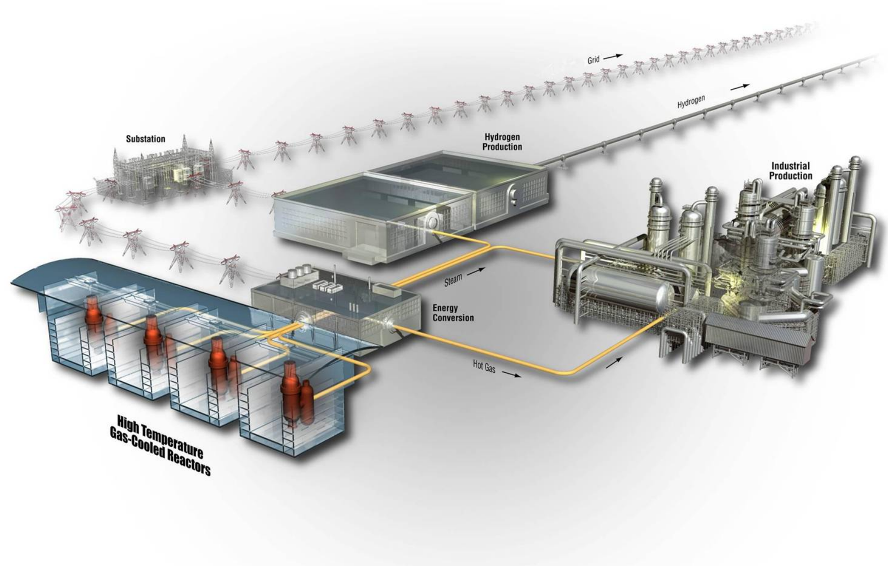
Document ID: TEV-672
Revision ID: 2
Effective Date: 09/30/2011

Technical Evaluation Study

Project No. 23843

HTGR-Integrated Coal and Gas to Liquids Production Analysis



The INL is a
U.S. Department of Energy
National Laboratory
operated by
Battelle Energy Alliance

Idaho National Laboratory

Approved by:

A. M. Gandrik per telecon
A. M. Gandrik
NGNP Engineering Support

9/30/2011
Date

Lee Nelson for MW Patterson
M. W. Patterson
NGNP Engineering Technical Manager

9/30/2011
Date

P. M. Mills
NGNP Engineering Director (Acting)

9/30/11
Date

REVISION LOG

Rev.	Date	Affected Pages	Revision Description
0	11/05/2009	All	Newly issued document.
1	05/15/2010	All	Added economic and GHG emissions sections.
2	09/30/2011	All	Updated process model to Aspen Plus version 7.3, updated economic analysis to include updated HTGR cost estimate.

EXECUTIVE SUMMARY

This technical evaluation (TEV) has been prepared as part of a study for the Next Generation Nuclear Plant (NGNP) Project to evaluate the integration of high-temperature gas-cooled reactor (HTGR) technology with conventional chemical processes. This TEV addresses the integration of HTGR heat and power into both coal to liquids (CTL) and gas to liquids (GTL) production; specifically, the technical and economic feasibility of the HTGR integration. The main liquid product produced in the CTL and GTL processes is diesel fuel. The economic results presented in this TEV are preliminary and should be refined as the design of the HTGR progresses, if the design of the HTGR is changed significantly, or if additional refinements of the HTGR and/or CTL and GTL capital and/or operating costs become available. The HTGR capital, operating and maintenance (O&M) costs, fuel, and decommissioning costs are based on the correlations and costs presented for an n^{th} of a kind HTGR in TEV-1196 (Idaho National Laboratory [INL] 2011a).

The following conclusions were drawn when evaluating the nuclear-integrated CTL process against the conventional process:

- One 664 MWt 850°C ROT HTGR for heat production and nine 604 MWt 700°C ROT HTGRs for power production would be required to support production of 50,000 bbl/day of liquid fuel products.

- Nuclear integration decreases coal consumption by 65% using an HTGR and high temperature steam electrolysis as the hydrogen source.

- Nuclear integration decreases CO_2 emissions by 83% if sequestration is assumed and 96% without sequestration.

- Economically, the nuclear-integrated CTL case provides a lower internal rate of return (IRR) than the conventional CTL case, either with or without CO_2 sequestration. Figure ES-1 presents the IRR versus the diesel selling price for the conventional and nuclear-integrated cases.

- The carbon tax results show that the nuclear-integrated CTL case outperforms the conventional case at a 12% IRR when the carbon tax is approximately \$120/ton-CO_2. Figure ES-2 presents the carbon tax results for the CTL cases analyzed.

- An economic sensitivity analysis was performed, it was determined the uncertainty in the HTGR TCI can have the largest impact on the required product selling price, followed by assumed IRR, the debt to equity ratio, and the assumed economic recovery period. Figure ES-3 presents a tornado diagram for nuclear-integrated CTL process, showing the resulting diesel price when varying the baseline economic assumptions.

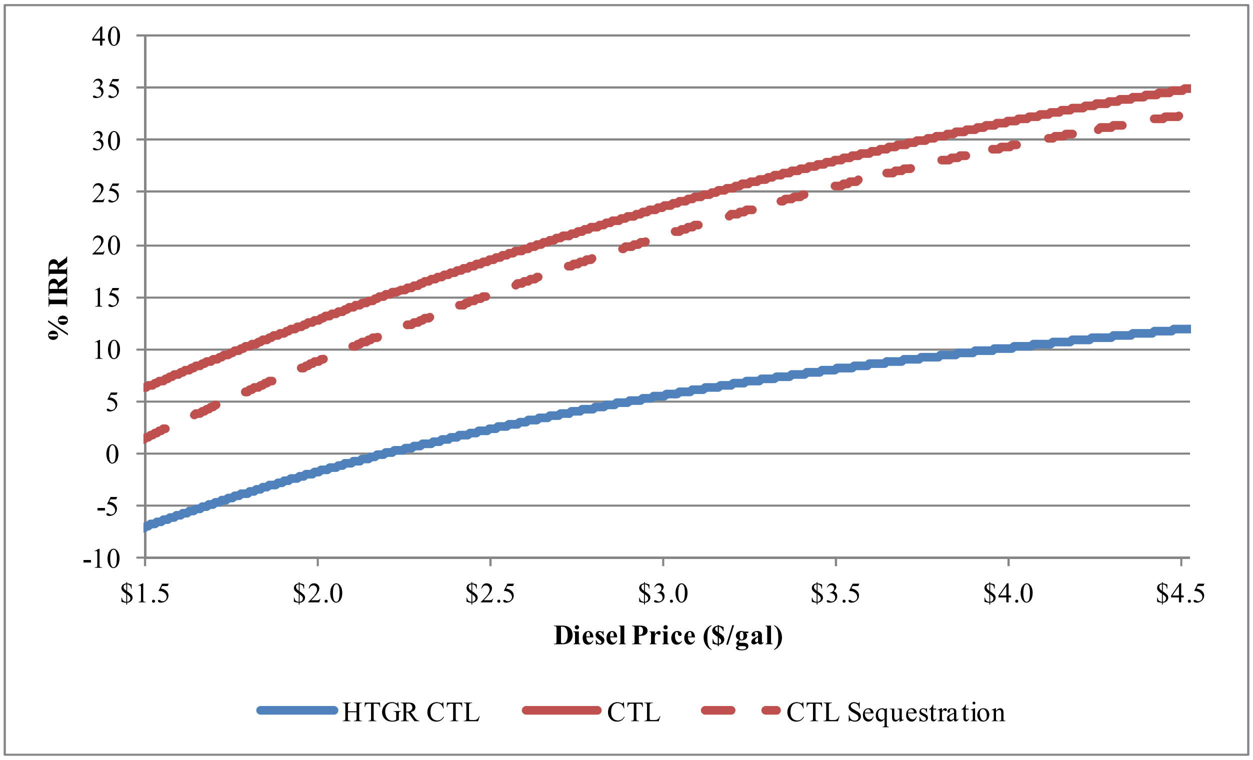


Figure ES-1. CTL cases, IRR as a function of the diesel selling price, 12% IRR.

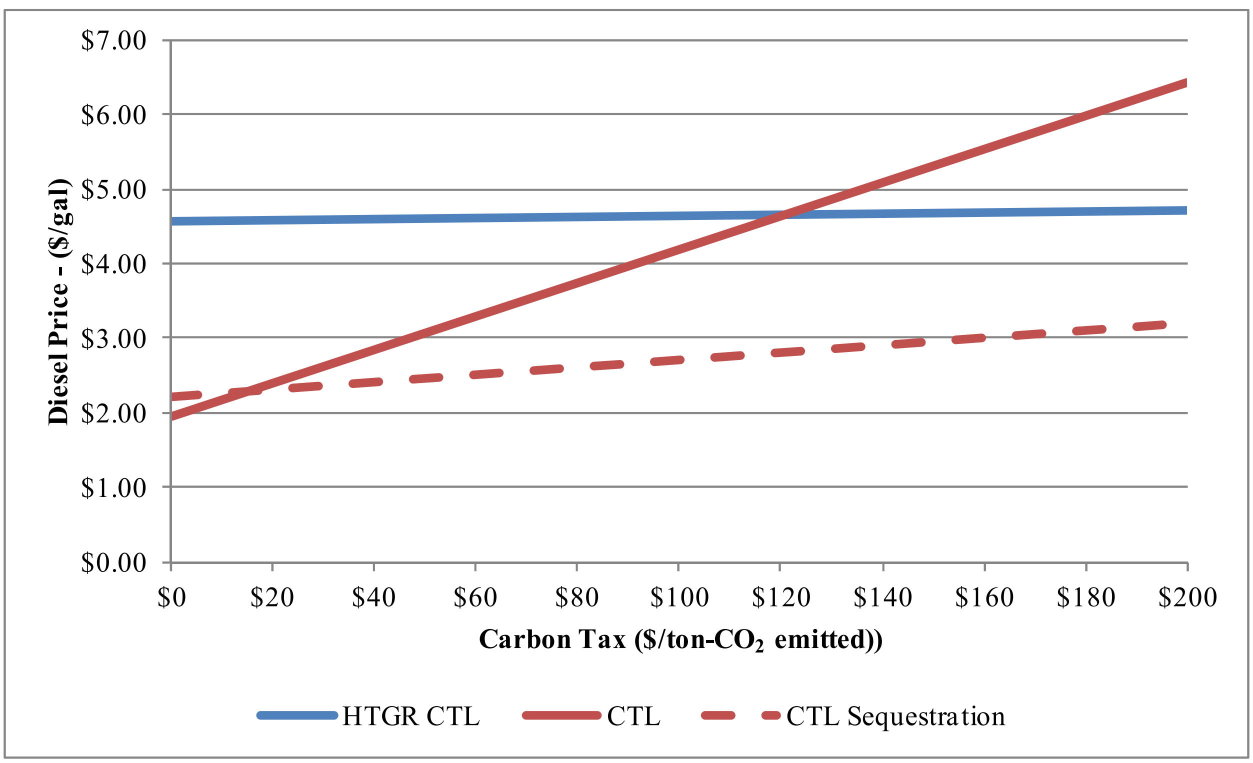


Figure ES-2. CTL cases, diesel price as a function of a carbon tax on CO_2 emissions, 12% IRR.

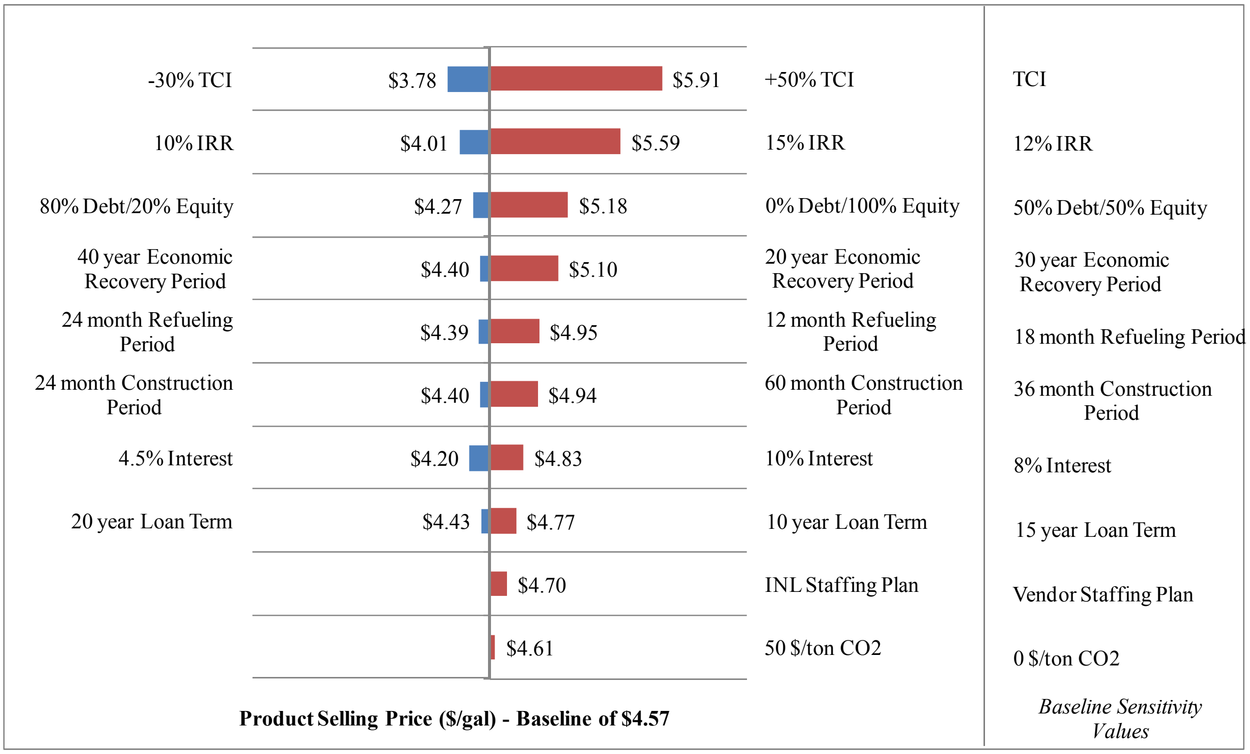


Figure ES-3. HTGR CTL tornado diagram.

The following conclusions were drawn when evaluating the nuclear-integrated GTL process against the conventional process:

- Approximately one 450 MW_t 700°C ROT HTGR would be required to support production of 50,000 bbl/day of liquid fuel products
- Nuclear integration decreases natural gas consumption by 9% using nuclear heat for gas combustion for preheating in the reforming and refining areas.
- Incorporating an HTGR into the GTL process decrease CO_2 emissions by 42% when sequestration is not assumed and by 88% if the pure CO_2 stream produced in the nuclear-integrated GTL process is sequestered.
- Economically, the nuclear-integrated GTL case, either with or without sequestration, requires a higher diesel selling price to achieve a 12% IRR than the conventional case, for natural gas prices less than approximately $14.00/MSCF. Figure ES-4 presents the diesel selling price versus the natural gas purchase price for the conventional and nuclear-integrated cases.
- The carbon tax results show that the nuclear-integrated GTL case without sequestration outperforms the conventional case at a 12% IRR for an average natural gas purchase price when the carbon tax is approximately $120/ton-$CO_2$. When sequestration is

assumed for the nuclear-integrated GTL case, the required CO_2 tax decreases to approximately \$70/ton-CO_2. Figure ES-5 presents the carbon tax results for the GTL cases analyzed.

- From the economic sensitivity analysis, the natural gas purchase price can have the largest impact on the required product selling price, followed by assumed IRR, the debt to equity ratio, and a \$50/ton CO_2 tax. Figure ES-6 presents a tornado diagram for nuclear-integrated GTL process, showing the resulting diesel price when varying the baseline economic assumptions.

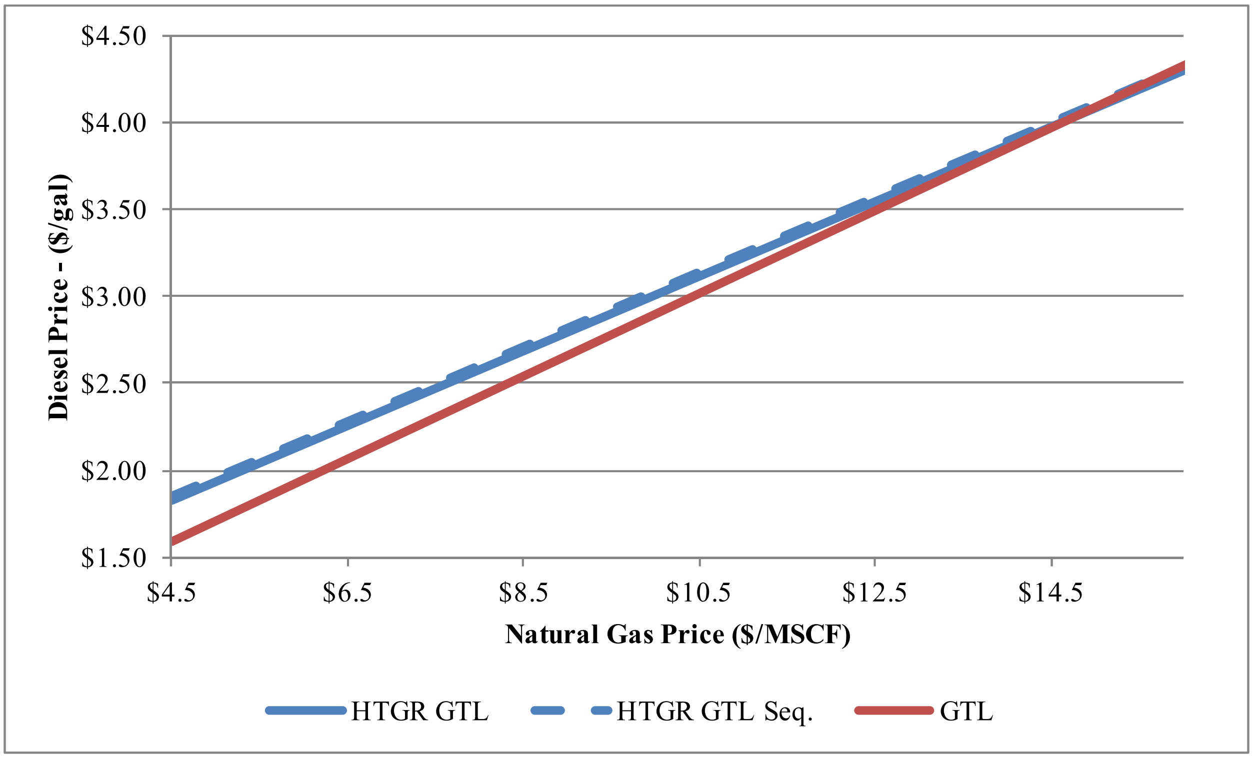


Figure ES-4. GTL cases, diesel selling price versus natural gas purchase price, 12% IRR.

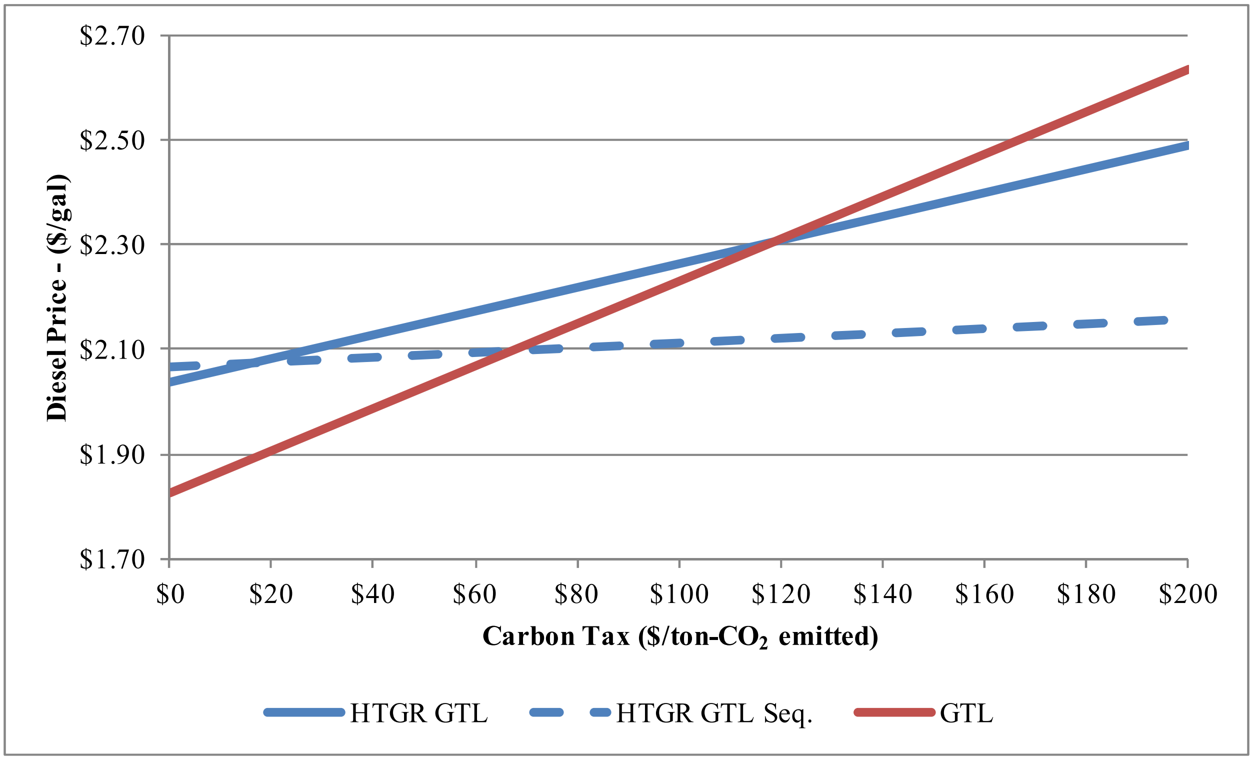


Figure ES-5. GTL cases, diesel price as a function of a carbon tax on CO_2 emissions, 12% IRR, average natural gas purchase price.

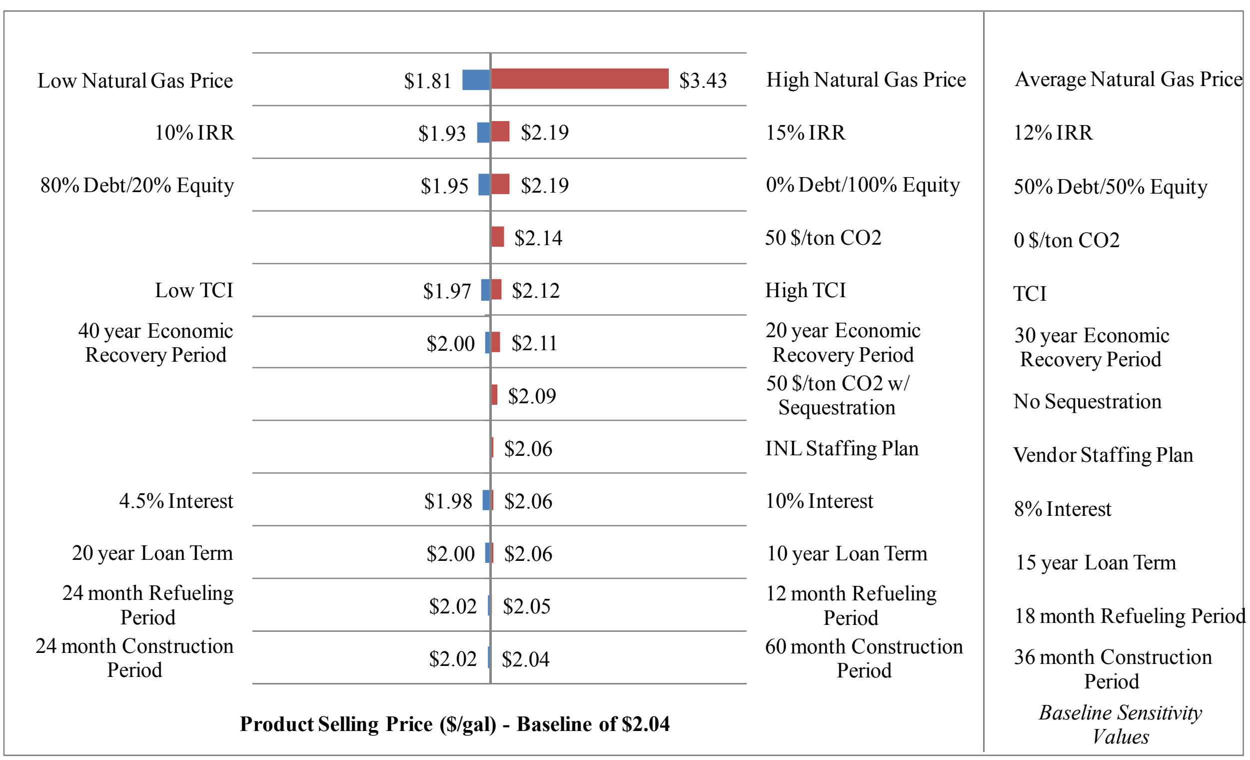


Figure ES-6. HTGR GTL tornado diagram.

CONTENTS

ACRONYMS AND NOMENCLATURE

AACE	Association for the Advancement of Cost Engineering
ASF	Anderson Schulz Flory
ASU	air separation unit
ATCF	after tax cash flow
BTCF	before tax cash flow
CEPCI	chemical engineering plant cost index
CTL	coal to liquids
DOE	Department of Energy
EIA	Energy Information Administration
FT	Fischer-Tropsch
GHG	greenhouse gas
GIF	GEN-IV International Forum
GTL	gas to liquids
GWP	global warming potential
HP	high pressure
HRSG	heat recovery steam generator
HTSE	high temperature steam electrolysis
HTGR	high temperature gas-cooled reactor
INL	Idaho National Laboratory
IPCC	Intergovernmental Panel on Climate Change
IRR	internal rate of return
LHV	lower heating value
LP	low pressure
LPG	liquefied petroleum gas
MACRS	modified accelerated cost recovery system
MARR	minimum annual rate of return
NETL	National Energy Technology Laboratory
NIBT	net income before taxes
NGNP	Next Generation Nuclear Plant
O&M	operations and maintenance

PSA	pressure swing absorption
PW	present worth
ROT	reactor outlet temperature
SMR	steam methane reforming
TCI	total capital investment
TEV	technical evaluation
WTW	well to wheel

C_1	cost of equipment with capacity q_1
C_2	cost of equipment with capacity q_2
C_k	capital expenditures
c_months	total number of months in the current modules construction period
CapF	capital breakdown per month
d_k	depreciation
E_k	cash outflows
i'	IRR
k	year
Mod	module/train being evaluated
ModF	capital fraction per module/train
month	current month in reactor/fossil construction period
Number	total number of reactor modules/fossil trains
n	exponential factor
q_1	equipment capacity
q_2	equipment capacity
R_k	revenues
t	tax rate
T_k	income taxes
y	exponent for current module/train

1. INTRODUCTION

This technical evaluation (TEV) has been prepared as part of a study for the Next Generation Nuclear Plant (NGNP) Project to evaluate the 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-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 high-temperature helium that can be used to produce electricity and/or process heat for export in the form of high-temperature helium or steam. A summary of these products and a brief description is shown in Table 1. This TEV specifically addresses HTGR integration opportunities for coal to liquids (CTL) and gas to liquids (GTL) production. For this study, an HTGR reactor outlet temperature (ROT) of up to 850°C is assumed. An ROT of 700°C is assumed for heat delivery to the GTL process based on a maximum process preheat temperature of 650°C and an assumed 25°C temperature approach for the gas to gas process heat exchangers. An ROT of 700°C was assumed for power generation, this reflects the economically optimal HTGR outlet temperature for a Rankine power cycle, as documented in TEV-988 (Idaho National Laboratory [INL] 2011c). Finally, an ROT of 850°C is assumed for heat delivery to the high-temperature steam electrolysis (HTSE) system for the CTL process, this ROT eliminates the need of co-firing fossil fuel in the HTSE process (INL 2010). In conventional chemical processes heat and power are generated by the combustion of fossil fuels such as coal and natural gas, resulting in significant emissions of greenhouse gases (GHGs), including 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.

Table 1. Projected outputs of the HTGR.

HTGR Product	Product Description
Process Heat	
High-Temperature Helium to HTSE	Delivered at 825°C and 7 MPa
High-Temperature Helium to GTL	Delivered at 675°C and 7 MPa
Electricity	Generated by a Rankine cycle, 43% efficiency

The HTGR would produce high-temperature heat and/or electricity and be physically located near the CTL or GTL production facility. A separate study has been conducted to assess heat losses associated with transporting HTGR heat long distances, using a variety of transport fluids, in TEV-1351 (INL 2011b). HTGR capital and operating costs used in the economic analysis are based on the detailed cost estimate presented in TEV-1196 (INL 2011a). A separate study should be conducted to assess the optimal siting of the HTGR with respect to the CTL and GTL facilities, balancing safety concerns associated with separation distance and heat losses associated with transporting high temperature heat long distances.

The CTL and GTL simulations were developed using version 7.3 of Aspen Plus, a state-of-the-art steady-state chemical process simulator (Aspen 2011). 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 both Aspen Plus and Excel. A detailed explanation of the software capabilities, of both Aspen Plus and Excel, is beyond the scope of this study. Similarly, this study assumes a familiarity with gasification, steam methane reforming (SMR), Fischer-Tropsch (FT) synthesis, product refining and upgrading, and common gas purification technologies. Hence, a thorough explanation of these technologies is considered to be beyond the scope of this TEV.

The TEV first presents an overview of the process modeling performed for the conventional and nuclear-integrated CTL and GTL cases. Afterwards, the process modeling results are presented for each case, specifically the impact of the HTGR integration. Next, the details of the economic model are discussed along with the analysis results. Following the economic modeling discussion, the method for calculating greenhouse gas emissions is discussed. Results for CTL, nuclear-integrated CTL, GTL, and nuclear-integrated GTL follow, with emphasis placed on impact of the HTGR integration. Finally, conclusions for CTL and GTL cases are discussed, separately.

2. PROCESS MODELING OVERVIEW

The plant models for the CTL and GTL processes were developed using version 7.3 of Aspen Plus (Aspen 2011). Because of the size and complexity of the processes modeled, the simulations were constructed using “hierarchy” blocks, a method for nesting one simulation within another simulation. In this fashion, submodels for each major plant section were constructed separately and then combined to represent the entire process. For the purpose of modeling, English units were used.

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. Utilities tracked in this manner for the CTL cases were electricity generated, electricity consumed, steam generated (medium pressure 700 psia, FT 300 psia, and low pressure 150 psia), steam consumed (medium pressure 700 psia, FT 300 psia, and low pressure 150 psia), and cooling water usage. Utilities tracked in this manner for the GTL cases were electricity generated, electricity consumed, steam generated (medium pressure 1500 psia, Fischer-Tropsch (FT) 300 psia, and low pressure 150 psia), steam consumed (medium pressure 1500 psia, FT 300 psia, and low pressure 150 psia), and cooling water usage.

Four cases were originally identified for modeling:

- Conventional CTL process
- Conventional GTL process with light gas recycle
- Nuclear-integrated CTL process
- Nuclear-integrated GTL process with light gas recycle

For the coal cases, a generic Illinois #6 coal was used as the feedstock. Illinois #6 was chosen as the coal type because it is a very commonly used and abundant coal. A dry-fed, entrained-flow, slagging gasifier (similar to a Shell, Uhde, or Siemens design) was selected as the gasification technology for this evaluation. Capacities for the coal cases were also set to produce 50,000 bpd of liquid products.

For the gas cases, natural gas composition was taken from data published by Northwest Gas Association. Capacity for the plant was set to produce 50,000 bpd of liquid products, including diesel, naphtha, and liquefied petroleum gas (LPG).

For the Aspen models described in this analysis, rigorous submodels of the nuclear power cycle and HTSE have not yet been integrated. Hence, in order to account for water usage, heat rejection for the HTSE system was calculated separately using the UNISIM modeling package. Cooling water requirements for this operation were then estimated and added to the overall Aspen model results. Water consumption for the HTGR 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 the liquid product is essentially unchanged between cases.

2.1 Conventional Coal to Liquids Case

The block flow diagram for the conventional CTL process is shown in Figure 1. The proposed process includes unit operations for air separation, coal milling and drying, coal gasification, syngas cleaning and conditioning, sulfur recovery, CO_2 compression/liquefaction, FT synthesis, product upgrading and refining, power production, cooling towers, and water treatment. Each unit operation is briefly described below.

