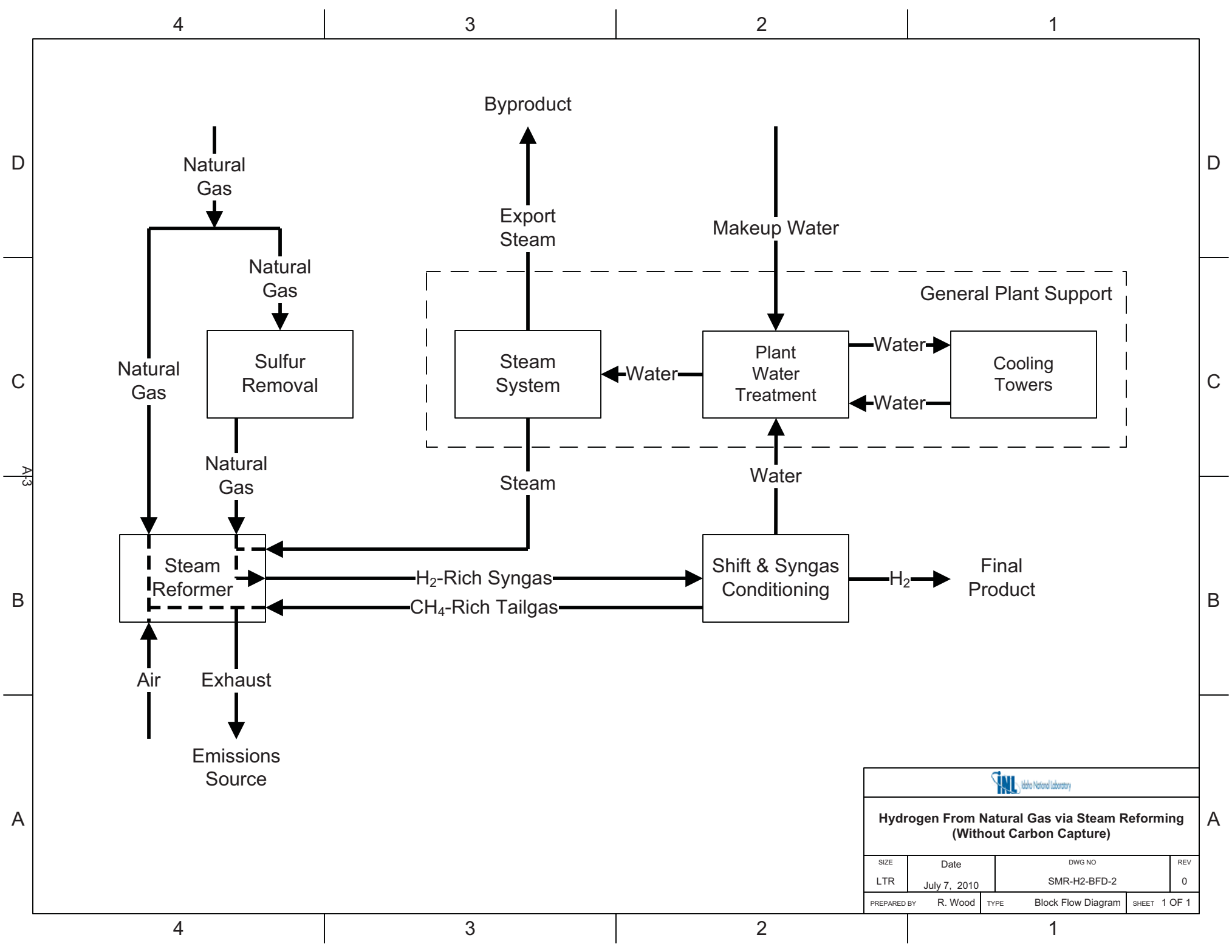
²Assumes a power generation efficiency of 33% for fossil-based power production and 40% for HTGR-based power production.


³Does not include heat rejection for the nuclear plant.

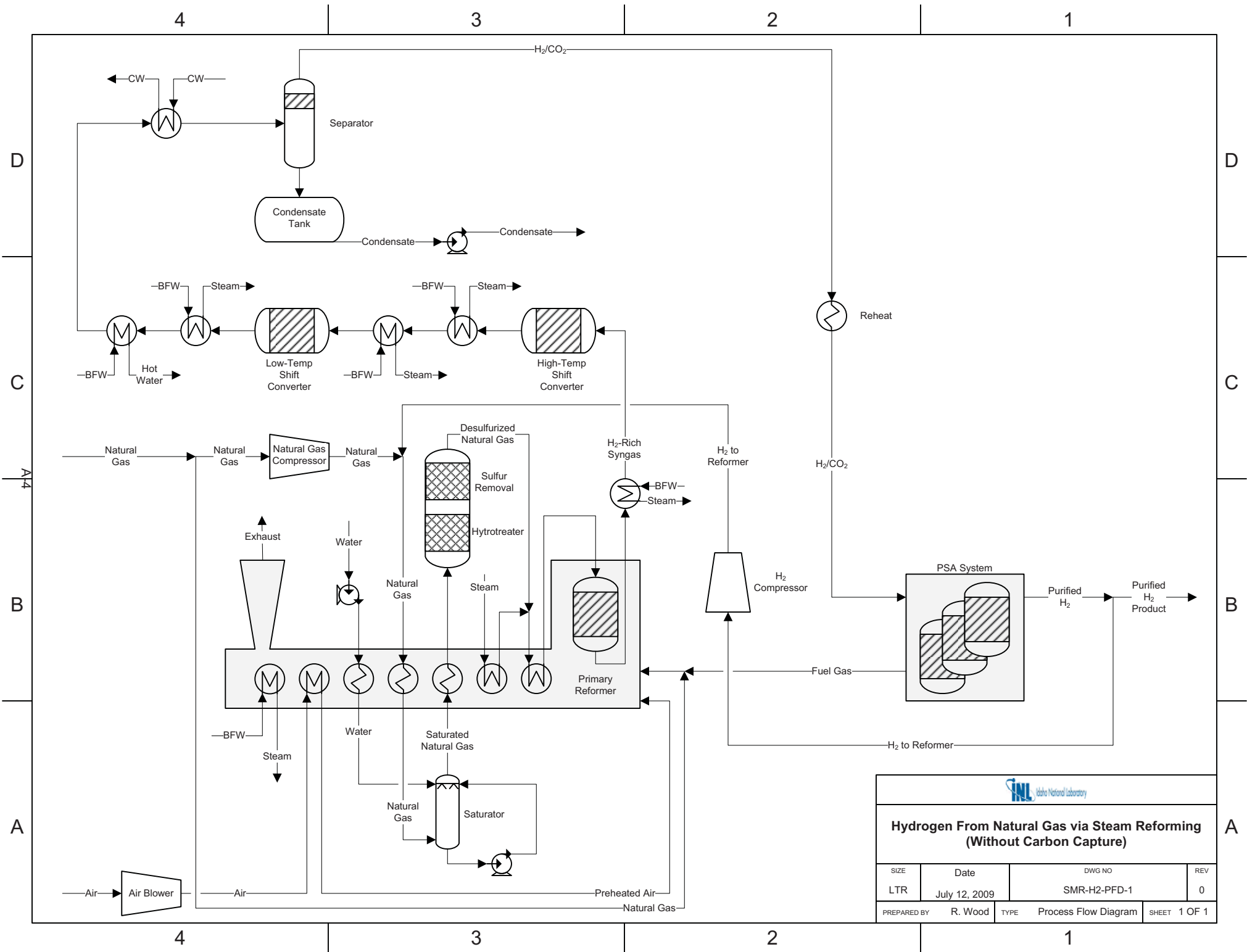


**SMR Hydrogen Production Modeling
Summary Comparison**

SIZE	Date	DWG NO	REV
LTR	July 8, 2010	SMR-SUM-1	0
PREPARED BY	R. Wood	TYPE	SHEET 1 OF 1
		Summary Diagram	



			
Hydrogen From Natural Gas via Steam Reforming (Without Carbon Capture)			
SIZE	Date	DWG NO	REV
LTR	July 7, 2010	SMR-H2-BFD-2	0
PREPARED BY	R. Wood	TYPE	Block Flow Diagram
			SHEET 1 OF 1



**Hydrogen From Natural Gas via Steam Reforming
(Without Carbon Capture)**

SIZE	Date	DWG NO	REV
LTR	July 12, 2009	SMR-H2-PFD-1	0
PREPARED BY	R. Wood	TYPE	Process Flow Diagram
		SHEET	1 OF 1

Conventional - Without Carbon Capture
 Calculator Block SUMMARY

FEED SUMMARY:

NATURAL GAS PROPERTIES:

MASS FLOW =	1189. TON/DY
VOLUME FLOW =	52.5 MMSCFD @ 60°F
HHV =	23063. BTU/LB
HHV =	1044. BTU/SCF @ 60°F
ENERGY FLOW =	54821. 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 =	362893. LB/HR
RAW SYNGAS VOLUME FLOW =	262. MMSCFD @ 60°F
RAW SYNGAS COMPOSITION:	
H2	47.7 MOL.%
CO	9.2 MOL.%
CO2	5.3 MOL.%
N2	0.2 MOL.%
H2O	34.2 MOL.%
CH4	3.5 MOL.%

FINAL PRODUCT SUMMARY:

HYDROGEN MASS FLOW =	29000. LB/HR
HYDROGEN VOLUME FLOW =	130. MMSCFD @ 60°F
HYDROGEN COMPOSITION:	
H2	99.9 MOL.%
N2	0.0 MOL.%
CH4	0.1 MOL.%
CO	11. PPMV
CO2	285. PPMV
H2O	0. PPMV

EXPORT HP STEAM FLOW =	150429. LB/HR
EXPORT HP STEAM CREDIT =	153.9 MMBTU/HR
STEAM PRESSURE =	615. PSIA
STEAM TEMPERATURE =	489. DEG. F
STEAM ENTHALPY (FOR CREDIT) =	1023. BTU/LB
CONDENSED TO LIQUID AT 1 ATM AT BOILING POINT	

EXPORT LP STEAM FLOW =	35212. LB/HR
EXPORT LP STEAM CREDIT =	35.3 MMBTU/HR
STEAM PRESSURE =	75. PSIA
STEAM TEMPERATURE =	307. DEG. F
STEAM ENTHALPY (FOR CREDIT) =	1002. BTU/LB
CONDENSED TO LIQUID AT 1 ATM AT BOILING POINT	

Conventional - Without Carbon Capture

POWER SUMMARY:

ELECTRICAL CONSUMERS:	
NG REFORMER POWER CONSUMPTION =	3.8 MW
GAS CLEANING POWER CONSUMPTION =	0.0 MW
STEAM SYSTEM POWER CONSUMPTION =	0.3 MW
COOLING TOWER POWER CONSUMPTION =	0.1 MW
WATER TREATMENT POWER CONSUMPTION =	2.1 MW
CONSUMER SUBTOTAL =	6.3 MW
NET PLANT POWER CONSUMPTION =	6.3 MW

WATER BALANCE:

EVAPORATIVE LOSSES:	
COOLING TOWER EVAPORATION =	135.9 GPM
ZLD SYSTEM EVAPORATION =	159.6 GPM
TOTAL EVAPORATIVE LOSSES =	295.5 GPM
WATER CONSUMED:	
BOILER FEED WATER MAKEUP =	904.4 GPM
COOLING TOWER MAKEUP =	143.1 GPM
TOTAL WATER CONSUMED =	1047.5 GPM
WATER GENERATED:	
GAS CLEANING CONDENSATE =	235.9 GPM
COOLING TOWER BLOWDOWN =	28.1 GPM
TOTAL WATER GENERATED =	264.1 GPM
PLANT WATER SUMMARY:	
NET MAKEUP WATER REQUIRED =	943.1 GPM
WATER CONSUMED / NG FED =	4.76 LB/LB

CO2 BALANCE:

CO2 EMITTED (TOTAL) =	3205. TON/DY
CO2 EMITTED (TOTAL) =	55. MMSCFD @ 60°F
CO2 EMITTED / NG FED =	2.70 LB/LB

EFFICIENCY CALCULATIONS:

OVERALL PROCESS THERMAL EFFICIENCY:
ASSUMPTIONS:
(1) FOR STEAM EXPORT - HEAT RECOVERY CREDIT
ASSUMES SATURATED LIQUID AT ATMOSPHERIC PRESSURE
(2) FOR ELECTRICITY IMPORT - ELECTRICAL
GENERATION EFFICIENCY OF 33%

HHV BASIS =	82.5%
LHV BASIS =	78.5%

OVERALL PROCESS COLD GAS EFFICIENCY:
(INCLUDES ONLY NATURAL GAS FEED AND HYDROGEN PRODUCT)

HHV BASIS =	76.6%
LHV BASIS =	71.8%

OTHER PERFORMANCE METRICS:

Conventional - Without Carbon Capture
HYDROGEN PRODUCT TO NATURAL GAS FEED = 2.47 SCF/SCF
(SHOULD BE AROUND 2.5)

STEAM EXPORT TO HYDROGEN PRODUCT = 6.40 TON/TON
(SHOULD BE BETWEEN 5 AND 18)

Calculator Block AIRPROPS

HUMIDITY DATA FOR STREAM PRI-AIR:
HUMIDITY RATIO = 43.5 GRAINS/LB
RELATIVE HUMIDITY = 39.0 %

Calculator Block NG-RFMR Hierarchy: REFORMER

SULFUR REMOVAL CONDITIONS:

INLET BED TEMPERATURE = 734. °F

PRIMARY REFORMER CONDITIONS:

INLET TEMPERATURE = 1292. °F
STEAM TO CARBON MOLAR RATIO = 3.00
NATURAL GAS BURNED FOR HEAT = 15.14 %
OUTLET TEMPERATURE = 1600. °F
METHANE CONVERSION = 78.1 %

Calculator Block GASCLEAN Hierarchy: GAS-CLN

WATER GAS SHIFT CONDITIONS:

HIGH-TEMPERATURE SHIFT REACTOR:

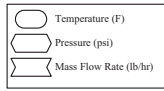
INLET TEMPERATURE = 635. °F
(MINIMUM ALLOWABLE = 600°F)
INLET STEAM:DRY GAS RATIO = 0.52
(MINIMUM ALLOWABLE = 0.4)
OUTLET TEMPERATURE = 769. °F
OUTLET CO (DRY BASIS) = 3.23 MOL.%

LOW-TEMPERATURE SHIFT REACTOR:

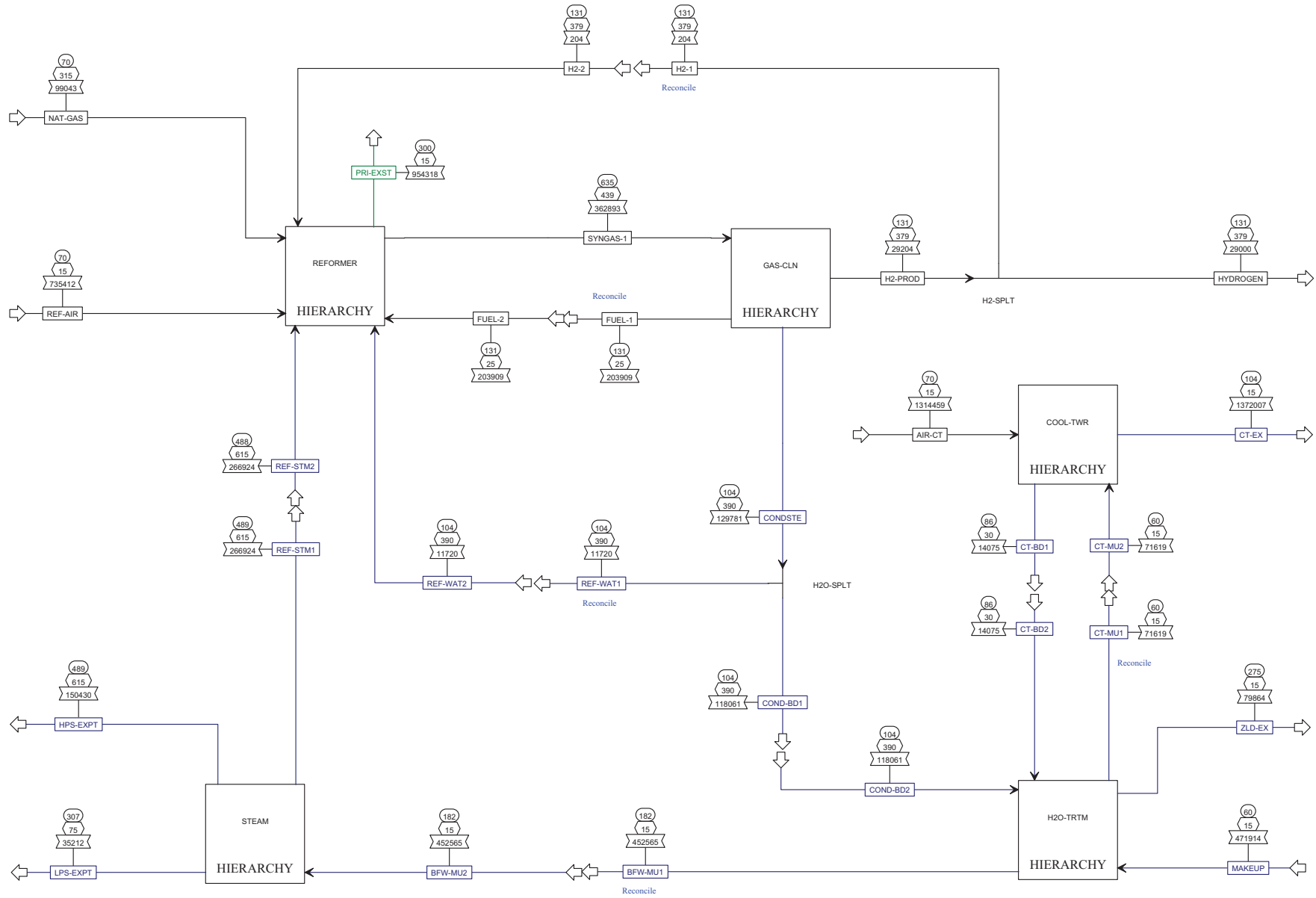
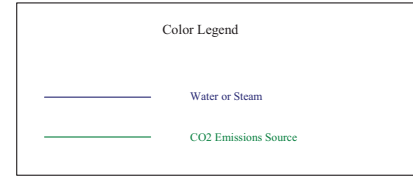
INLET DEW TEMPERATURE = 332. °F
INLET TEMPERATURE = 375. °F
(SHOULD BE AT LEAST 375°F)
(SHOULD BE AT LEAST 27°F ABOVE DEW TEMPERATURE)
INLET STEAM:DRY GAS RATIO = 0.38
(MINIMUM ALLOWABLE = 0.3)
OUTLET TEMPERATURE = 418. °F
OUTLET CO (DRY BASIS) = 0.35 MOL.%

PSA PERFORMANCE:

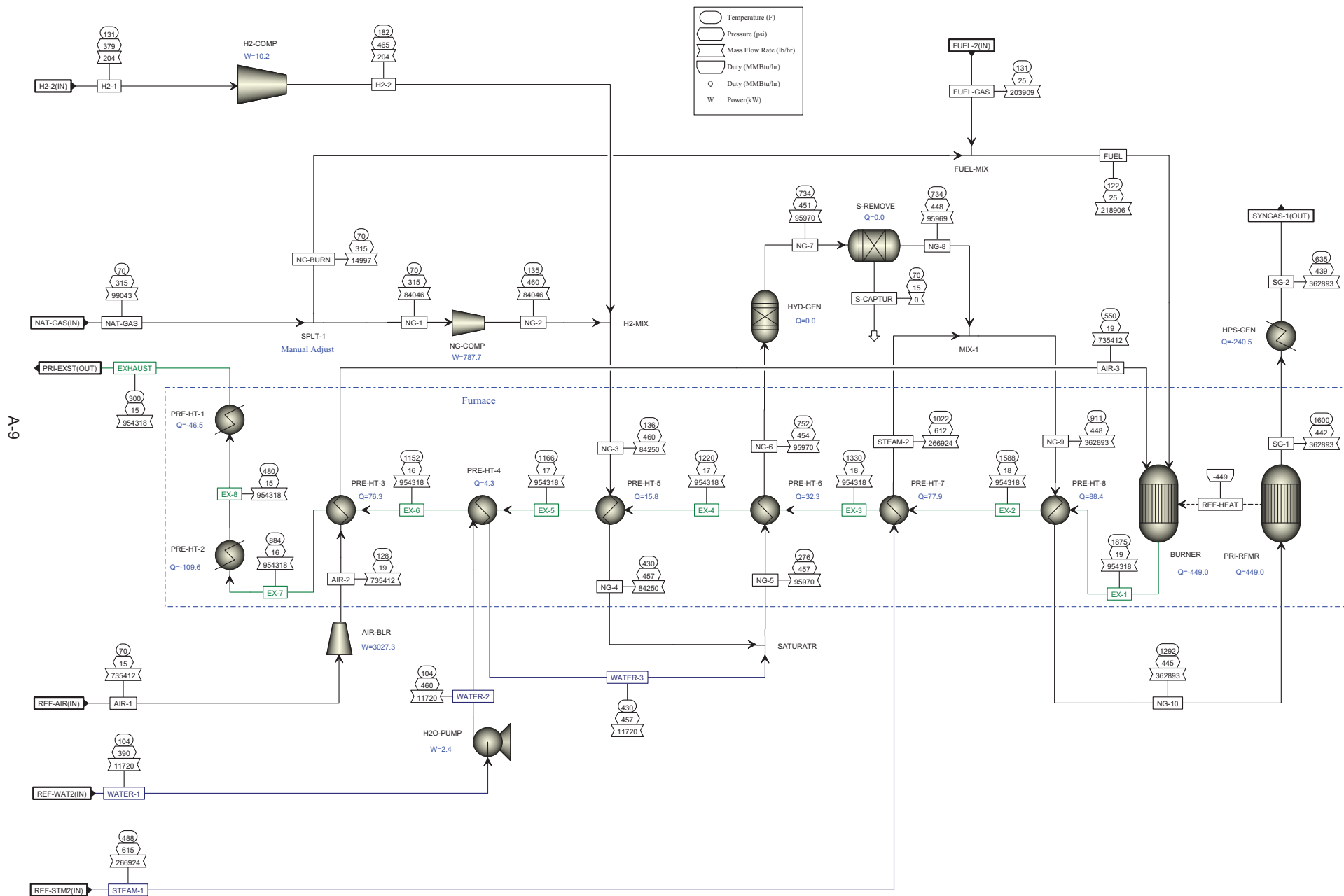
HYDROGEN RECOVERY = 88.0 %
HYDROGEN PURITY = 99.90 %



Hydrogen from Natural Gas (Without Carbon Capture)

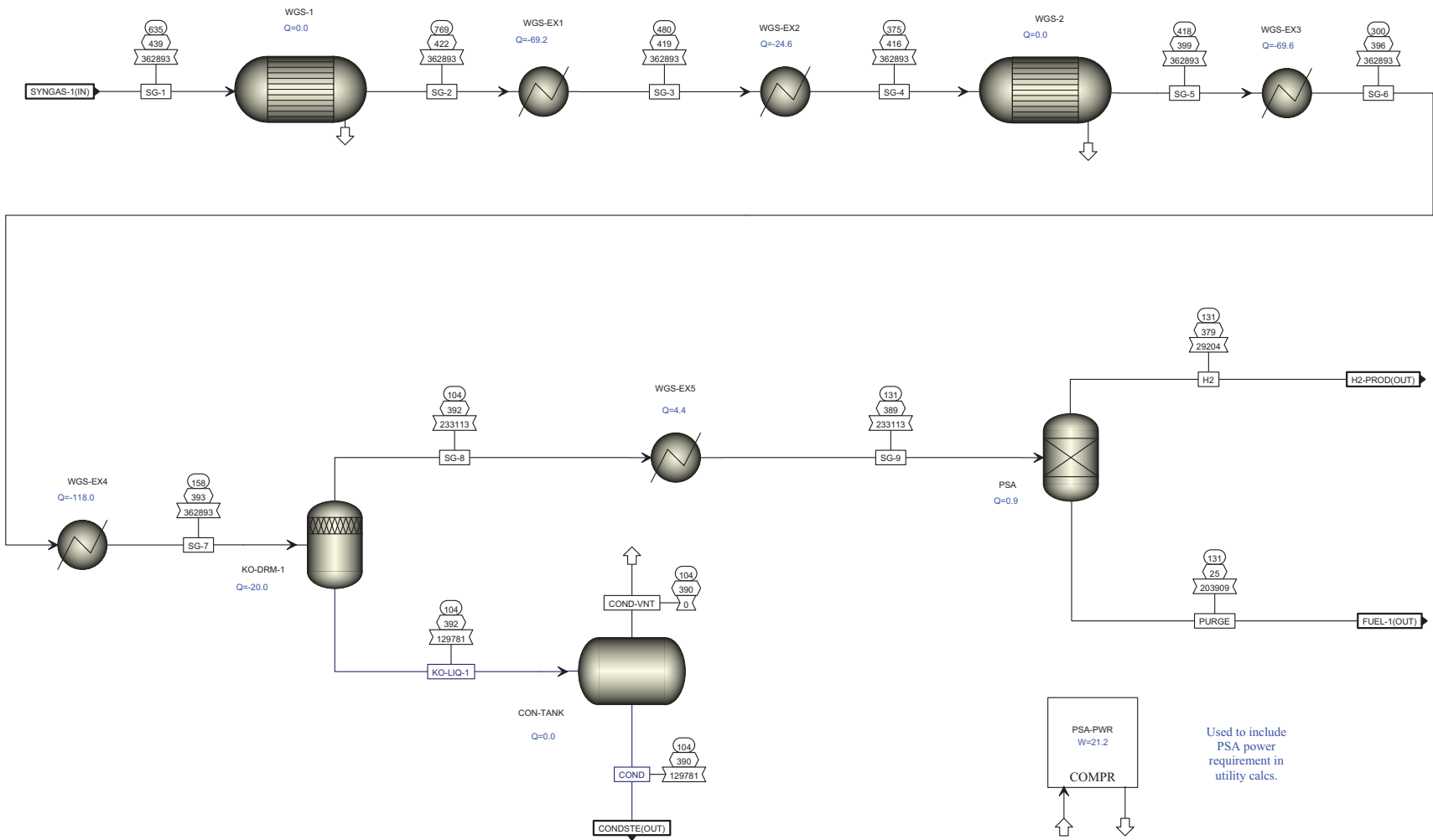
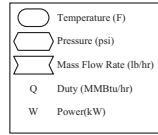


Natural Gas Steam Reforming



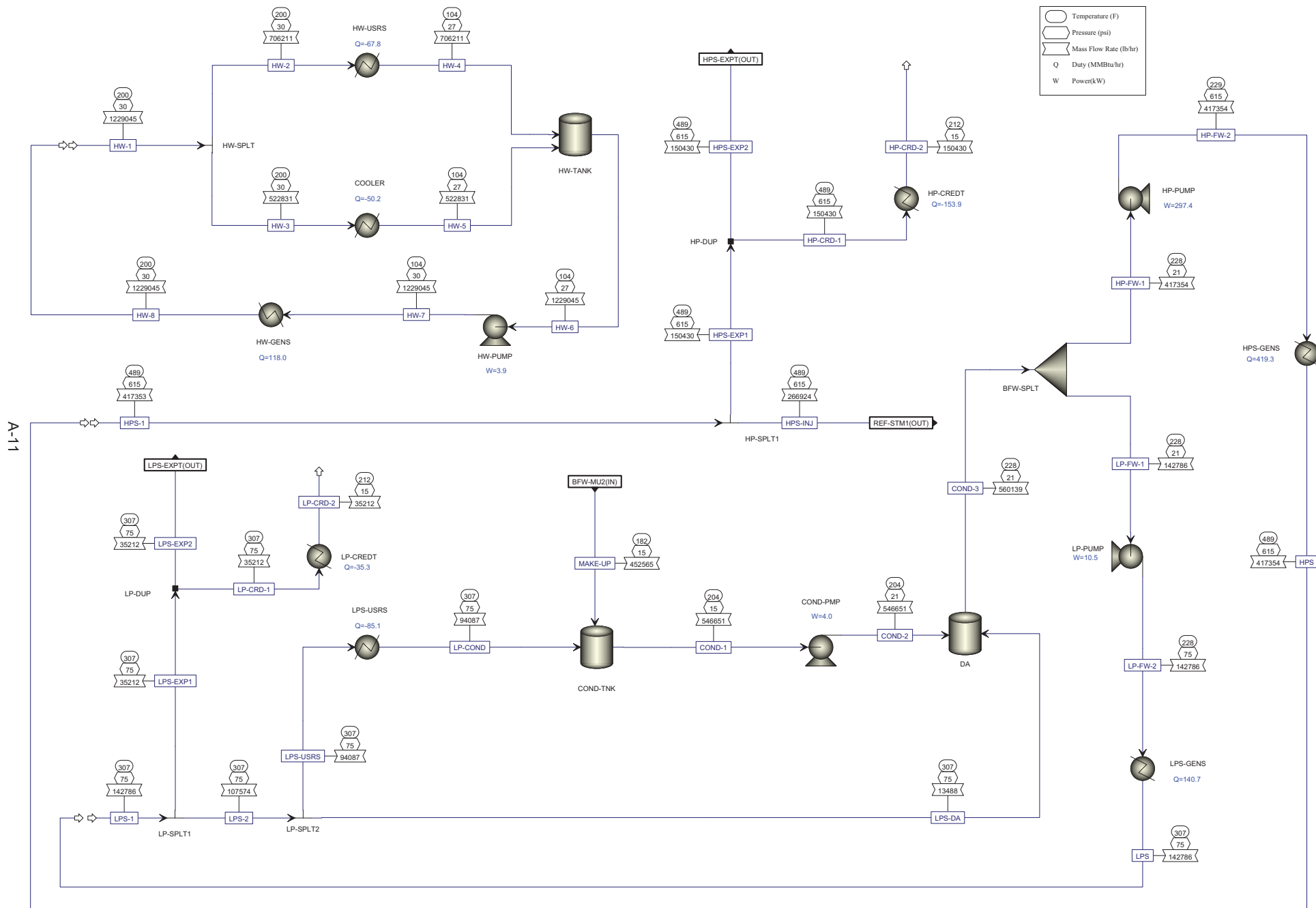
A-6

Syngas Cleaning & Conditioning

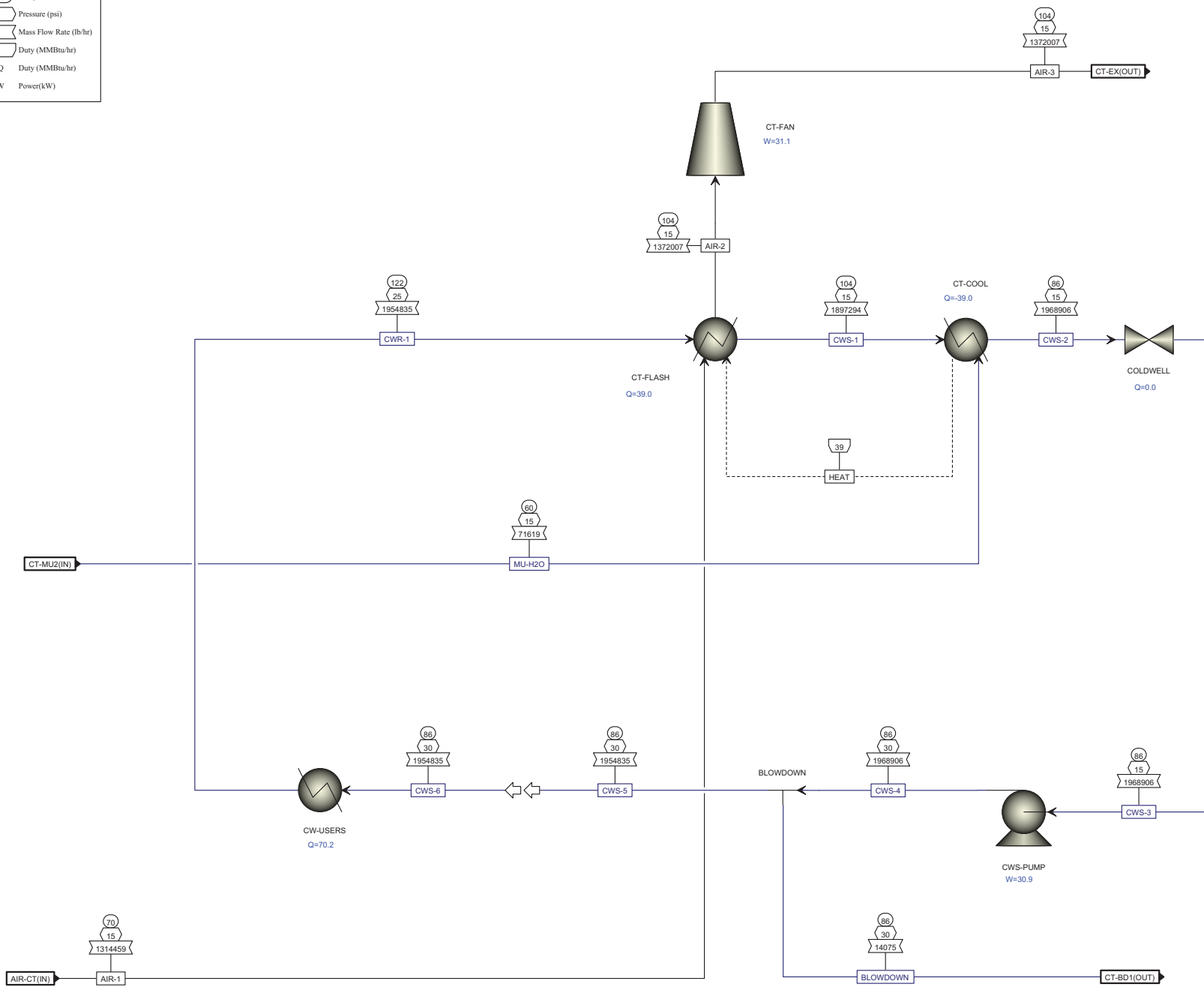
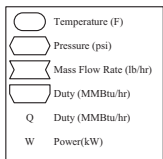


Used to include PSA power requirement in utility calcs.

Steam System

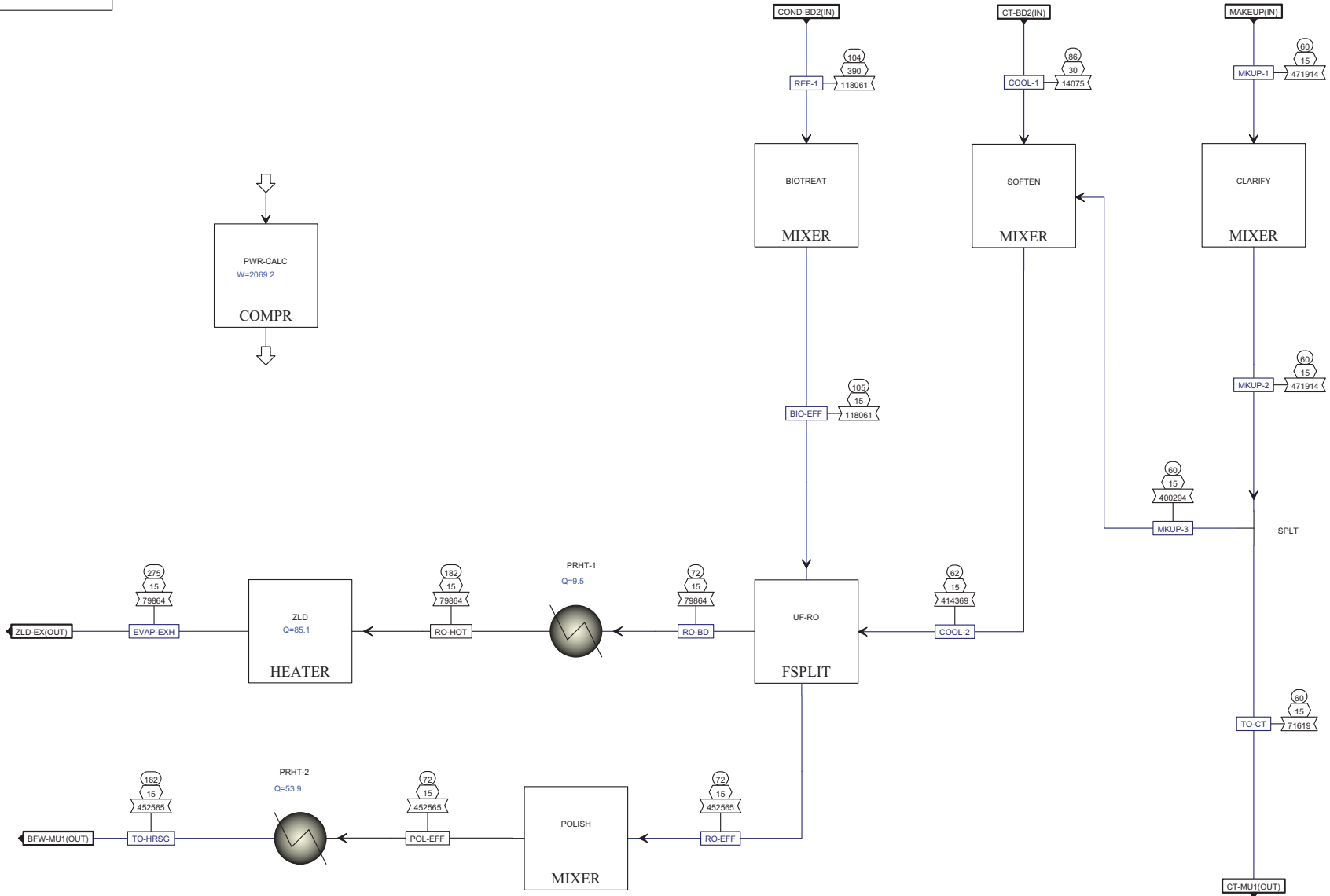
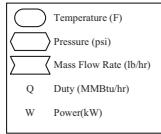