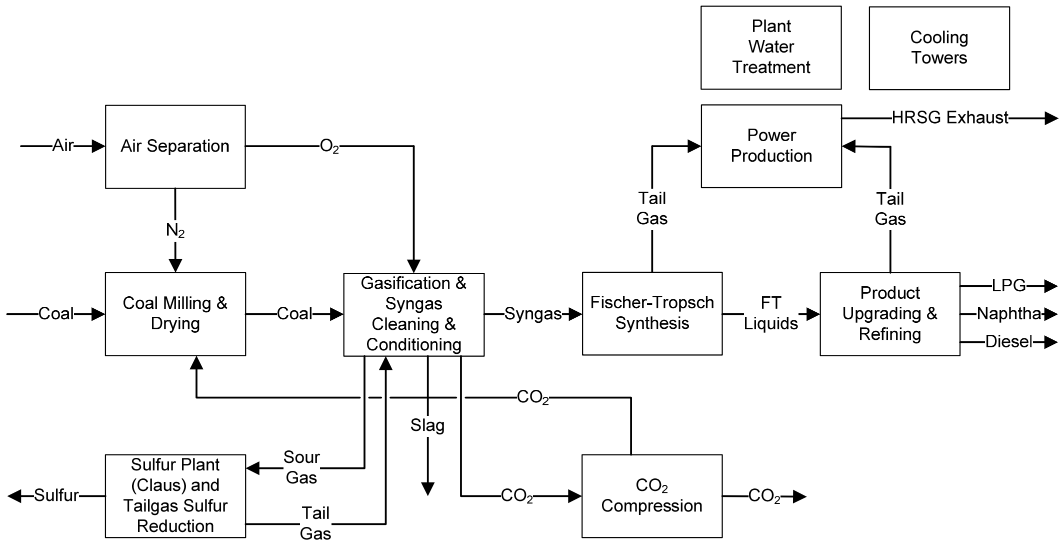


Figure 1. Block flow diagram for the conventional CTL process.

- **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. In order to reduce the inert content in the synthesis gas, an O_2 purity of 99.5% is specified. It should be noted that lower oxygen purity could be specified, such as 95%; however, the high purity oxygen is desired to minimize diluent nitrogen in the fuel synthesis loops. The nitrogen co-product from the ASU can be used for coal drying and transport, and as an inert gas to be used throughout the plant. The waste stream from the ASU is an O_2-enriched air stream. A portion of the enriched air stream is used as feed to the Claus unit in place of air (WorleyParsons 2002).

- **Coal Milling & 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 using heat recovered throughout the process. The 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 carbon dioxide, from the Rectisol unit, is then 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, for both the nitrogen and carbon dioxide transport gases. The amount of CO_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 Shell's SCGP technology (entrained-flow, dry-fed, slagging, oxygen-blown, upflow gasifier). Oxygen is fed to the gasifier to achieve an outlet temperature of 2,800°F, while steam (700 psia) is fed such that the molar 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 medium and FT pressure steam (Shell 2004). The syngas is further cooled by a water quench. A portion of the quenched syngas is returned to the top of the gasifier 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. 2.5% of the water from the slag quench loop is assumed to be sent to water treatment to avoid any buildup of contaminants.

- **Syngas Cleaning & Conditioning (GAS-CLN)** – After gasification, a fraction of the syngas is passed through sour shift reactors and then remixed with unshifted syngas to provide the optimal H_2:CO ratio to the FT reactors which utilize a cobalt catalyst; a ratio of approximately 2.1 H_2:CO. Steam (700 lb) 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. Five percent of the water recycled around the water gas shift loop is sent to water treatment to avoid high concentrations of ammonia and chloride compounds. Heat is further recovered from the syngas after shifting and used for nitrogen heating for coal drying and Rectisol heat requirements. Elemental mercury is then captured in a mercury guard bed. The syngas is further treated in an absorber with refrigerated methanol which acts as a physical solvent for the removal of CO_2, H_2S, and COS (Rectisol process). It is assumed that 1.5% CO_2 and less than 1 ppm of H_2S are present in the clean syngas stream. The H_2S rich stream is assumed to contain approximately 55% H_2S, with the remainder being CO_2 (Lurgi 2006). Gas containing H_2S from the sulfur reduction unit is also sent to the Rectisol process for sulfur removal, the nitrogen and argon contained in this stream are assumed to pass through to the CO_2 rich stream. It is also assumed that a steam reboiler, rather than nitrogen flow, is used for stripping in order to ensure a sufficiently pure CO_2 stream for sequestration or enhanced oil recovery. Utility usage is calculated based on values presented in literature for the Rectisol process (Cover 1986). However, confidence in the predicted utility usage is low due to the substitution of steam for nitrogen stripping. This substitution may significantly increase the power requirement for refrigeration and steam usage. Because of the extreme sulfur intolerance of the Fischer-Tropsch catalyst, guard beds are included as an added measure of protection against poisoning. A portion of syngas is sent to a pressure swing absorption unit (PSA), where a pure hydrogen stream is produced for use in the refinery, for hydrocracking and hydrotreating, and the sulfur reduction unit, to reduce sulfur compounds to H_2S. A portion of the PSA tailgas is sent to the sulfur reduction unit, where it is fired to provide heat for the reduction reactions, the remaining PSA tailgas is fired to provide heat in the refinery.

- **Sulfur Plant (CLAUS & S-REDUCT) –** Sulfur recovery is based on the Claus process. The Illinois coal has a sufficiently high sulfur content, which can create a sour 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 (Kohl 1997). 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 Rectisol 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 Compression (CO2-COMP) –** Carbon dioxide is removed from the syngas in the Rectisol process. By properly designing the solvent regeneration scheme, a pure stream of CO_2 is produced. The resulting stream is then compressed, along with the CO_2 recycle from coal transport, and liquefied prior to being pumped to the required pressure for use in enhanced oil recovery or sequestration. CO_2 for filtration is split from the CO_2 pressurization scheme at 700 psia, while the CO_2 for coal slurrying is split from the CO_2 pressurization scheme at 1,160 psia. Eight stages are assumed for the CO_2 compression scheme resulting in an overall efficiency of 84.4%. At 2,005 psia CO_2 should be liquid; however, Aspen's physical property methods do not predict the proper phase of the CO_2 stream because a small quantity of inert gas is present. The number of stages, stage efficiencies, and resulting power requirement were tuned to commercial CO_2 compression turbines; thus, the incorrect phase prediction will not impact the resulting power requirement.

- **Fischer-Tropsch Synthesis (FT) –** Syngas is converted to liquid synthetic crude in a slurry bubble column reactor utilizing a cobalt catalyst, a chain growth factor of 0.92 was assumed for the catalyst. Syngas flow to the reactor is preheated to the reaction temperature of 428°F. FT steam (300 psia) is generated from the exothermic FT reactions. The resulting product is primarily paraffinic, but also contains some olefins and oxygenates. The product distribution is estimated using a modified version of the Anderson Schulz Flory (ASF) distribution (Dry 2001). Modifications are required to the classical ASF distribution to better match actual performance of FT catalysts, especially for carbon numbers between one and four. Carbon chain length in the product stream varies from one (methane) to more than 100; hence, separations are performed to fractionate the product into light gas, crude naphtha, middle distillate, and molten wax. To improve conversion a light gas recycle is implemented. Currently a single-stage slurry bubble column reactor is modeled; however, a two-stage reactor may improve conversion and reduce the amount of light gas recycled. In addition, depending on column design, the steam pressure generated may have to be reduced below 300 psia.

- **Product Upgrading & Refining (REFINERY, HYDTREAT, HYDCRACK) –** The middle distillate product is hydrotreated to saturate olefinic bonds. The hydrotreated product is refined via a combination of pressurized and vacuum distillation into naphtha and diesel fuel products. The bottoms product from vacuum distillation and the molten wax stream are hydrocracked to improve overall yield of the diesel and naphtha fractions (Parkash 2003). The hydrotreating and hydrocracking operations are modeled as separate hierarchies within the refinery hierarchy. Hydrogen for hydrotreating and hydrocracking is supplied using pressure swing absorption, modeled in the gas cleaning hierarchy. A fraction of the light gasses produced are combusted to provide the heat required in the refining section, the remaining light gases are sent to LPG recovery. At present, no attempt is made to refine the naphtha fraction.

- **Power Production (GAS-TURB, ST-HRSG) –** Light gas from FT synthesis and refining areas is used to fire gas turbines to produce electricity. The gas turbine model is tuned to reflect actual turbine performance as modeled in GT-Pro (Thermoflow 2009). To increase power production, a combined cycle is utilized. Hot exhaust from the gas turbine is routed to the heat recovery steam generator (HRSG) to produce superheated steam. This steam is used in conventional condensing turbines to produce additional power. To further maximize power production, the medium (700 psia), FT (300 psia), and low pressure (150 psia) steam generated throughout the plant are sent to the power production block where they are passed through three saturated steam turbines. The efficiencies of the turbines for the various steam pressures were calculated using Steam Pro, steam turbine modeling software from Thermoflow (2009). It was found that even given low quality steam at 150 psia, efficiencies for the saturated steam turbines remain constant at approximately 80%. The condensed steam 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. FT steam is added to the deaerator to achieve the appropriate dew point temperature. 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 use 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. It is anticipated that energy consumption for the water treatment portion of the plant could change considerably based on input from a water treatment vendor. 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 Nuclear-Integrated Coal to Liquids Case

The block flow diagram for the nuclear-integrated CTL case is shown in Figure 2. The proposed process includes the same unit operations as the conventional process with the following exceptions: the cryogenic ASU and water gas shift reactors are replaced by high-temperature steam electrolysis to provide oxygen and hydrogen for the process.

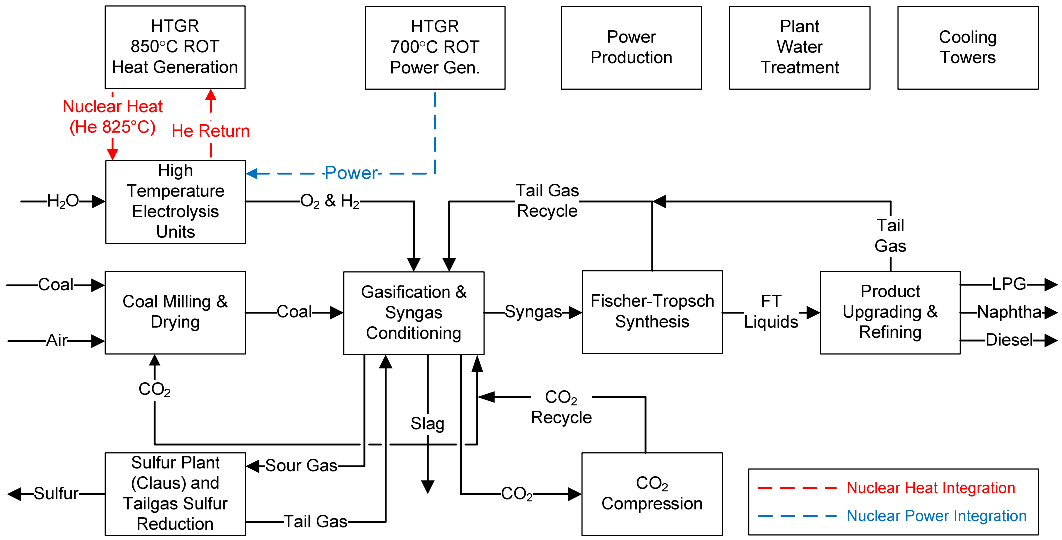


Figure 2. Block flow diagram for the nuclear-integrated CTL process.

While developing the nuclear-integrated case, opportunities for heat integration between the nuclear, electrolysis, and fossil plants were evaluated; however, very few opportunities were identified. The primary reason for this conclusion was that the fossil plant produced an excess of heat that could provide for the heat requirements within the fossil portion of the plant. In a few instances (notably product refining), it was believed that nuclear heat could displace burning of light gas to reduce overall plant greenhouse gas emissions. However, the modeling analysis indicated that full light gas recycle would lead to unacceptable buildup of inert gases in the process. Hence, it was deemed practical to use this gas as fuel rather than develop complex schemes to separate inerts from the light gas.

In the previous revision of this TEV an upper limit on the HTGR ROT of 750°C required syngas firing for topping heat in the HTSE process. However, this upper limit was lifted for this revision, and topping heat is no longer required.

With the ASU and water gas shift reactors removed from the flowsheet, an unexpected result was observed. A shortage of inert gas for use in coal drying, transport, and feeding was created. To overcome this issue, air was selected for use in coal drying and transport, rather than nitrogen.

Each unit operation in the nuclear-integrated CTL flowsheet is briefly described below. Because the majority of unit operations remain unchanged from the conventional CTL flowsheet, emphasis is placed on differences in configuration between the two cases.

- **Electrolysis (ELEC)** – Water is converted to hydrogen and oxygen utilizing high-temperature electrolysis units. Helium at 1,517°F, provided by the HTGR, is used to convert the water to steam and raise the temperature to 1,472°F for electrolysis. Conversion and power consumption are based on information presented in TEV-981 (INL 2010). The oxygen generated is used for gasification and air enrichment for the Claus and sulfur reduction units, the hydrogen is used to adjust the hydrogen to carbon monoxide ratio for the FT reactions, in place of sour shift reactors.

- **Coal Milling & Drying (CMD) –** Coal milling and drying for the nuclear-integrated case is similar to the conventional case. However, because nitrogen is not readily available in this scenario, coal drying is accomplished using air; the airflow for drying is specified to be 2.5 times the coal flowrate (Mullinger 2008). Air is also used as transport gas for the pulverized coal. Although air is used industrially for coal drying and transport, it introduces additional flammability issues compared to using an inert gas for this purpose. Transport of coal into the gasifier is accomplished using CO_2 recovered from the Rectisol unit. The air for drying is heated using heat recovered throughout the process.

- **Gasification (GASIFIER) –** Gasification for the nuclear-integrated case is similar to the conventional case. However, because hydrogen is supplied externally from the electrolyzers rather than shifting the syngas, the gasification island throughput is reduced to 35% of the conventional design to produce the same amount of liquid fuel product.

- **Syngas Cleaning & Conditioning (GAS-CLN)** – Syngas cleaning is greatly simplified for the nuclear-integrated case, because the water gas shift reactors are eliminated. Hydrogen from the electrolyzers is added to the syngas to achieve the optimal H_2:CO of approximately 2.1 for the cobalt FT catalyst. When the shift reactors are eliminated, the CO_2 concentration entering the Rectisol unit is reduced from 30 mol.% in the conventional case to 10 mol.% in the nuclear-integrated case. Similarly, CO_2 concentration in the purified syngas is reduced from 1.3 mol.% in the conventional case to 0.1 mol.% in the nuclear-integrated case. Rectisol capacity and utility usage are reduced by more than half in the nuclear-integrated case as compared to the conventional case.

- **Sulfur Plant (CLAUS & S-REDUCT) –** The Claus and sulfur reduction plants for the nuclear-integrated case are similar to those in the conventional case. However, as with the gasification island, the required capacity of these units is approximately less than half that of the conventional case configuration.

- **CO_2 Compression (CO2-COMP) –** CO_2 compression for the nuclear-integrated case is similar to CO_2 compression in the conventional case. However, when the shift reactors are eliminated, required capacity and utility usage are reduced by a factor of approximately seven. Additionally, the last stage of compression is removed, as all CO_2 is recycled to the gasifier to increase carbon conversion to the liquid product.

- **Fischer-Tropsch Synthesis (FT) –** The FT synthesis plant remains unchanged between the conventional and nuclear-integrated cases. Inlet gas composition is slightly different between the cases because of increased N_2 in the nuclear-integrated case from the recycle of CO_2 back to the gasifier.

- **Product Upgrading & Refining (REFINERY, HYDTREAT, HYDCRACK) –** The product refining and upgrading process in the nuclear-integrated case remains essentially unchanged from the process in the conventional case.

- **Power Production (ST) –** Power production in the nuclear-integrated case changes because the gas turbine system is removed, since the light gases are recycled to the gasification island. As a result there is no longer hot tailgas to superheat steam to use in the condensing steam turbines. Only the saturated turbines remain, being fed the medium pressure (700 psia), Fischer-Tropsch (300 psia), and low pressure (150 psia) steam generated throughout the plant. Due to size reductions in some portions of the plant, the capacity of the steam system in the nuclear-integrated case is approximately 60% of the conventional case. The saturated steam turbines are also smaller in the nuclear-integrated case, approximately 80% of the conventional case capacity.

- **Cooling Towers (COOL-TWR) –** The cooling water system requirements are similar for both cases. Again, cooling water requirements for the HTGR are not included in this analysis.

- **Water Treatment (H2O-TRTM) –** The water treatment system in the nuclear-integrated case is similar to the conventional case. No further comparison will be made on water treatment between the two cases until the water treatment hierarchy has been refined.

2.3 Conventional Natural Gas to Liquids Case

The block flow diagram for the conventional GTL process is shown in Figure 3. The proposed process includes unit operations for air separation, natural gas purification and reforming, FT synthesis, product upgrading and refining, power production, cooling towers, and water treatment. Because many unit operations remain unchanged from the conventional CTL flowsheet, emphasis is placed on differences in configuration between the natural gas and coal cases.

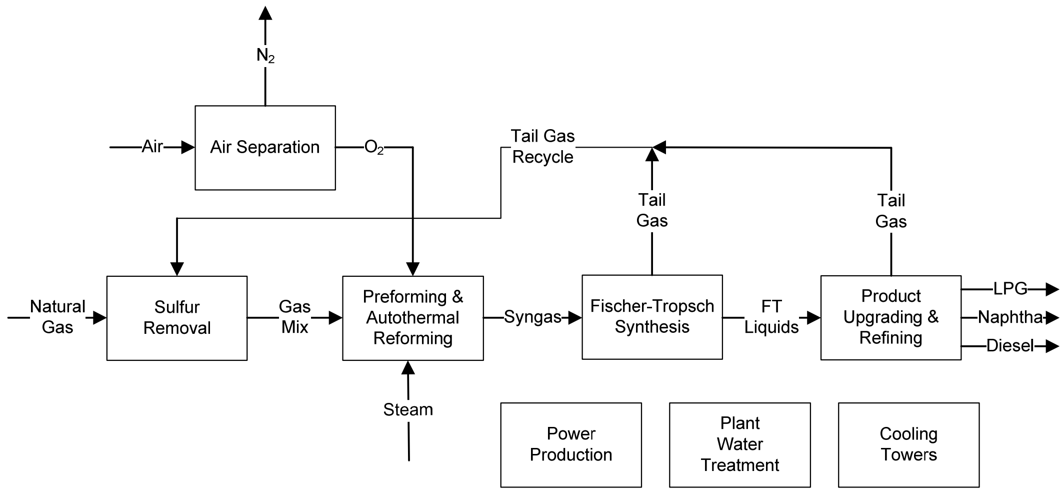


Figure 3. Block flow diagram for the conventional GTL process.

- **Air Separation (ASU) –** Air separation in the conventional GTL case is identical to that of the conventional CTL case. However, because the natural gas flowsheets do not require coal drying, the N_2 product from the ASU could be available for sale as a byproduct. However, the amount of nitrogen produced in the GTL scenarios would potentially saturate the industrial nitrogen market; as a result revenues from sales were not included in the economic model.

- **Natural Gas Purification and Reforming (NG-RFMR)** – Two reforming scenarios were considered: autothermal reforming and two-step reforming consisting of primary steam reforming followed by secondary autothermal reforming. Although two-step reforming appears to offer the best opportunity for nuclear heat integration, the steam to carbon ratio entering the primary reformer is too low for commercial operation and whisker carbon formation would occur (Pedersen 2010). As a result, only autothermal reforming was assumed for the GTL process. The desired syngas H_2:CO ratio for the FT reactors, which utilize a cobalt catalyst, is approximately 2.1. This ratio was achieved by setting the steam to carbon inlet molar ratio to 0.92 and the exit temperature of the autothermal reformer to 1,870°F (1,021°C).

  Natural gas and the light gas recycle are first compressed to 500 psia, saturated with water, then preheated to 750°F and passed through a hydrotreater and sulfur removal bed. Hydrotreating will break down any olefins present in the light gas recycle, which would cause operational issues in the preformer. The gas is then heated further and mixed with steam (1,500 psia) to achieve the desired H_2:CO ratio downstream of the autothermal reformer (Pedersen 2010). The hot natural gas stream is then fed to a preformer that irreversibly converts C_2+ hydrocarbons to CH_4, CO, H_2, and CO_2. The preforming step is required, as further heating of the natural gas and steam could result in steam cracking of the C_2+ components to olefins, which tend to form carbon in the autothermal reformer. Carbon formation is detrimental to long-term operation, as it deactivates the reforming catalyst.

The effluent from the preformer is preheated to 1,202°F (650°C) mixed with oxygen and fed to an autothermal reformer. The outlet temperature is set at 1,870°F, which results in an oxygen to carbon molar ratio of 0.57 and a steam to carbon ratio of 0.94. The steam to carbon ratio in the autothermal reformer is sufficiently high to avoid the formation of whisker carbon. The hot gas from the outlet of the autothermal reformer is quickly cooled and produces medium and FT pressure steam, followed by water removal in a quench. Finally, a portion of syngas is sent to a pressure swing absorption unit, where a pure hydrogen stream is produced to use in the refinery for hydrocracking and hydrotreating. The tailgas stream is remixed with the main syngas stream. The resulting syngas has a H_2 to CO ratio of 2.1, contains 8.0 mol.% CO_2, and contains 8.8 mol.% inerts.

A portion of the light gas recycled is fired and used for preheating the inlet syngas, water, and steam for hydrotreating, preforming, and autothermal reforming.

- **Fischer-Tropsch Synthesis (FT) –** FT synthesis in the conventional GTL case is identical to that of the conventional CTL case.

- **Product Upgrading & Refining (REFINERY, HYDTREAT, HYDCRACK) –** Product upgrading and refining in the conventional GTL case is identical to that of the conventional CTL case

- **Power Production (ST) –** Power production in the conventional GTL case differs slightly from the conventional CTL case. Since light gases are recycled to the steam methane reformer tailgas is no longer fired in a gas turbine, and therefore no longer produces hot tailgas used to superheat steam for the condensing steam turbines. Only the saturated steam turbines are used to generate power. Furthermore, the medium pressure steam generated in the GTL case is 1,500 psia, rather than 700 psia.

- **Cooling Towers (COOL-TWR) –** The cooling towers in the conventional GTL case are modeled identically to those in the conventional CTL case.

- **Water Treatment (H2O-TRTM) –** Water treatment in the conventional GTL case is identical to that of the conventional CTL case.

2.4 Nuclear-Integrated Natural Gas to Liquids Case

The block flow diagram for the nuclear-integrated GTL case is shown in Figure 4. The proposed process includes the same unit operations as the conventional process with the following except nuclear heat is used for preheating in the reforming section and reboiler duty in the refining section rather than burning light gas.

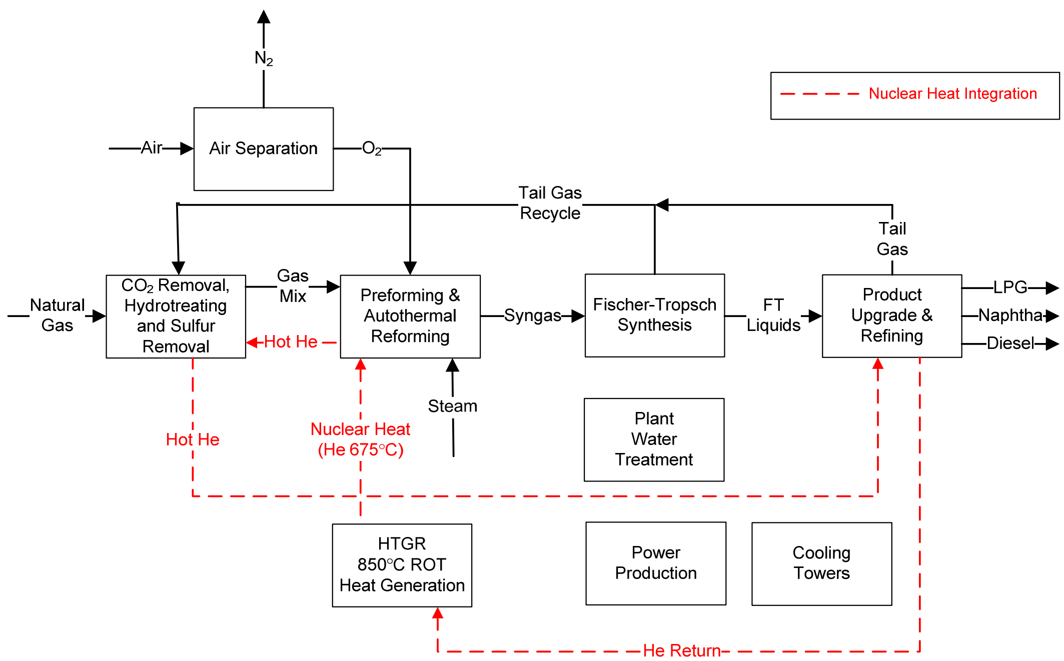


Figure 4. Block flow diagram for the nuclear-integrated GTL process.

It should be noted, that a full light gas recycle would lead to unacceptable buildup of inert gases in the process. Hence, it was deemed practical to fire a small portion of the tailgas recycle to minimize inert gas buildup. The fraction fired, was too small to adequately displace the heat provided by the HTGR, as a result steam is generated instead.

Each unit operation in the nuclear-integrated GTL flowsheet is briefly described below. Because the majority of unit operations remain unchanged from the conventional GTL flowsheet, emphasis is placed on differences in configuration between the two cases.

- **Air Separation (ASU)** – Air separation in the nuclear-integrated cases is identical to that of the conventional case.

- **Natural Gas Purification and Reforming (NG-RFMR)** – Conditions in the reforming section of the plant are nearly identical to those of the conventional case, excluding the fact that nuclear heat is used to provide the heat for all preheat streams and the light gas recycle must be treated for CO_2 removal to avoid a buildup of inert gases. CO_2 is partially removed using Fluor's propylene carbonate solvent given its low solubility of light hydrocarbons and nitrogen (BRE 2008). The steam to carbon ratio is 0.50 for the autothermal reformer. To achieve the 1,870°F outlet temperature on the autothermal reformer, an oxygen to carbon molar ratio of 0.54 was required. The resulting syngas has a H_2:CO ratio of 2.1, contains 4.5 mol.% CO_2, and contains 9.8 mol.% inerts.

- **CO_2 Compression (CO2-COMP) –** CO_2 compression is not present in the conventional GTL case; however it is required if the pure CO_2 stream is to be sequestered. CO_2 compression for the nuclear-integrated case is similar to CO_2 compression in the conventional CTL case. However, the required capacity and utility usage are reduced significantly. Additionally, the CO_2 off-take splits are removed as the natural gas reforming section does not require CO_2.

- **Fischer-Tropsch Synthesis (FT) –** The FT synthesis plant remains unchanged between the conventional and nuclear-integrated cases. Inlet gas composition is slightly different between the cases due to the substitution of nuclear heat in the reforming section.

- **Product Upgrading & Refining (REFINERY, HYDTREAT, HYDCRACK) –** The product refining and upgrading process in the nuclear-integrated case remains unchanged from the process in the conventional case, except that nuclear heat provides the reboiler heat duties.

- **Power Production (ST) –** Steam generation and power production in the nuclear-integrated case is identical to that of the conventional case.

- **Cooling Towers (COOL-TWR) –** The cooling towers in the nuclear-integrated case is identical to that of the conventional case.

- **Water Treatment (H2O-TRTM) –** The water treatment system in the nuclear-integrated case is similar to the conventional case.

3. PROCESS MODELING RESULTS

Analysis of the conventional CTL case indicated a potential need for hydrogen supplementation from HTSE. By supplementing the process with an external hydrogen source, the need to "shift" the syngas using conventional water-gas shift reactors was eliminated. The primary benefit of this change is a reduction in greenhouse gas emissions from the process. It was also determined that the conventional CTL case produced heat beyond what was needed to support demands of the plant. Based on these observations, a nuclear-integrated model was developed which focuses primarily on integrating nuclear hydrogen rather than nuclear heat.

Analysis of the conventional GTL case indicates a strong potential heat integration opportunity for an HTGR. In the conventional case, light gases are burned to provide heat to the reforming and refinery processes. Both the conventional and nuclear-integrated cases assume recycling of light gas back to the reformer.

Results from the nuclear-integrated CTL case indicate that integration of nuclear hydrogen can improve carbon utilization and reduce GHG emissions. Coal consumption is decreased by 65% using electrolysis and nuclear power as the hydrogen source. Similarly, with nuclear-integration the fraction of carbon in the coal partitioned to the liquid fuel products increases from 32 to 92%. Integrating nuclear power and high temperature steam electrolysis also decreases CO_2 emissions from the plant. If carbon capture and sequestration are assumed for the conventional configuration, CO_2 emissions decrease by 83% when electrolysis and nuclear power are utilized. However, if carbon capture and sequestration are not assumed for the conventional configuration, CO_2 emissions decrease by 96% when electrolysis and nuclear power are utilized. In the nuclear-integrated case, nuclear energy is used to offset a portion of the energy requirement derived from coal. This is evident, as power consumption is increased from producing 220 MWe to consuming 2,348 MWe. It is estimated that one 664 MWt 850°C ROT HTGR for heat production and nine 604 MWt 700°C ROT HTGRs for power production would be required to support production of 50,000 bbl/day of liquid fuel products. Water consumption for the HTGR has not been included, as a detailed water balance for the HTGR has not been completed.

Results for the nuclear-integrated natural gas to liquids case look promising. Approximately one 450 MW_t 700°C ROT HTGR would be required to support this configuration. In addition, the reactor would supply only heat to the fossil process, as more power is generated in the process than is required. By substituting nuclear heat for light gas combustion for preheat in the reformer and reboiler duty in the refinery; natural gas consumption is decreased by 9%. Power production for the plant decreases by 8% for the nuclear-integrated case. CO_2 emissions from the plant also decrease by integrating HTGRs into the flowsheet. CO_2 emissions decrease by 42% when sequestration is not assumed and by 88% if the pure CO_2 stream is sequestered in the nuclear-integrated GTL case. Water consumption for the HTGR has not been included, as a detailed water balance for the HTGR has not been completed.

A summary of the modeling results for all cases is presented in Table 2. A high-level material and energy balance summary for each case is graphically presented in Figure 5. The conventional coal and natural gas cases serve as a basis for comparison with the nuclear-integrated cases. For the complete Aspen stream results for the CTL and nuclear-integrated CTL cases, see Appendixes B and C, for GTL and nuclear-integrated GTL see Appendixes D and E.

Table 2. CTL and GTL modeling case study results.

	Conventional CTL	Nuclear Integration CTL	Conventional GTL	Nuclear Integration GTL
Inputs				
Coal Feed rate (ton/day)	26,941	9,354	N/A	N/A
Natural Gas Feed Rate (MMSCFD)[1]	N/A	N/A	427	390
% Carbon to Liquid Product	31.8%	91.7%	71.9%	79.3%
# HTGRs (600 MW_t)	N/A	10.17	N/A	0.75
Outputs				
Total Liquid Products (bbl/day)t	*50,002*	*50,002*	*49,994*	*49,998*
Diesel	35,587	35,194	34,581	35,410
Naphtha	12,259	11,810	11,892	11,674
LPG	2,156	2,998	3,521	2,914
Utility Summary				
Total Power (MW)	*220.3*	*-2,347.8*	*66.6*	*69.7*
Power Consumed	-739.7	-2,749.4	-330.1	-402.3
Electrolyzers	N/A	-2,511.8	N/A	N/A
Secondary Helium Circulator	N/A	-23.0	N/A	-48.4
ASU	-301.3	N/A	-132.7	-131.3
Coal Milling and Drying	-13.8	-9.5	N/A	N/A
Natural Gas Reforming	N/A	N/A	-68.0	-68.9
Gasification and Gas Cleanup	-174.7	-82.1	N/A	N/A
CO_2 Compression/Liquefaction	-140.8	-19.6	N/A	-11.7
Fischer Tropsch & Refining Processes	-40.9	-45.7	-53.8	-60.3
Refrigeration	-24.0	-26.2	-41.5	-47.1
Cooling Tower	-26.6	-18.5	-18.8	-20.8
Water Treatment	-17.6	-13.0	-15.4	-13.9
Power Generated	960.0	401.7	396.7	471.9
Gas Turbine	300.0	N/A	N/A	N/A
Condensing Turbines	178.6	N/A	N/A	N/A
Saturated Turbines	481.4	401.7	396.7	471.9
Water Requirements[2]				
Water Consumed (gpm)	20,856	15,454	13,790	14,552
Water Consumed/lb Feed (lb/lb)	4.65	9.92	8.55	9.86
Water Consumed/bbl Product (bbl/bbl)	14.3	10.6	9.5	10.0

Table 2. CTL and GTL modeling case study results.

	Conventional CTL	Nuclear Integration CTL	Conventional GTL	Nuclear Integration GTL
CO_2 Summary				
Total CO_2 Produced (ton/day)	*40,046*	*1,473*	*7,164*	*4,190*
Emitted	8,803	1,473	7,164	841
Capturable	31,243	N/A	N/A	3,349
Nuclear Integration Summary				
Electricity (MW)	*N/A*	*-2,643.0*	*N/A*	*-13.9*
HTSE	N/A	-2,511.8	N/A	N/A
HTGR House Loads	N/A	-295.2	N/A	-13.9
Balance of Fossil Plant	N/A	164.0	N/A	N/A
Electrolysis Heat (MMBTU/hr)	*N/A*	*2408.7*	*N/A*	*N/A*
From Nuclear Plant	N/A	2330.2	N/A	N/A
From Secondary Circulator	N/A	78.5	N/A	N/A
Electrolysis Products				
Total Hydrogen (ton/day)	N/A	1,957	N/A	N/A
Total Oxygen (ton/day)	N/A	15,430	N/A	N/A
Used in Plant (ton/day)	N/A	9,198	N/A	N/A
Excess (ton/day)	N/A	6,232	N/A	N/A
HTGR Heat Use (MMBTU/hr)	*N/A*	*N/A*	*N/A*	*1,633*
Reformer	N/A	N/A	N/A	1,057
Refinery	N/A	N/A	N/A	741
From Secondary Circulator	N/A	N/A	N/A	-165

[1]Standard temperature of 60 degrees F.

[2]Does not include water usage for HTGR.

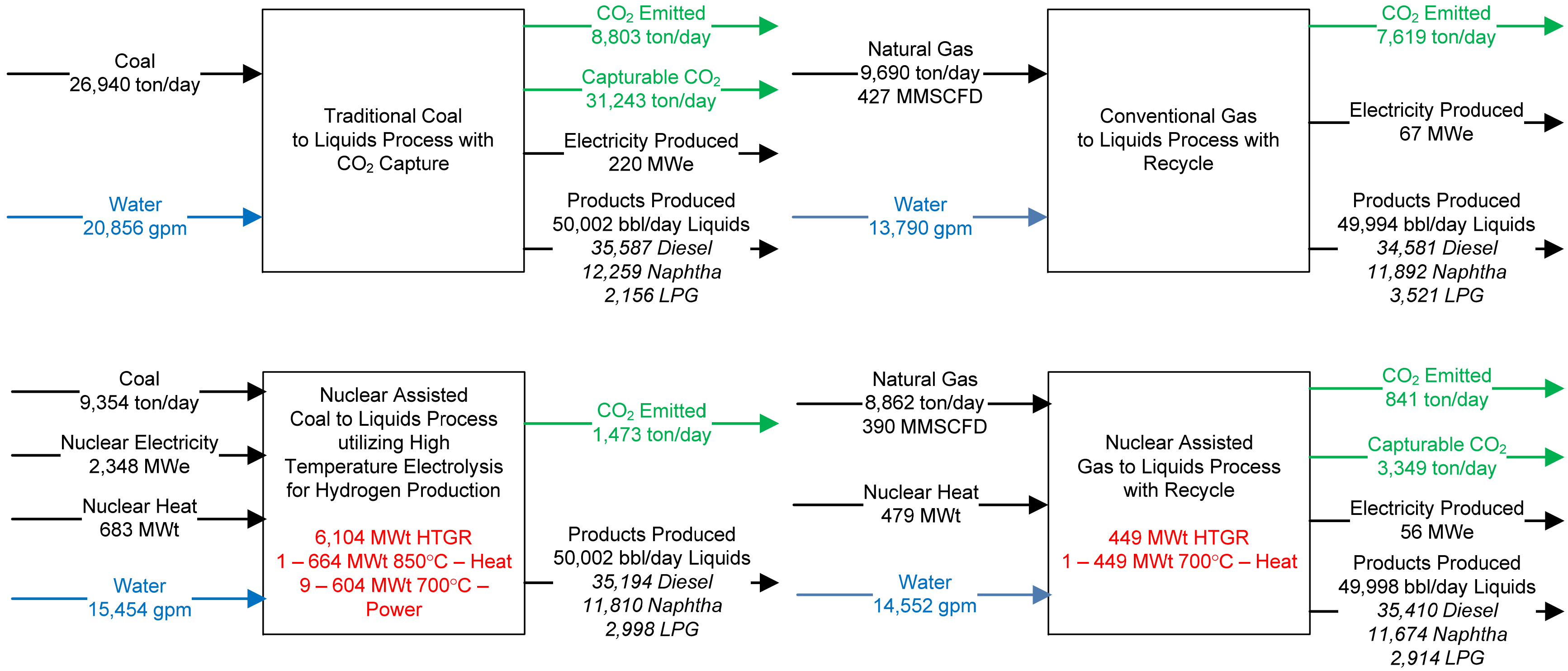


Figure 5. CTL and GTL modeling case material balance summary.

4. ECONOMIC MODELING OVERVIEW

The economic viability of the CTL and GTL processes was assessed using standard economic evaluation methods, specifically the internal rate of return (IRR). The economics were evaluated for the conventional and nuclear-integrated cases 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 was then calculated based on the annual after tax cash flows. The following sections describe the methods used to calculate the capital costs, annual revenues, annual manufacturing costs, and the resulting economic results. For the economics it is assumed that the primary selling product is diesel. The economics were analyzed for multiple owner operator scenarios, with the HTGR and synthetic fuel facilities operated by independent organizations or a single owner/operator. The economic results are preliminary and should be refined as the design of the HTGR progresses, if the design of the HTGR is changed significantly, or if additional refinements of the HTGR and/or CTL/GTL capital and/or operating costs become available.

4.1 Capital Cost Estimation

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 4 estimate, which has a probable error of -30% and +50%, as described in TEV-1196 (INL 2011a). A Class 4 estimate is associated with a feasibility study or top-down cost estimate and has one to fifteen percent of full project definition (AACE 2005).

Equipment items for this study were not individually priced. Rather, cost estimates were based on scaled costs for major plant processes from published literature. Cost estimates were generated for coal preparation, the ASU, gasification, gas cleanup, FT synthesis, product refining and upgrading, gas turbines, steam turbines, the HRSG, cooling towers, HTSE electrolysis, and the HTGRs for the CTL scenarios. Cost estimates were generated for SMR, the ASU, FT synthesis, product refining and upgrading, steam turbines, the HRSG, and the HTGR for the GTL scenarios. In some instances, several costs were averaged. Gas cleanup includes costs for water-gas-shift reactors, the Rectisol process, sulfur recovery, and CO_2 compression/liquefaction for CTL. Gas cleanup is not necessary in the GTL flowsheets, except for CO_2 compression/liquefaction when sequestration is assumed for the nuclear-integrated case.

The installed 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:

$$C_2 = C_1\left(\frac{q_2}{q_1}\right)^n \tag{1}$$

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.

The HTGR installed capital costs are based on the capital cost correlations presented in Section 2.6 of TEV-1196 for an n^{th} of a kind HTGR, a mature commercial installation. Preconstruction costs, balance of equipment costs, indirect costs, and project contingencies were added in accordance with the costs outlined in Sections 2.1 through 2.5 of TEV-1196 (INL 2011a).

Cost indices were used to adjust equipment prices from previous years to 2010 values using the Chemical Engineering Plant Cost Index (CEPCI) as depicted in Table 3.

Table 3. CEPCI data.

Year	CEPCI	Year	CEPCI
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	2009	521.9
2000	394.1	2010	550.8

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, based on information provided in studies performed by the National Energy Technology Laboratory (NETL) (2000). These factors were not added to the cost of the HTGR, as the cost basis for the HTGR 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 for the fossil processes. The capital cost correlations used for the HTGR includes all engineering fees and contingencies; therefore, these factors were not applied to this cost.

Based on the AACE International contingency guidelines 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%. 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%. Eighteen percent was selected based on similar studies conducted by NETL (2007).

Table 5 and Figure 6 presents the capital cost estimate breakdown for the conventional CTL case, Table 6 and Figure 7 for the nuclear-integrated CTL case, Table 7 and Figure 8 for the conventional GTL case, and Table 8 and Figure 9 for the nuclear-integrated GTL case. Capital costs are presented assuming no CO_2 sequestration; however, cases that have sequestration as an option list the differential TCI that would be required to include CO_2 sequestration, i.e. compression and/or liquefaction equipment.

Table 5. Total capital investment, conventional CTL case.

	Installed Cost	Engineering Fee	Contingency	Total Capital Cost
Coal Preparation	$294,826,984	$29,482,698	$58,375,743	$382,685,426
ASU	$412,284,613	$41,228,461	$81,632,353	$535,145,428
Gasification	$948,158,150	$94,815,815	$187,735,314	$1,230,709,279
Gas Cleaning	$811,266,409	$81,126,641	$160,630,749	$1,053,023,798
FT Reactors & Refining	$355,434,504	$35,543,450	$70,376,032	$461,353,986
Gas Turbines	$76,258,421	$7,625,842	$15,099,167	$98,983,430
Steam Turbines	$143,343,132	$14,334,313	$28,381,940	$186,059,385
HRSG	$51,579,237	$5,157,924	$10,212,689	$66,949,850
Cooling Towers	$9,985,833	$998,583	$1,977,195	$12,961,611
Water Systems	$220,322,747	$22,032,275	$43,623,904	$285,978,926
Piping	$220,322,747	$22,032,275	$43,623,904	$285,978,926
I&C	$80,681,569	$8,068,157	$15,974,951	$104,724,677
Electrical Systems	$248,250,983	$24,825,098	$49,153,695	$322,229,775
Buildings & Structures	$285,488,630	$28,548,863	$56,526,749	$370,564,242
Total Capital Investment				**$5,397,348,737**
Differential for Adding CO_2 Sequestration				*$33,564,727*

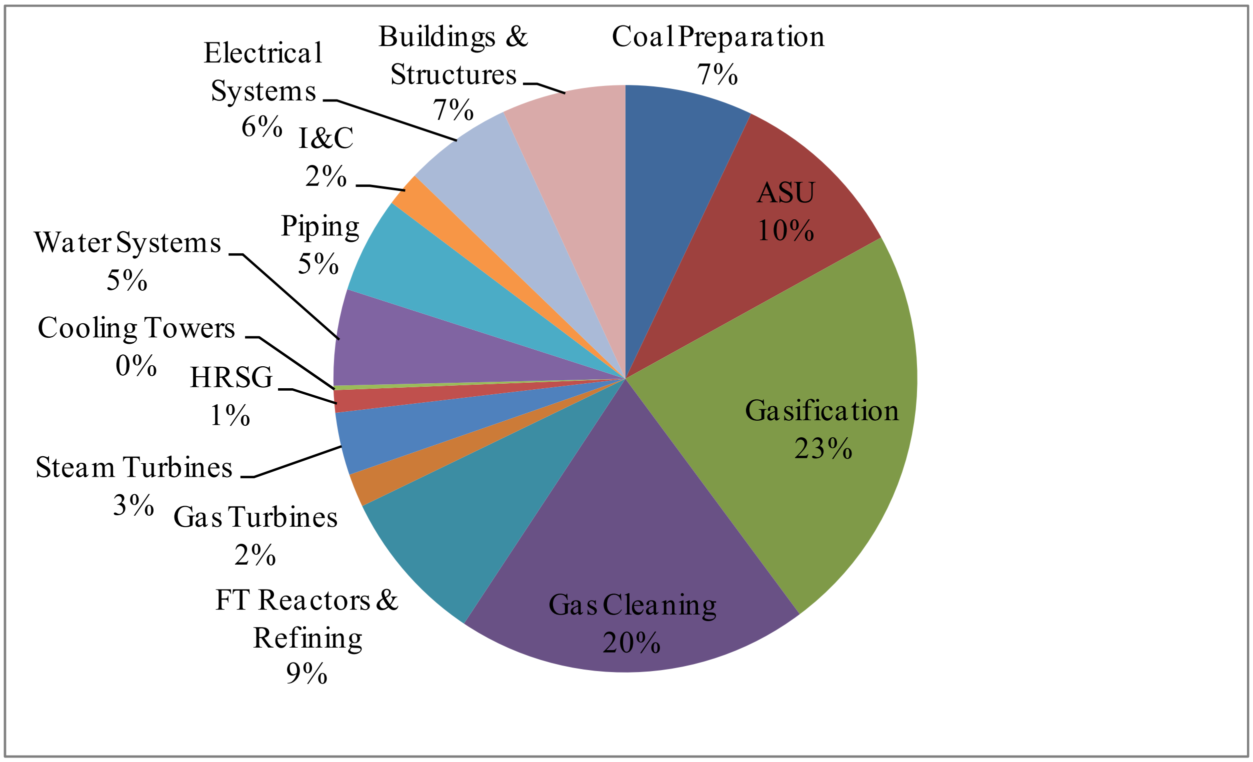


Figure 6. Total capital investment, conventional CTL case, no sequestration.

Table 6. Total capital investment, nuclear-integrated CTL case.

	Installed Cost	Engineering Fee	Contingency	Total Capital Cost
850°C ROT HTGR(s)	$858,289,406	Included	Included	$858,289,406
700°C ROT HTGR(s)	$6,673,774,875	Included	Included	$6,673,774,875
Power Cycles	$2,575,261,279	Included	Included	$2,575,261,279
HTSE	$742,126,119	$74,212,612	$146,940,972	$963,279,703
Coal Preparation	$111,361,310	$11,136,131	$22,049,539	$144,546,980
Gasification	$360,189,281	$36,018,928	$71,317,478	$467,525,687
Gas Cleaning	$355,702,237	$35,570,224	$70,429,043	$461,701,504
FT Reactors and Refining	$362,827,302	$36,282,730	$71,839,806	$470,949,838
Steam Turbines	$106,441,282	$10,644,128	$21,075,374	$138,160,784
HRSG	$9,315,065	$931,507	$1,844,383	$12,090,955
Cooling Towers	$25,254,070	$2,525,407	$5,000,306	$32,779,783
Water Systems	$147,198,383	$14,719,838	$29,145,280	$191,063,502
Piping	$147,198,383	$14,719,838	$29,145,280	$191,063,502
I&C	$53,903,633	$5,390,363	$10,672,919	$69,966,916
Electrical Systems	$165,857,333	$16,585,733	$32,839,752	$215,282,819
Buildings and Structures	$190,735,933	$19,073,593	$37,765,715	$247,575,242
Total Capital Investment				**$13,713,312,773**
HTGR and Power Cycle				$3,605,987,213
CTL Process				$10,107,325,559

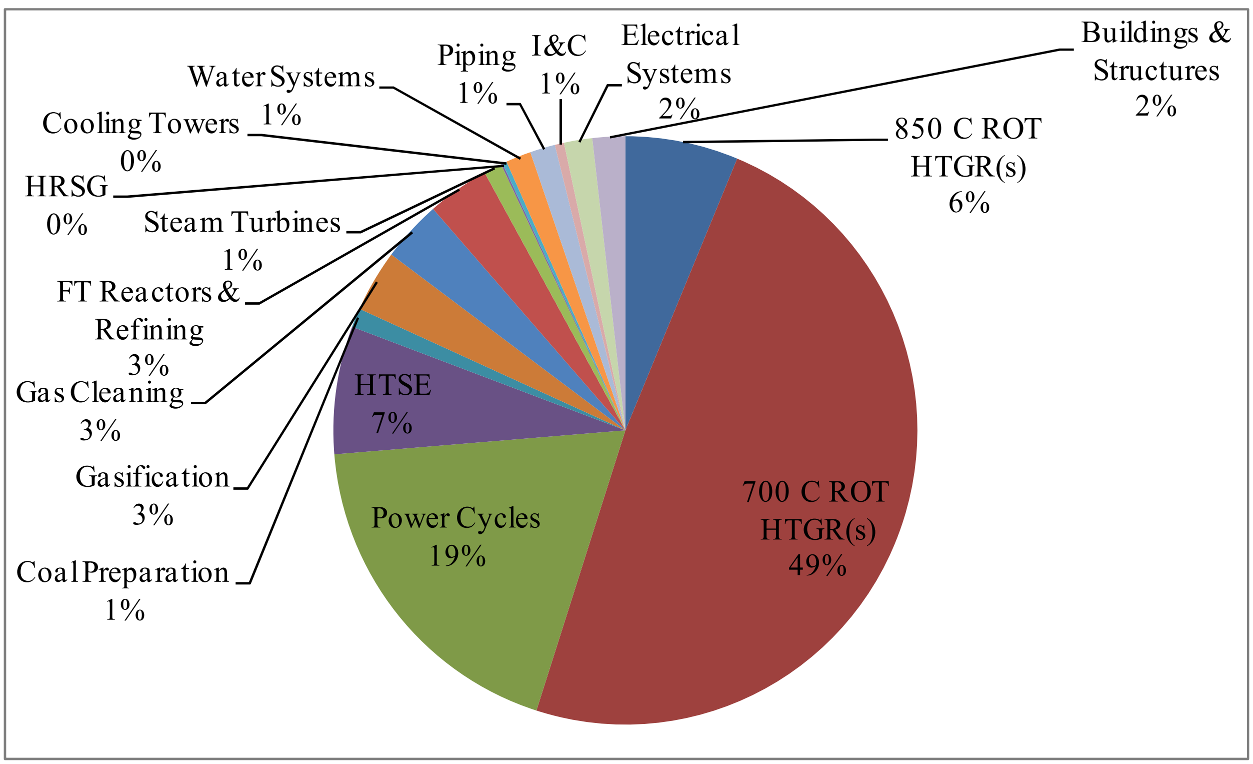


Figure 7. Total capital investment, nuclear-integrated CTL case.

Table 7. Total capital investment, conventional GTL case.

	Installed Cost	Engineering Fee	Contingency	Total Capital Cost
ASU	$258,831,117	$25,883,112	$51,248,561	$335,962,790
Autothermal Reforming	$349,828,953	$34,982,895	$69,266,133	$454,077,981
FT Reactors & Refining	$414,248,152	$41,424,815	$82,021,134	$537,694,101
Steam Turbines	$105,644,360	$10,564,436	$20,917,583	$137,126,380
HRSG	$9,848,554	$984,855	$1,950,014	$12,783,423
Cooling Towers	$25,355,761	$2,535,576	$5,020,441	$32,911,778
Water Systems	$82,626,740	$8,262,674	$16,360,094	$107,249,508
Piping	$82,626,740	$8,262,674	$16,360,094	$107,249,508
I&C	$30,257,679	$3,025,768	$5,991,021	$39,274,468
Electrical Systems	$93,100,552	$9,310,055	$18,433,909	$120,844,516
Buildings & Structures	$107,065,635	$10,706,563	$21,198,996	$138,971,194
Total Capital Investment				**$2,024,145,646**

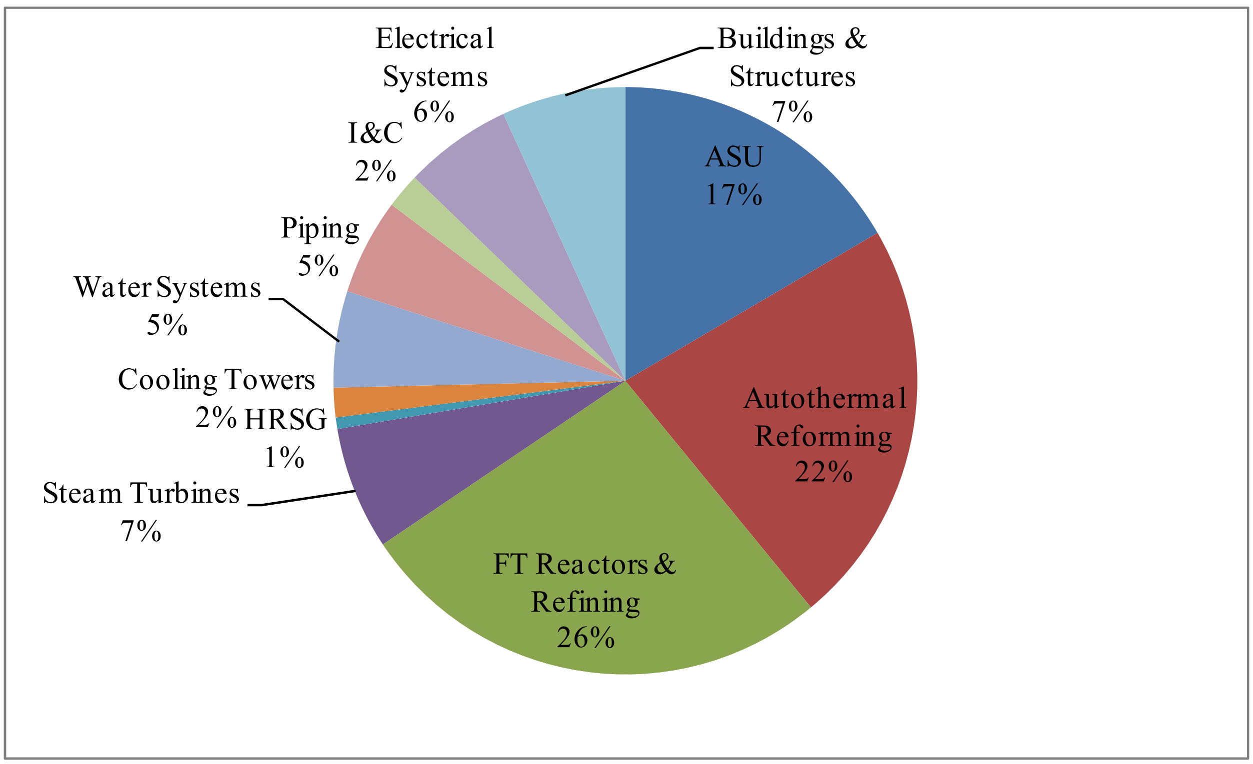


Figure 8. Total capital investment, conventional GTL case.

Table 8. Total capital investment, nuclear-integrated GTL case.

	Installed Cost	Engineering Fee	Contingency	Total Capital Cost
700°C ROT HTGR(s)	$1,006,875,557	Included	Included	$1,006,875,557
ASU	$257,209,828	$25,720,983	$50,927,546	$333,858,357
Autothermal Reforming	$355,019,247	$35,501,925	$70,293,811	$460,814,982
CO_2 Removal	$41,008,243	$4,100,824	$8,119,632	$53,228,699
FT Reactors & Refining	$430,422,340	$43,042,234	$85,223,623	$558,688,198
Steam Turbines	$117,240,857	$11,724,086	$23,213,690	$152,178,632
HRSG	$5,575,514	$557,551	$1,103,952	$7,237,017
Cooling Towers	$29,243,593	$2,924,359	$5,790,231	$37,958,184
Water Systems	$87,736,093	$8,773,609	$17,371,746	$113,881,449
Piping	$87,736,093	$8,773,609	$17,371,746	$113,881,449
I&C	$32,128,710	$3,212,871	$6,361,485	$41,703,066
Electrical Systems	$98,857,570	$9,885,757	$19,573,799	$128,317,126
Buildings & Structures	$113,686,205	$11,368,621	$22,509,869	$147,564,694
Total Capital Investment				**$3,156,187,410**
HTGR and Power Cycle				$2,149,311,853
GTL Process				$1,006,875,557
Differential for Adding CO_2 Sequestration				*$16,394,475*

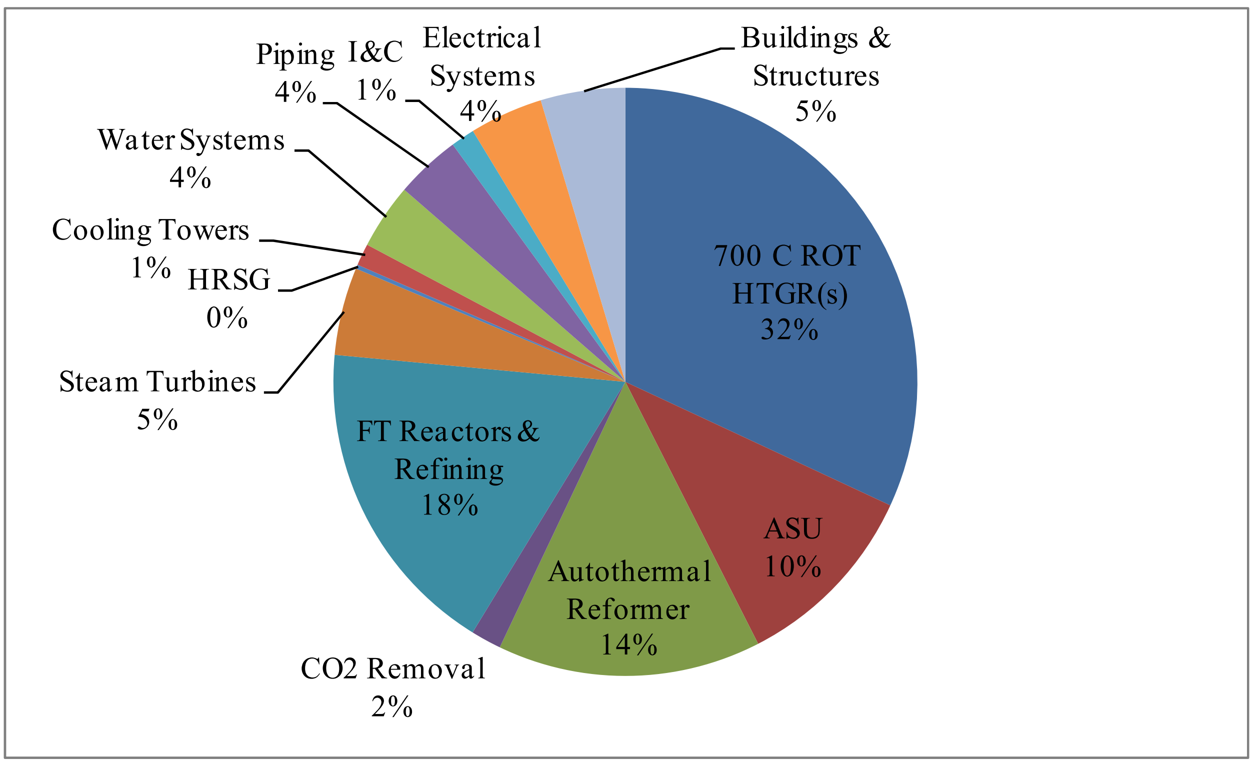


Figure 9. Total capital investment, nuclear-integrated GTL case, no sequestration.

4.2 Estimation of Revenue

Yearly revenues were estimated for all cases based on recent price data for the various products generated. When a separate owner operator configuration is assumed, the HTGR collects revenues from the heat and electricity supplied to the CTL/GTL processes. When heat is exported from the HTGR, the selling price is assumed to be related to electricity price based on the HTGR power generation efficiency as follows:

$$Heat\ Price = Electricity\ Price * Power\ Generation\ \ Efficiency \quad (2)$$

An HTGR power generation efficiency of 43% is assumed, regardless of the power cycle configuration. This allows for an equal comparison for cases where cycle efficiencies may be higher due to power cycle type and/or steam extraction.

Revenues were estimated for low, average, and high prices for diesel and naphtha. High prices correspond to values from July 2008, low prices are from March 2009, and average prices were the average of the high and low values (EIA 2011a). Diesel prices were gathered from the Energy Information Administration (EIA) and represent wholesale prices and do not include taxes. Naphtha prices were scaled based on diesel prices. Selling prices for LPG, electricity, slag, and sulfur were not varied in the study; this was a reasonable assumption since these prices historically follow the standard rate of inflation and do not vary widely during the year, unlike liquid fuel products. The electricity selling price to the industrial process is based on the current industrial market price of electricity, \$67.90/MWe-hr (EIA 2011b). When electricity is sold to the grid, the price is based on 60%[1] of the current average market price of electricity, \$59.28/MWe-hr (EIA 2011b). Revenues were also calculated to determine the necessary selling prices of diesel and heat and electricity, for the separate owner operator scenario, to achieve a specific rate of return; however, these revenues are not presented in the following tables. Additionally, revenues are only presented for the non-sequestration cases; however, cases that have sequestration as an option list the differential revenue that would result from including CO_2 sequestration, i.e. revenue losses associated with electricity use from compression and/or liquefaction equipment.

Oxygen and nitrogen are generated in the CTL and GTL cases. However, it was determined that the volume produced would saturate the U.S. industrial gas market for both commodities if several plants were constructed. Therefore, revenues for these streams are not included in the analysis.

---

1 The current average market price for electricity is \$98.80/MWe-hr, 60% represents the fraction of the power price that accounts for generation.

The revenues presented for the fossil portion are for selling diesel at the low, average, and high product prices. When intermediate revenues for the HTGR are presented, for the independent owner operator scenarios, the heat and electricity prices are presented at the market price. A stream factor of 90% is assumed for both the fossil and nuclear plants. Table 9 presents the revenues for the conventional CTL case and Table 10 presents the revenues for the HTGR-integrated CTL case. Table 11 presents the revenues for the conventional GTL case and Table 12 presents the revenues for the HTGR-integrated GTL case.

Table 9. Annual revenues, conventional CTL case.

	Price		Generated		Annual Revenue
LPG	1.52	$/gal	90,552	gal/day	$45,253,392
Electricity	59.28	$/MWe-hr	248	MWe	$115,725,080
Slag	20.00	$/ton	1,924	ton/day	$12,640,680
Sulfur	40.00	$/ton	847	ton/day	$11,129,580
Diesel, low	1.54	$/gal	1,494,654	gal/day	$753,675,543
Naphtha, low	1.41	$/gal	514,878	gal/day	$238,597,480
Diesel, average	2.80	$/gal	1,494,654	gal/day	$1,373,064,271
Naphtha, average	2.57	$/gal	514,878	gal/day	$434,682,640
Diesel, high	4.06	$/gal	1,494,654	gal/day	$1,992,452,999
Naphtha, high	3.73	$/gal	514,878	gal/day	$115,725,080
Annual Revenue, low					**$1,177,021,755**
Annual Revenue, average					**$1,992,495,643**
Annual Revenue, high					**$2,807,969,531**
Differential for Adding CO_2 Sequestration					*-$12,718,160*

Table 10. Annual revenues, nuclear-integrated CTL case.

	Price		Generated		Annual Revenue
LPG	1.52	$/gal	125,916	gal/day	$62,926,563
Slag	20.00	$/ton	668	ton/day	$4,388,760
Sulfur	40.00	$/ton	298	ton/day	$3,915,720
Diesel, low	1.54	$/gal	1,478,148	gal/day	$745,352,434
Naphtha, low	1.41	$/gal	496,020	gal/day	$229,858,572
Diesel, average	2.80	$/gal	1,478,148	gal/day	$1,357,901,030
Naphtha, average	2.57	$/gal	496,020	gal/day	$418,761,887
Diesel, high	4.06	$/gal	1,478,148	gal/day	$1,970,449,626
Naphtha, high	3.73	$/gal	496,020	gal/day	$607,665,203
Annual Revenue – Fossil, low					**$1,046,442,049**
Annual Revenue – Fossil, average					**$1,847,893,960**
Annual Revenue – Fossil, high					**$2,649,345,872**
Heat	29.20	$/MWt-hr	683	MWt	$157,219,188
Electricity	69.70	$/MWe-hr	2,348	MWe	$1,256,832,748
Annual Revenue – HTGR (separate owner operator)					**$1,414,051,936**

Table 11. Annual revenues, conventional GTL case.

	Price		Generated		Annual Revenue
LPG	1.52	$/gal	147,882	gal/day	$73,904,079
Electricity	59.28	$/MWe-hr	67	MWe	$31,126,410
Diesel, low	1.54	$/gal	1,452,402	gal/day	$732,370,077
Naphtha, low	1.41	$/gal	499,464	gal/day	$231,454,542
Diesel, average	2.80	$/gal	1,452,402	gal/day	$1,334,249,460
Naphtha, average	2.57	$/gal	499,464	gal/day	$421,669,464
Diesel, high	4.06	$/gal	1,452,402	gal/day	$1,936,128,843
Naphtha, high	3.73	$/gal	499,464	gal/day	$611,884,385
Annual Revenue, low					**$1,068,855,109**
Annual Revenue, average					**$1,860,949,413**
Annual Revenue, high					**$2,653,043,718**

Table 12. Annual revenues, nuclear-integrated GTL case.