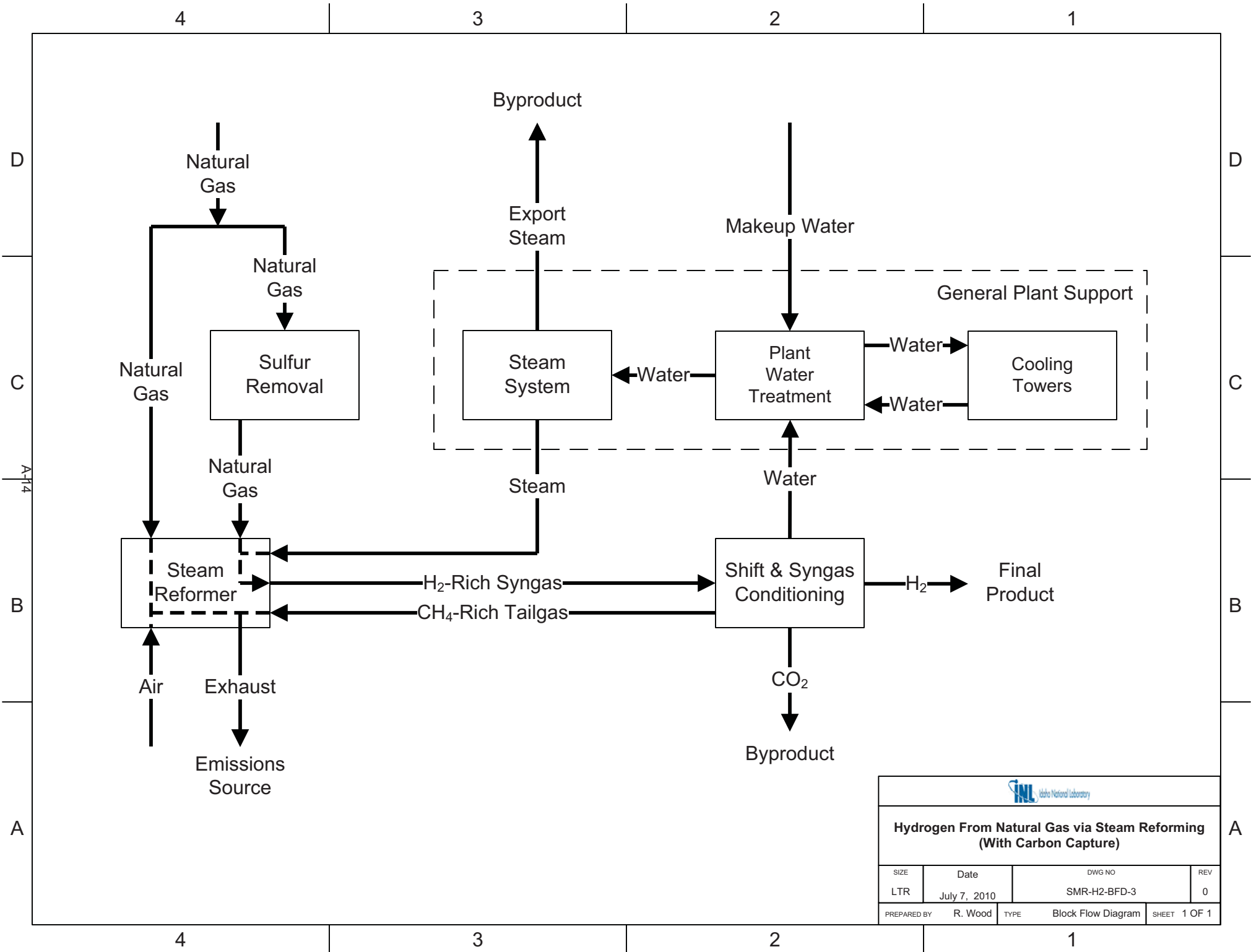
Cooling Tower




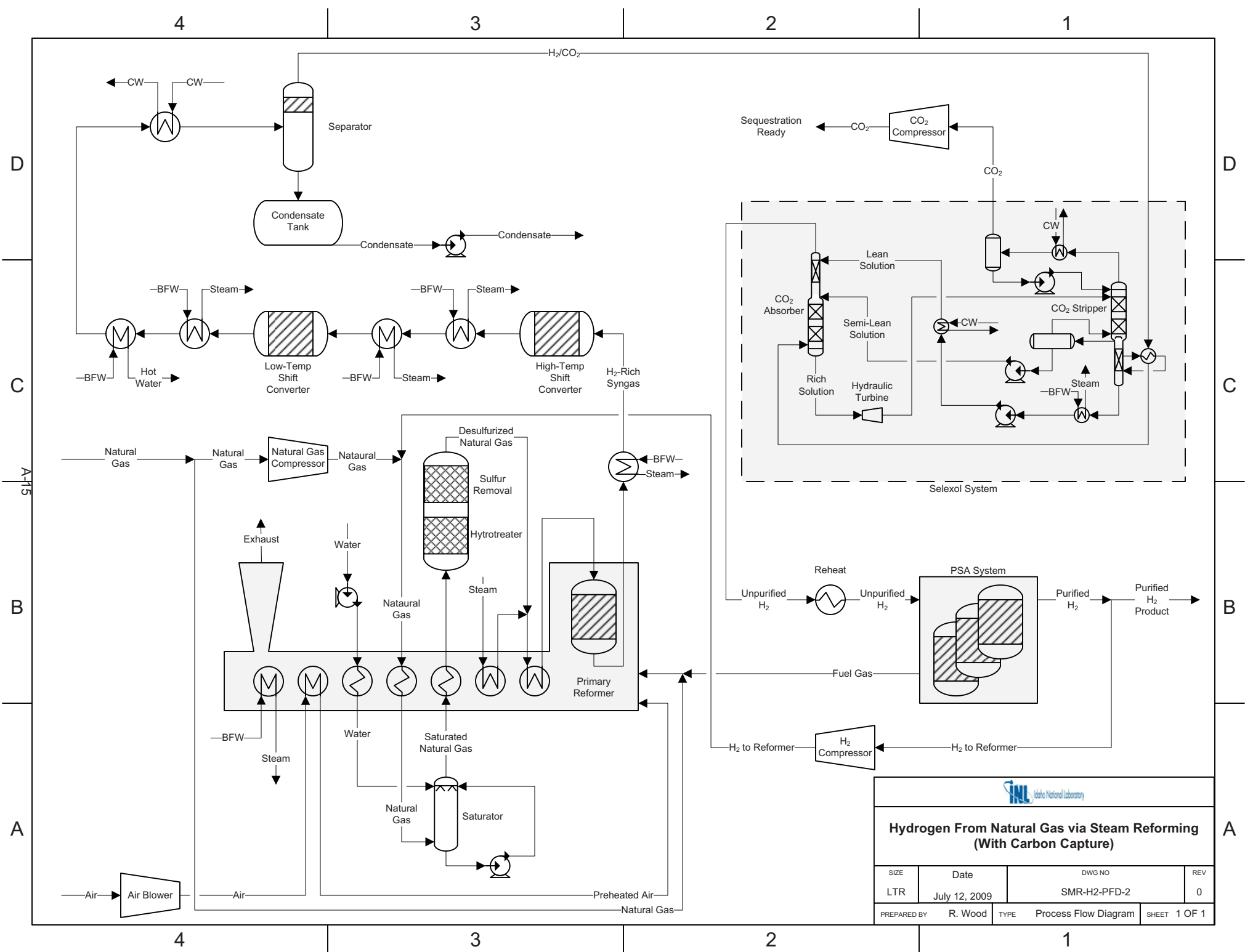
A-12


Simplified Water Treatment





 Hydrogen From Natural Gas via Steam Reforming (With Carbon Capture)			
SIZE	Date	DWG NO	REV
LTR	July 7, 2010	SMR-H2-BFD-3	0
PREPARED BY	R. Wood	TYPE	Block Flow Diagram
		SHEET	1 OF 1





**Hydrogen From Natural Gas via Steam Reforming
(With Carbon Capture)**

SIZE	Date	DWG NO	REV
LTR	July 12, 2009	SMR-H2-PFD-2	0
PREPARED BY	R. Wood	TYPE	Process Flow Diagram
		SHEET	1 OF 1

Conventional - with Carbon Capture
 Calculator Block SUMMARY

FEED SUMMARY:

NATURAL GAS PROPERTIES:

MASS FLOW =	1115. TON/DY
VOLUME FLOW =	49.2 MMSCFD @ 60°F
HHV =	23063. BTU/LB
HHV =	1044. BTU/SCF @ 60°F
ENERGY FLOW =	51411. 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 =	364878. LB/HR
RAW SYNGAS VOLUME FLOW =	263. MMSCFD @ 60°F
RAW SYNGAS COMPOSITION:	
H2	47.7 MOL.%
CO	9.2 MOL.%
CO2	5.3 MOL.%
N2	0.2 MOL.%
H2O	34.2 MOL.%
CH4	3.5 MOL.%

FINAL PRODUCT SUMMARY:

HYDROGEN MASS FLOW =	29000. LB/HR
HYDROGEN VOLUME FLOW =	130. MMSCFD @ 60°F
HYDROGEN COMPOSITION:	
H2	99.9 MOL.%
N2	0.0 MOL.%
CH4	0.1 MOL.%
CO	11. PPMV
CO2	3. PPMV
H2O	0. PPMV

EXPORT HP STEAM FLOW =	23097. LB/HR
EXPORT HP STEAM CREDIT =	23.6 MMBTU/HR
STEAM PRESSURE =	615. PSIA
STEAM TEMPERATURE =	489. DEG. F
STEAM ENTHALPY (FOR CREDIT) =	1023. BTU/LB
CONDENSED TO LIQUID AT 1 ATM AT BOILING POINT	

CARBON DIOXIDE MASS FLOW =	178876. LB/HR
CARBON DIOXIDE VOLUME FLOW =	37. MMSCFD @ 60°F
CARBON DIOXIDE COMPOSITION:	
CO2	99.1 MOL.%
H2	0.6 MOL.%
N2	0. PPMV
CH4	2458. PPMV

Conventional - with Carbon Capture

CO 187. PPMV
H2O 221. PPMV

POWER SUMMARY:

ELECTRICAL CONSUMERS:
NG REFORMER POWER CONSUMPTION = 3.3 MW
GAS CLEANING POWER CONSUMPTION = 2.3 MW
STEAM SYSTEM POWER CONSUMPTION = 0.2 MW
CO2 COMPRESSION POWER CONSUMPTION = 7.1 MW
COOLING TOWER POWER CONSUMPTION = 0.1 MW
WATER TREATMENT POWER CONSUMPTION = 1.4 MW
CONSUMER SUBTOTAL = 14.4 MW

NET PLANT POWER CONSUMPTION = 14.4 MW

WATER BALANCE:

EVAPORATIVE LOSSES:
COOLING TOWER EVAPORATION = 238.3 GPM
ZLD SYSTEM EVAPORATION = 102.8 GPM
TOTAL EVAPORATIVE LOSSES = 341.1 GPM

WATER CONSUMED:
BOILER FEED WATER MAKEUP = 582.5 GPM
COOLING TOWER MAKEUP = 250.9 GPM
TOTAL WATER CONSUMED = 833.5 GPM

WATER GENERATED:
GAS CLEANING CONDENSATE = 237.2 GPM
COOLING TOWER BLOWDOWN = 49.3 GPM
TOTAL WATER GENERATED = 286.5 GPM

PLANT WATER SUMMARY:
NET MAKEUP WATER REQUIRED = 649.7 GPM
WATER CONSUMED / NG FED = 3.50 LB/LB

CO2 BALANCE:

CO2 EMITTED (TOTAL) = 855. TON/DY
CO2 EMITTED (TOTAL) = 15. MMSCFD @ 60°F

SEQUESTRATION READY CO2 = 2143. TON/DY
SEQUESTRATION READY CO2 = 37. MMSCFD @ 60°F

CO2 EMITTED / NG FED = 0.77 LB/LB

EFFICIENCY CALCULATIONS:

OVERALL PROCESS THERMAL EFFICIENCY:
ASSUMPTIONS:
(1) FOR STEAM EXPORT - HEAT RECOVERY CREDIT
ASSUMES SATURATED LIQUID AT ATMOSPHERIC PRESSURE
(2) FOR ELECTRICITY IMPORT - ELECTRICAL
GENERATION EFFICIENCY OF 33%

HHV BASIS = 77.8%
LHV BASIS = 72.6%

OVERALL PROCESS COLD GAS EFFICIENCY:

Conventional - with Carbon Capture
(INCLUDES ONLY NATURAL GAS FEED AND HYDROGEN PRODUCT)

HHV BASIS = 82.1%
LHV BASIS = 76.9%

OTHER PERFORMANCE METRICS:

HYDROGEN PRODUCT TO NATURAL GAS FEED = 2.64 SCF/SCF
(SHOULD BE AROUND 2.5)

STEAM EXPORT TO HYDROGEN PRODUCT = 0.80 TON/TON
(SHOULD BE BETWEEN 5 AND 18)

calculator block AIRPROPS

HUMIDITY DATA FOR STREAM PRI-AIR:

HUMIDITY RATIO = 43.5 GRAINS/LB
RELATIVE HUMIDITY = 39.0 %

calculator block NG-RFMR Hierarchy: REFORMER

SULFUR REMOVAL CONDITIONS:

INLET BED TEMPERATURE = 734. °F

PRIMARY REFORMER CONDITIONS:

INLET TEMPERATURE = 1292. °F
STEAM TO CARBON MOLAR RATIO = 3.00
NATURAL GAS BURNED FOR HEAT = 9.02 %
OUTLET TEMPERATURE = 1600. °F
METHANE CONVERSION = 78.1 %

calculator block GASCLEAN Hierarchy: GAS-CLN

WATER GAS SHIFT CONDITIONS:

HIGH-TEMPERATURE SHIFT REACTOR:

INLET TEMPERATURE = 635. °F
(MINIMUM ALLOWABLE = 600°F)
INLET STEAM:DRY GAS RATIO = 0.52
(MINIMUM ALLOWABLE = 0.4)
OUTLET TEMPERATURE = 769. °F
OUTLET CO (DRY BASIS) = 3.23 MOL.%

LOW-TEMPERATURE SHIFT REACTOR:

INLET DEW TEMPERATURE = 332. °F
INLET TEMPERATURE = 375. °F
(SHOULD BE AT LEAST 375°F)
(SHOULD BE AT LEAST 27°F ABOVE DEW TEMPERATURE)
INLET STEAM:DRY GAS RATIO = 0.38
(MINIMUM ALLOWABLE = 0.3)
OUTLET TEMPERATURE = 418. °F

Conventional - with Carbon Capture

OUTLET CO (DRY BASIS) = 0.35 MOL.%

SELEXOL PERFORMANCE:

INLET CO₂ CONCENTRATION = 18.9 %
OUTLET CO₂ CONCENTRATION = 0.2 %

CO₂ CAPTURE = 99.0 %
CO₂ PURITY = 99.1 %

SELEXOL UTILITY REQUIREMENTS:

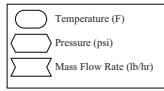
ELECTRICITY USAGE = 2.3 MW
STEAM USAGE = 72959. LB/HR

CO₂ COMPRESSION:

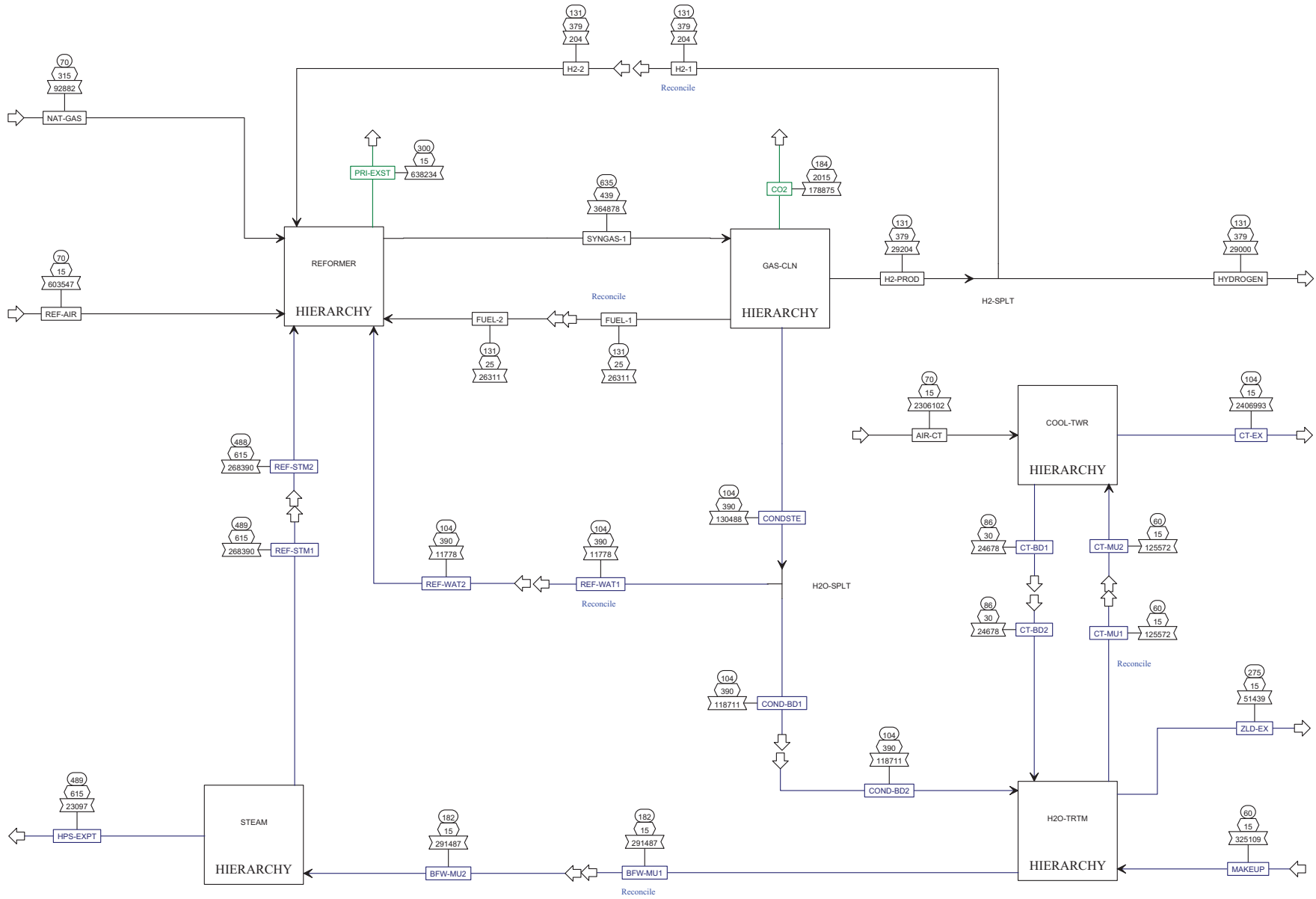
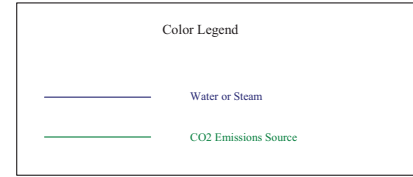
PRESSURE = 2015. PSIA
ELECTRICITY USAGE = 7.1 MW