	Price		Generated		Annual Revenue
LPG	1.52	$/gal	122,388	gal/day	$61,163,444
Electricity	59.28	$/MWe-hr	81	MWe	$38,043,391
Diesel, low	1.54	$/gal	1,487,220	gal/day	$749,926,967
Naphtha, low	1.41	$/gal	490,308	gal/day	$227,211,598
Diesel, average	2.80	$/gal	1,487,220	gal/day	$1,366,235,025
Naphtha, average	2.57	$/gal	490,308	gal/day	$413,939,566
Diesel, high	4.06	$/gal	1,487,220	gal/day	$1,982,543,083
Naphtha, high	3.73	$/gal	490,308	gal/day	$600,667,534
Annual Revenue – Fossil, low					**$1,076,345,399**
Annual Revenue – Fossil, average					**$1,879,381,425**
Annual Revenue – Fossil, high					**$2,682,417,451**
Differential for Adding CO_2 Sequestration					*-$5,468,153*
Heat	29.20	$/MWt-hr	479	MWt	$110,191,545
Annual Revenue – HTGR (separate owner operator)					**$110,191,545**

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 TCI for both cases. This percentage is based on staffing requirements for a conventional 50,000 bbl/day CTL plant, which is assumed to adequately represent the labor for the fossil portion of the nuclear-integrated CTL plant and the GTL plants. Maintenance costs were assumed to be 3% of the TCI 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, this value was assumed based on information presented in the *Handbook of Petroleum Processing* (Jones 2006). Table 13 and Table 14 provide the manufacturing costs for the conventional CTL case and the nuclear-integrated CTL case, respectively. Table 15 and Table 16 provide the manufacturing costs for the conventional GTL case and the nuclear-integrated GTL case, respectively. Again, availability of both the fossil and nuclear plants was assumed to be 90%. The conventional CTL and nuclear-integrated GTL annual manufacturing costs presented do not include costs for CO_2 sequestration; however, the differential manufacturing costs that would result from including CO_2 sequestration are presented, i.e. costs associated with CO_2 pipeline transport and injection.

Table 13. Annual manufacturing costs, conventional CTL case.

	Price		**Consumed**		**Annual Cost**
Direct Costs					
Materials					
Coal	34.35	\$/ton	26,941	ton/day	\$304,000,353
Fly Ash Disposal	34.20	\$/ton	807	ton/day	\$9,065,343
Rectisol Solvent	1.03	\$/gal	7,830	gal/day	\$2,649,169
Wastewater Treatment	1.37	\$/k-gal	6,668	k-gal/day	\$3,010,079
Makeup H_2O Clarifying	0.03	\$/k-gal	30,032	k-gal/day	\$252,192
Carbon, Hg Guard Bed	5.56	\$/lb	35	lb/day	\$64,605
Zinc Oxide	300	\$/ft^3	10.72	ft^3/day	\$1,056,784
Sour Shift Catalyst	825	\$/ft^3	4.42	ft^3/day	\$1,198,267
Claus Catalyst	21	\$/ft^3	6.46	ft^3/day	\$44,573
Sulfur Reduction Catalyst	275	\$/ft^3	1.33	ft^3/day	\$120,537
FT Catalyst	37.50	\$/lb	856	lb/day	\$10,547,297
Hydrocracking Catalyst	850	\$/ft^3	10	ft^3/day	\$2,819,344
Hydrotreating Catalyst	360	\$/ft^3	3	ft^3/day	\$387,644
CO_2 Sequestration	15.19	\$/ton	0	ton/day	\$0
Utilities					
Water	0.05	\$/k-gal	30,032	k-gal/day	\$467,427
Royalties					\$3,040,004
Labor and Maintenance					\$223,989,973
Indirect Costs					
Overhead					\$145,593,482
Insurance and Taxes					\$80,960,231
Manufacturing Costs					**\$789,267,303**
Differential for Adding CO_2 Sequestration					*\$158,746,570*

Manufacturing costs for the nuclear plant were based on information presented in TEV-1196. HTGR manufacturing costs include operations and maintenance (O&M) costs, fuel costs, and decommissioning costs. The O&M, fuel, and decommissioning costs are based on the total thermal rating of the plant (INL 2011a). O&M and decommissioning costs are presented on an annual basis, fuel costs are presented as the total refueling cost per core. The nuclear-integrated case is presented for the single owner operator scenario only. When the HTGR is operated independently, the CTL process would purchase heat and electricity as specified in the HTGR revenues table presented previously (Table 10) and the manufacturing costs would be comprised of the nuclear fuel, O&M, and decommissioning costs presented below (Table 14). Again, availability was assumed to be 90%.

The decommissioning fund payment is calculated using the decommissioning cost in dollars per MWt presented in TEV-1196, which is based on NUREG-1307 (NRC 2010). That cost is multiplied by the total reactor power level to determine the total decommissioning cost and then inflated to the year decommissioning will occur, which is based on the economic recovery period. The sinking fund payment is calculated based on the estimated decommissioning cost and a 5% discount rate (GIF 2007).

It is recognized that the HTGR may operate longer than the specified economic recovery period. However, assuming that the reactor is decommissioned at the end of the recovery period is an economically conservative assumption.

Table 14. Annual manufacturing costs, nuclear-integrated CTL case.

	Price		Consumed		Annual Cost
Direct Costs					
Materials					
Coal	34.35	\$/ton	9,354	ton/day	\$105,552,967
Fly Ash Disposal	34.20	\$/ton	280	ton/day	\$3,145,348
Rectisol Solvent	1.03	\$/gal	3,023	gal/day	\$1,022,706
Wastewater Treatment	1.37	\$/k-gal	5,714	k-gal/day	\$2,579,464
Makeup H_2O Clarifying	0.03	\$/k-gal	22,253	k-gal/day	\$186,871
Carbon, Hg Guard Bed	5.56	\$/lb	12	lb/day	\$22,119
Zinc Oxide	300	\$/ft^3	9.40	ft^3/day	\$925,949
Claus Catalyst	21	\$/ft^3	2.44	ft^3/day	\$16,824
Sulfur Reduction Catalyst	275	\$/ft^3	0.50	ft^3/day	\$45,397
FT Catalyst	37.50	\$/lb	856	lb/day	\$10,547,297
Hydrocracking Catalyst	850	\$/ft^3	10	ft^3/day	\$2,769,742
Hydrotreating Catalyst	360	\$/ft^3	3	ft^3/day	\$379,396
HTSE Cell Replacement	0.025	\$/lb H_2	3,914	k-lb/hr H_2	\$32,742,109
Utilities					
Water	0.05	\$/k-gal	22,253	k-gal/day	\$346,356
Royalties					\$1,055,530
Labor and Maintenance					\$132,598,419
Indirect Costs					
Overhead					\$86,188,972
Insurance and Taxes					\$54,089,808
Nuclear Costs					
O&M	4.31	\$/MWt-hr	6,104	MWt	\$207,640,555
Decommissioning Fund Payment					\$46,536,365
Annual Manufacturing Costs					**\$688,392,193**

	Cost Per Core
Refueling Cost	**\$52,608,619**

The natural gas purchase price for the GTL cases was varied to account for the large fluctuations seen in the market. Costs were calculated for a low (\$4.50/MSCF), average (\$5.50/MSCF), and high (\$12.00/MSCF) industrial natural gas price. 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 prices (EIA 2011c). Only average natural gas prices are presented in the tables that follow.

Table 15. Annual manufacturing costs, conventional GTL case.

	Price		Consumed		Annual Cost
Direct Costs					
Materials					
Average Natural Gas	5.50	\$/MSCF	427,000	MSCFD	\$771,482,250
Wastewater Treatment	1.37	\$/k-gal	7,741	k-gal/day	\$3,494,846
Makeup H_2O Clarifying	0.03	\$/k-gal	19,857	k-gal/day	\$166,754
Zinc Oxide	300	\$/ft^3	7.33	ft^3/day	\$722,837
Hydrolysis Catalyst	450	\$/ft^3	2	ft^3/day	\$238,856
Preforming Catalyst	2,350	\$/ft^3	2	ft^3/day	\$1,630,522
Reforming Catalyst	650	\$/ft^3	1	ft^3/day	\$135,581
FT Catalyst	37.50	\$/lb	856	lb/day	\$10,545,609
Hydrocracking Catalyst	850	\$/ft^3	10	ft^3/day	\$2,657,422
Hydrotreating Catalyst	360	\$/ft^3	3	ft^3/day	\$409,280
Utilities					
Water	0.05	\$/k-gal	19,857	k-gal/day	\$309,070
Royalties					\$7,714,823
Labor and Maintenance					\$84,002,044
Indirect Costs					
Overhead					\$54,601,329
Insurance and Taxes					\$30,362,185
Manufacturing Costs, Average Natural Gas					**\$968,473,408**

When the HTGR is operated independently, the GTL process would purchase heat and electricity as specified in the HTGR revenues table presented previously (Table 12) and the manufacturing costs would be comprised of the nuclear fuel, O&M, and decommissioning costs presented below (Table 16). Again, availability was assumed to be 90%.

Table 16. Annual manufacturing costs, nuclear-integrated GTL case.

	Price		Consumed		Annual Cost
Direct Costs					
Materials					
Average Natural Gas	5.50	\$/MSCF	390,000	MSCFD	\$704,632,500
Wastewater Treatment	1.37	\$/k-gal	6,297	k-gal/day	\$2,842,748
Makeup H_2O Clarifying	0.03	\$/k-gal	20,955	k-gal/day	\$175,973
Zinc Oxide	300	\$/ft^3	7.79	ft^3/day	\$767,293
Hydrolysis Catalyst	450	\$/ft^3	2	ft^3/day	\$270,709
Preforming Catalyst	2,350	\$/ft^3	2	ft^3/day	\$1,445,513
Propylene Carbonate	1.64	\$/lb	186	lb/day	\$100,330
Reforming Catalyst	650	\$/ft^3	1	ft^3/day	\$119,203
FT Catalyst	37.50	\$/lb	856	lb/day	\$10,546,453
Hydrocracking Catalyst	850	\$/ft^3	10	ft^3/day	\$2,717,156
Hydrotreating Catalyst	360	\$/ft^3	3	ft^3/day	\$409,340
CO_2 Sequestration	15.19	\$/ton	0	ton/day	\$0
Utilities					
Water	0.05	\$/k-gal	20,955	k-gal/day	\$326,158
Royalties					\$7,046,325
Labor and Maintenance					\$89,196,442
Indirect Costs					
Overhead					\$57,977,687
Insurance and Taxes					\$32,239,678
Nuclear Costs					
O&M	9.83	\$/MWt-hr	479	MWt	\$34,820,406
Decommissioning Fund Payment					\$3,895,985
Manufacturing Costs, Average Natural Gas					**\$949,529,898**
Differential for Adding CO_2 Sequestration					*\$18,084,561*

	Cost Per Core
Refueling Cost	**\$38,716,117**

4.4 Economic Comparison

Several economic indicators were calculated for each case to assess the economic desirability of synthetic diesel production. For all cases the IRR was calculated for the CTL and GTL[2] processes at low, average, and high diesel prices, as well as for multiple owner operator scenarios for the nuclear-integrated cases. In addition, the diesel price necessary for a return of 12% was calculated for all cases, as well as the heat and electricity prices for a 12% rate of return for the separate owner operator nuclear configurations. Table 17 lists the economic assumptions used for the analyses.

2 For low, average, and high natural gas prices for the GTL scenarios.

Table 17. Economic assumptions.

	Assumption
Year Construction Begins	2012
Construction Information	
Preconstruction Period	6 months
Nuclear Construction Period – per Reactor	36 months
Reactor Startup Staggering	6 months
Fossil Construction Period – per Train	36 months
Train Startup Staggering	6 months
Percent Capital Invested Each Year	S-Curve Distribution
Plant Startup Information	
Startup Time	12 months
Operating Costs Multiplier	1.2
Revenue Multiplier	0.65
Economic Analysis Period	30 years
Availability	90%
Inflation Rate	3%
Debt to Equity Ratio	50%/50%
Loan Information	
Interest Rate on Debt	8%
Interest on Debt During Construction	8%
Loan Repayment Term	15 years
Tax Information	
Effective Tax Rate	35.9%
State Tax Rate	6%
Federal Tax Rate	35%
MACRS Depreciation Term	15 year life
IRR	12%

4.4.1 Cash Flow

To assess the IRR and present worth (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 outflows and noncash costs. Therefore the income taxes per year are defined as follows (Sullivan 2003):

$$T_k = t(R_k - E_k - d_k) \tag{3}$$

Depreciation for the economic calculations was calculated using a standard Modified Accelerated Cost Recovery System (MACRS) depreciation method with a property class of 15 years. Depreciation was assumed for the TCI for each reactor module and fossil process train with the first charge occurring the year the corresponding HTGR/process train comes online, i.e. when initial revenues are received. Table 18 presents the recovery rates for a 15-year property class (Perry 2008).

Table 18. 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. Figure 10 presents the yearly ATCFs for the nuclear-integrated CTL case for a 12% IRR.

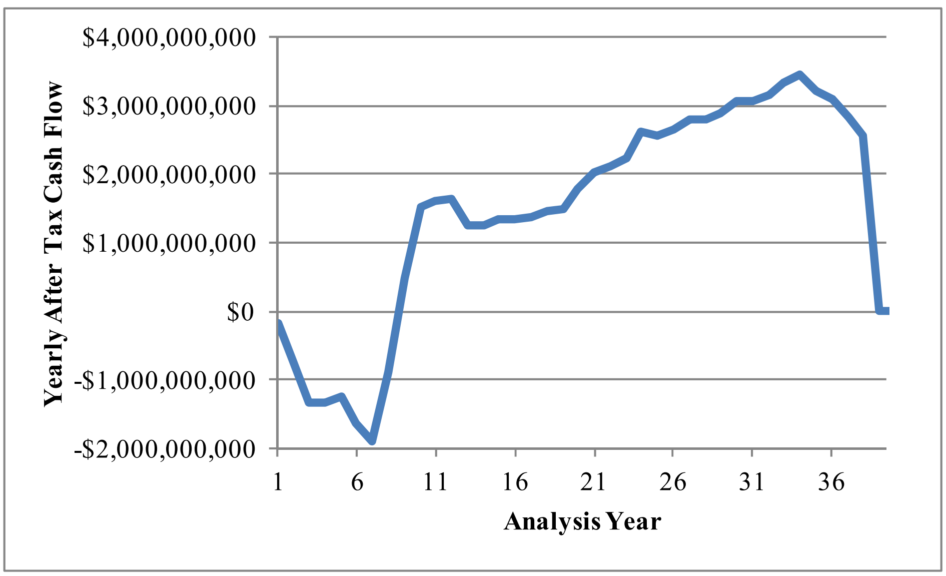


Figure 10. ATCFs, HTGR-integrated CTL process, 12% IRR.

The BTCF is defined as follows (Sullivan 2003):

$$BTCF_k = R_k - E_k - C_k \tag{4}$$

The ATCF can then be defined as:

$$ATCF_k = BTCF_k - T_k \tag{5}$$

4.4.1.1 Capital Cash Flows during Construction

Capital cash flows for the HTGR and fossil processes during construction were calculated for each year of construction based on two separate correlations. First, the percentage of capital assigned to each module or train was calculated based on an exponential correlation (Demick 2011). The exponent for the correlation is calculated based on the current module/train number, such that:

$$y(Mod) = 0.102 \times \ln(Mod + 0.963) - 0.402 \tag{6}$$

where *y* is the exponent for the current module/train and *Mod* is the module/train being evaluated. The capital fraction is then determined for each module/train:

$$ModF(Mod) = \left(1 - \sum_{i=1}^{i=Mod} ModF(i-1)\right) \times (Number - (Mod - 1))^{y(Mod)} \tag{7}$$

where *Number* is the total number of reactor modules or process trains. The yearly fractional breakdown for each module's/train's capital is calculated by applying a generic standard cumulative distribution, the S-Curve, as recommended by the GEN-IV International Forum (GIF) (2007). The capital breakdown per month is calculated as follows:

$$CapF(month) = 0.5 \times \left(\sin\left(\frac{\pi}{2} + \frac{\pi \times month}{c\_months}\right) + 1\right) - CapF(month - 1) \tag{8}$$

where *month* is the current month in the reactor/fossil construction period and *c_months* is the total number of months in the current module's/train's construction period. The capital fraction for each year is calculated by summing the capital fraction for the corresponding months. The yearly capital fractions are then multiplied by the module/train fraction to determine to overall yearly capital fractional breakdown per module/train. Figure 11 presents the percentage of the TCI spent each year of construction for the HTGR-integrated CTL case.

Figure 11. Percentage of TCI spent each year of construction, HTGR-integrated CTL process.

4.4.1.2 Reactor Refueling Cash Flows

Reactor refueling charges occur in the year a refueling is scheduled. The occurrences are determined based on the total number of reactor modules, when the modules come online, and the specified refueling period.

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), which was assumed to be 12% (Sullivan 2003).

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

IRR calculations were performed for the calculated TCI for all cases. In addition, the price of diesel and heat and electricity, for the separate owner/operator scenario, necessary for an IRR of 12% and a PW of zero was calculated for each case. All calculations were performed using Excel (Excel 2007).

Finally, a CO_2 tax was included into the calculations to determine the price of diesel necessary in all cases for a 12% IRR and a CO_2 tax of

$0/ton to $200/ton. The tax calculated was added to the existing yearly tax liability.

5. ECONOMIC MODELING RESULTS

Table 19 presents the results for the conventional CTL case, presenting the IRR for selling diesel at low, average, and high product prices, and the diesel selling price required for a 12% IRR. The nuclear-integrated CTL results are presented in Table 20, for both the single and independent owner/operator scenarios. A value of "N/A" indicates that the manufacturing costs exceeded the revenues.

Table 19. Conventional CTL economic results.

	TCI	
	% IRR	Product Price
Conventional CTL Process	*$5,402,509,707*	
	6.1	$1.54/gal
	21.1	$2.80/gal
	31.9	$4.06/gal
	12.0	$1.95/gal
Conventional CTL Process with Sequestration	*$5,430,913,464*	
	23.3	$1.54/gal
	44.9	$2.80/gal
	56.9	$4.06/gal
	12.0	$2.22/gal

Table 20. Nuclear-integrated CTL economic results.

	TCI	
	% IRR	Product Price
HTGR CTL Process **Single Owner/Operator**	*$13,713,312,773*	
	N/A	$1.54/gal
	4.1	$2.80/gal
	10.2	$4.06/gal
	12.0	$4.57/gal
HTGR CTL Process **Independent Owner/Operator** **Heat/Power at Market Price**	*$10,107,325,559*	
	6.7	$67.90/MWe-hr
	6.7	$29.20/MWt-hr
	$3,605,987,213	
	N/A	$1.54/gal
	N/A	$2.80/gal
	18.5	$4.06/gal
	12.0	$1.95/gal
HTGR CTL Process **Independent Owner/Operator** **Heat/Power at 12% IRR**	*$10,107,325,559*	
	12.0	$92.77/MWe-hr
	12.0	$39.89/MWt-hr
	$3,605,987,213	
	N/A	$1.54/gal
	N/A	$2.80/gal
	5.1	$4.06/gal
	12.0	$4.47/gal

From the nuclear-integrated results, it is apparent that selling heat and power at the market price provides for the largest return on investment for the CTL process. However, the HTGR only has a 7% IRR selling heat and power at the market price to the fossil process; therefore, this case will not be included in the results comparison. Considering the two remaining cases, it is economically beneficial to have an independent owner operator for the CTL and HTGR facilities at an IRR of 12%; however, the single owner/operator scenario is more economical for a variety of diesel selling prices. As a result, the single owner operator scenario will be presented for the breakeven analyses. Figure 12 presents a graphical comparison of the IRR versus the diesel selling price for the convention and nuclear-integrated CTL cases, the nuclear-integrated case presented is for the single owner/operator scenario. The results demonstrate that the nuclear-integrated CTL case provides a lower IRR than the conventional case, either with or without CO_2 sequestration.

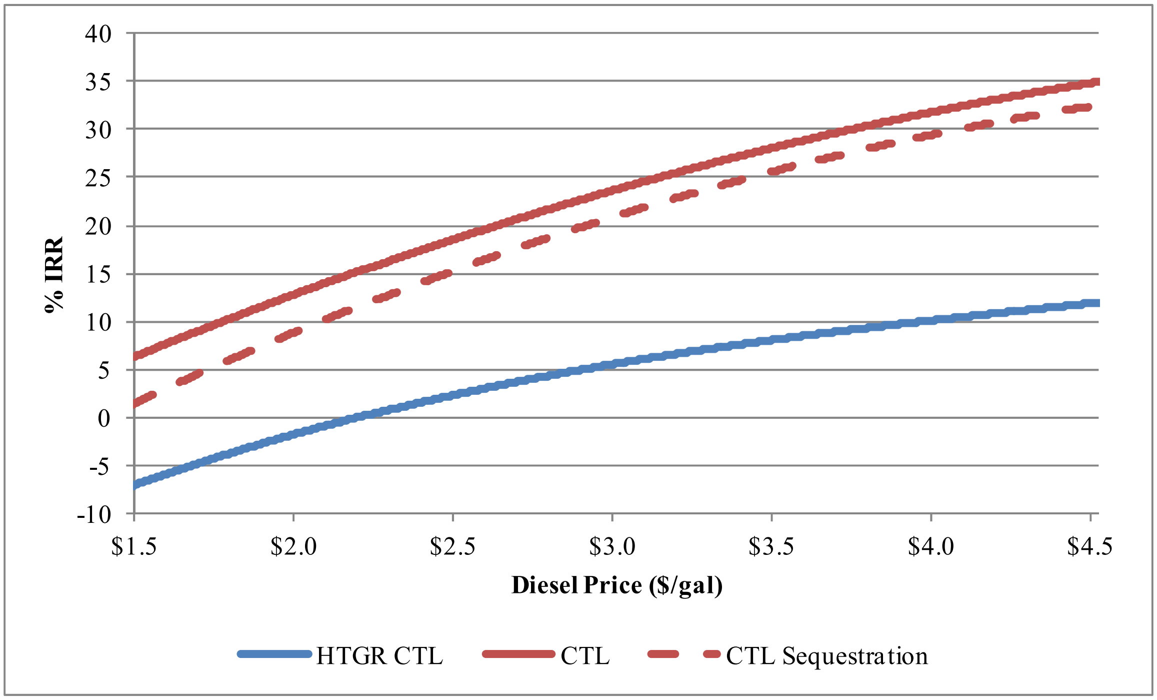


Figure 12. Conventional and nuclear-integrated CTL, IRR as a function of diesel selling price, single owner/operator for the nuclear-integrated process.

Table 21 presents the carbon tax results for the conventional and nuclear-integrated CTL cases, excluding the separate owner/operator scenario where heat and electricity are sold at the market price. Figure 13 depicts the carbon tax results for the conventional and nuclear-integrated CTL cases for the single owner/operator scenario and a 12% IRR.

Table 21. Conventional and nuclear-integrated CTL carbon tax results at 12% IRR.

	Carbon Tax $/ton	Diesel Price ($/gal)
Conventional CTL	0	1.95
	50	3.06
	100	4.18
	150	5.31
	200	6.43
Conventional CTL with Sequestration	0	2.22
	50	2.46
	100	2.70
	150	2.95
	200	3.20
HTGR CTL Single Owner/Operator	0	4.57
	50	4.61
	100	4.64
	150	4.68
	200	4.72
HTGR CTL Independent Owner/Operator	0	4.47
	50	4.50
	100	4.53
	150	4.57
	200	4.60

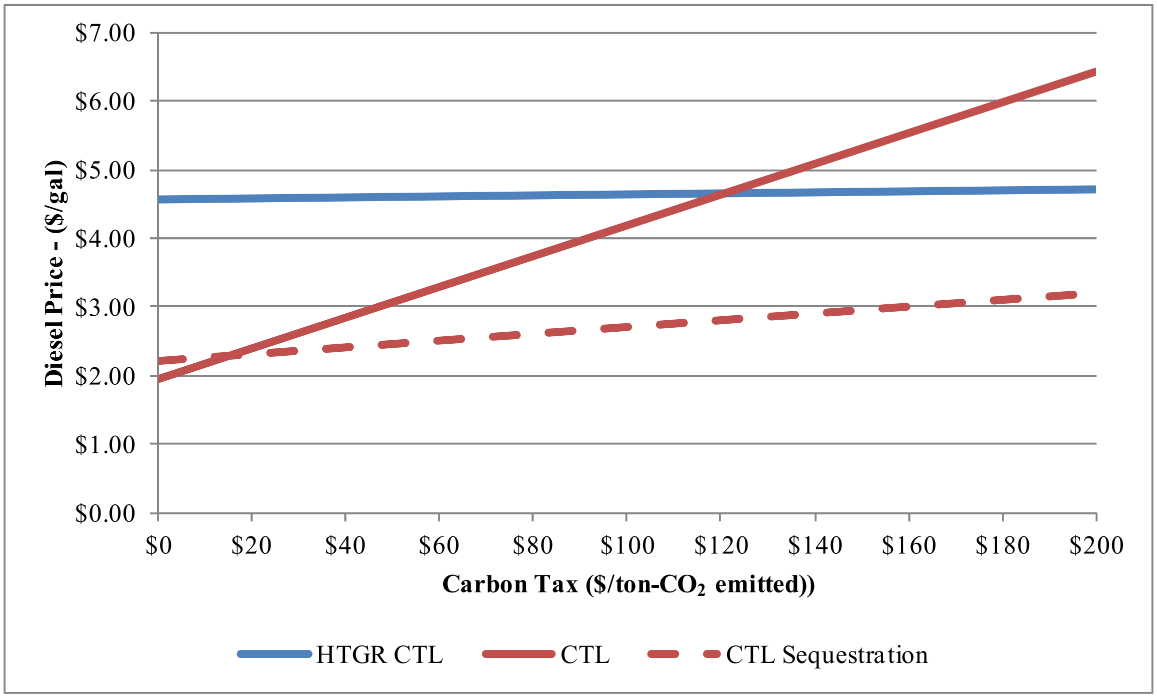


Figure 13. Conventional and nuclear-integrated CTL as a function of a carbon tax, 12% IRR, single owner/operator for the nuclear-integrated process.

The carbon tax results show that the nuclear-integrated CTL case outperforms the conventional case at a 12% IRR when the carbon tax is approximately \$120/ton-CO_2.

Table 22 presents the results for the conventional GTL case, presenting the IRR for selling diesel at low, average, and high product prices, and the diesel selling price required for a 12% IRR for low, average, and high natural gas purchase prices. The nuclear-integrated GTL results are presented in Table 23, for both the single and independent owner/operator scenarios. A value of "N/A" indicates that the manufacturing costs exceeded the revenues.

Table 22. Conventional GTL economic results.

	TCI	
	% IRR	Product Price
Conventional GTL Process **Low Natural Gas Price ($4.50/MSCF)**	*$2,024,145,646*	
	10.5	$1.54/gal
	38.1	$2.80/gal
	57.3	$4.06/gal
	12.0	$1.59/gal
Conventional GTL Process **Average Natural Gas Price ($5.50/MSCF)**	*$2,024,145,646*	
	2.1	$1.54/gal
	33.3	$2.80/gal
	53.5	$4.06/gal
	12.0	$1.83/gal
Conventional GTL Process **High Natural Gas Price ($12.00/MSCF)**	*$2,024,145,646*	
	N/A	$1.54/gal
	N/A	$2.80/gal
	25.8	$4.06/gal
	12.0	$3.38/gal

Table 23. Nuclear-integrated GTL economic results.

		TCI – no Sequestration		TCI – with Sequestration	
		% IRR	Product Price	% IRR	Product Price
Low Natural Gas Price ($4.50/MSCF)	**HTGR GTL Process** **Single Owner/Operator**	*$3,156,187,410*		*$3,172,581,885*	
		5.4	$1.54/gal	4.6	$1.54/gal
		27.7	$2.80/gal	27.2	$2.80/gal
		42.9	$4.06/gal	42.5	$4.06/gal
		12.0	$1.82/gal	12.0	$1.85/gal
	HTGR GTL Process **Independent Owner/Operator** **Heat at Market Price**	*$1,006,875,557*		*$1,006,875,557*	
		1.9	$29.20/MWt-hr	1.9	$29.20/MWt-hr
		$2,149,311,853		*$2,165,706,328*	
		6.4	$1.54/gal	5.4	$1.54/gal
		35.2	$2.80/gal	34.6	$2.80/gal
		54.9	$4.06/gal	54.4	$4.06/gal
		12.0	$1.72/gal	12.0	$1.75/gal
	HTGR GTL Process **Independent Owner/Operator** **Heat/Power at 12% IRR**	*$1,006,875,557*		*$1,006,875,557*	
		12.0	$49.41/MWt-hr	12.0	$49.41/MWt-hr
		$2,149,311,853		*$2,165,706,328*	
		1.8	$1.54/gal	0.7	$1.54/gal
		32.7	$2.80/gal	32.1	$2.80/gal
		52.9	$4.06/gal	52.3	$4.06/gal
		12.0	$1.84/gal	12.0	$1.87/gal

Table 23. Nuclear-integrated GTL economic results.

Natural Gas Price	Case	TCI – no Sequestration % IRR	TCI – no Sequestration Product Price	TCI – with Sequestration % IRR	TCI – with Sequestration Product Price
Average Natural Gas Price ($5.50/MSCF)	**HTGR GTL Process Single Owner/Operator**	*$3,156,187,410*		*$3,172,581,885*	
		-0.7	$1.54/gal	-2.2	$1.54/gal
		24.5	$2.80/gal	24.0	$2.80/gal
		40.3	$4.06/gal	40.0	$4.06/gal
		12.0	$2.04/gal	12.0	$2.07/gal
	HTGR GTL Process Independent Owner/Operator Heat/Power at Market Price	*$1,006,875,557*		*$1,006,875,557*	
		1.9	$29.20/MWt-hr	1.9	$29.20/MWt-hr
		$2,149,311,853		*$2,165,706,328*	
		-2.8	$1.54/gal	-5.6	$1.54/gal
		30.9	$2.80/gal	30.3	$2.80/gal
		51.5	$4.06/gal	50.9	$4.06/gal
		12.0	$1.93/gal	12.0	$1.96/gal
	HTGR GTL Process Independent Owner/Operator Heat/Power at 12% IRR	*$1,006,875,557*		*$1,006,875,557*	
		12.0	$49.41/MWt-hr	12.0	$49.41/MWt-hr
		$2,149,311,853		*$2,165,706,328*	
		N/A	$1.54/gal	N/A	$1.54/gal
		28.3	$2.80/gal	27.6	$2.80/gal
		49.4	$4.06/gal	48.9	$4.06/gal
		12.0	$2.06/gal	12.0	$2.09/gal
High Natural Gas Price ($12.00/MSCF)	**HTGR GTL Process Single Owner/Operator**	*$3,156,187,410*		*$3,172,581,885*	
		N/A	$1.54/gal	N/A	$1.54/gal
		-5.3	$2.80/gal	N/A	$2.80/gal
		21.6	$4.06/gal	21.1	$4.06/gal
		12.0	$3.44/gal	12.0	$3.47/gal
	HTGR GTL Process Independent Owner/Operator Heat/Power at Market Price	*$1,006,875,557*		*$1,006,875,557*	
		1.9	$29.20/MWt-hr	1.9	$29.20/MWt-hr
		$2,149,311,853		*$2,165,706,328*	
		N/A	$1.54/gal	N/A	$1.54/gal
		N/A	$2.80/gal	N/A	$2.80/gal
		26.5	$4.06/gal	25.9	$4.06/gal
		12.0	$3.33/gal	12.0	$3,36/gal
	HTGR GTL Process Independent Owner/Operator Heat/Power at 12% IRR	*$1,006,875,557*		*$1,006,875,557*	
		12.0	$49.41/MWt-hr	12.0	$49.41/MWt-hr
		$2,149,311,853		*$2,165,706,328*	
		N/A	$1.54/gal	N/A	$1.54/gal
		N/A	$2.80/gal	N/A	$2.80/gal
		24.0	$4.06/gal	23.4	$4.06/gal
		12.0	$3.46/gal	12.0	$3.49/gal

From the nuclear-integrated results, it is apparent that selling heat and power at the market price provides for the largest return on investment for the GTL process.

However, the HTGR only has a 2% IRR selling heat at the market price to the fossil process; therefore, this case will not be included in the results comparison. Considering the two remaining cases, it is economically beneficial to have a single owner operator for the GTL and HTGR facilities at an IRR of 12%; additionally, the single owner/operator scenario is more economical for a variety of diesel selling prices. As a result, the single owner operator scenario will be presented for the breakeven analyses. Figure 14 presents a graphical comparison of the diesel price versus the natural gas purchase price for the convention and nuclear-integrated GTL cases, the nuclear-integrated case presented is for the single owner/operator scenario. The results demonstrate that the nuclear-integrated GTL case, either with or without sequestration, requires a higher diesel selling price to achieve a 12% IRR than the conventional case, for natural gas prices less than approximately $14.00/MSCF.

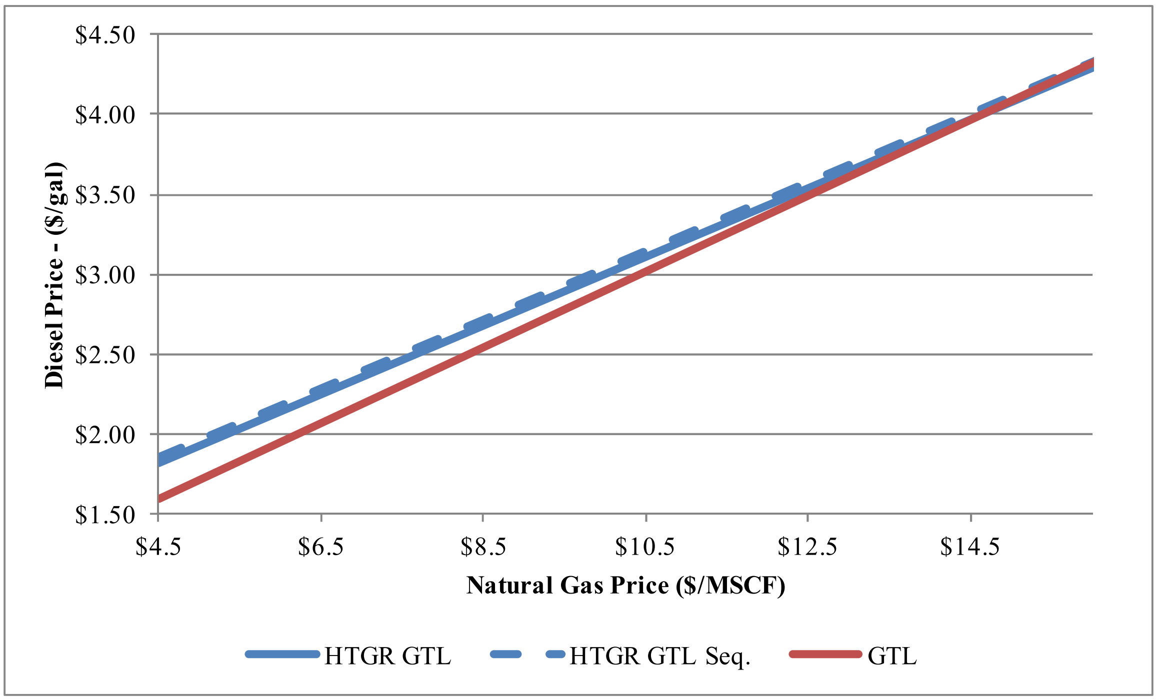


Figure 14. Conventional and nuclear-integrated GTL, diesel price as a function of natural gas purchase price.

Table 24 presents the carbon tax results for the conventional and nuclear-integrated GTL cases, excluding the separate owner/operator scenario where heat and electricity are sold at the market price. Figure 15 depicts the carbon tax results for the conventional and nuclear-integrated CTL cases for the single owner/operator scenario and a 12% IRR.

Table 24. Conventional and nuclear-integrated GTL carbon tax results at 12% IRR.

	Carbon Tax $/ton	Low Natural Gas Price	Average Natural Gas Price	High Natural Gas Price
		Diesel Price ($/gal)		
Conventional GTL	0	1.59	1.83	3.38
	50	1.79	2.02	3.58
	100	1.99	2.23	3.78
	150	2.20	2.43	3.98
	200	2.40	2.64	4.18
HTGR GTL Single Owner/Operator	0	1.82	2.04	3.44
	50	1.93	2.15	3.55
	100	2.05	2.26	3.66
	150	2.16	2.38	3.77
	200	2.28	2.49	3.89
HTGR GTL Independent Owner/Operator	0	1.84	2.06	3.46
	50	1.95	2.17	3.57
	100	2.07	2.28	3.68
	150	2.18	2.40	3.79
	200	2.30	2.51	3.91
HTGR GTL with Sequestration Single Owner/Operator	0	1.85	2.07	3.47
	50	1.87	2.09	3.49
	100	1.90	2.11	3.51
	150	1.92	2.13	3.53
	200	1.94	2.16	3.56
HTGR GTL with Sequestration Independent Owner/Operator	0	1.87	2.09	3.49
	50	1.89	2.11	3.51
	100	1.92	2.13	3.53
	150	1.94	2.15	3.55
	200	1.96	2.18	3.58

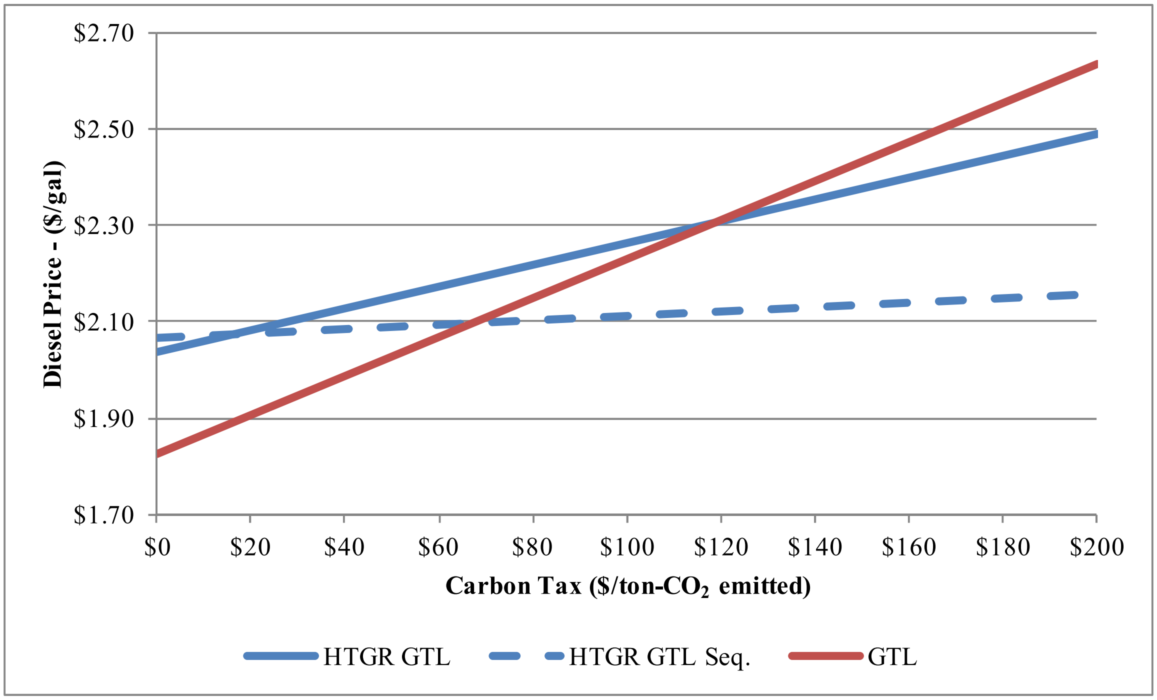


Figure 15. Conventional and nuclear-integrated GTL as a function of a carbon tax, 12% IRR, single owner/operator for the nuclear-integrated process, average natural gas price.

The carbon tax results show that the nuclear-integrated GTL case without sequestration outperforms the conventional case at a 12% IRR for an average natural gas purchase price when the carbon tax is approximately \$120/ton-CO_2. When sequestration is assumed for the nuclear-integrated GTL case, the required CO_2 tax decreases to approximately \$70/ton-CO_2.

6. SENSITIVITY ANALYSIS

A sensitivity analysis was conducted for the nuclear-integrated CTL and GTL cases, for the single owner operator scenario only. The sensitivity analysis assesses the impact on the required product selling price for various changes in the baseline economic assumptions; the result of this sensitivity analysis is a tornado diagram. A tornado diagram is useful in comparing the relative importance of variables, where the sensitive variable is varied while all other variables are held at baseline values.

For the economic assumptions sensitivity analysis, the baseline economic assumptions were varied to determine the effect on the product selling price for the HTGR-integrated cases only. Table 25 lists the values used in the economic sensitivity analysis.

Table 25. Lower, baseline, and upper values used in the economic sensitivity analysis.

	Lower Value	Baseline Value	Upper Value
IRR (%)	10	12	15
Debt Ratio (%)	80	50	0
Debt Interest Rate (%)[3]	4.5	8	10
Loan Term (years)	20	15	10
Construction Period per HTGR (months)	24	36	60
HTGR Staffing Level		Design Supplier	INL Staffing[4]
Economic Recovery Period (years)	40	30	20
HTGR TCI	-30%	TCI	+50%
HTGR Refueling Period (months)	24	18	12
CO_2 Tax		\$0/ton	\$50/ton
Sequestration[5]		No	Yes
Natural Gas Price[5]	Low	Average	High

Again, the sensitivity analysis was only conducted for the single owner operator scenario. Table 26 summarizes the results of the sensitivity analysis listing the required product selling prices for the nuclear-integrated CTL case as well as the percent change in the product selling price versus the baseline case. The tornado plot is presented in Figure 16. Table 27 summarizes the results of the sensitivity analysis listing the required product selling prices for the nuclear-integrated GTL case as well as the percent change in the product selling price versus the baseline case. The tornado plot is presented in Figure 17.

3 The debt interest rate selected in the sensitivity analysis is also used for the interest on debt during construction.

4 The INL staffing level is outlined in TEV-1196. It assumes 595 employees for a four-pack facility versus the design supplier estimate of 418 employees (INL 2011a).

5 Variation only assessed for the nuclear-integrated GTL case.

Table 26. Results from the economic sensitivity analysis, nuclear-integrated CTL, single owner/operator scenario.

	Nuclear-Integrated CTL	
	$/gal	% Change
Baseline Product Price	**4.57**	
IRR		
10%	4.01	-12
15%	5.59	22
Debt Ratio		
80%	4.27	-7
0%	5.18	13
Debt Interest Rate		
4.5%	4.20	-8
10%	4.83	6
Loan Term		
20 years	4.43	-3
10 years	4.77	4
Construction Period		
24 months per HTGR	4.40	-4
60 months per HTGR	4.94	8
Staffing Level		
INL Staffing	4.70	3
Economic Recovery Period		
40 years	4.40	-4
20 years	5.10	12
HTGR TCI		
-30% TCI	3.78	-17
+50% TCI	5.91	29
Refueling Period		
24 months	4.39	-4
12 months	4.95	8
CO_2 Tax		
\$50/ton CO_2	4.61	1

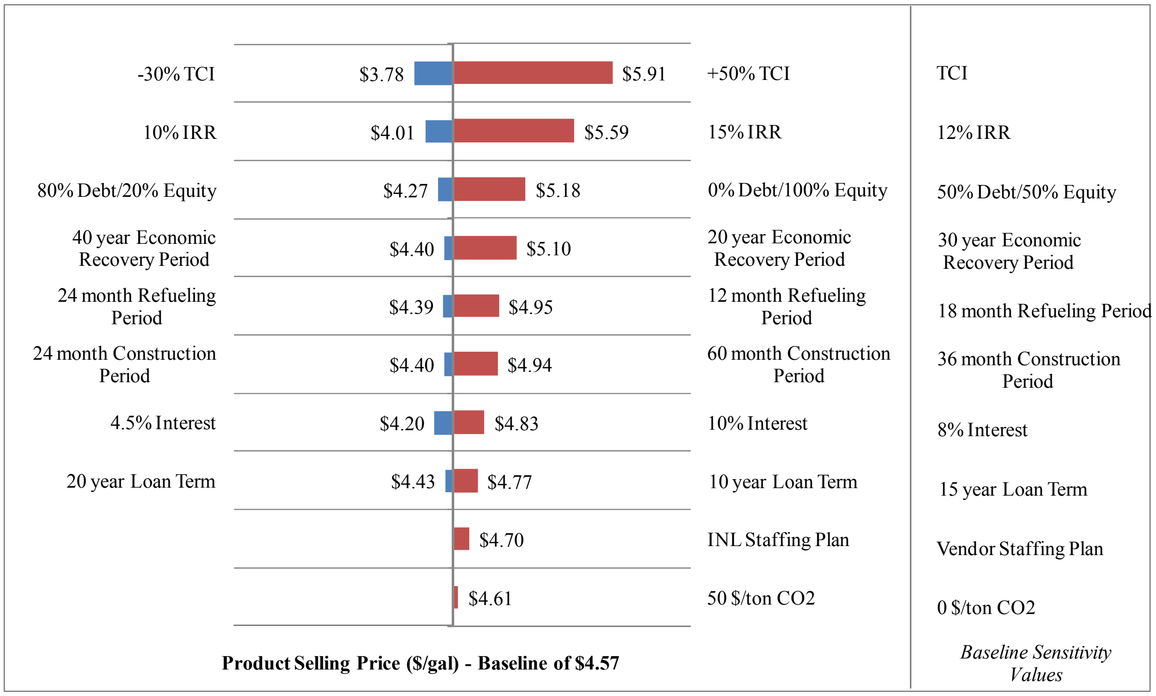


Figure 16. HTGR CTL sensitivity analysis.

From the economic sensitivity analysis, the uncertainty in the HTGR TCI (AACE Class 4) can have the largest impact on the required product selling price, followed by assumed IRR, the debt to equity ratio, and the assumed economic recovery period.

Table 27. Results from the economic sensitivity analysis, nuclear-integrated GTL, single owner/operator scenario.

	Nuclear-Integrated GTL	
	$/gal	% Change
Baseline Product Price	**2.04**	
Natural Gas Price		
Low Natural Gas Price	1.81	-11
High Natural Gas Price	3.43	68
IRR		
10%	1.93	-5
15%	2.19	7
Debt Ratio		
80%	1.95	-4
0%	2.19	7
Debt Interest Rate		
4.5%	1.98	-3
10%	2.06	1
Loan Term		
20 years	2.00	-2
10 years	2.06	1
Construction Period		
24 months per HTGR	2.02	-1
60 months per HTGR	2.04	0
Staffing Level		
INL Staffing	2.06	1
Economic Recovery Period		
40 years	2.00	-2
20 years	2.11	3
HTGR TCI		
-30% TCI	1.97	-3
+50% TCI	2.12	4
Refueling Period		
24 months	2.02	-1
12 months	2.05	0
CO_2 Tax		
\$50/ton CO_2	2.14	5
Sequestration with CO_2 Tax		
\$50/ton CO_2 with Seq.	2.09	2

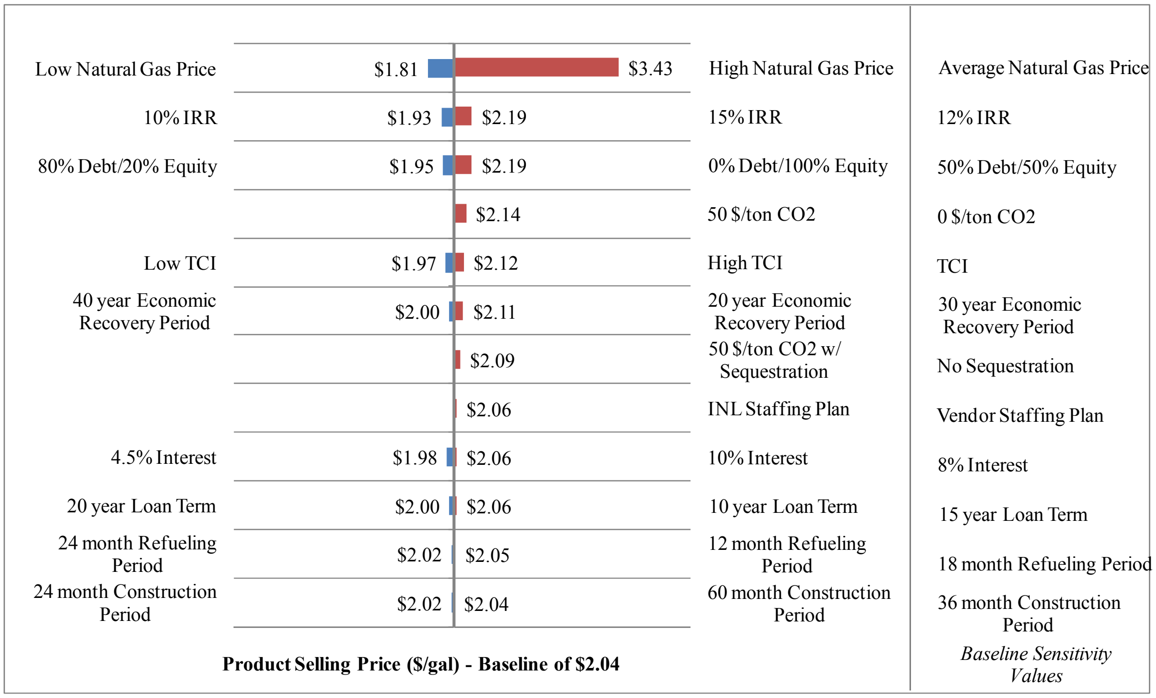


Figure 17. HTGR GTL sensitivity analysis.

From the economic sensitivity analysis, the natural gas purchase price can have the largest impact on the required product selling price, followed by assumed IRR, the debt to equity ratio, and a $50/ton CO_2 tax.

7. GHG MODELING OVERVIEW

This section presents a full life-cycle inventory or well-to-wheel (WTW) analysis of greenhouse gas emissions for the production of synthetic diesel using the conventional and nuclear CTL and GTL processes described in the preceding sections. The WTW analysis conducted for this study was based on the formal methodology presented by NETL in the "Life-Cycle Greenhouse-Gas Emissions Inventory for Fischer-Tropsch Fuels," and categorizes GHG emissions according to the following sources (NETL 2001):

1. Resource extraction
2. Transportation of the resources to the plant
3. Conversion and refining of the product
4. Transportation and distribution of the product
5. End use combustion of the product

Results from the WTW analysis for FT diesel were compared to WTW emissions for the U.S. baseline and average imported WTW emissions for conventional diesel fuel to determine the environmental impact of the synthetic fuels in comparison to standard petroleum fuels. The U.S. baseline and average imported WTW emissions for diesel were derived from a 2009 NETL refinery report (NETL 2009).

7.1 GHG Methodology

The following sections outline the methodology used for calculating GHG emissions for the conventional and nuclear CTL and GTL cases analyzed. For this study, all results are scaled for the diesel, naphtha, LPG, and/or electricity products. This is accomplished by ratioing the lower heating values of the products along with the electricity, if produced in the plant, to determine the emissions assignment, or the percentage of the total energy content for the diesel, naphtha, LPG, and/or electricity product. LPG, naphtha, and diesel all have similar heating values on a mass basis; thus, including the LPG and naphtha with the diesel product has no appreciable impact on overall WTW emissions. The emissions for the diesel product are converted to a gram per mile basis using a vehicle fuel economy of 25.8 miles per gallon. The fuel economy was adjusted to account for the heating value of the synthetic fuel versus traditional petroleum derived products (SAE 1999). The vehicle fuel economy represents the average mileage of a diesel powered SUV.

The GHG emissions considered in this report include carbon dioxide (CO_2), methane (CH_4), and nitrous oxide (N_2O). Emissions for CH_4 and N_2O are converted into CO_2 equivalents using their global warming potentials (GWP). CO_2 equivalents are the amount of carbon dioxide by weight emitted into the atmosphere that would produce the same radiative force as a given weight of another radiatively active gas. The GWPs used in this report are referenced from the Intergovernmental Panel on Climate Change's (IPCC) climate study in 2006. The 100-year GWP for CH_4 and N_2O are 23 and 296, respectively (IPCC 2006).

7.2 Resource Extraction and Production

GHG emissions for resource extraction are calculated for the two feeds considered in this study, coal and natural gas. Coal extraction emissions include emissions from fuel usage associated with coal mining and coal bed methane. Natural gas production emissions include emissions associated with natural gas extraction, natural gas processing, and natural gas transport from the wellhead to the processing facility. Natural gas production emissions include all vents and leaks from the wellhead through the processing phase.

7.2.1 Coal Extraction

The CTL plant is intended to operate using Illinois #6 bituminous coal. The majority of this coal will be mined in the state of Illinois. According to the Energy Information Administration in 2007 approximately 82% of the coal mined in Illinois was mined using underground mining methods, the remainder was surface mined (EIA 2009). Fuel usage per ton of coal mined for both surface and underground mining were calculated based on the most recent U.S. Census data available, either 2002 or 1997 depending upon data released to the public. Based on this census data, power, coal, diesel, residual fuel oil, natural gas, and gasoline usage for mining activities were calculated. The associated CO_2 emissions were calculated based on the lower heating values (LHV) and carbon contents of the various fuel types, for power the emissions for the average U.S. energy mixed were assumed. Emissions for CH_4 and N_2O were calculated assuming either mobile or stationary combustion emission factors from the 2006 IPCC report (IPCC 2006). Emissions for mining support activities were calculated in a similar fashion. Finally, coal bed methane emissions are calculated for the methane released during Illinois mining operations based on the 2009 EPA report, Inventory of U.S. Greenhouse Gas Emissions and Sinks (U.S. EPA 2009).

7.2.2 Natural Gas Production

Methane and non-combustion CO_2 emissions from natural gas systems are generally process related, with normal operations, routine maintenance, and system upsets being the primary contributors. Emissions from normal operations include: natural gas engines and turbine uncombusted exhaust, bleed and discharge emissions from pneumatic devices, and fugitive emissions from system components. Routine maintenance emissions originate from pipelines, equipment, and wells during repair and maintenance activities. Pressure surge relief systems and accidents can lead to system upset emissions. The total CO_2 equivalent emissions were calculated for 2007 in the 2009 EPA report, "Inventory of U.S. Greenhouse Gas Emissions and Sinks" (U.S. EPA 2009). To determine the CO_2 equivalent emissions per MMSCF of natural gas utilized, the equivalent emissions were divided by the amount of natural gas processed in 2007, which is available from the EIA website (EIA 2010).

7.3 Transportation and Distribution

All scenarios considered in this study include transportation of resources and products over large distances. The mode of transportation depends upon the location and destination of the products as well as the type of product being transported. For instance, dry materials being transported short distances would utilize trucks as the main mode of transportation, while dry materials being transported long distances would take advantage of rail transportation. Table 28 lists the distances and modes of transportation assumed for the various resources and products.

Table 28. Transportation information for resources and products.

Product Transported	Miles Transported	Mode of Transport
Petroleum Products to Mine	50	Rail
Natural Gas to Mine	50	Pipeline
Natural Gas to GTL Plant	50	Pipeline
Coal to CTL Plant	100	Rail
Petroleum Products to CTL/MTG Plant	50	Rail
CO_2 to Sequestration Area	50	Pipeline
Diesel – Plant to Distribution Point	200	Pipeline
Diesel – Distribution Point to Pump	200	Truck

The modes of transportation were assumed based on the amount of product being transported, the product state, the distance transported, and the available transportation methods. The emissions associated with the various transportation methods include the combustion of fuel necessary for the transportation (or electricity use) as well as the upstream emissions associated with producing the fuel or electricity. Fuel use per mode of transportation was developed based on information provided by the U.S. EPA "Inventory of U.S. Greenhouse Gas Emissions and Sinks" (U.S. EPA 2009), the *Transportation Energy Databook* (Davis 2009), and the "Freight in America" report (U.S. DOT 2006).

7.4 Conversion and Refining

GHG emissions are generated from several sources within the conventional and nuclear-integrated CTL and GTL plants, including: emissions from importing power, emissions associated with nuclear power use, upstream emissions associated with methanol use, emissions from coal milling and drying, SMR emissions, Rectisol plant emissions, HRSG stack emissions, fired heater emissions, high pressure (HP) and low pressure (LP) flare systems, and fugitive emissions. Fugitive emissions are emissions from leaking equipment, such as valves and pumps, storage tanks, and wastewater treatment facilities. Emissions for the HP and LP flare systems were assumed based on generalized plant startup parameters and fugitive emissions were calculated based on recommendations from the 2006 IPCC guidelines (IPCC 2006). All other emissions were taken from the Aspen modeling results. Emissions were calculated for CH_4 and N_2O for all sources based on IPCC emission factors for CH_4 and N_2O.

7.5 End Use Combustion

Emissions for the end use combustion of the fuel were estimated from the carbon content of the synthetic diesel. It was assumed that all carbon present in the fuel is completely combusted to form CO_2. Based on the fuel density, this would provide the emissions of CO_2 per barrel of fuel. Again, emissions for CH_4 and N_2O were added based on IPCC guidelines for mobile combustion sources.

8. GREENHOUSE GAS MODELING RESULTS

A summary of the GHG results for the cases analyzed is presented in Table 29 for conventional and nuclear CTL diesel and Table 30 for conventional and nuclear GTL diesel. GHG emissions results are presented on a gram CO_2 equivalent per barrel of diesel fuel (g CO_2-eq/bbl) basis, a gram CO_2 equivalent per LHV (g CO_2-eq/MMBTU), and a gram CO_2 equivalent per mile (g CO_2-eq/mile). GHG emissions results are presented in Figure 18 for the CTL diesel cases and Figure 19 for GTL cases.

Table 29. CTL fuels GHG case study results.

	CTL	CTL w/ Seq	HTGR CTL	Baseline Diesel	Imported Diesel
gCO_2-eq/bbl diesel					
Extraction and Production	41,786	42,167	15,729	35,894	45,683
Transportation to Plant	1,493	1,507	562	7,070	9,245
Conversion and Refining	696,731	159,355	64,372	51,666	57,104
Transportation to Pump	4,359	5,953	4,360	4,895	4,351
End Use Combustion	360,375	360,375	360,375	439,910	439,910
Total Fuel Chain	*1,104,744*	*569,357*	*445,398*	*539,434*	*556,293*
gCO_2-eq/MMBTU diesel					
Extraction and Production	8,652	8,730	3,256	6,600	8,400
Transportation to Plant	309	312	116	1,300	1,700
Conversion and Refining	144,255	32,994	13,325	9,500	10,500
Transportation to Pump	902	1,233	903	900	800
End Use Combustion	74,614	74,614	74,599	80,888	80,888
Total Fuel Chain	*228,732*	*117,883*	*92,199*	*99,188*	*102,288*
gCO_2-eq/mile					
Extraction and Production	43	44	16	33	42
Transportation to Plant	2	2	1	7	9
Conversion and Refining	724	166	67	48	53
Transportation to Pump	5	6	5	5	4
End Use Combustion	375	375	375	406	406
Total Fuel Chain	***1,149***	***592***	***463***	***498***	***513***

Table 30. GTL fuels GHG case study results.

	GTL	HTGR GTL	HTGR GTL w/ Seq.	Baseline Diesel	Imported Diesel
gCO_2-eq/bbl diesel					
Extraction and Production	74,879	67,470	67,750	35,894	45,683
Transportation to Plant	39	35	35	7,070	9,245
Conversion and Refining	136,467	79,572	31,221	51,666	57,104
Transportation to Pump	4,365	4,365	4,547	4,895	4,351
End Use Combustion	360,375	360,375	360,375	439,910	439,910
Total Fuel Chain	*576,124*	*511,816*	*463,927*	*539,434*	*556,293*
gCO_2-eq/MMBTU diesel					
Extraction and Production	15,483	13,950	14,008	6,600	8,400
Transportation to Plant	8	7	7	1,300	1,700
Conversion and Refining	28,219	16,452	6,455	9,500	10,500
Transportation to Pump	903	903	940	900	800
End Use Combustion	74,518	74,512	74,512	80,888	80,888
Total Fuel Chain	*119,130*	*105,824*	*95,922*	*99,188*	*102,288*
gCO_2-eq/mile					
Extraction and Production	78	70	70	33	42
Transportation to Plant	0	0	0	7	9
Conversion and Refining	142	83	32	48	53
Transportation to Pump	5	5	5	5	4
End Use Combustion	375	375	375	406	406
Total Fuel Chain	***599***	***532***	***482***	***498***	***513***

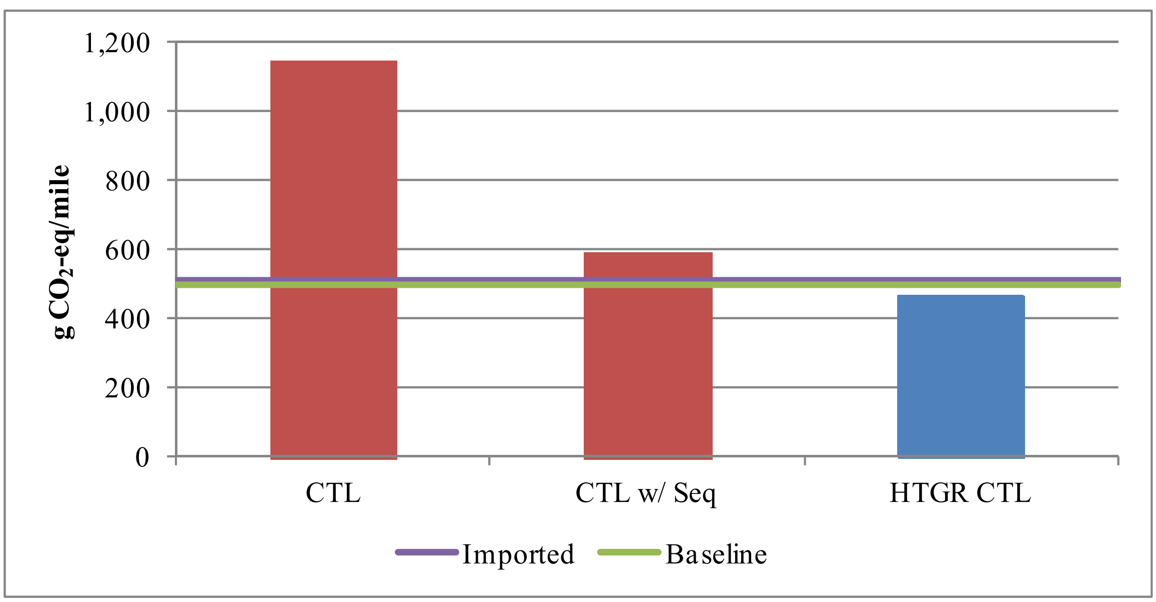


Figure 18. CTL fuels WTW GHG results.

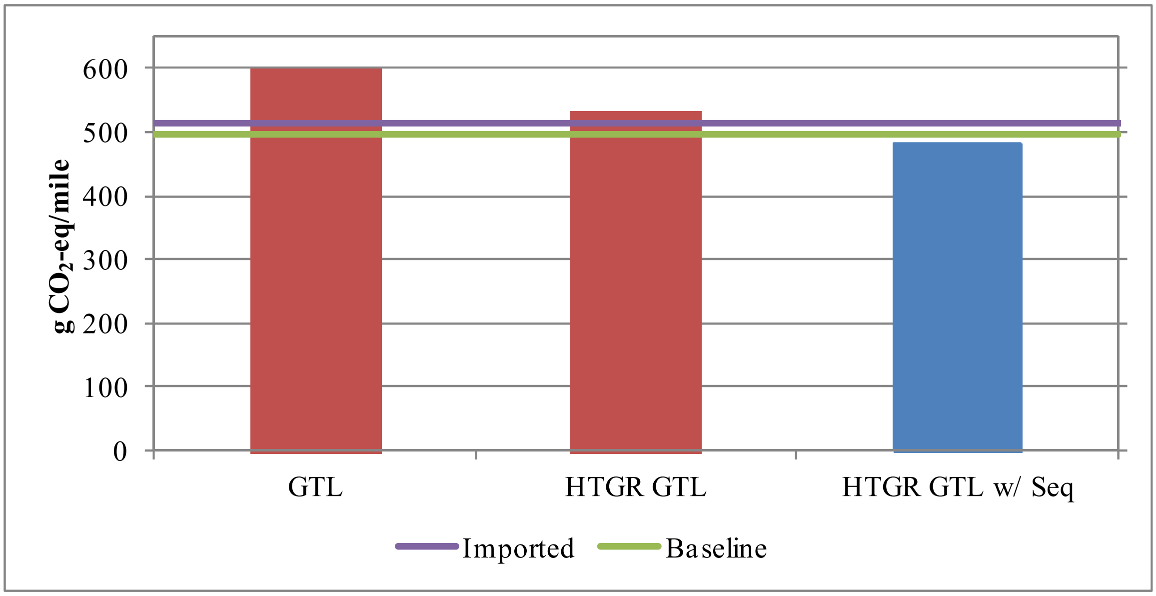


Figure 19. GTL fuels WTW GHG results.

From the results presented in the tables and figures above, integration of an HTGR into CTL and GTL processes can reduce WTW GHG emissions to levels below imported and/or baseline conventional diesel. Conventional CTL WTW emissions are significantly higher than conventional diesel, even with incorporation of sequestration. Nuclear integration into the CTL process reduces WTW GHG below conventional diesel without CO_2 sequestration.

Incorporation of an HTGR with a GTL process reduces WTW GHG emissions when compared to the conventional case; however, they are still slightly higher than baseline and imported diesel. In order to reduce emissions below conventional fuels the pure CO_2 stream produced in the CO_2 removal process in the reforming section must be sequestered.

9. CTL CONCLUSIONS

Results from the nuclear-integrated CTL case indicate that integration of nuclear hydrogen can drastically improve carbon utilization and reduce GHG emissions:

- Coal consumption is decreased by 65% using an HTGR and high temperature steam electrolysis as the hydrogen source.
- Integrating nuclear power and HTSE decreases CO_2 emissions from the plant:
  - If carbon capture and sequestration are assumed for the conventional configuration, CO_2 emissions decrease by 83% when electrolysis and nuclear power are utilized.

- If carbon capture and sequestration are not assumed for the conventional configuration, CO_2 emissions decrease by 96% when electrolysis and nuclear power are utilized.