PSA PERFORMANCE:

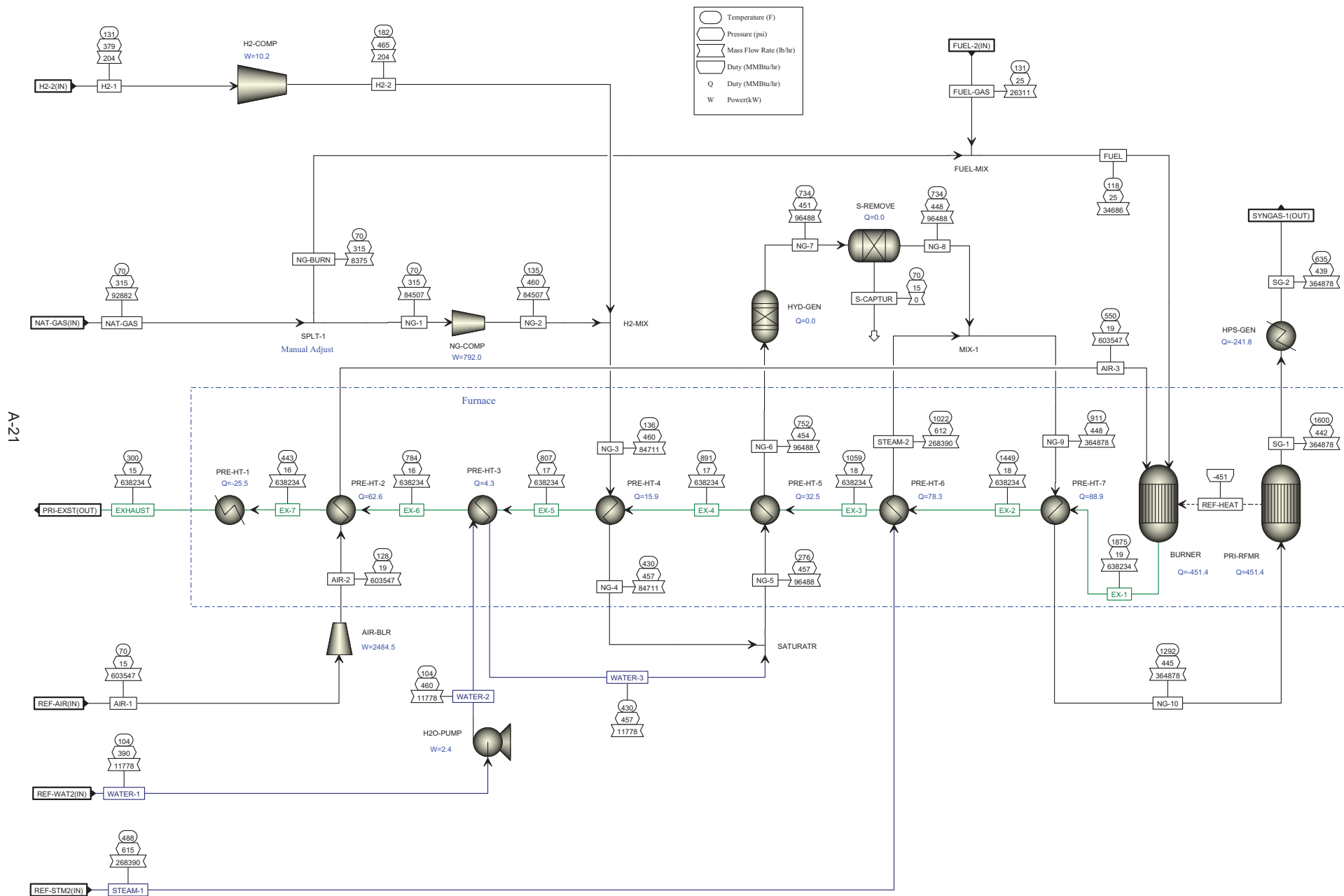
HYDROGEN RECOVERY = 88.0 %
HYDROGEN PURITY = 99.90 %



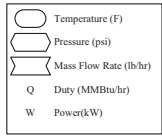
Hydrogen from Natural Gas (With Carbon Capture)



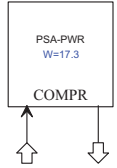
Natural Gas Steam Reforming



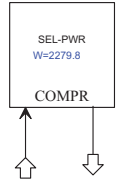
Syngas Cleaning & Conditioning



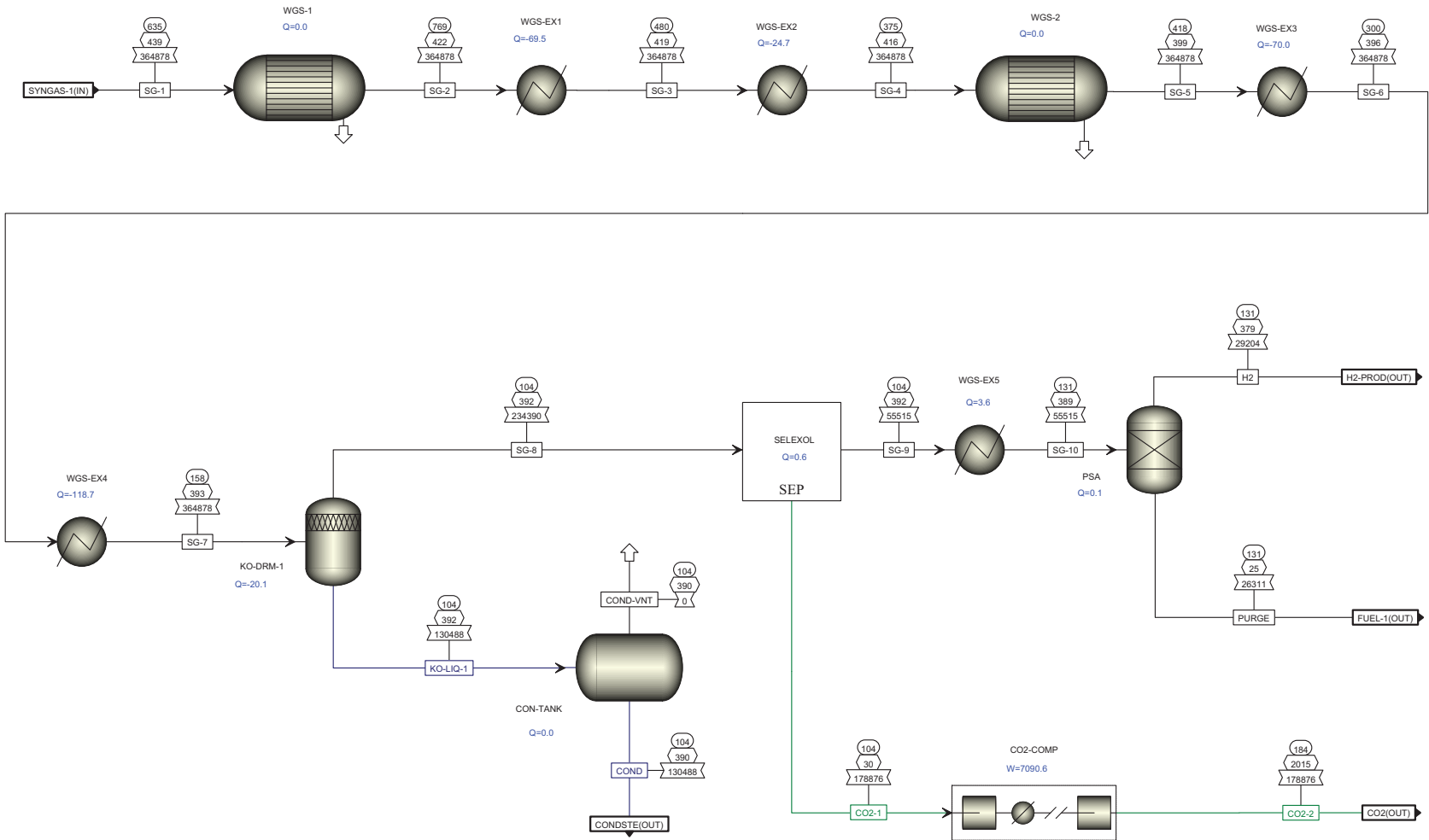
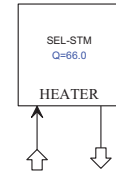
Used to include
PSA power
requirement in
utility calcs.



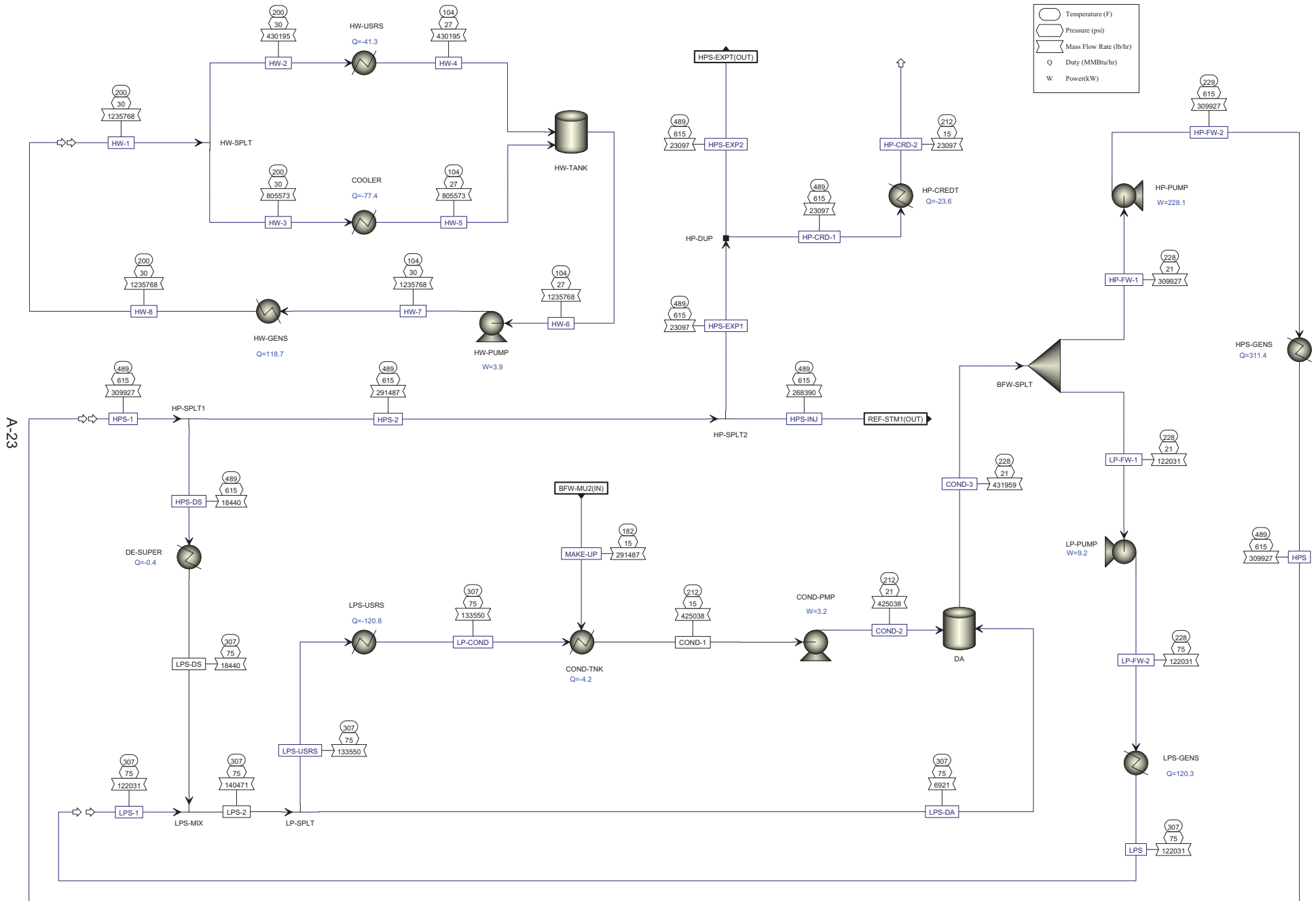
Used to include
Selexol power
requirement in
utility calcs.



Used to include
Selexol steam
usage in utility
calcs.

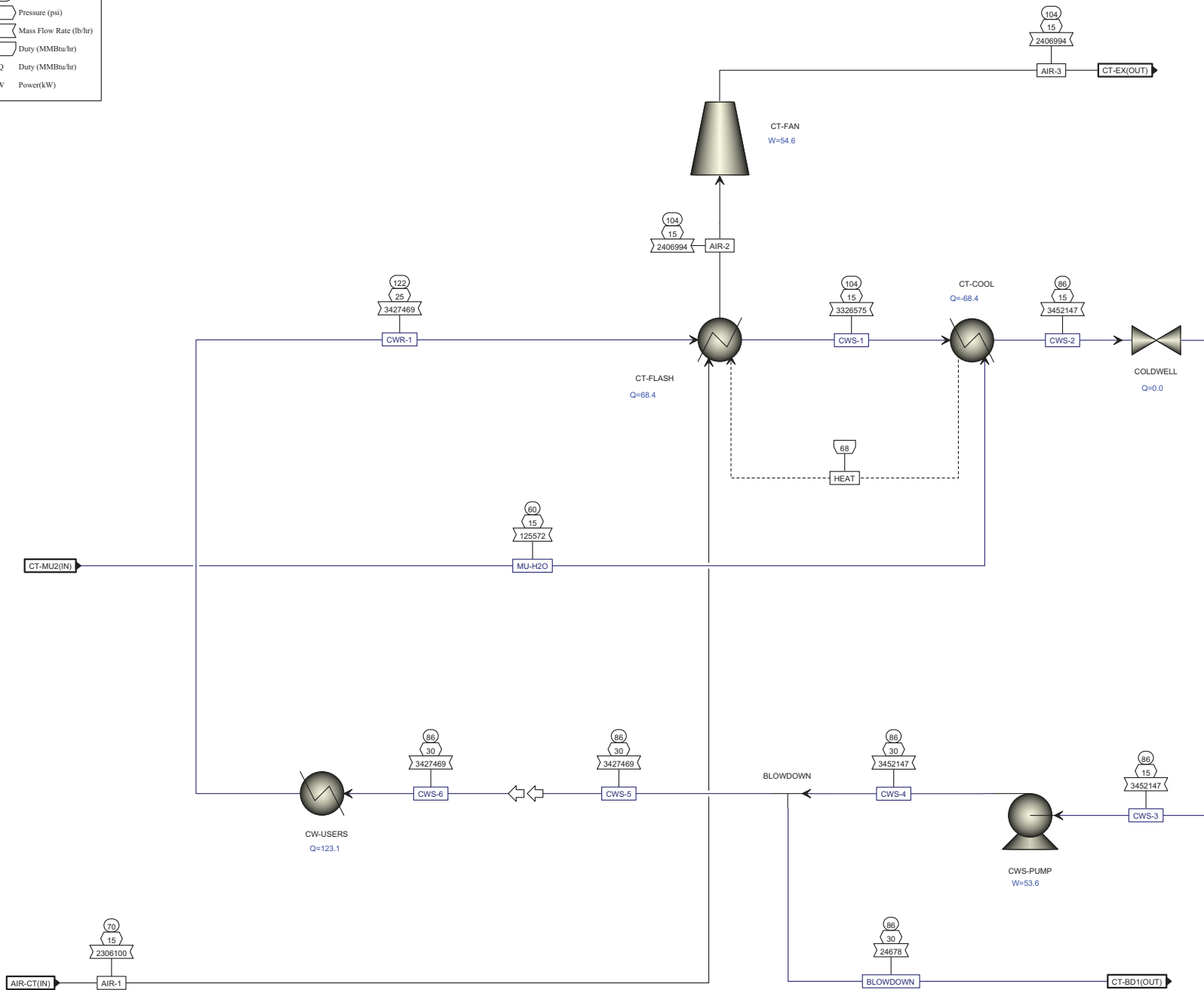
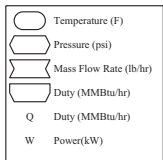


Steam System

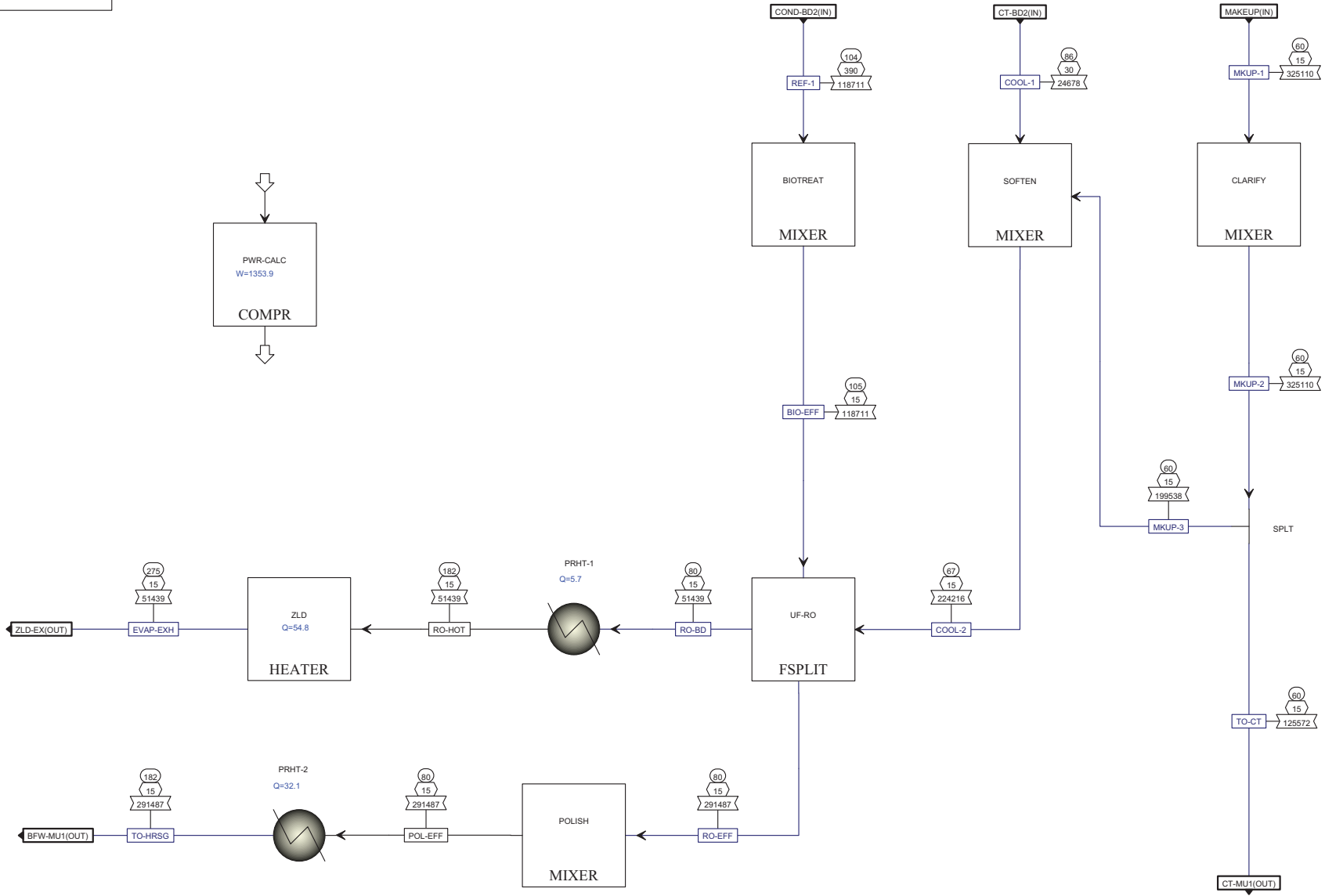
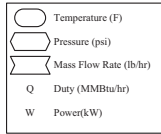