- It is estimated that one 664 MWt 850°C ROT HTGR for heat production and nine 604 MWt 700°C ROT HTGRs for power production would be required to support production of 50,000 bbl/day of liquid fuel products.

Economically, the incorporation of 10 HTGRs and the associated HTSEs significantly impacts the expected return on investment, when compared to conventional CTL with or without sequestration:

- The required selling price of diesel to achieve a 12% IRR for the nuclear-integrated case is more than two times the selling price required for the conventional CTL case, with or without sequestration.

- In a carbon constrained scenario where CO_2 emissions are taxed and sequestration is not an option, a CO_2 tax of $120/ton CO_2 equates the economics of the nuclear-integrated CTL case with the conventional CTL case.

- From the economic sensitivity analysis, the uncertainty in the HTGR TCI can have the largest impact on the required product selling price, followed by assumed IRR, the debt to equity ratio, and the assumed economic recovery period.

Integration of the HTGR reduces WTW GHG emissions to levels below imported and/or baseline conventional diesel:

- Conventional CTL WTW emissions are significantly higher than conventional diesel and even with incorporation of sequestration emissions are greater than conventional fuels.

- Nuclear-integration is an option where WTW GHG emissions of coal based synthetic fuels are lower than conventional fuels without CO_2 sequestration.

- If there is policy enacted which legislates that synthetically produced diesel fuels must meet or beat current fuel WTW GHG emissions; HTGR incorporation provides a solution with less risk than options which employ CO_2 sequestration.

10. GTL CONCLUSIONS

Results for the nuclear-integrated natural gas to liquids case look promising:

- Approximately one 450 MW_t 700°C ROT HTGR would be required to support this configuration.

- The reactor would supply only heat to the fossil process, as more power is generated in the process than is required.

- By substituting nuclear heat for gas combustion for preheating in the reforming and refining areas, natural gas consumption is decreased by 9%.

- Incorporating an HTGR into the GTL process decrease CO_2 emissions by 42% when sequestration is not assumed and by 88% if the pure CO_2 stream is sequestered.

Economically, the nuclear-integrated GTL option provides economic stability with respect to fluctuations in natural gas prices:

- Though the IRR is slightly lower at higher diesel selling prices, it is still significantly above 12%, indicating a sizable return on investment.

- The nuclear-integrated case requires a higher diesel selling price to achieve a 12% IRR than the conventional case, for natural gas prices less than approximately $14.00/MSCF.

- In a carbon constrained scenario where CO_2 emissions are taxed and sequestration is not an option, a CO_2 tax of $120/ton CO_2 equates the economics of the nuclear-integrated GTL case with the conventional GTL case. When CO_2 is sequestered for the nuclear-integrated GTL case, the necessary CO_2 tax decreases to $70/ton.

- From the economic sensitivity analysis, the natural gas purchase price can have the largest impact on the required product selling price, followed by assumed IRR, the debt to equity ratio, and a $50/ton CO_2 tax.

Integration of the HTGR reduces WTW GHG emissions when compared to the conventional case:

- Conventional GTL WTW emissions are slightly higher than conventional diesel.

- It would be possible reduce nuclear-integrated GTL emissions below conventional fuels with sequestration of the pure CO_2 stream produced in the propylene carbonate process in the reforming section.

11. FUTURE WORK AND RECOMMENDATIONS

As the design of the HTGR progresses towards finalization, this TEV should be updated if the design of the HTGR is changed significantly or if additional refinements of the capital, O&M, fuel, and decommissioning costs become available.

The costs utilized in this study were developed for the prismatic block reactor configuration. Costs for the pebble bed reactor configuration will be included in a future revision of the TEV, when TEV-1196 is updated; however, the capital costs are roughly equivalent and the difference does not affect the overall accuracy of the estimates for both prismatic and pebble bed configurations (INL 2011a).

The capital and operating costs for the CTL and GTL process are based on scaled estimates. If costs come down significantly in the near term or if refined costs become available, this TEV should be updated.

12. REFERENCES

AACE, 2005, *Cost Estimate Classification System – As Applied in Engineering, Procurement, and Construction for the Process Industries*, AACE International Recommended Practice No. 18R-97.

Aspen Plus, Version 7.3, Burlington, Massachusetts: Aspen Tech, 2011.

BRE, 2008, *A Comparison of Physical Solvents for Acid Gas Removal.*

Cover, E., D. A. Hubbar, S. K. Jain, and K. V. Shah, 1986, *Advanced Coal Gasificatio Technical Analyses,* DOE Report.

Davis, Stacy C., Susan W. Diegel, and G. Boundy Robert, 2009, *Tansportation Energy Data Book (Edition 2),* ORNL-6984.

Demick, Larry, 2011, *HTGR Cost Information,* personal communications.

Dry, Mark E, 2001, "High Quality Diesel via the Fischer-Tropsch Process - A Review," *Journal of Chemical Technology & Biotechnology,* Vol. 77, pp. 43-50.

EIA, 2011a, *U.S. No 2 Diesel Low Sulfur 15-500 ppm Retail Sales by Refiners*, http://www.eia.gov/dnav/pet/hist/LeafHandler.ashx?n=PET&s=EMA_EPD2DM10_PTG_NUS_DPG&f=M, July 1, 2011, July 19, 2011.

EIA, 2011b, *Average Retail Price of Electricity to Ultimate Customers: Total by End-Use Sector*, http://www.eia.doe.gov/cneaf/electricity/epm/table5_3.html, April 14, 2011, April 29, 2011.

EIA, 2011c, *Monthly United States Natural Gas Industrial Price*, http://www.eia.gov/dnav/ng/hist/n3035us3m.htm, June 29, 2011, July 6, 2011.

EIA, 2010, *Natural Gas Gross Withdrawals and Production,* http://tonto.eia.doe.gov/dnav/ng/ng_prod_sum_dcu_NUS_a.htm (accessed March 15, 2009), September 18, 2009, April 16, 2010.

EIA, 2009, *Coal Production and Number of Mines by State and Mine Type*, http://www.eia.doe.gov/cneaf/coal/page/acr/table1.html, September 18, 2009, March 1, 2010.

Excel 2007, Version 12.0, Redmond, Washington: Microsoft Corporation, 2007.

GIF, 2007, *Cost Estimating Guidelines for Generaiton IV Nuclear Energy Systems*, GIF/EMWG/2007/004, Rev. 4.2.

INL, 2011a, "Assessment of High Temperature Gas-Cooled (HTGR) Capital and Operating Costs," Idaho National Laboratory, TEV-1196, Rev. 0, April 29, 2011.

INL, 2011b, "An Analysis of Fluids for the Transport of Heat with HTGR-Integrated Steam Assisted Gravity Drainage," Idaho National Laboratory, TEV-1351, DRAFT.

INL, 2011c, "Sensitivity of HTGR Heat and Power Production to Reactor Outlet Temperature, Economic Analysis," Idaho National Laboratory, TEV-988, Rev. 1, June 30, 2011.

INL, 2010, "An Analysis of the Effect of Reactor Outlet Temperature of a High Temperature Reactor on Electric Power Generation, Hydrogen Production, and Process Heat," Idaho National Laboratory, TEV-981, Rev. 0, September 14, 2010.

IPCC, 2006, *2006 IPCC Guidelines for National Greenhouse Gas Inventories,* IPCC National Greenhouse Gas Inventories Programme.

Jones, David S. J., and Peter R. Rujado, 2006, *Handbook of Petroleum Processing*, Dordrecht: Springer.

Kohl, Arthur L., and Richard B. Nielsen, 1997, *Gas Purificatin (5th Edition),* Houston: Elsevier.

Leeper, C. Stephen A, 1981, *Wet Cooling Towers: 'Rule-of-Thumb' Design and Simulation,* EGG-GTH-5775.

Linde, 2008, *Cryogenic Air Separation History and Technological Process,* Product Brochure.

Lurgi, 2006, *The Rectisol Process Lurgi's Leading Technology for Purification and Conditioning of Synthesis Gas,* Product Brochure.

Mullinger, Peter and Barrie Jenkins, 2008, *Industrial and Process Furnaces*, Oxford: Butterworth-Hienemann.

NETL, 2009, *An Evaluation of the Extraction, Tansport and Refining of Imported Crude Oils and the Impact of Life Cycle Greenhouse Gas Emissions,* DOE/NETL-2009/1362.

NETL, 2007, *Cost and Performance Baseline for Fossil Energy Plants,* DOE/NETL-2007/1281.

NETL, 2001, *Life-Cycle Greenhouse-Gas Emissions Inventory for Fischer-Tropsch Fuels.*

NETL, 2000, *Shell Gasifier IGCC Base Cases.* PED-IGCC-98-002.

NRC, 2010, *Report of Waste Burial Charges: Changes in Decommissioning Waste Disposal Costs at Low-Level Waste Burial Facilities*, NUREG-1307, Rev. 14.

Parkash, Surinder, 2003, *Refining Processes Handbook,* Amsterdam: Elsevier.

Pedersen, Peter, 2009, *Steam Reforming Information*, Personal Communication with Anastasia Gribik.

Perry, Robert H., and Don W. Green, 2008, *Perry's Chemical Engineers' Handbook (8th Edition)*, New York: McGraw Hill.

Peters, Max S., and Klaus D. Timmerhaus, 2002, *Plant Design and Economics for Chemical Engineers (5th Edition)*, New York: McGraw Hill.

SAE, 1999, *Emissions from Buses with DDC 6V92 Engines Using Synthetic Diesel Fuel,* 199-01-1512.

Shell, 2005 "Shell Coal Gasification Process Using Low Rank Coal," *Gasification Technologies Conference,San Francisco, CA, October 2005.*

Shell, 2004, "The Shell Coal Gasification Process for the U.S. Industry," *Gasification Technologies Conference, Washington D.C.,October, 2004.*

Sullivan, William G., Elin M. Wicks, and James T. Luxhoj, 2003, *Engineering Economy.* Upper Saddle River: Printice Hall.

Thermoflow Suite, Version 19, Sudbury, Massachusetts: Thermoflow, 2009.

U.S. DOT, 2006 *Freight in America - A New National Picture,* Bureau of Trasportation Statistics.

U.S. EPA, 2009, *Inventory of U.S. Greenhouse Gas Emissions and Sinks*.

WorleyParsons, 2002, "Cost Effective Options to Expand SRU Capacity Using Oxygen," *Sulfur Recovery Symposium, Banff, Alberta, Canada, May 6 – 10, 2002.*

13. APPENDIXES

Appendix A, Detailed Modeling Results and Flowsheets

Appendix B, [Electronic] CTL Baseline Stream Results.xlsx

Appendix C, [Electronic] CTL HTGR Stream Results.xlsx

Appendix D, [Electronic] GTL Baseline Stream Results.xlsx

Appendix E, [Electronic] GTL HTGR Stream Results.xlsx

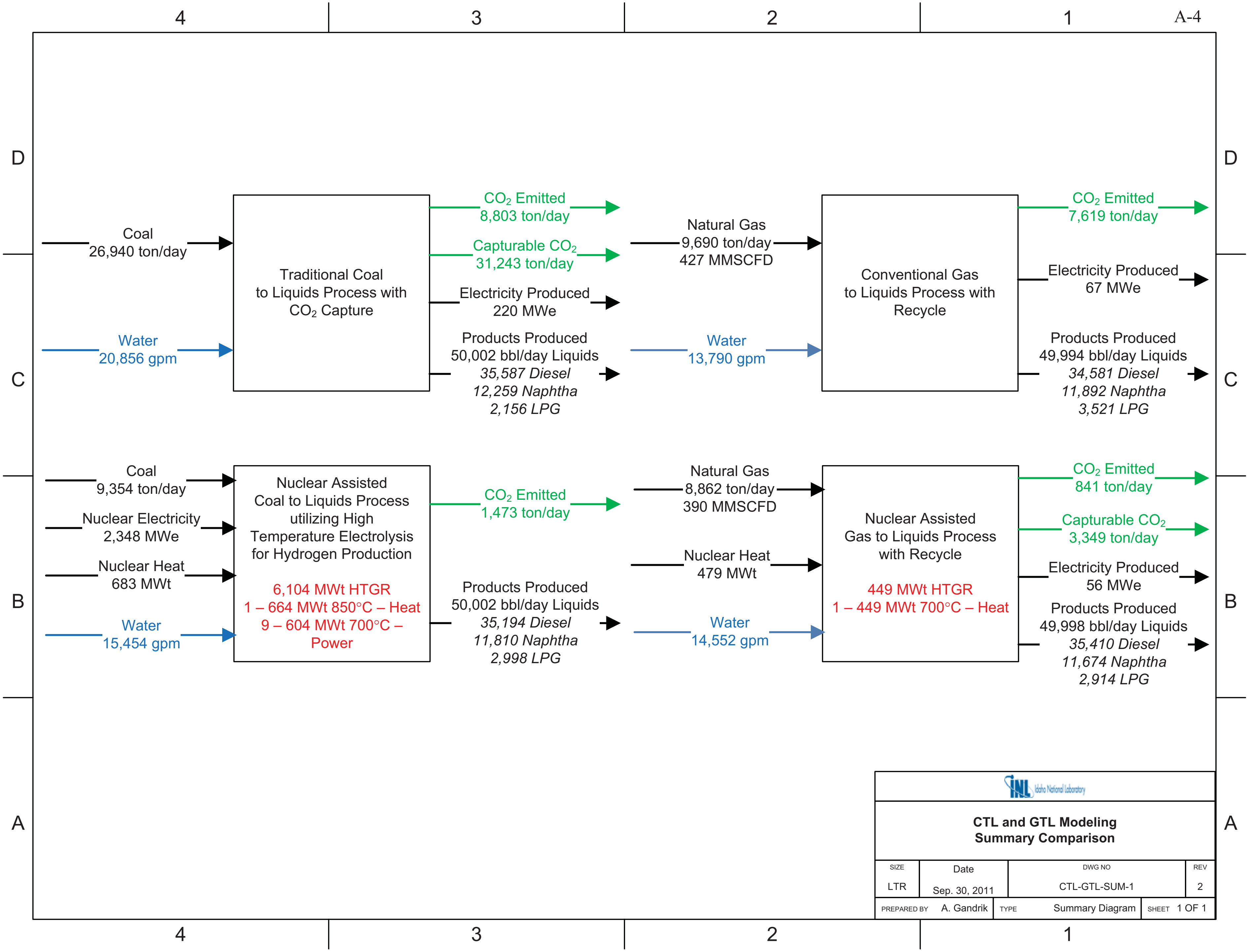
Appendix A
Detailed Modeling Results and Flowsheets

	Conventional CTL	Nuclear Integration CTL	Conventional GTL	Nuclear Integration GTL
Inputs				
Coal Feed rate (ton/day)	26,941	9,354	N/A	N/A
Natural Gas Feed Rate (MMSCFD)[1]	N/A	N/A	427	390
% Carbon to Liquid Product	31.8%	91.7%	71.9%	79.3%
# HTGRs (600 MW_t)	N/A	10.17	N/A	0.75
Outputs				
Total Liquid Products (bbl/day)t	*50,002*	*50,002*	*49,994*	*49,998*
Diesel	35,587	35,194	34,581	35,410
Naphtha	12,259	11,810	11,892	11,674
LPG	2,156	2,998	3,521	2,914
Utility Summary				
Total Power (MW)	*220.3*	*-2,347.8*	*66.6*	*69.7*
Power Consumed	-739.7	-2,749.4	-330.1	-402.3
Electrolyzers	N/A	-2,511.8	N/A	N/A
Secondary Helium Circulator	N/A	-23.0	N/A	-48.4
ASU	-301.3	N/A	-132.7	-131.3
Coal Milling and Drying	-13.8	-9.5	N/A	N/A
Natural Gas Reforming	N/A	N/A	-68.0	-68.9
Gasification and Gas Cleanup	-174.7	-82.1	N/A	N/A
CO_2 Compression/Liquefaction	-140.8	-19.6	N/A	-11.7
Fischer Tropsch & Refining Processes	-40.9	-45.7	-53.8	-60.3
Refrigeration	-24.0	-26.2	-41.5	-47.1
Cooling Tower	-26.6	-18.5	-18.8	-20.8
Water Treatment	-17.6	-13.0	-15.4	-13.9
Power Generated	960.0	401.7	396.7	471.9
Gas Turbine	300.0	N/A	N/A	N/A
Condensing Turbines	178.6	N/A	N/A	N/A
Saturated Turbines	481.4	401.7	396.7	471.9
Water Requirements[2]				
Water Consumed (gpm)	20,856	15,454	13,790	14,552
Water Consumed/lb Feed (lb/lb)	4.65	9.92	8.55	9.86
Water Consumed/bbl Product (bbl/bbl)	14.3	10.6	9.5	10.0

	Conventional CTL	Nuclear Integration CTL	Conventional GTL	Nuclear Integration GTL
CO_2 Summary				
Total CO_2 Produced (ton/day)	*40,046*	*1,473*	*7,164*	*4,190*
Emitted	8,803	1,473	7,164	841
Capturable	31,243	N/A	N/A	3,349
Nuclear Integration Summary				
Electricity (MW)	*N/A*	*-2,643.0*	*N/A*	*-13.9*
HTSE	N/A	-2,511.8	N/A	N/A
HTGR House Loads	N/A	-295.2	N/A	-13.9
Balance of Fossil Plant	N/A	164.0	N/A	N/A
Electrolysis Heat (MMBTU/hr)	*N/A*	*2408.7*	*N/A*	*N/A*
From Nuclear Plant	N/A	2330.2	N/A	N/A
From Secondary Circulator	N/A	78.5	N/A	N/A
Electrolysis Products				
Total Hydrogen (ton/day)	N/A	1,957	N/A	N/A
Total Oxygen (ton/day)	N/A	15,430	N/A	N/A
Used in Plant (ton/day)	N/A	9,198	N/A	N/A
Excess (ton/day)	N/A	6,232	N/A	N/A
HTGR Heat Use (MMBTU/hr)	*N/A*	*N/A*	*N/A*	*1,633*
Reformer	N/A	N/A	N/A	1,057
Refinery	N/A	N/A	N/A	741
From Secondary Circulator	N/A	N/A	N/A	-165

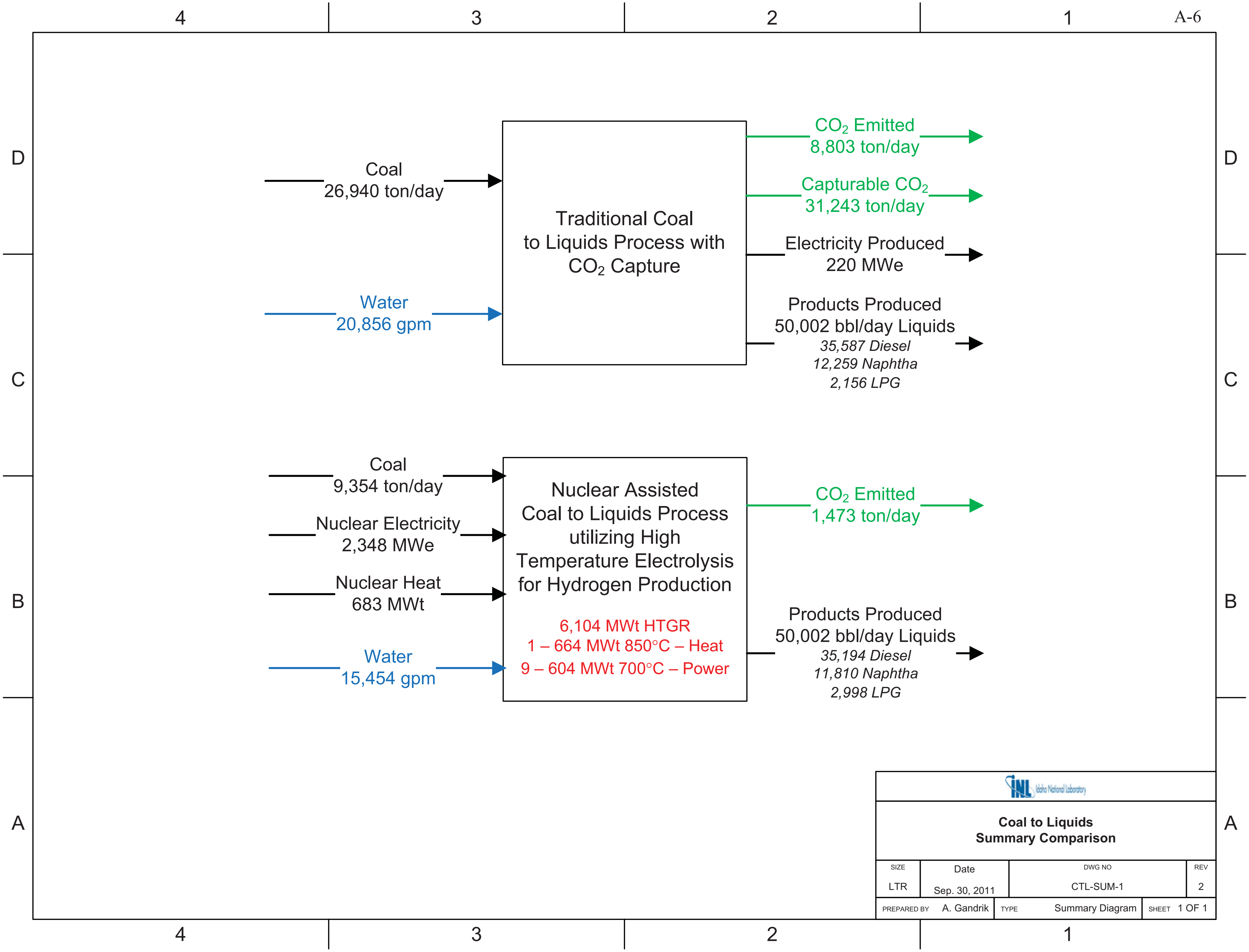
[1]Standard temperature of 60 degrees F.

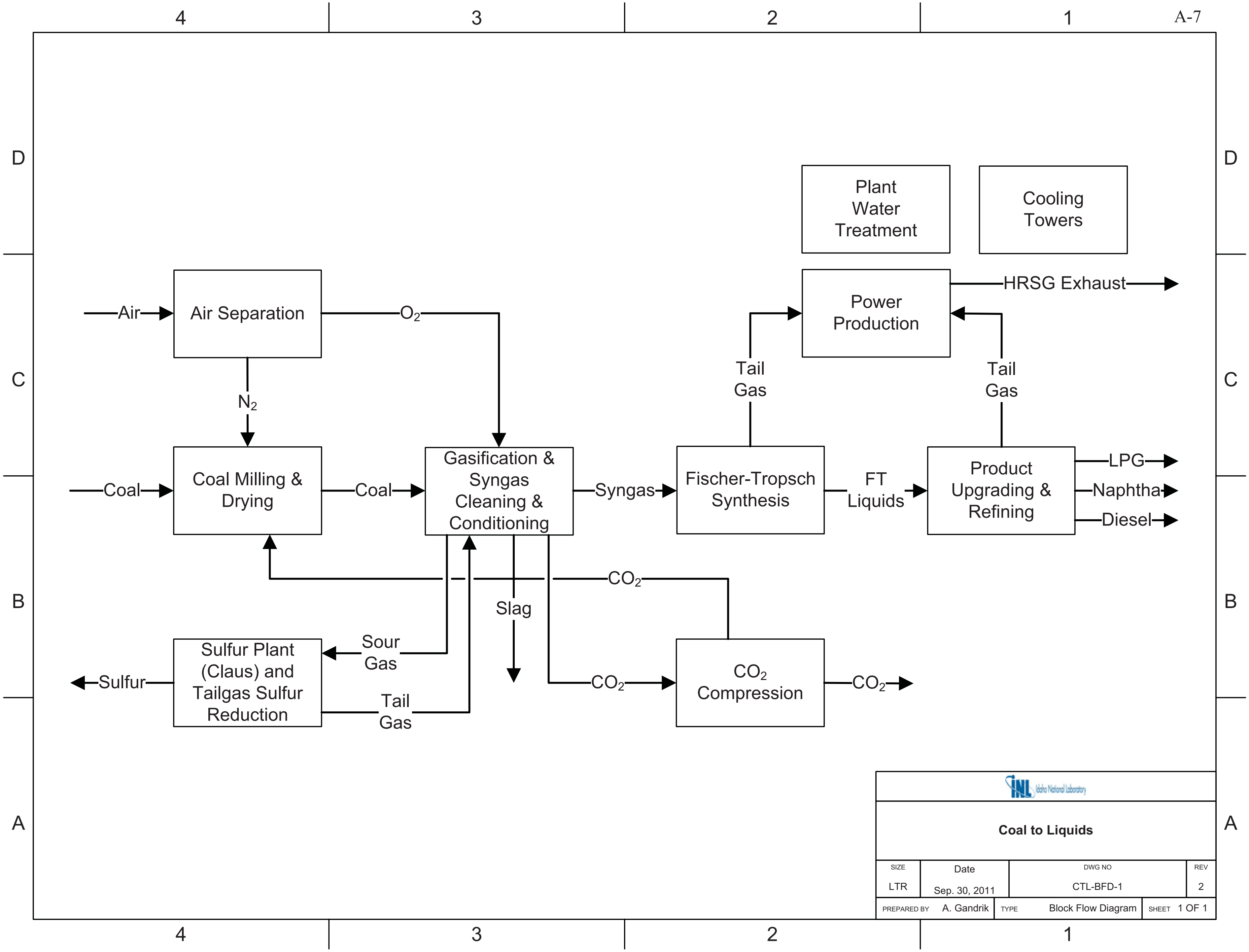
[2]Does not include water usage for HTGR.

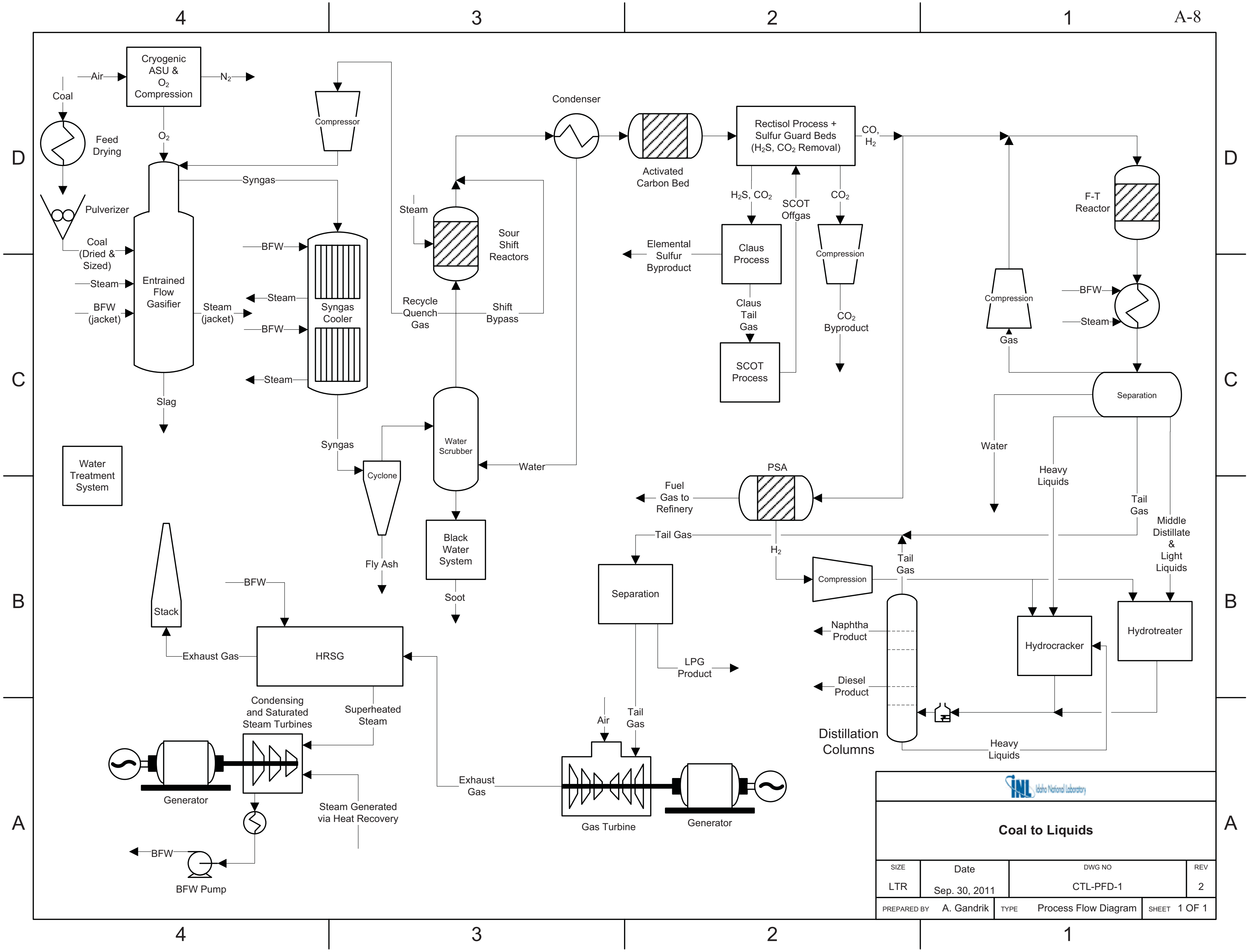

Coal
26,940 ton/day
Water
20,856 gpm
Traditional Coal to Liquids Process with CO_2 Capture
CO_2 Emitted
8,803 ton/day
Capturable CO_2
31,243 ton/day
Electricity Produced
220 MWe
Products Produced
50,002 bbl/day Liquids
35,587 Diesel
12,259 Naphtha
2,156 LPG
Natural Gas
9,690 ton/day
427 MMSCFD
Water
13,790 gpm
Conventional Gas to Liquids Process with Recycle
CO_2 Emitted
7,619 ton/day
Electricity Produced
67 MWe
Products Produced
49,994 bbl/day Liquids
34,581 Diesel
11,892 Naphtha
3,521 LPG
Coal
9,354 ton/day
Nuclear Electricity
2,348 MWe
Nuclear Heat
683 MWt
Water
15,454 gpm
Nuclear Assisted Coal to Liquids Process utilizing High Temperature Electrolysis for Hydrogen Production
6,104 MWt HTGR
1 – 664 MWt 850°C – Heat
9 – 604 MWt 700°C – Power
CO_2 Emitted
1,473 ton/day
Products Produced
50,002 bbl/day Liquids
35,194 Diesel
11,810 Naphtha
2,998 LPG
Natural Gas
8,862 ton/day
390 MMSCFD
Nuclear Heat
479 MWt
Water
14,552 gpm
Nuclear Assisted Gas to Liquids Process with Recycle
449 MWt HTGR
1 – 449 MWt 700°C – Heat
CO_2 Emitted
841 ton/day
Capturable CO_2
3,349 ton/day
Electricity Produced
56 MWe
Products Produced
49,998 bbl/day Liquids
35,410 Diesel
11,674 Naphtha
2,914 LPG
INL Idaho National Laboratory
CTL and GTL Modeling Summary Comparison
SIZE LTR
Date Sep. 30, 2011
DWG NO CTL-GTL-SUM-1
REV 2
PREPARED BY A. Gandrik
TYPE Summary Diagram
SHEET 1 OF 1

	Conventional CTL	Nuclear Integration CTL
Inputs		
Coal Feed rate (ton/day)	26,941	9,354
% Carbon to Liquid Product	31.8%	91.7%
# HTGRs (600 MW_t)	N/A	10.17
Outputs		
Total Liquid Products (bbl/day)t	*50,002*	*50,002*
Diesel	35,587	35,194
Naphtha	12,259	11,810
LPG	2,156	2,998
Utility Summary		
Total Power (MW)	*220.3*	*-2,347.8*
Power Consumed	-739.7	-2,749.4
Electrolyzers	N/A	-2,511.8
Secondary Helium Circulator	N/A	-23.0
ASU	-301.3	N/A
Coal Milling and Drying	-13.8	-9.5
Gasification and Gas Cleanup	-174.7	-82.1
CO_2 Compression/Liquefaction	-140.8	-19.6
Fischer Tropsch & Refining Processes	-40.9	-45.7
Refrigeration	-24.0	-26.2
Cooling Tower	-26.6	-18.5
Water Treatment	-17.6	-13.0
Power Generated	960.0	401.7
Gas Turbine	300.0	N/A
Condensing Turbines	178.6	N/A
Saturated Turbines	481.4	401.7
Water Requirements[1]		
Water Consumed (gpm)	20,856	15,454
Water Consumed/lb Feed (lb/lb)	4.65	9.92
Water Consumed/bbl Product (bbl/bbl)	14.3	10.6
CO_2 Summary		
Total CO_2 Produced (ton/day)	*40,046*	*1,473*
Emitted	8,803	1,473
Capturable	31,243	N/A
Nuclear Integration Summary		
Electricity (MW)	*N/A*	*-2,643.0*
HTSE	N/A	-2,511.8
HTGR House Loads	N/A	-295.2
Balance of Fossil Plant	N/A	164.0
Electrolysis Heat (MMBTU/hr)	*N/A*	*2408.7*
From Nuclear Plant	N/A	2330.2
From Secondary Circulator	N/A	78.5
Electrolysis Products		
Total Hydrogen (ton/day)	N/A	1,957
Total Oxygen (ton/day)	N/A	15,430
Used in Plant (ton/day)	N/A	9,198
Excess (ton/day)	N/A	6,232

[1]Does not include water usage for HTGR.