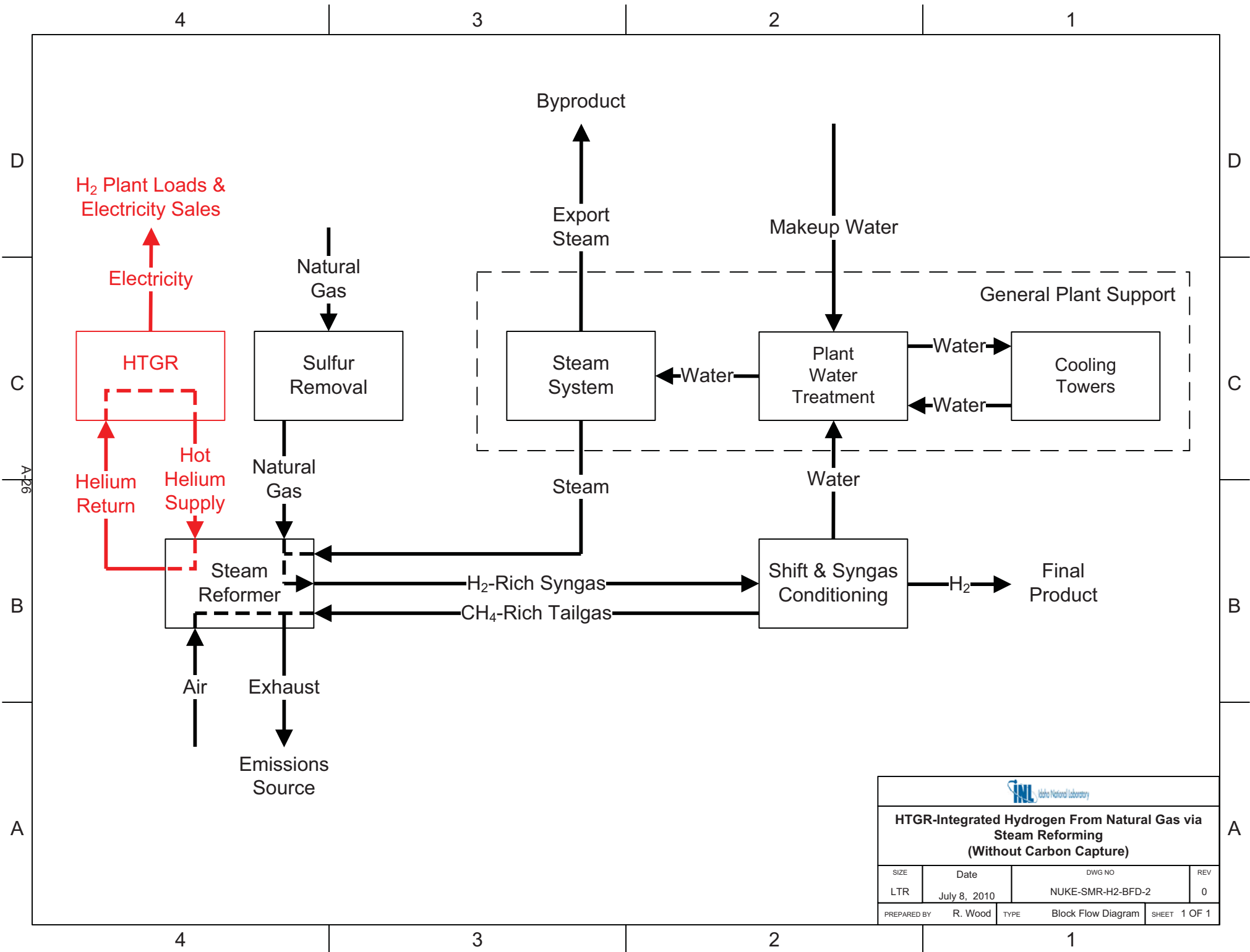
A-23


Cooling Tower

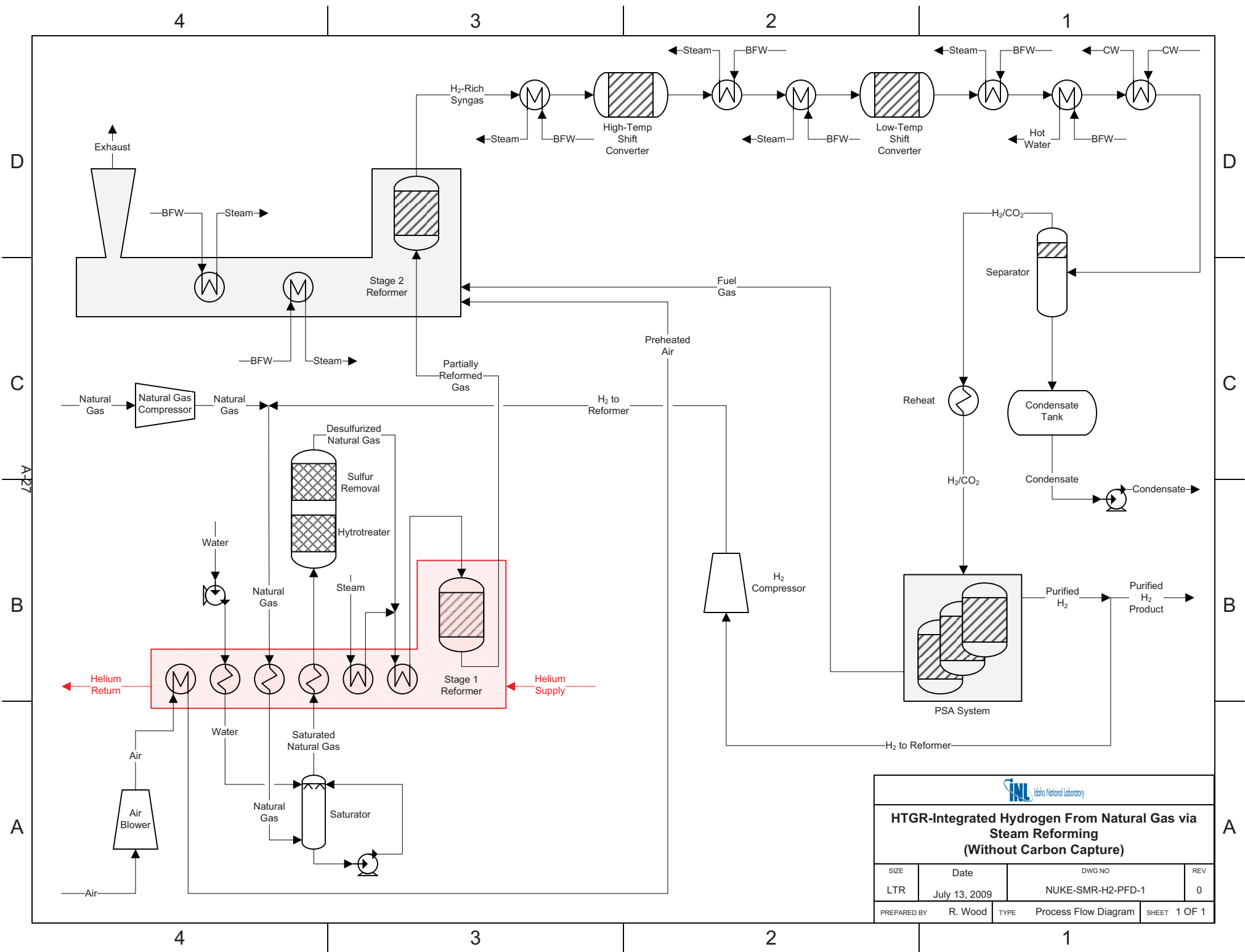



Simplified Water Treatment





 Idaho National Laboratory			
HTGR-Integrated Hydrogen From Natural Gas via Steam Reforming (Without Carbon Capture)			
SIZE	Date	DWG NO	REV
LTR	July 8, 2010	NUKE-SMR-H2-BFD-2	0
PREPARED BY	R. Wood	TYPE	Block Flow Diagram
			SHEET 1 OF 1



 HTGR-Integrated Hydrogen From Natural Gas via Steam Reforming (Without Carbon Capture)			
SIZE	Date	DWG NO	REV
LTR	July 13, 2009	NUKE-SMR-H2-PFD-1	0
PREPARED BY	R. Wood	TYPE	Process Flow Diagram
		SHEET	1 OF 1

HTGR-Integrated - without CC

Calculator Block SUMMARY

FEED SUMMARY:

NATURAL GAS PROPERTIES:

MASS FLOW =	1007. TON/DY
VOLUME FLOW =	44.5 MMSCFD @ 60°F
HHV =	23063. BTU/LB
HHV =	1044. BTU/SCF @ 60°F
ENERGY FLOW =	46442. 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 =	362281. LB/HR
RAW SYNGAS VOLUME FLOW =	261. MMSCFD @ 60°F
RAW SYNGAS COMPOSITION:	
H2	47.7 MOL.%
CO	9.2 MOL.%
CO2	5.3 MOL.%
N2	0.2 MOL.%
H2O	34.1 MOL.%
CH4	3.5 MOL.%

FINAL PRODUCT SUMMARY:

HYDROGEN MASS FLOW =	29000. LB/HR
HYDROGEN VOLUME FLOW =	130. MMSCFD @ 60°F
HYDROGEN COMPOSITION:	
H2	99.9 MOL.%
N2	0.0 MOL.%
CH4	0.1 MOL.%
CO	11. PPMV
CO2	285. PPMV
H2O	0. PPMV

EXPORT HP STEAM FLOW =	266567. LB/HR
EXPORT HP STEAM CREDIT =	272.8 MMBTU/HR
STEAM PRESSURE =	615. PSIA
STEAM TEMPERATURE =	489. DEG. F
STEAM ENTHALPY (FOR CREDIT) =	1023. BTU/LB
CONDENSED TO LIQUID AT 1 ATM AT BOILING POINT	

EXPORT LP STEAM FLOW =	370. LB/HR
EXPORT LP STEAM CREDIT =	0.4 MMBTU/HR
STEAM PRESSURE =	75. PSIA
STEAM TEMPERATURE =	307. DEG. F
STEAM ENTHALPY (FOR CREDIT) =	1002. BTU/LB
CONDENSED TO LIQUID AT 1 ATM AT BOILING POINT	

HTGR-Integrated - without CC

POWER SUMMARY:

ELECTRICAL CONSUMERS:

NG REFORMER POWER CONSUMPTION =	1.7 MW
GAS CLEANING POWER CONSUMPTION =	0.0 MW
STEAM SYSTEM POWER CONSUMPTION =	0.4 MW
COOLING TOWER POWER CONSUMPTION =	0.1 MW
WATER TREATMENT POWER CONSUMPTION =	2.4 MW
CONSUMER SUBTOTAL =	4.6 MW
NET PLANT POWER CONSUMPTION =	4.6 MW

WATER BALANCE:

EVAPORATIVE LOSSES:

COOLING TOWER EVAPORATION =	112.3 GPM
ZLD SYSTEM EVAPORATION =	188.1 GPM
TOTAL EVAPORATIVE LOSSES =	300.4 GPM

WATER CONSUMED:

BOILER FEED WATER MAKEUP =	1066.0 GPM
COOLING TOWER MAKEUP =	118.3 GPM
TOTAL WATER CONSUMED =	1184.3 GPM

WATER GENERATED:

GAS CLEANING CONDENSATE =	235.0 GPM
COOLING TOWER BLOWDOWN =	23.2 GPM
TOTAL WATER GENERATED =	258.3 GPM

PLANT WATER SUMMARY:

NET MAKEUP WATER REQUIRED =	1114.1 GPM
WATER CONSUMED / NG FED =	6.64 LB/LB

CO2 BALANCE:

CO2 EMITTED (TOTAL) =	2713. TON/DY
CO2 EMITTED (TOTAL) =	47. MMSCFD @ 60°F
CO2 EMITTED / NG FED =	2.69 LB/LB

NUCLEAR INTEGRATION REQUIREMENTS:

TOTAL ELECTRICITY DEMAND =	4.6 MW
TOTAL HEAT DEMAND =	119.2 MW
TOTAL HEAT DEMAND =	406.7 MMBTU/HR
HELIUM FLOWRATE REQUIRED =	670625. LB/HR
HELIUM FLOWRATE REQUIRED =	84.50 KG/S
HELIUM SUPPLY TEMPERATURE =	1292. DEG. F.
HELIUM SUPPLY TEMPERATURE =	700. DEG. C.
HELIUM RETURN TEMPERATURE =	803. DEG. F.
HELIUM RETURN TEMPERATURE =	428. DEG. C.

EFFICIENCY CALCULATIONS:

OVERALL PROCESS THERMAL EFFICIENCY:

HTGR-Integrated - without CC

ASSUMPTIONS:

- (1) FOR STEAM EXPORT - HEAT RECOVERY CREDIT
ASSUMES SATURATED LIQUID AT ATMOSPHERIC PRESSURE
- (2) FOR ELECTRICITY IMPORT - ELECTRICAL
GENERATION EFFICIENCY OF 40%

HHV BASIS = 85.0%
LHV BASIS = 80.0%

OVERALL PROCESS COLD GAS EFFICIENCY:
(INCLUDES ONLY NATURAL GAS FEED AND HYDROGEN PRODUCT)

HHV BASIS = 90.4%
LHV BASIS = 84.8%

OTHER PERFORMANCE METRICS:

HYDROGEN PRODUCT TO NATURAL GAS FEED = 2.91 SCF/SCF
(CONVENTIONAL PROCESS IS AROUND 2.5)

STEAM EXPORT TO HYDROGEN PRODUCT = 9.20 TON/TON
(CONVENTIONAL PROCESS IS TYPICALLY BETWEEN 5 AND 18)

calculator block AIRPROPS

HUMIDITY DATA FOR STREAM PRI-AIR:

HUMIDITY RATIO = 43.5 GRAINS/LB
RELATIVE HUMIDITY = 39.0 %

calculator block NG-RFMR Hierarchy: REFORMER

SULFUR REMOVAL CONDITIONS:

INLET BED TEMPERATURE = 734. °F

PRIMARY REFORMER CONDITIONS:

STAGE 1:

INLET TEMPERATURE = 1000. °F
STEAM TO CARBON MOLAR RATIO = 3.00
OUTLET TEMPERATURE = 1247. °F
METHANE CONVERSION = 28.0 %

STAGE 2:

INLET TEMPERATURE = 1247. °F
STEAM TO CARBON MOLAR RATIO = 2.37
OUTLET TEMPERATURE = 1600. °F
METHANE CONVERSION = 69.8 %

OVERALL METHANE CONVERSION = 78.2 %

calculator block GASCLEAN Hierarchy: GAS-CLN

WATER GAS SHIFT CONDITIONS:

HIGH-TEMPERATURE SHIFT REACTOR:

INLET TEMPERATURE = 635. °F
(MINIMUM ALLOWABLE = 600 °F)

HTGR-Integrated - without CC

INLET STEAM:DRY GAS RATIO = 0.52
(MINIMUM ALLOWABLE = 0.4)

OUTLET TEMPERATURE = 769. °F

OUTLET CO (DRY BASIS) = 3.24 MOL.%

LOW-TEMPERATURE SHIFT REACTOR:

INLET DEW TEMPERATURE = 332. °F

INLET TEMPERATURE = 375. °F
(SHOULD BE AT LEAST 375°F)
(SHOULD BE AT LEAST 27°F ABOVE DEW TEMPERATURE)

INLET STEAM:DRY GAS RATIO = 0.38
(MINIMUM ALLOWABLE = 0.3)

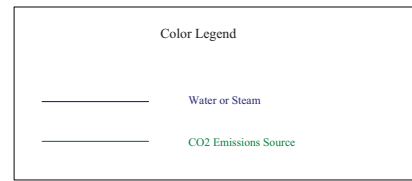
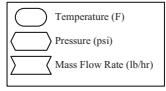
OUTLET TEMPERATURE = 418. °F

OUTLET CO (DRY BASIS) = 0.36 MOL.%

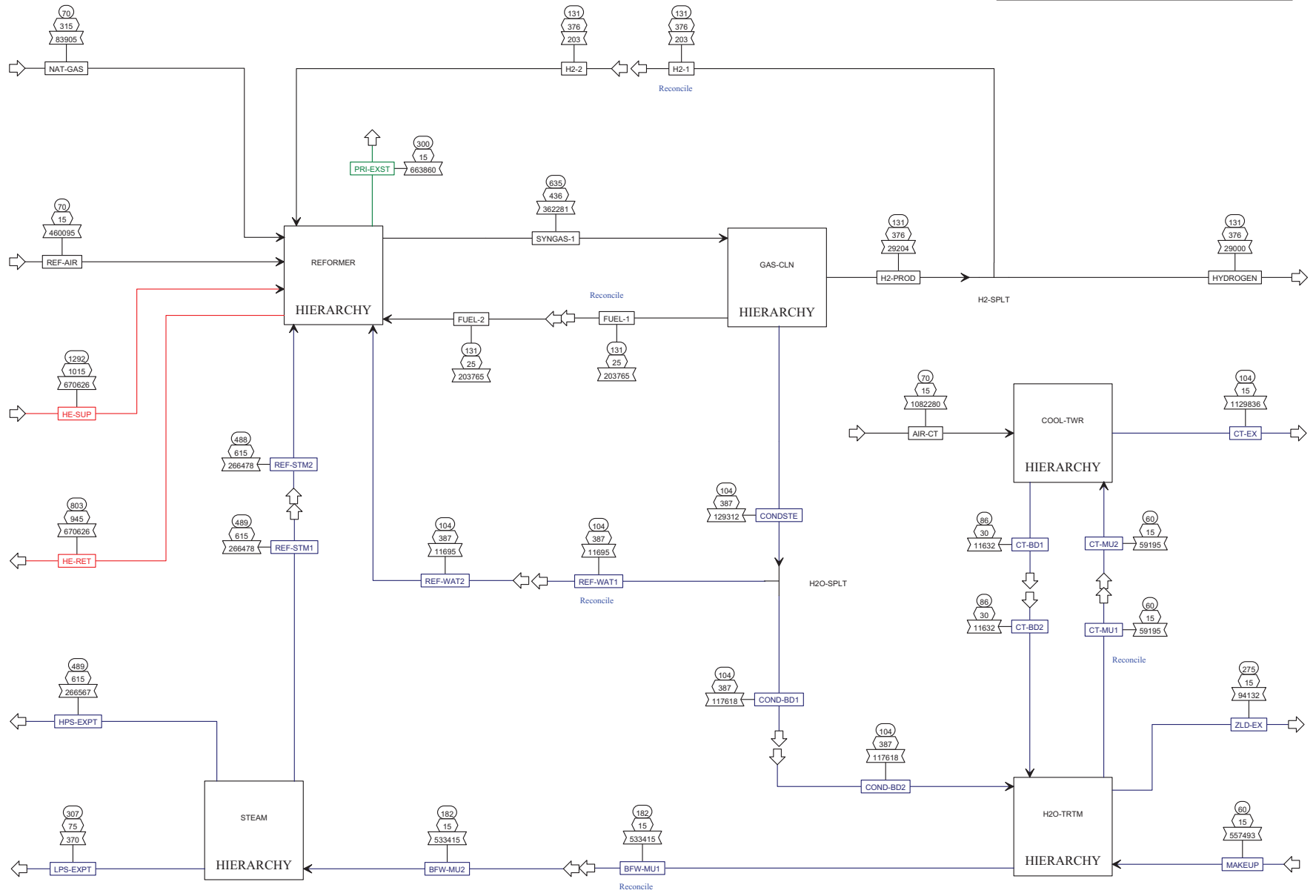
PSA PERFORMANCE:

HYDROGEN RECOVERY = 88.0 %
HYDROGEN PURITY = 99.90 %

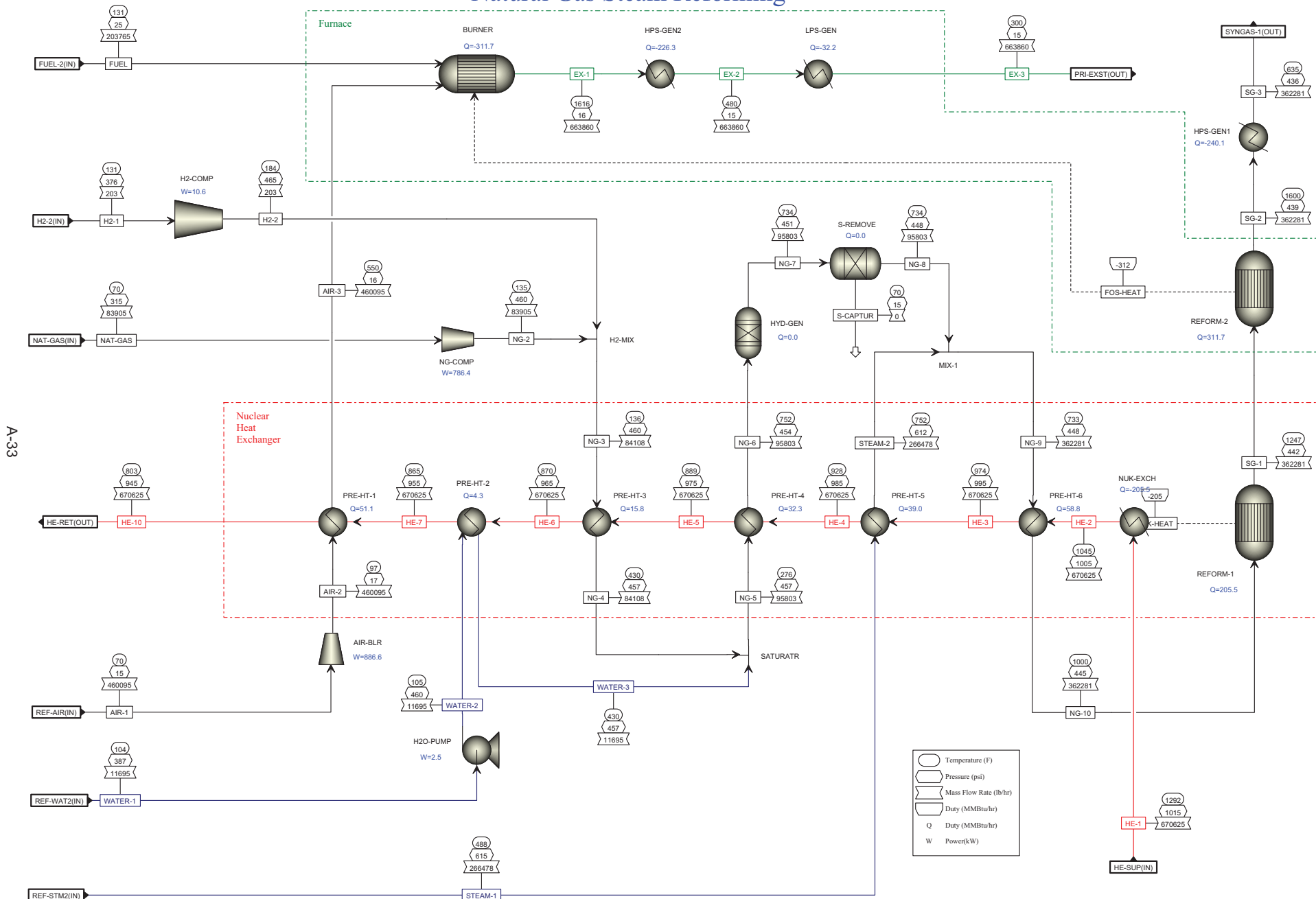
HTGR-Integrated Hydrogen from Natural Gas (Without Carbon Capture)



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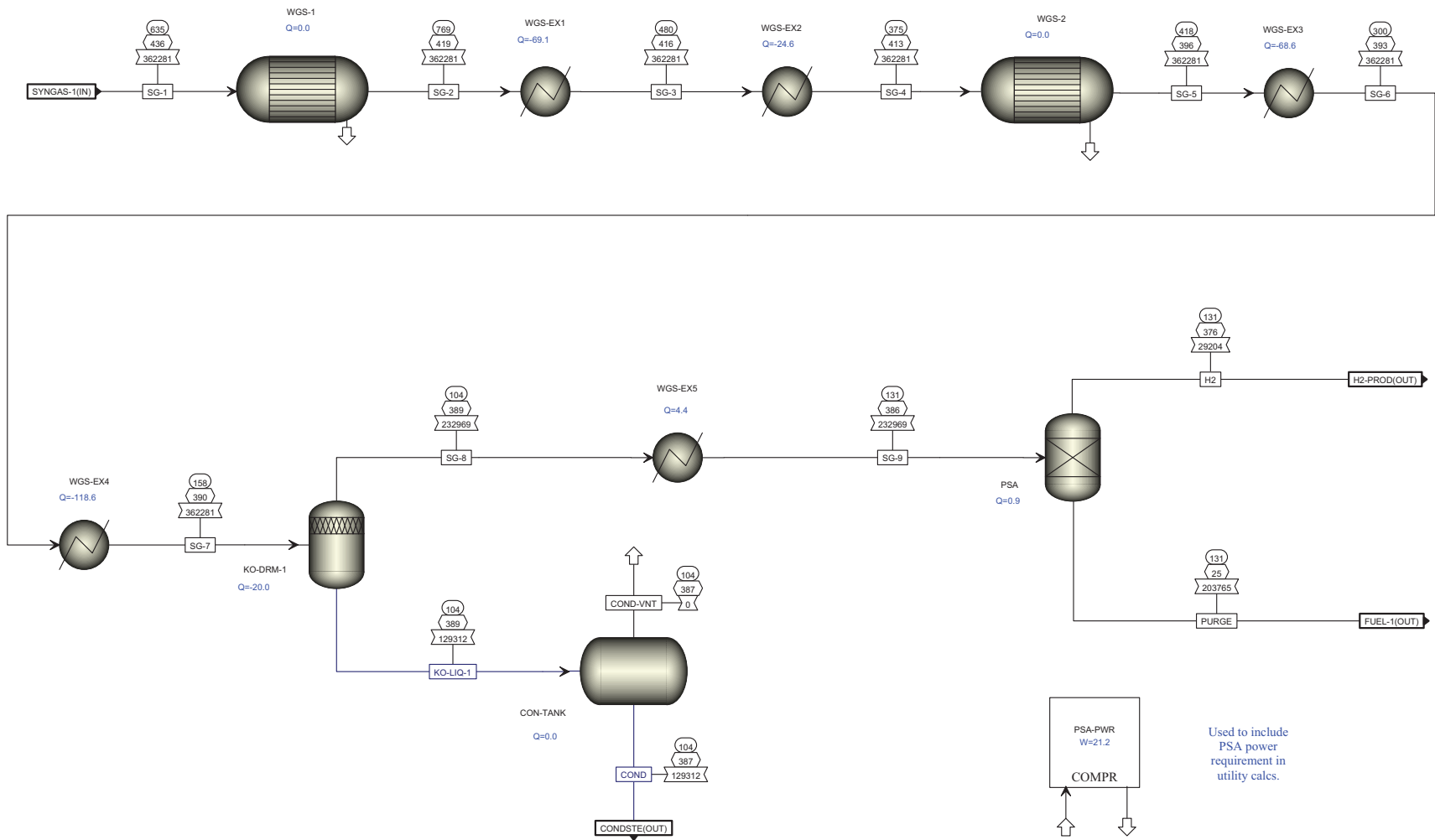
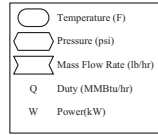


Natural Gas Steam Reforming



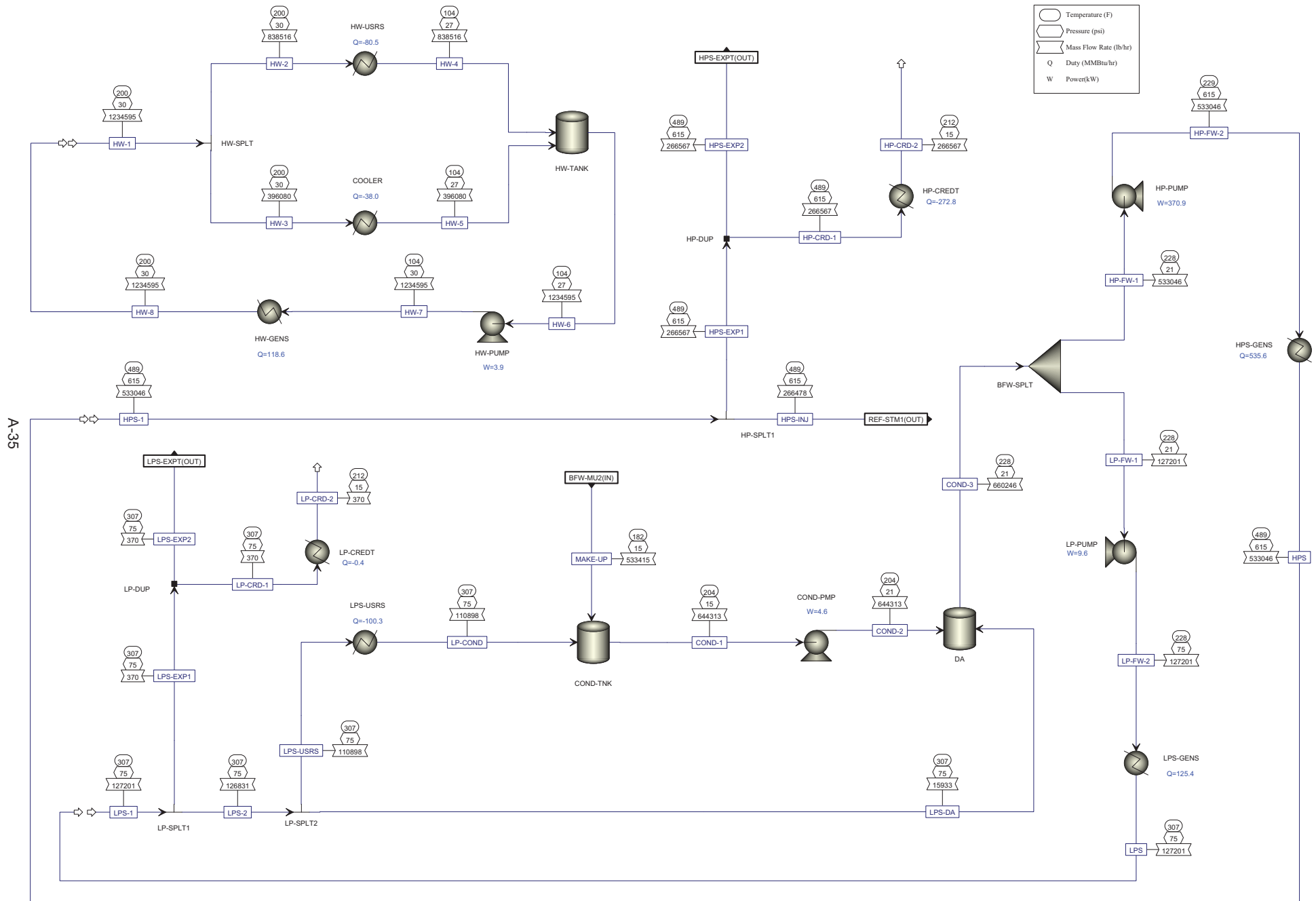
A-33

Syngas Cleaning & Conditioning

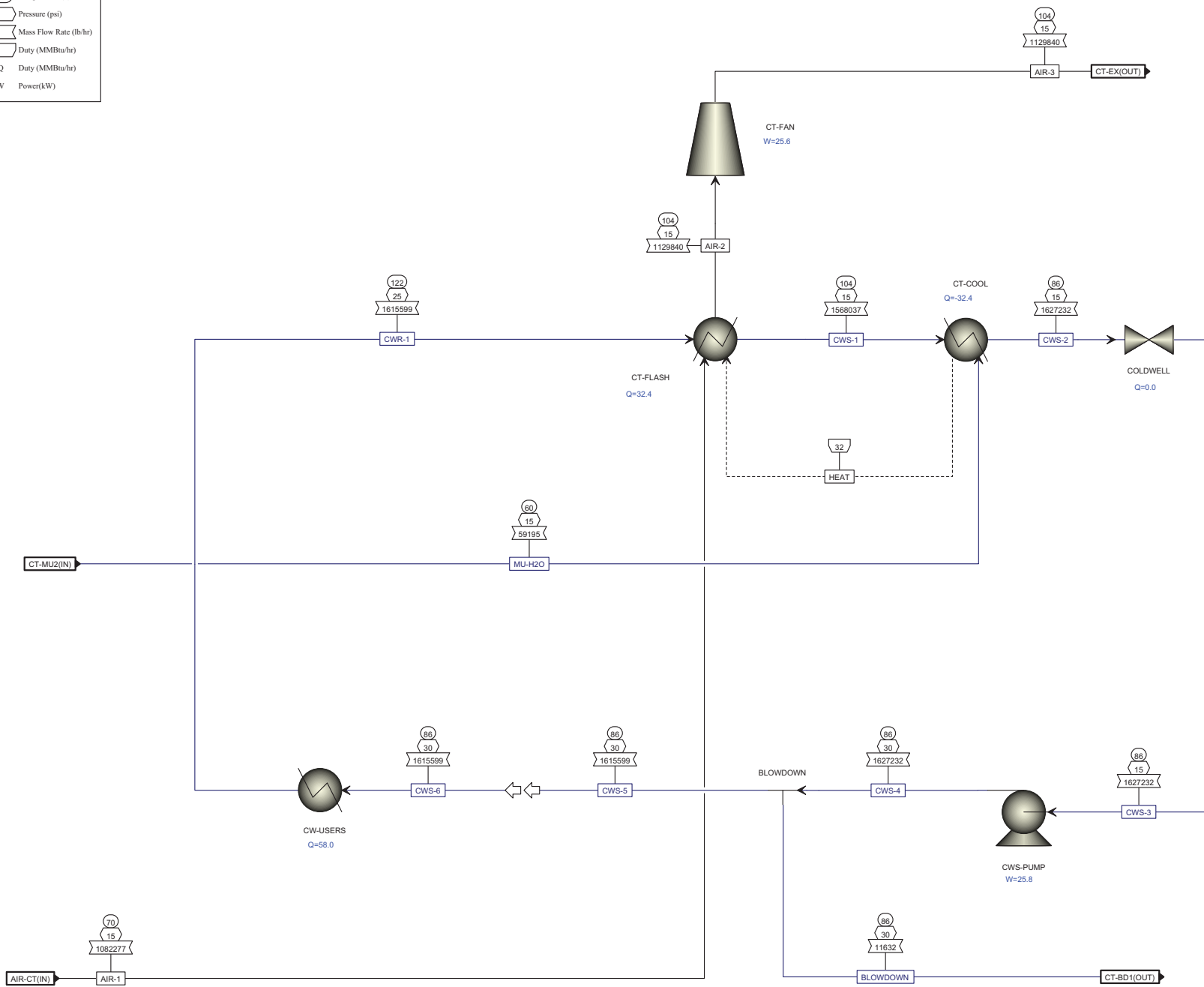
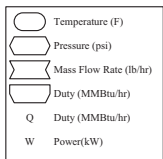


Used to include PSA power requirement in utility calcs.