Coal
26,940 ton/day
Water
20,856 gpm
Traditional Coal to Liquids Process with CO_2 Capture
CO_2 Emitted
8,803 ton/day
Capturable CO_2
31,243 ton/day
Electricity Produced
220 MWe
Products Produced
50,002 bbl/day Liquids
35,587 Diesel
12,259 Naphtha
2,156 LPG
Coal
9,354 ton/day
Nuclear Electricity
2,348 MWe
Nuclear Heat
683 MWt
Water
15,454 gpm
Nuclear Assisted Coal to Liquids Process utilizing High Temperature Electrolysis for Hydrogen Production
6,104 MWt HTGR
1 – 664 MWt 850°C – Heat
9 – 604 MWt 700°C – Power
CO_2 Emitted
1,473 ton/day
Products Produced
50,002 bbl/day Liquids
35,194 Diesel
11,810 Naphtha
2,998 LPG
Coal to Liquids
Summary Comparison
SIZE LTR
Date Sep. 30, 2011
DWG NO CTL-SUM-1
REV 2
PREPARED BY A. Gandrik
TYPE Summary Diagram
SHEET 1 OF 1


Plant Water Treatment
Cooling Towers
Air
Air Separation
O_2
N_2
Power Production
HRSG Exhaust
Tail Gas
Tail Gas
Coal
Coal Milling & Drying
Coal
Gasification & Syngas Cleaning & Conditioning
Syngas
Fischer-Tropsch Synthesis
FT Liquids
Product Upgrading & Refining
LPG
Naphtha
Diesel
CO_2
Slag
Sour Gas
Sulfur Plant (Claus) and Tailgas Sulfur Reduction
Sulfur
Tail Gas
CO_2
CO_2 Compression
CO_2
Idaho National Laboratory
Coal to Liquids
SIZE
LTR
Date
Sep. 30, 2011
DWG NO
CTL-BFD-1
REV
2
PREPARED BY
A. Gandrik
TYPE
Block Flow Diagram
SHEET
1 OF 1


4
3
2
1
D
C
B
A
Air
Cryogenic ASU & O_2 Compression
N_2
Coal
O_2
Compressor
Feed Drying
Syngas
Pulverizer
Coal (Dried & Sized)
Steam
BFW (jacket)
Entrained Flow Gasifier
Steam (jacket)
BFW
Steam
Syngas Cooler
BFW
Steam
Slag
Condenser
Activated Carbon Bed
Rectisol Process + Sulfur Guard Beds (H_2S, CO_2 Removal)
CO, H_2
Steam
Sour Shift Reactors
Recycle Quench Gas
Shift Bypass
H_2S, CO_2
SCOT Offgas
CO_2
Elemental Sulfur Byproduct
Claus Process
Compression
Claus Tail Gas
CO_2 Byproduct
SCOT Process
F-T Reactor
Compression
Gas
BFW
Steam
Separation
Water
Water Scrubber
Syngas
Cyclone
Water
Water Treatment System
Fly Ash
Black Water System
Soot
PSA
Fuel Gas to Refinery
Tail Gas
H_2
Heavy Liquids
Tail Gas
Middle Distillate & Light Liquids
Tail Gas
Compression
Separation
Stack
BFW
HRSG
Exhaust Gas
Naphtha Product
Diesel Product
LPG Product
Hydrocracker
Hydrotreater
Distillation Columns
Heavy Liquids
Condensing and Saturated Steam Turbines
Superheated Steam
Air
Tail Gas
Generator
Steam Generated via Heat Recovery
Exhaust Gas
Gas Turbine
Generator
BFW
BFW Pump
Idaho National Laboratory
Coal to Liquids
SIZE
LTR
Date
Sep. 30, 2011
DWG NO
CTL-PFD-1
REV
2
PREPARED BY
A. Gandrik
TYPE
Process Flow Diagram
SHEET
1 OF 1

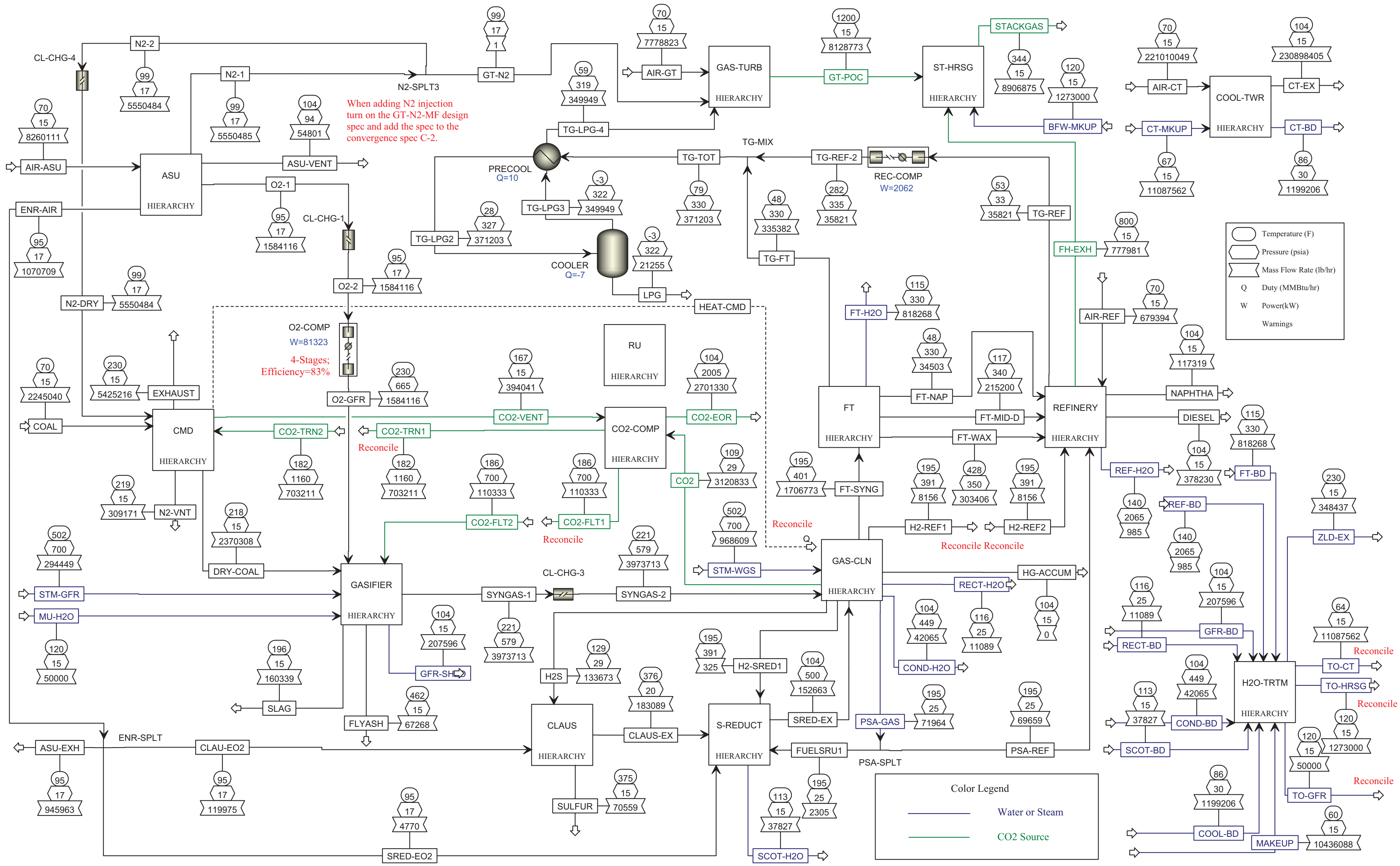
```
CALCULATOR BLOCK SUMMARY

   FEED & PRODUCT SUMMARY:

     FEEDS:

       RAW COAL FEED RATE =                 26940.5 TON/DY
         COAL HHV AS FED =                  10934. BTU/LB
         COAL MOISTURE AS FED =                13.70 %

       PROXIMATE ANALYSIS:
          MOISTURE                             13.70 %
          FIXED CARBON                         40.12 %
          VOLATILE MATTER                      49.28 %
          ASH                                  10.60 %

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

       SULFANAL ANALYSIS:
          PYRITIC                               1.94 %
          SULFATE                               0.08 %
          ORGANIC                               1.70 %

     INTERMEDIATES:

       COAL FEED RATE AFTER DRYING =        24733.7 TON/DY
         COAL HHV AFTER DRYING =            11910. BTU/LB
         COAL MOISTURE AFTER DRYNG =            6.00 %

       RAW SYNGAS MASS FLOW =             4041893. LB/HR
       RAW SYNGAS VOLUME FLOW =              1737. MMSCFD
       RAW SYNGAS HHV (WET) =                 280.8 BTU/SCF
       RAW SYNGAS HHV (DRY) =                 305.2 BTU/SCF
       RAW SYNGAS COMPOSITION:
          H2                                   27.4 MOL.%
          CO                                   56.6 MOL.%
          CO2                                   5.8 MOL.%
          N2                                    0.6 MOL.%
          H2O                                   8.0 MOL.%
          CH4                                  51. PPMV
          H2S                               10664. PPMV

       QUENCHED SYNGAS MASS FLOW =        3973714. LB/HR
       QUENCHED SYNGAS VOLUME FLOW =         1675. MMSCFD
       QUENCHED SYNGAS HHV (WET) =            290.2 BTU/SCF
       QUENCHED SYNGAS HHV (DRY) =            299.8 BTU/SCF
       QUENCHED SYNGAS COMPOSITION:
          H2                                   28.5 MOL.%
          CO                                   58.9 MOL.%
          CO2                                   7.4 MOL.%
          N2                                    0.7 MOL.%
          H2O                                   3.2 MOL.%
          CH4                                  53. PPMV
          H2S                               11092. PPMV

       CLEANED SYNGAS MASS FLOW =         1706773. LB/HR
       CLEANED SYNGAS VOLUME FLOW =          1422. MMSCFD
       CLEANED SYNGAS HHV (WET) =             315.6 BTU/SCF
```

```
   CLEANED SYNGAS HHV (DRY) =           315.6 BTU/SCF
   CLEANED SYNGAS COMPOSITION:
      H2                                 66.6 MOL.%
      CO                                 31.1 MOL.%
      CO2                                 1.3 MOL.%
      N2                                  0.8 MOL.%
      H2O                                 0.0 MOL.%
      CH4                                56. PPMV
      H2S                                 0. PPMV

 PRODUCTS:

   LIQUID PRODUCTS PRODUCED =        516804. LB/HR
   LIQUID PRODUCTS PRODUCED =          6201.6 TON/DY
     DIESEL =                        378230. LB/HR
     DIESEL =                          4539. TON/DY
     NAPHTHA =                       117319. LB/HR
     NAPHTHA =                         1408. TON/DY
     LPG =                            21255. LB/HR
     LPG =                              255. TON/DY
   LIQUID PRODUCTS PRODUCED =         50002. BBL/DY
     DIESEL =                         35587. BBL/DY
     NAPHTHA =                        12259. BBL/DY
     LPG =                             2156. BBL/DY
   LIQUIDS PRODUCED / COAL FED =          0.23 LB/LB
   LIQUIDS PRODUCED / COAL FED =          1.86 BBL/TON

 FUEL PROPERTIES:
                              DIESEL       NAPHTHA        LPG
   PROD. RATE, BBL/DAY     35587.        12259.         2156.
   LHV RATE, MMBTU/DAY    171880.        52590.         9285.
   MW                        187.8          79.6          58.2
   API GRAVITY                54.3          84.9
   DENSITY, LB/GAL             6.07          5.47           5.63
   CETANE NO.                 93.9          29.2
   HHV CONTENT, BTU/LB     20369.        20161.        19667.
   LHV CONTENT, BTU/LB     18935.        18678.        18202.
   % CARBON                   84.7          81.7          79.2
   D86T CURVE, DEG. C:
       0%                   147.         -107.
       10%                  182.           20.
       20%                  200.           49.
       50%                  247.           80.
       90%                  327.          119.
       100%                 355.          161.

POWER CALCULATIONS:

 POWER GENERATORS:
   GAS TURBINE POWER OUTPUT =           300.0 MW
   CONDENSING TURBINE POWER OUTPUT =    178.6 MW
   SATURATED TURBINE POWER OUTPUT =     481.4 MW
 TOTAL POWER GENERATED =                960.0 MW

 POWER CONSUMERS:
   COAL PROCESSING POWER CONSUMPTION =   13.8 MW
   ASU POWER CONSUMPTION =              301.3 MW
   GASIFIER POWER CONSUMPTION =          17.4 MW
   GAS CLEANING POWER CONSUMPTION =     146.9 MW
   SCOT PROCESS POWER CONSUMPTION =       9.3 MW
   CLAUS POWER CONSUMPTION =              1.1 MW
   CO2 LIQUEF. POWER CONSUMPTION =      140.8 MW
   FISHER TROPSCH POWER CONSUMPTION =    24.2 MW
   REFINERY POWER CONSUMPTION =          10.8 MW
   POWER BLOCK POWER CONSUMPTION =        5.9 MW
```

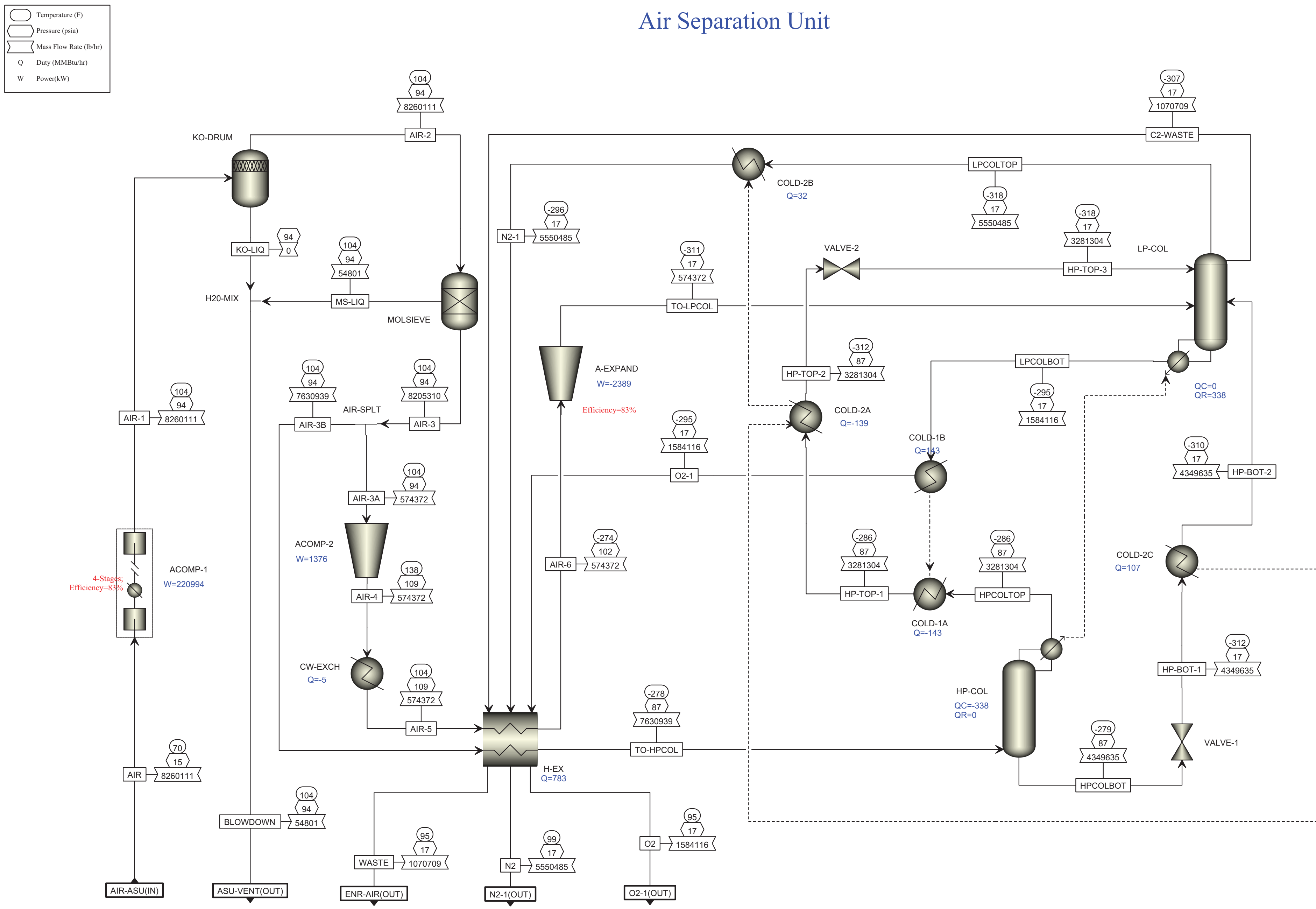
```
    REFRIGERATION POWER CONSUMPTION =      24.0 MW
    COOLING TOWER POWER CONSUMPTION =      26.6 MW
    WATER TREATMENT POWER CONSUMPTION =    17.6 MW
  TOTAL POWER CONSUMED =                  739.7 MW

  NET PLANT POWER (+ GEN, - CONS)=        220.3 MW

WATER BALANCE:

  EVAPORATIVE LOSSES:
    CMD WATER NOT RECOVERED =             367.5 GPM
    COOLING TOWER EVAPORATION =         24194.3 GPM
    ZLD SYSTEM EVAPORATION =              696.3 GPM
  TOTAL EVAPORATIVE LOSSES =            25258.1 GPM

  WATER CONSUMED:
    GASIFIER ISLAND MAKEUP =               99.9 GPM
    BOILER FEED WATER MAKEUP =           2544.0 GPM
    COOLING TOWER MAKEUP =              22157.4 GPM
  TOTAL WATER CONSUMED =                24801.3 GPM

  WATER GENERATED:
    GASIFIER ISLAND BLOWDOWN =            414.9 GPM
    SYNGAS CONDENSER BLOWDOWN =            84.1 GPM
    RECTISOL BLOWDOWN =                    22.2 GPM
    SULFUR REDUCTION BLOWDOWN =            75.6 GPM
    FT PROCESS BLOWDOWN =                1635.2 GPM
    REFINERY PROCESS BLOWDOWN =             2.0 GPM
    COOLING TOWER BLOWDOWN =             2396.5 GPM
  TOTAL WATER GENERATED =                4630.4 GPM

  PLANT WATER SUMMARY:
    NET MAKEUP WATER REQUIRED =         20855.5 GPM
    WATER CONSUMED / COAL FED =            4.65 LB/LB
    WATER CONSUMED / LIQUID PRODUCT =      14.3 BBL/BBL

BYPRODUCTS SUMMARY:

  SLAG =                                 1924. TON/DY
  FLYASH =                                807. TON/DY
  SULFUR =                                847. TON/DY

CARBON BALANCE SUMMARY:

  % CARBON TO LIQUID FUEL =               31.8 %
  % CARBON TO SLAG & FLYASH =              0.4 %
  % CARBON TO SEQ OR EOR =                52.9 %
  % CARBON TO CMD VENT =                   0.0 %
  % CARBON TO HRSG TAILGAS =              14.7 %
  % UNACCOUNTED CARBON =                   0.2 %

  CO2 CAPTURED (SEQ OR EOR) =           31243. TON/DY
  CO2 CAPTURED (SEQ OR EOR) =             547. MMSCFD
    CO2 PURITY =                          94.3 %
  CO2 CAPTURED / LIQ PROD  =               5.04 LB/LB
  CO2 CAPTURED / LIQ PROD  =               0.01 MMSCF/BBL
  CO2 CAPTURED / COAL FED =                1.16 LB/LB

  CO2 EMITTED =                          8803. TON/DY
  CO2 EMITTED =                           154. MMSCFD
    FROM GT =                            6662. TON/DY
      LHV TO GT =                       84684. MMBTU/DY
    FROM CMD =                              0. TON/DY
      LHV TO CMD =                          0. MMBTU/DY
    FROM REFINERY =                      2141. TON/DY
```

```
    LHV TO REFINERY =                 21902. MMBTU/DY
  CO2 EMMITED / LIQ PROD  =               1.42 LB/LB
  CO2 EMMITED / COAL FED =                0.33 LB/LB

STARTUP FLARE SUMMARY:

  CO2 FROM FLARE =                      326. TON/DY
  LHV TO FLARE =                       2380. MMBTU/DY

EFFICIENCY CALCULATIONS:

  HEAT IN (HHV BASED):
    COAL HEAT CONTENT =               24547.7 MMBTU/HR

  HEAT OUT (HHV BASED):
    NET POWER =                         751.8 MMBTU/HR
    LIQUID HEAT CONTENT =             10487.6 MMBTU/HR

  PLANT EFFICIENCY (HHV BASED):
    EFFICIENCY =                         45.8 %

CALCULATOR BLOCK GAS-TURB HIERARCHY: GAS-TURB

GAS TURBINE CALCULATIONS:

    TAILGAS FLOW =                     349949.  LB/HR
        GAS HEAT CONTENT (60 DEG F) =     534.1 BTU/SCF

    N2 FLOW =                              1.  LB/HR

    FUEL + DILUENT TOTAL FLOW =        349950.  LB/HR
        GAS HEAT CONTENT (60 DEG F) =     534.1 BTU/SCF

    GAS TURBINE AIR FLOW =            7778823.  LB/HR
        COOLING FRACTION =                 10.8 %

    COMBUSTION TEMPERATURE =             2321.  DEG F
        (A LITTLE HIGH - TUNED TO MATCH POWER OUTPUT)
    EXHAUST TEMPERATURE =                1200.  DEG F

    AIR COMPRESSOR LOAD =                 360.2 MW
    TURBINE GROSS POWER =                 671.3 MW
    GENERATOR LOSSES =                      8.6 MW
    FUEL COMPRESSOR LOSSES =                2.6 MW
        NET GAS TURBINE POWER =           300.0 MW
```