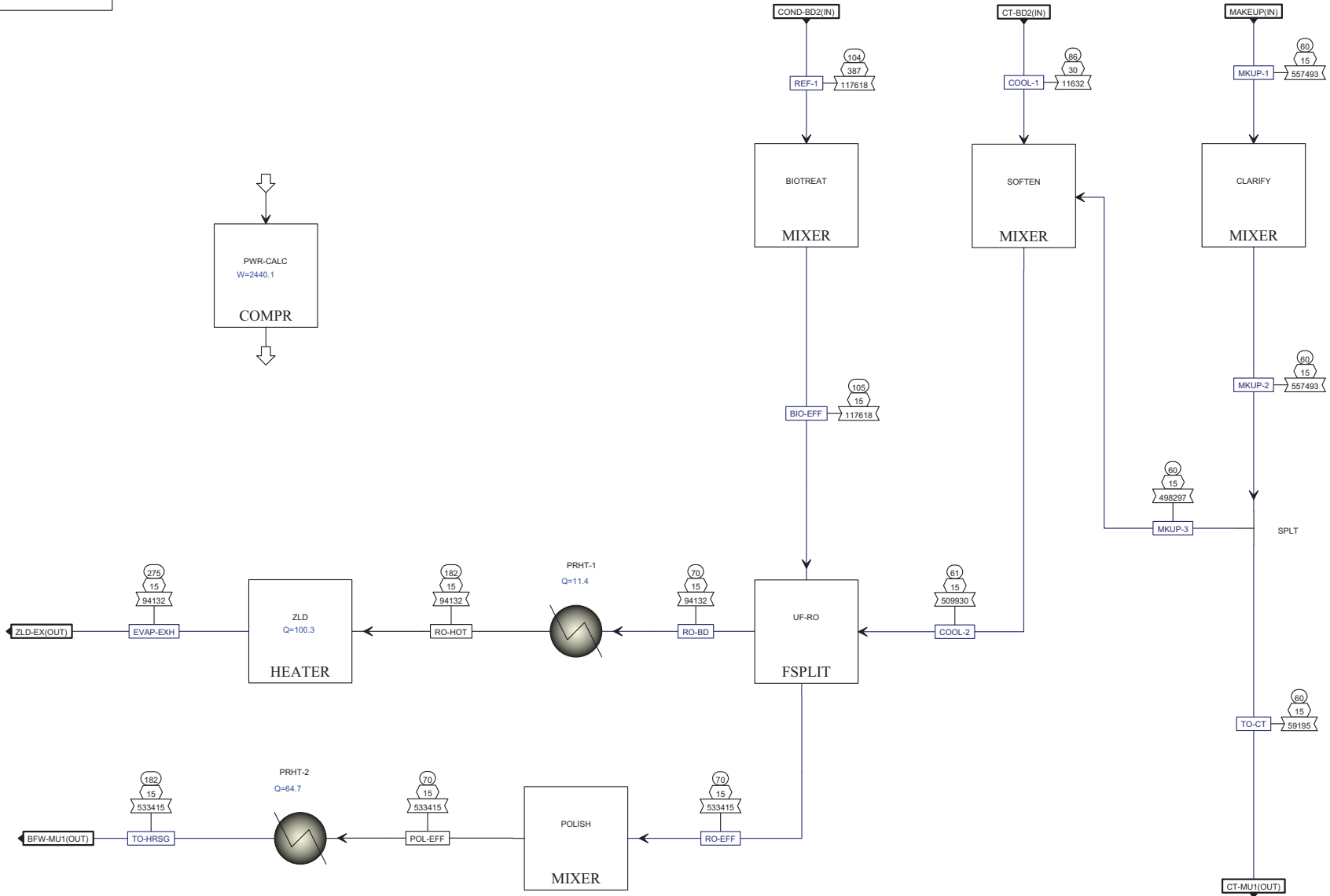
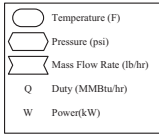
Steam System

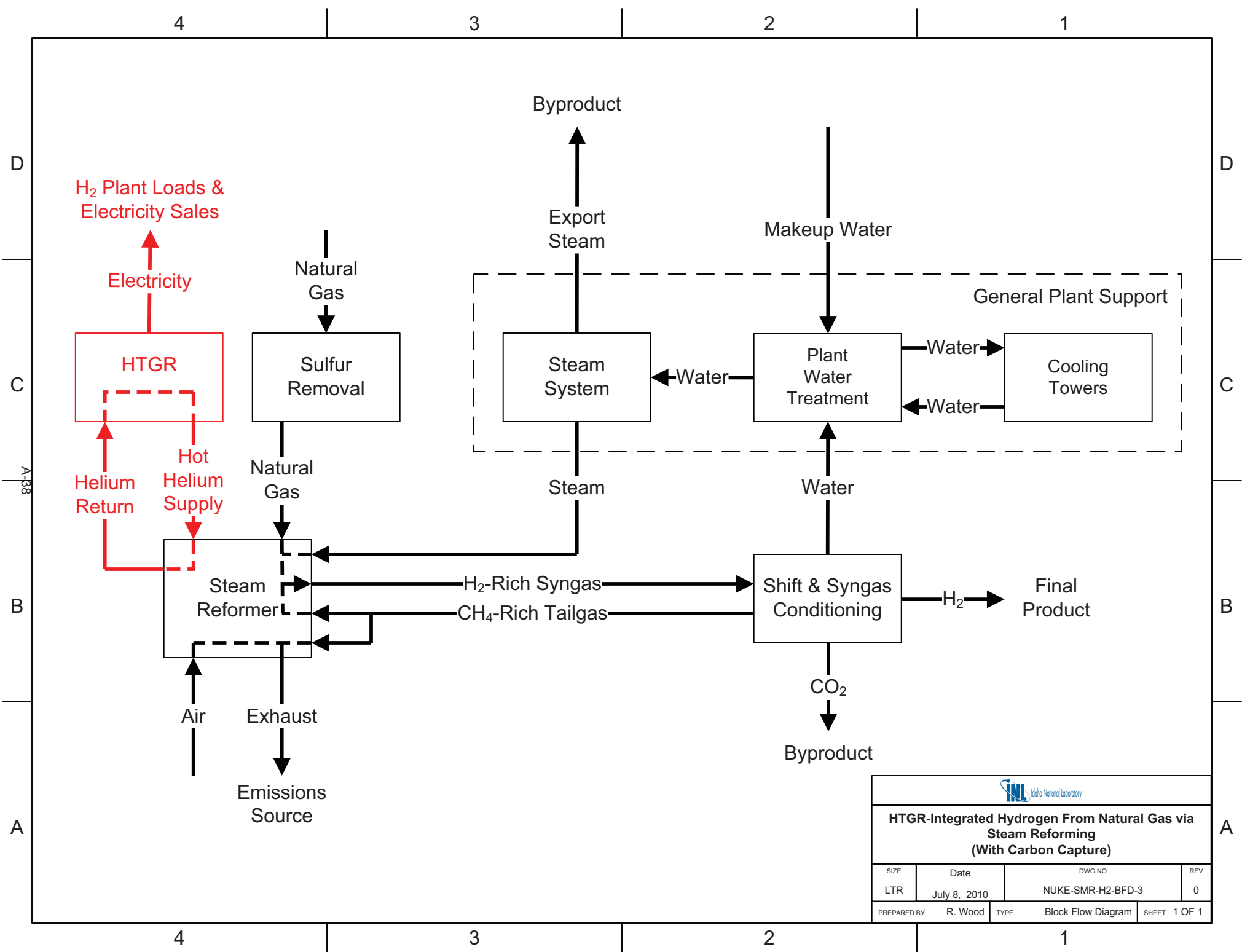



Cooling Tower

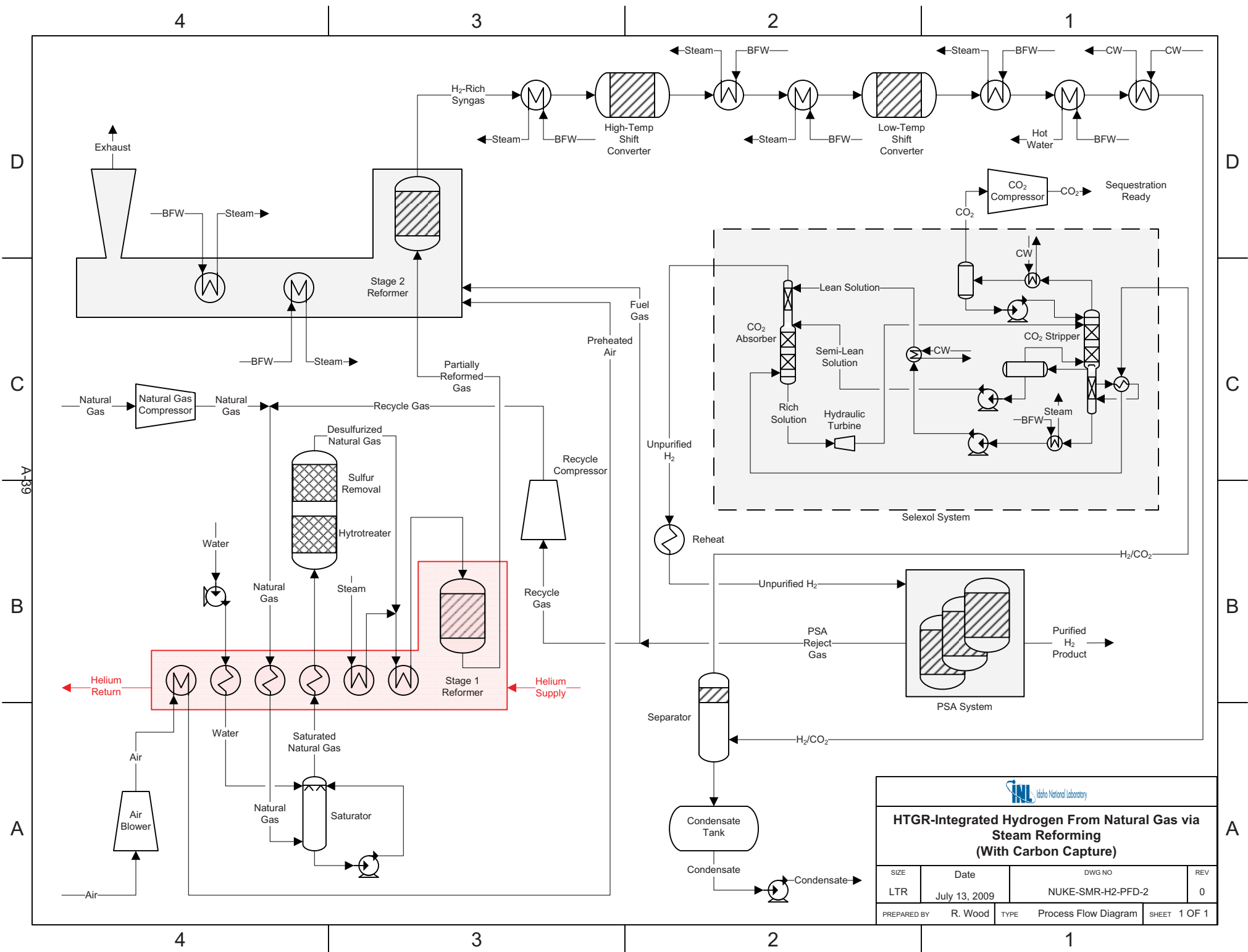



Simplified Water Treatment





 HTGR-Integrated Hydrogen From Natural Gas via Steam Reforming (With Carbon Capture)			
SIZE	Date	DWG NO	REV
LTR	July 8, 2010	NUKE-SMR-H2-BFD-3	0
PREPARED BY	R. Wood	TYPE	Block Flow Diagram
			SHEET 1 OF 1



 HTGR-Integrated Hydrogen From Natural Gas via Steam Reforming (With Carbon Capture)			
SIZE	Date	DWG NO	REV
LTR	July 13, 2009	NUKE-SMR-H2-PFD-2	0
PREPARED BY	R. Wood	TYPE	Process Flow Diagram
		SHEET	1 OF 1

HTGR-Integrated - With CC

Calculator Block SUMMARY

FEED SUMMARY:

NATURAL GAS PROPERTIES:

MASS FLOW =	985. TON/DY
VOLUME FLOW =	43.5 MMSCFD @ 60°F
HHV =	23063. BTU/LB
HHV =	1044. BTU/SCF @ 60°F
ENERGY FLOW =	45434. 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 =	361400. LB/HR
RAW SYNGAS VOLUME FLOW =	261. MMSCFD @ 60°F
RAW SYNGAS COMPOSITION:	
H2	47.8 MOL.%
CO	9.2 MOL.%
CO2	5.3 MOL.%
N2	0.2 MOL.%
H2O	34.1 MOL.%
CH4	3.5 MOL.%

FINAL PRODUCT SUMMARY:

HYDROGEN MASS FLOW =	29000. LB/HR
HYDROGEN VOLUME FLOW =	130. MMSCFD @ 60°F
HYDROGEN COMPOSITION:	
H2	99.9 MOL.%
N2	0.0 MOL.%
CH4	0.1 MOL.%
CO	11. PPMV
CO2	3. PPMV
H2O	0. PPMV

EXPORT HP STEAM FLOW =	172213. LB/HR
EXPORT HP STEAM CREDIT =	176.2 MMBTU/HR
STEAM PRESSURE =	615. PSIA
STEAM TEMPERATURE =	489. DEG. F
STEAM ENTHALPY (FOR CREDIT) =	1023. BTU/LB
CONDENSED TO LIQUID AT 1 ATM AT BOILING POINT	

CARBON DIOXIDE MASS FLOW =	177159. LB/HR
CARBON DIOXIDE VOLUME FLOW =	37. MMSCFD @ 60°F
CARBON DIOXIDE COMPOSITION:	
CO2	99.1 MOL.%
H2	0.6 MOL.%
N2	0. PPMV
CH4	2442. PPMV

	HTGR-Integrated - With CC	
CO		188. PPMV
H2O		223. PPMV

POWER SUMMARY:

ELECTRICAL CONSUMERS:		
NG REFORMER POWER CONSUMPTION =		2.0 MW
GAS CLEANING POWER CONSUMPTION =		2.3 MW
STEAM SYSTEM POWER CONSUMPTION =		0.4 MW
CO2 COMPRESSION POWER CONSUMPTION =		7.0 MW
COOLING TOWER POWER CONSUMPTION =		0.1 MW
WATER TREATMENT POWER CONSUMPTION =		2.0 MW
CONSUMER SUBTOTAL =		13.7 MW
NET PLANT POWER CONSUMPTION =		13.7 MW

WATER BALANCE:

EVAPORATIVE LOSSES:		
COOLING TOWER EVAPORATION =		193.3 GPM
ZLD SYSTEM EVAPORATION =		154.3 GPM
TOTAL EVAPORATIVE LOSSES =		347.7 GPM
WATER CONSUMED:		
BOILER FEED WATER MAKEUP =		874.5 GPM
COOLING TOWER MAKEUP =		203.7 GPM
TOTAL WATER CONSUMED =		1078.2 GPM
WATER GENERATED:		
GAS CLEANING CONDENSATE =		234.7 GPM
COOLING TOWER BLOWDOWN =		40.0 GPM
TOTAL WATER GENERATED =		274.7 GPM
PLANT WATER SUMMARY:		
NET MAKEUP WATER REQUIRED =		957.7 GPM
WATER CONSUMED / NG FED =		5.84 LB/LB

CO2 BALANCE:

CO2 EMITTED (TOTAL) =		525. TON/DY
CO2 EMITTED (TOTAL) =		9. MMSCFD @ 60°F
SEQUESTRATION READY CO2 =		2123. TON/DY
SEQUESTRATION READY CO2 =		37. MMSCFD @ 60°F
CO2 EMITTED / NG FED =		0.53 LB/LB

NUCLEAR INTEGRATION REQUIREMENTS:

TOTAL ELECTRICITY DEMAND =		13.7 MW
TOTAL HEAT DEMAND =		116.9 MW
TOTAL HEAT DEMAND =		398.8 MMBTU/HR
HELIUM FLOWRATE REQUIRED =		663671. LB/HR
HELIUM FLOWRATE REQUIRED =		83.62 KG/S
HELIUM SUPPLY TEMPERATURE =		1292. DEG. F.
HELIUM SUPPLY TEMPERATURE =		700. DEG. C.

HTGR-Integrated - With CC

HELIUM RETURN TEMPERATURE =	807. DEG. F.
HELIUM RETURN TEMPERATURE =	431. DEG. C.

EFFICIENCY CALCULATIONS:

OVERALL PROCESS THERMAL EFFICIENCY:

ASSUMPTIONS:

- (1) FOR STEAM EXPORT - HEAT RECOVERY CREDIT
ASSUMES SATURATED LIQUID AT ATMOSPHERIC PRESSURE
- (2) FOR ELECTRICITY IMPORT - ELECTRICAL
GENERATION EFFICIENCY OF 40%

HHV BASIS =	80.3%
LHV BASIS =	74.8%

OVERALL PROCESS COLD GAS EFFICIENCY:

(INCLUDES ONLY NATURAL GAS FEED AND HYDROGEN PRODUCT)

HHV BASIS =	92.9%
LHV BASIS =	87.1%

OTHER PERFORMANCE METRICS:

HYDROGEN PRODUCT TO NATURAL GAS FEED = 2.99 SCF/SCF
(CONVENTIONAL PROCESS IS AROUND 2.5)

STEAM EXPORT TO HYDROGEN PRODUCT = 5.94 TON/TON
(CONVENTIONAL PROCESS IS TYPICALLY BETWEEN 5 AND 18)

calculator Block AIRPROPS

HUMIDITY DATA FOR STREAM PRI-AIR:

HUMIDITY RATIO =	43.5 GRAINS/LB
RELATIVE HUMIDITY =	39.0 %

calculator Block NG-RFMR Hierarchy: REFORMER

SULFUR REMOVAL CONDITIONS:

INLET BED TEMPERATURE =	729. °F
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PRIMARY REFORMER CONDITIONS:

STAGE 1:

INLET TEMPERATURE =	1000. °F
STEAM TO CARBON MOLAR RATIO =	3.00
OUTLET TEMPERATURE =	1247. °F
METHANE CONVERSION =	27.8 %

STAGE 2:

INLET TEMPERATURE =	1247. °F
STEAM TO CARBON MOLAR RATIO =	2.37
OUTLET TEMPERATURE =	1600. °F
METHANE CONVERSION =	69.7 %

OVERALL METHANE CONVERSION =	78.2 %
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calculator Block GASCLEAN Hierarchy: GAS-CLN

HTGR-Integrated - With CC

WATER GAS SHIFT CONDITIONS:

HIGH-TEMPERATURE SHIFT REACTOR:

INLET TEMPERATURE = 635. °F
(MINIMUM ALLOWABLE = 600°F)
INLET STEAM:DRY GAS RATIO = 0.52
(MINIMUM ALLOWABLE = 0.4)
OUTLET TEMPERATURE = 769. °F
OUTLET CO (DRY BASIS) = 3.23 MOL.%

LOW-TEMPERATURE SHIFT REACTOR:

INLET DEW TEMPERATURE = 332. °F
INLET TEMPERATURE = 375. °F
(SHOULD BE AT LEAST 375°F)
(SHOULD BE AT LEAST 27°F ABOVE DEW TEMPERATURE)
INLET STEAM:DRY GAS RATIO = 0.38
(MINIMUM ALLOWABLE = 0.3)
OUTLET TEMPERATURE = 418. °F
OUTLET CO (DRY BASIS) = 0.36 MOL.%

SELEXOL PERFORMANCE:

INLET CO2 CONCENTRATION = 18.9 %
OUTLET CO2 CONCENTRATION = 0.2 %
CO2 CAPTURE = 99.0 %
CO2 PURITY = 99.1 %

SELEXOL UTILITY REQUIREMENTS:

ELECTRICITY USAGE = 2.3 MW
STEAM USAGE = 72398. LB/HR

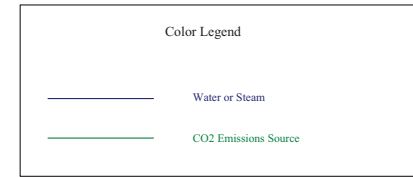
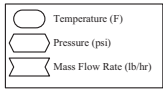
CO2 COMPRESSION:

PRESSURE = 2015. PSIA
ELECTRICITY USAGE = 7.0 MW

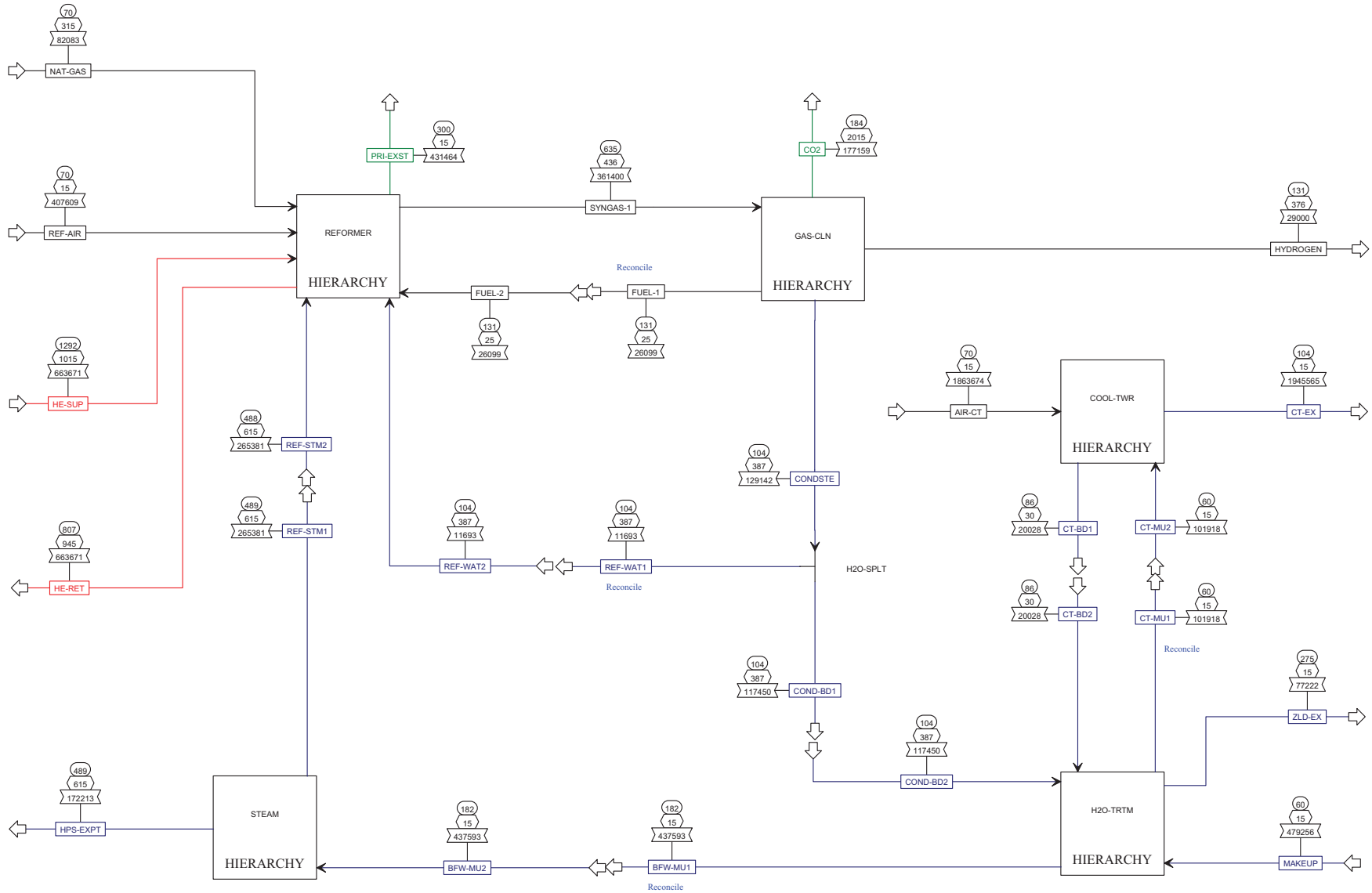
PSA PERFORMANCE:

HYDROGEN RECOVERY = 88.0 %
HYDROGEN PURITY = 99.90 %

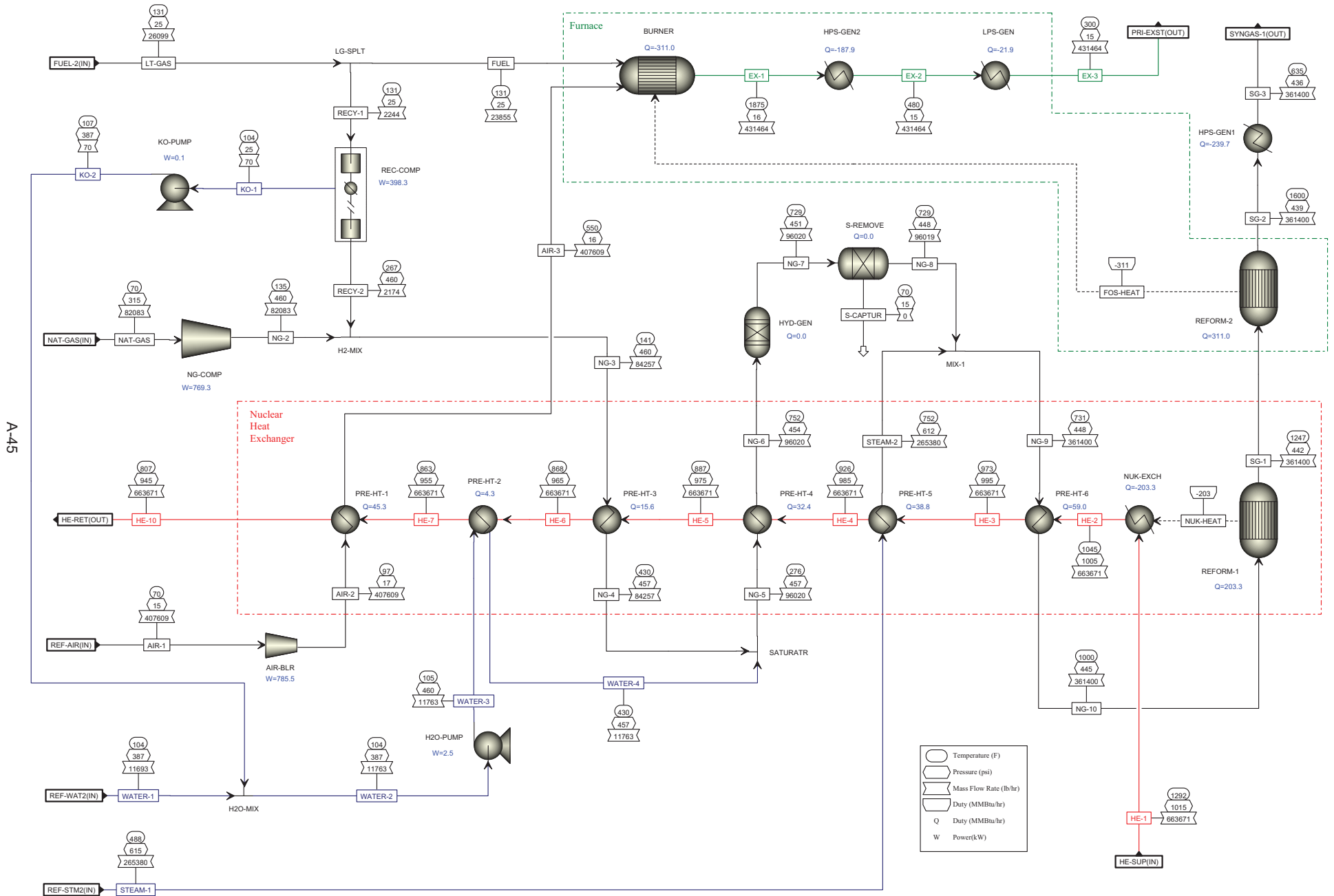
Hydrogen from Natural Gas (With Carbon Capture; With Recycle)



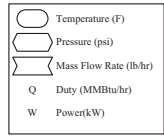
A-44



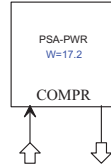
Natural Gas Steam Reforming



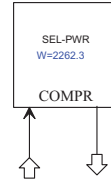
Syngas Cleaning & Conditioning



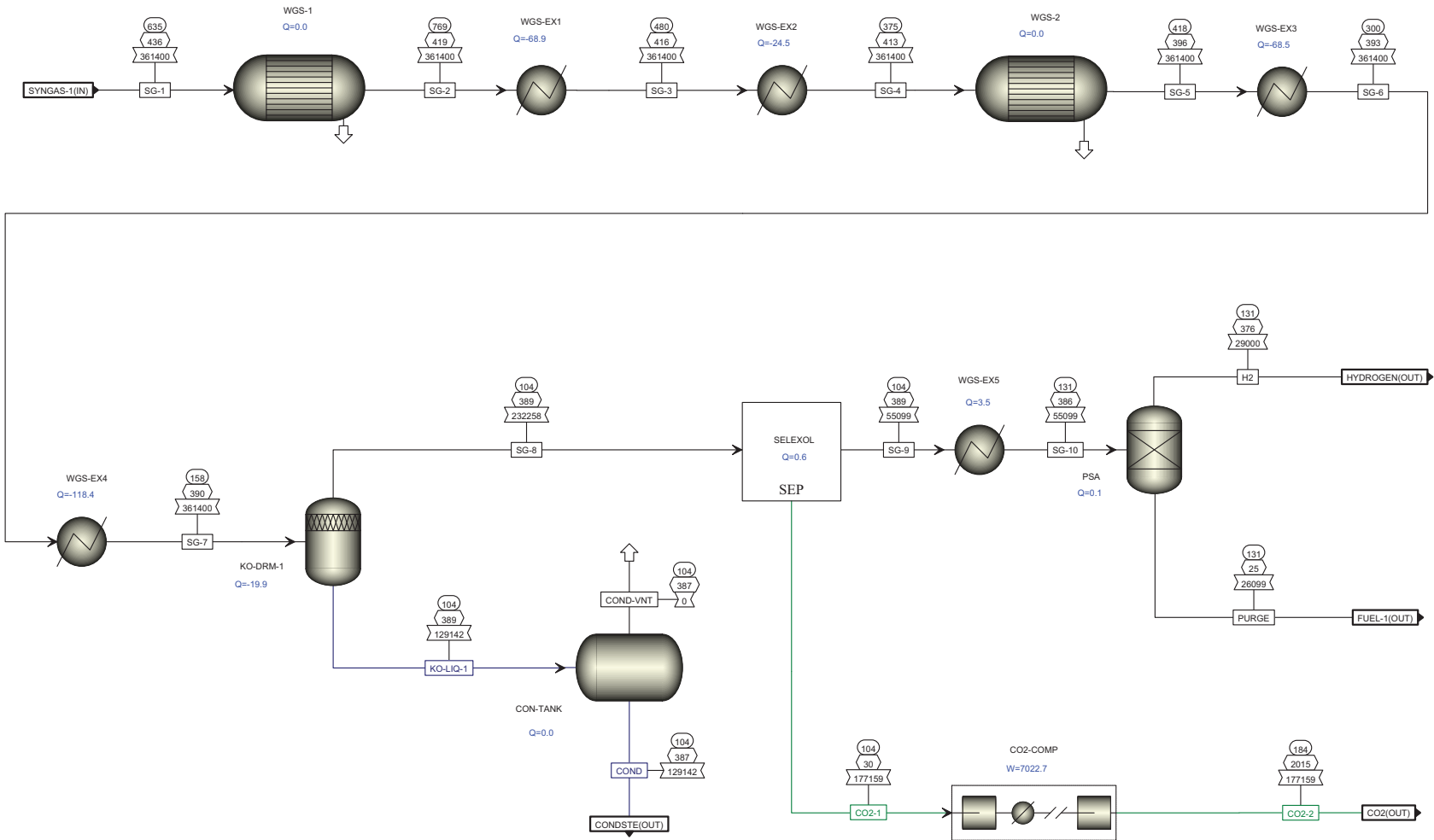
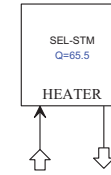
Used to include
PSA power
requirement in
utility calcs.



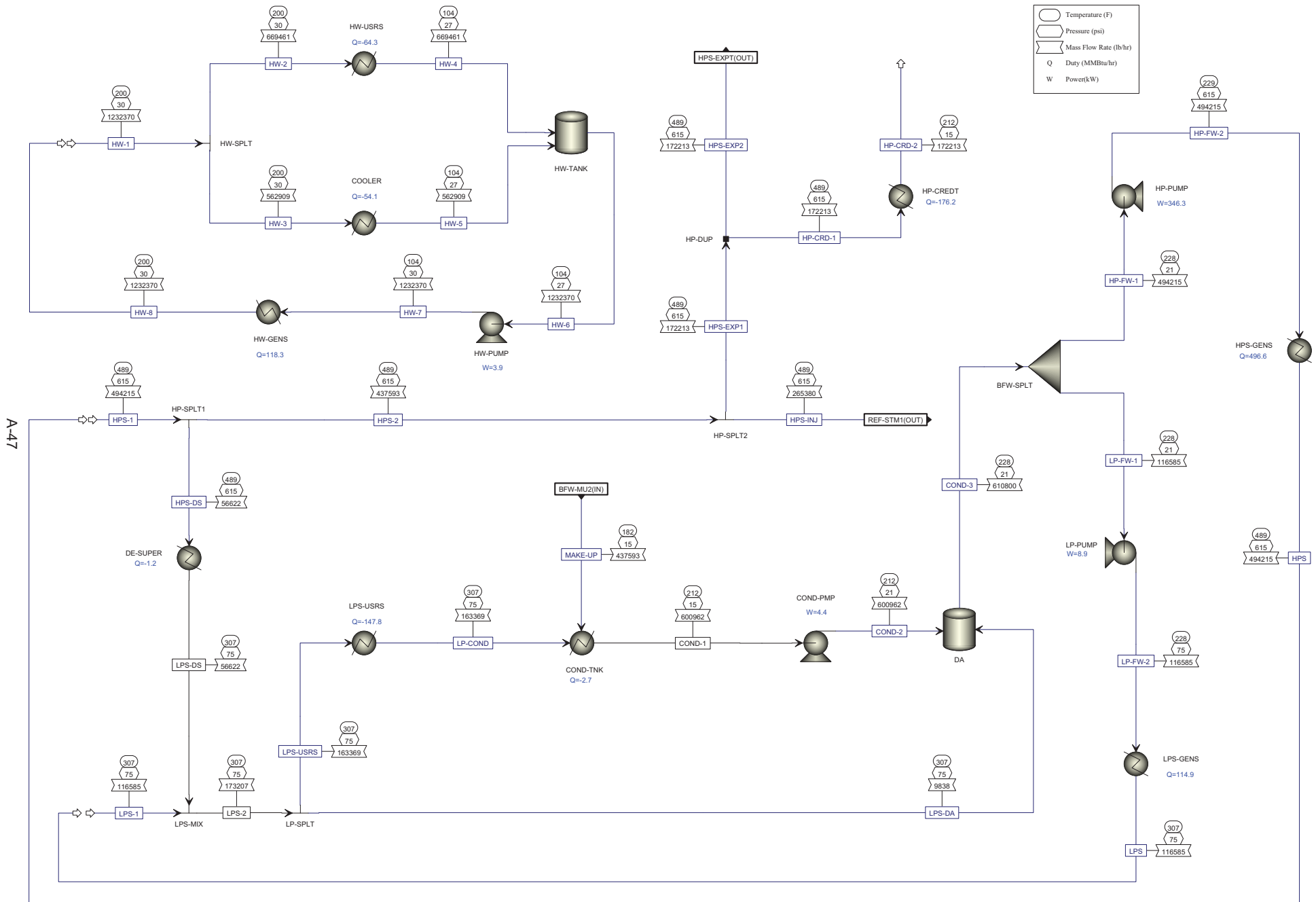
Used to include
Selexol power
requirement in
utility calcs.



Used to include
Selexol steam
usage in utility
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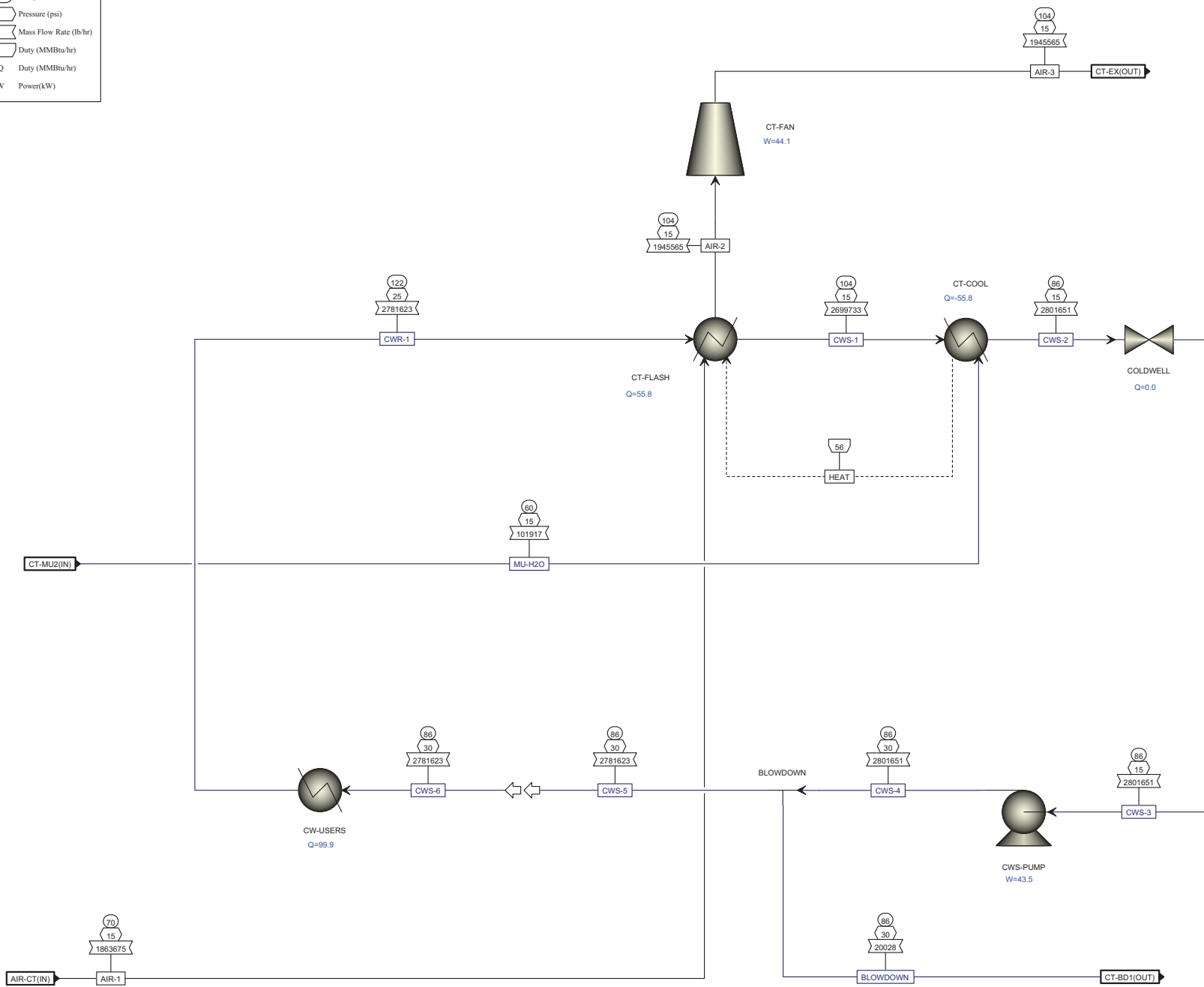
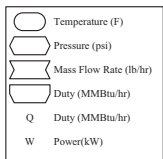


Steam System



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Cooling Tower



Simplified Water Treatment