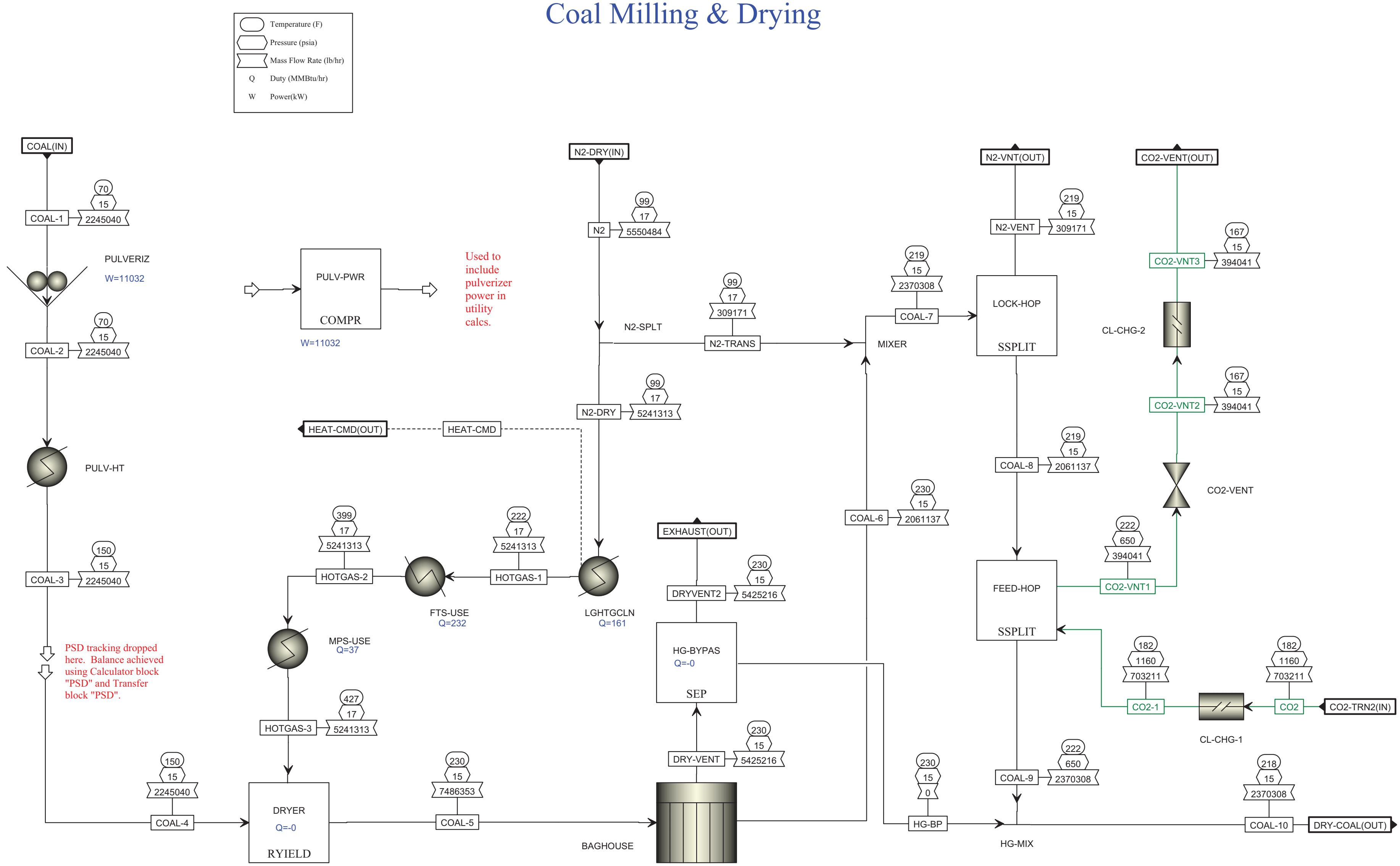
Conventional Coal to Liquid Fuels



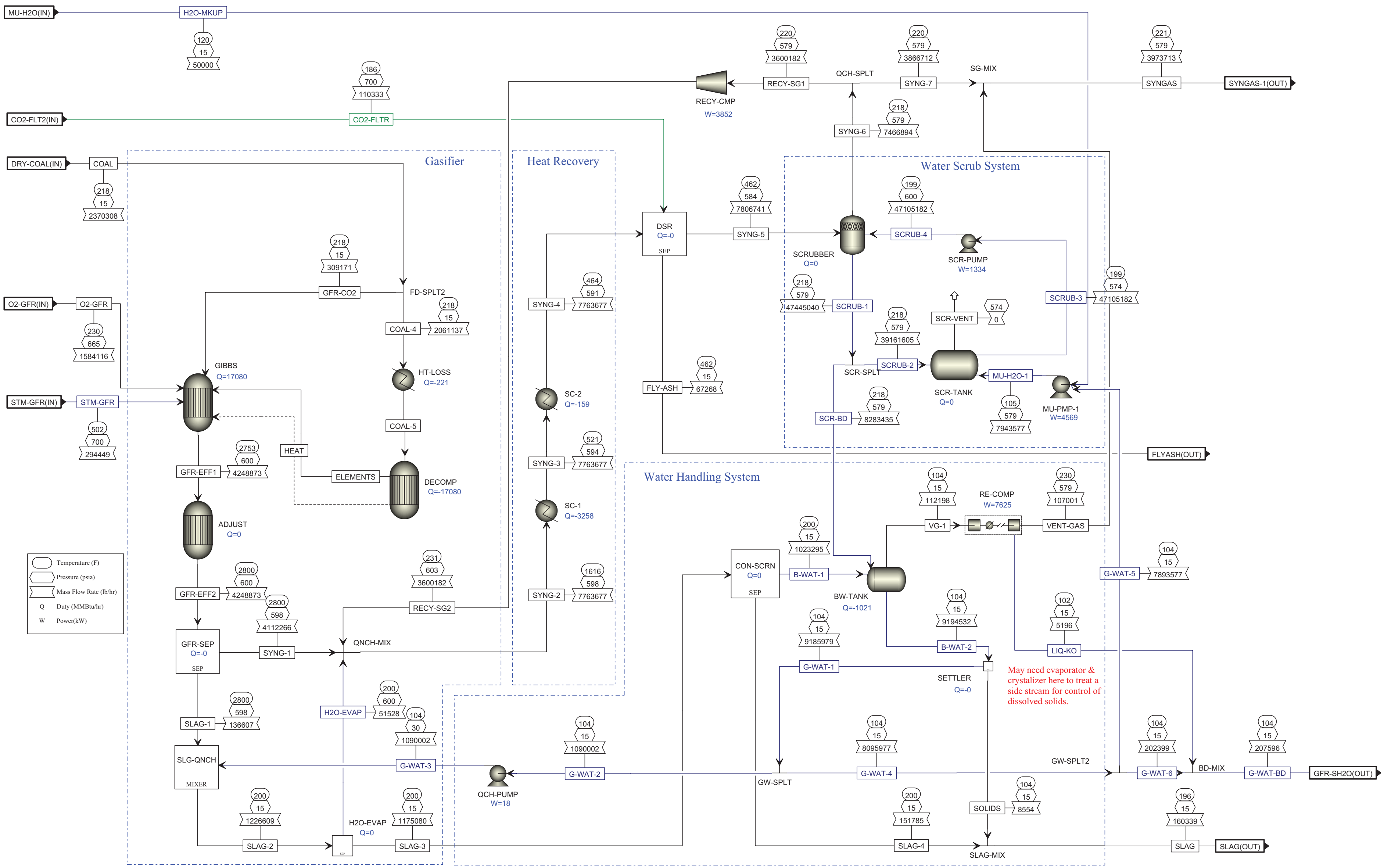
Air Separation Unit

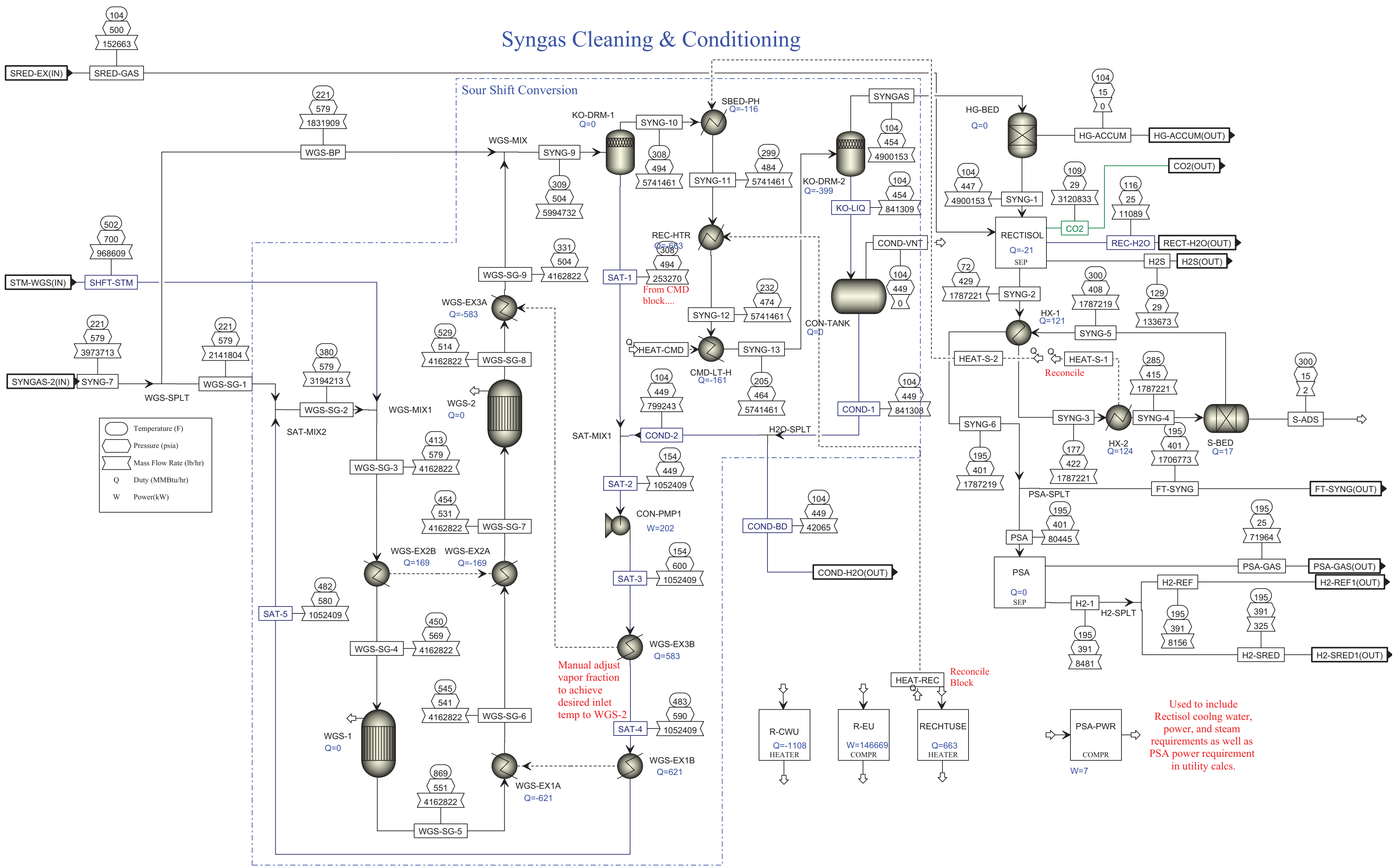


Coal Milling & Drying



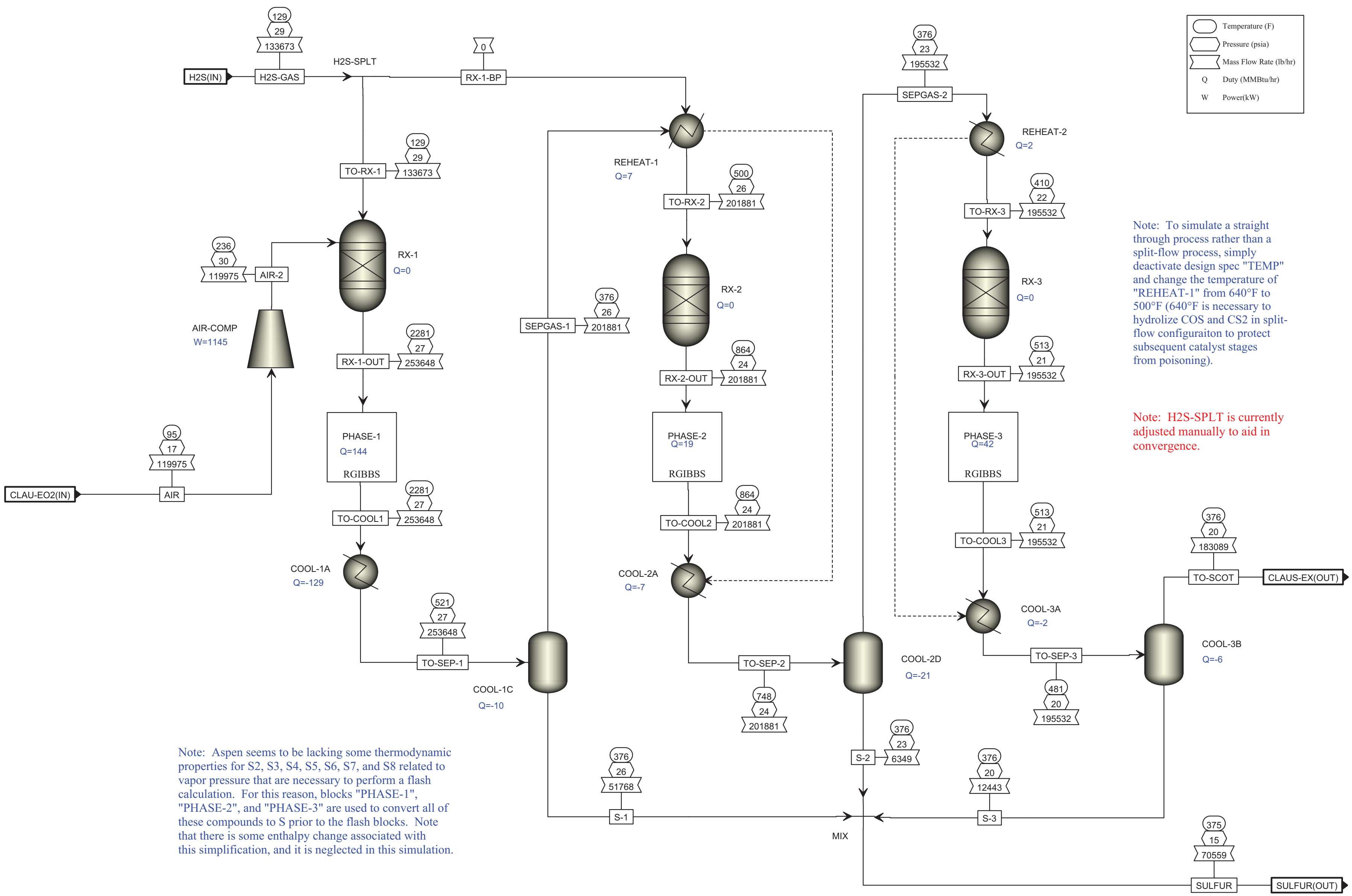
Shell Gasifier w/ Heat Recovery





Syngas Cleaning & Conditioning
Sour Shift Conversion
SRED-EX(IN)
SRED-GAS
STM-WGS(IN)
SHFT-STM
SYNGAS-2(IN)
SYNG-7
WGS-SPLT
WGS-SG-1
WGS-BP
WGS-MIX
SYNG-9
KO-DRM-1
Q=0
SYNG-10
SBED-PH
Q=-116
SYNG-11
REC-HTR
From CMD block....
SAT-1
SYNG-12
HEAT-CMD
CMD-LT-H
Q=-161
SYNG-13
SAT-MIX1
COND-2
H2O-SPLT
SAT-2
CON-PMP1
W=202
SAT-3
WGS-EX3B
Q=583
Manual adjust vapor fraction to achieve desired inlet temp to WGS-2
SAT-4
WGS-EX1B
Q=621
WGS-EX1A
Q=-621
WGS-SG-5
WGS-1
Q=0
WGS-SG-6
WGS-SG-4
WGS-EX2B
Q=169
WGS-EX2A
Q=-169
WGS-SG-7
WGS-SG-3
WGS-MIX1
WGS-SG-2
SAT-MIX2
SAT-5
WGS-2
Q=0
WGS-SG-8
WGS-EX3A
Q=-583
WGS-SG-9
SYNGAS
KO-DRM-2
Q=-399
KO-LIQ
COND-VNT
CON-TANK
Q=0
COND-1
COND-BD
COND-H2O(OUT)
HG-BED
Q=0
HG-ACCUM
HG-ACCUM(OUT)
CO2(OUT)
SYNG-1
RECTISOL
Q=-21
SEP
CO2
REC-H2O
RECT-H2O(OUT)
H2S
H2S(OUT)
SYNG-2
HX-1
Q=121
SYNG-5
HEAT-S-2
HEAT-S-1
Reconcile
SYNG-3
HX-2
Q=124
SYNG-4
S-BED
Q=17
S-ADS
SYNG-6
FT-SYNG
FT-SYNG(OUT)
PSA-SPLT
PSA
Q=0
SEP
PSA-GAS
PSA-GAS(OUT)
H2-REF
H2-REF1(OUT)
H2-1
H2-SPLT
H2-SRED
H2-SRED1(OUT)
HEAT-REC
Reconcile Block
R-CWU
Q=-1108
HEATER
R-EU
W=146669
COMPR
RECHTUSE
Q=663
HEATER
PSA-PWR
COMPR
W=7
Used to include Rectisol coolng water, power, and steam requirements as well as PSA power requirement in utility calcs.
Temperature (F)
Pressure (psia)
Mass Flow Rate (lb/hr)
Q Duty (MMBtu/hr)
W Power(kW)

Claus Process

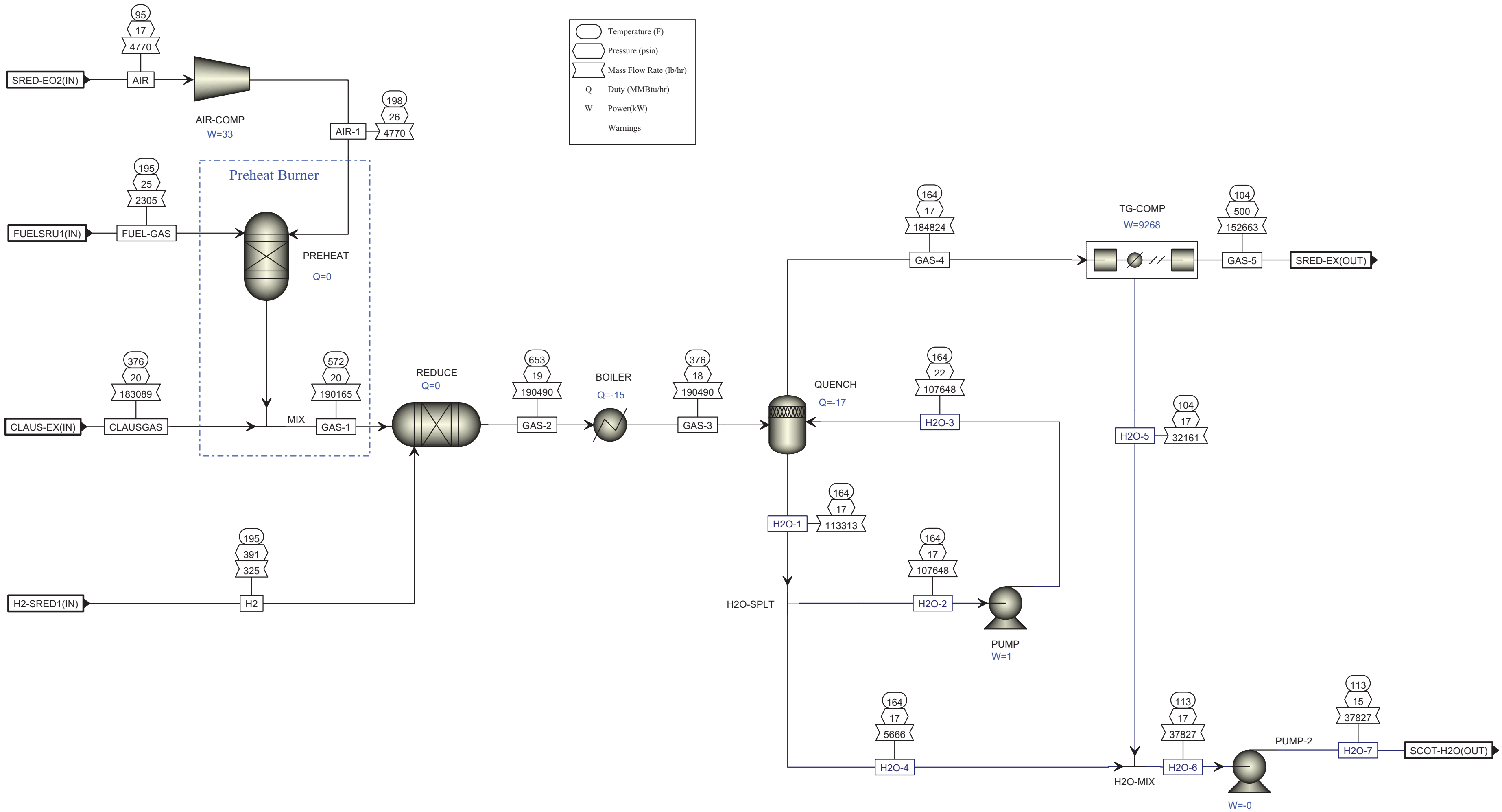


Note: To simulate a straight through process rather than a split-flow process, simply deactivate design spec "TEMP" and change the temperature of "REHEAT-1" from 640°F to 500°F (640°F is necessary to hydrolize COS and CS2 in split-flow configuraiton to protect subsequent catalyst stages from poisoning).

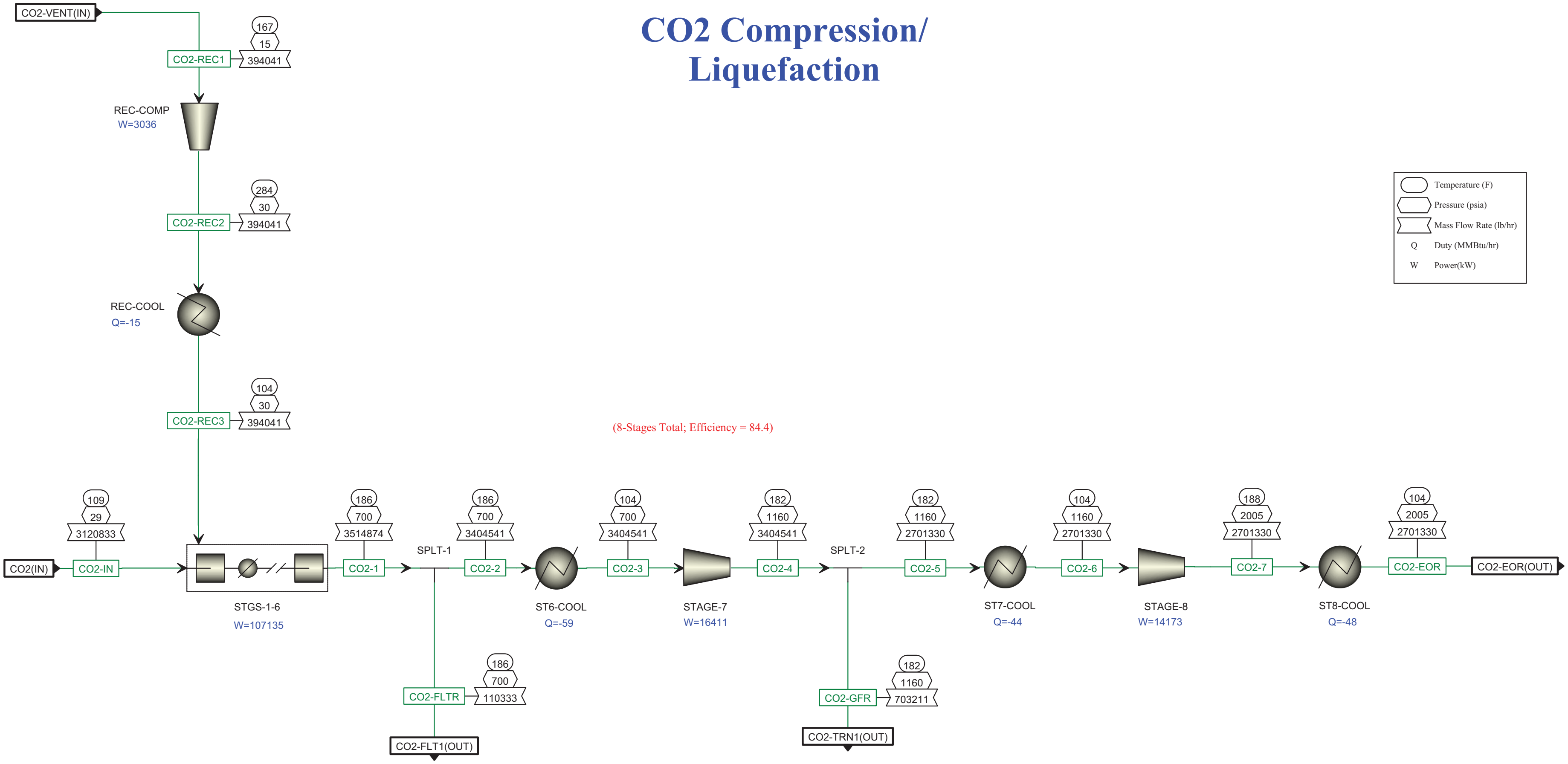
Note: H2S-SPLT is currently adjusted manually to aid in convergence.

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.

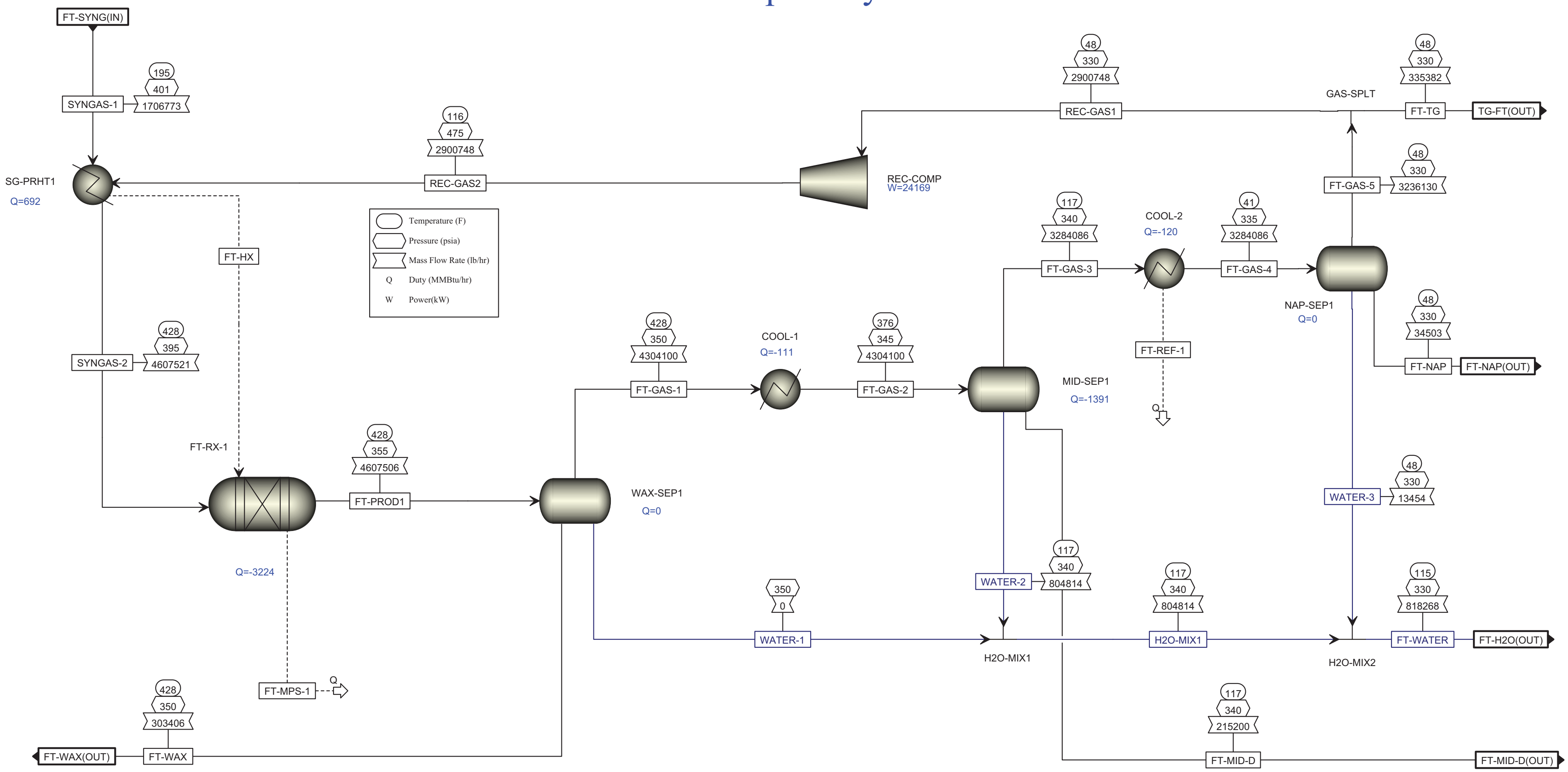
Catalytic Sulfur Reduction (SCOT or Beavon Process w/o H2S Absorber)



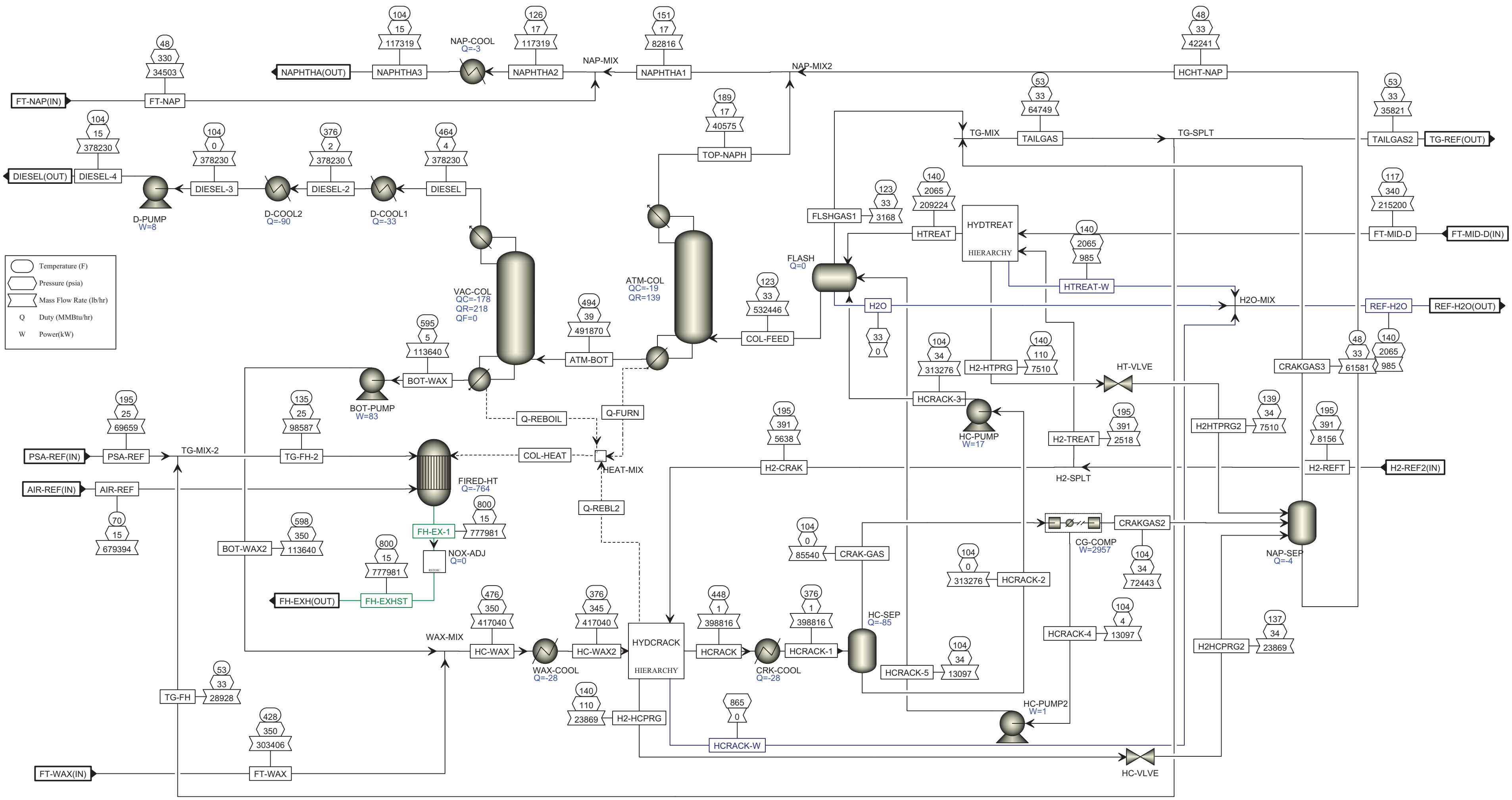
CO2 Compression/ Liquefaction



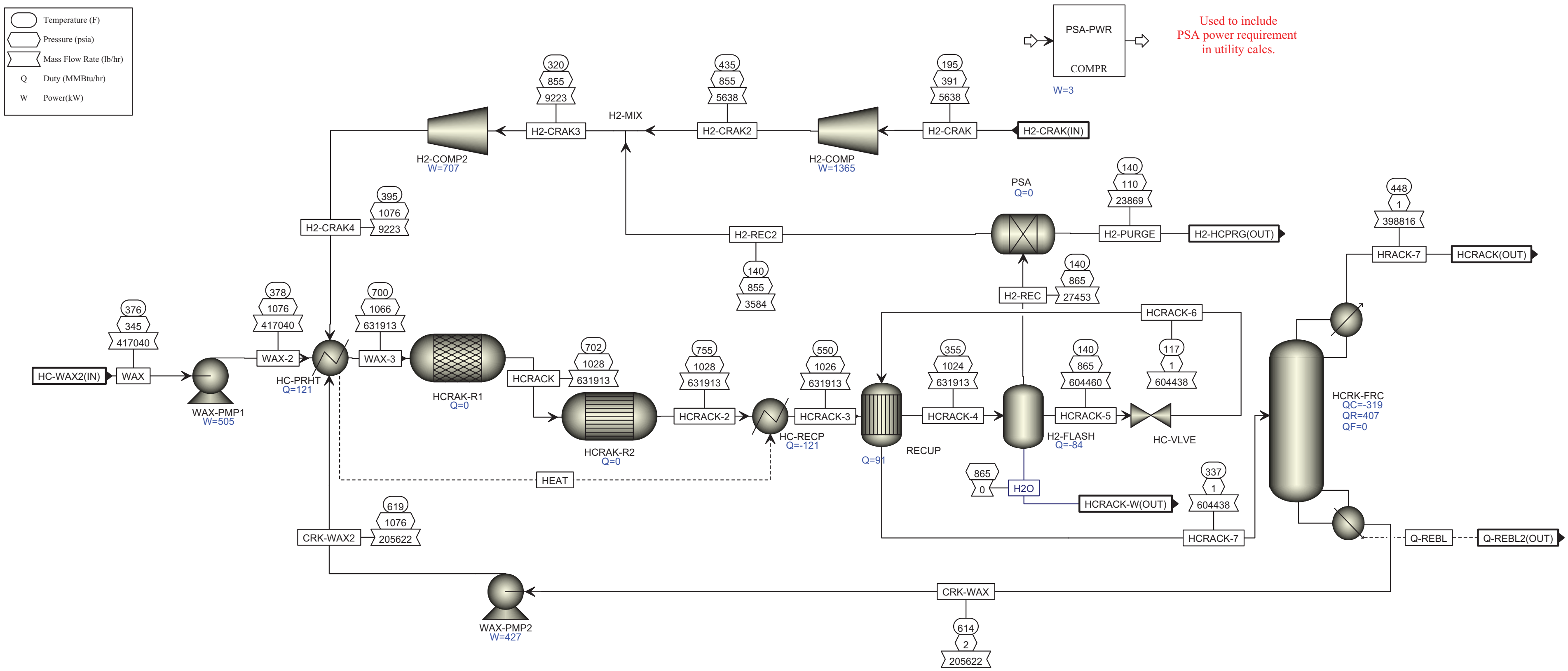
Fischer Tropsch Synthesis



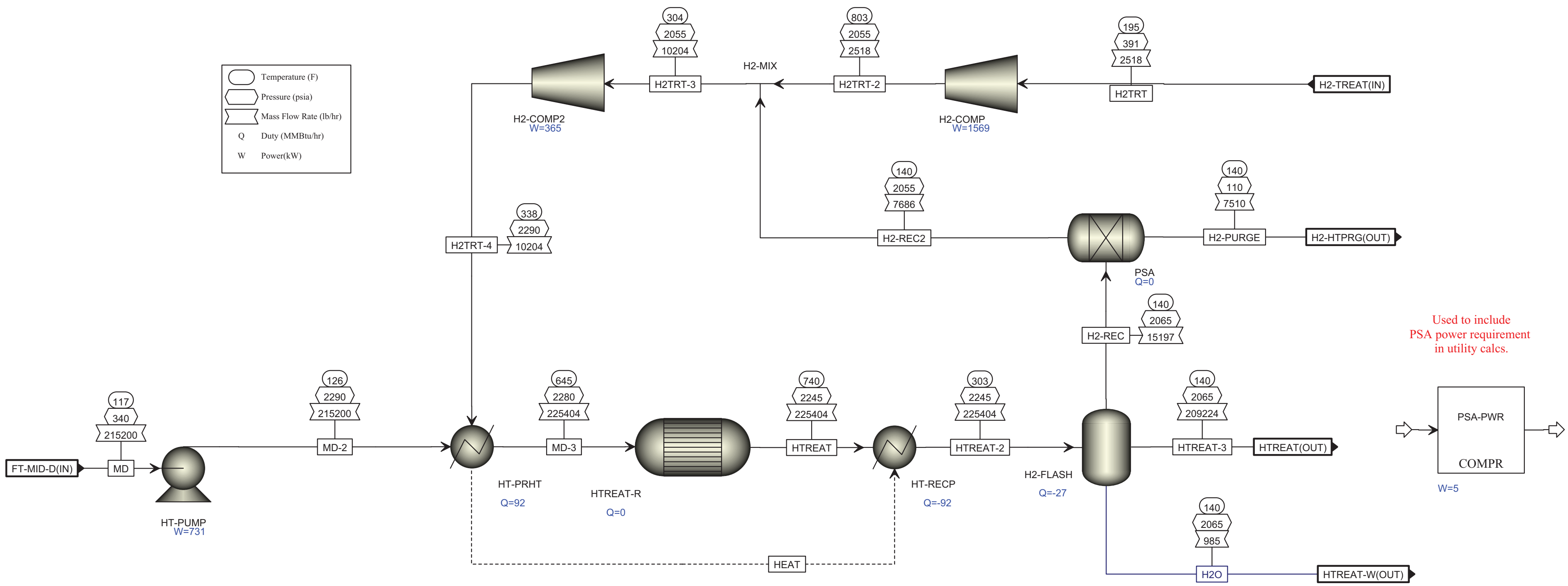
Product Upgrading and Refining



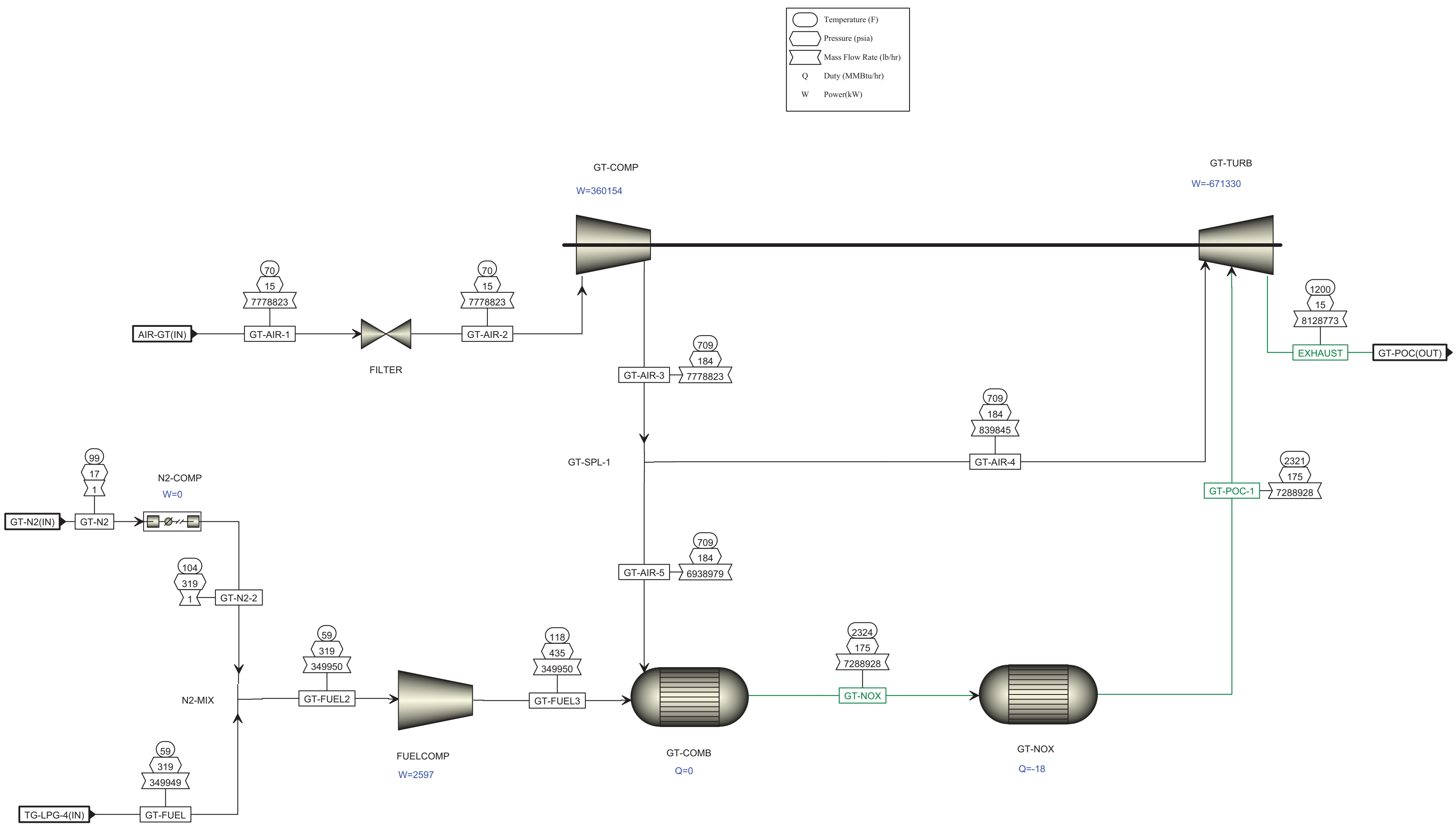
Hydrocracking



Hydrotreating

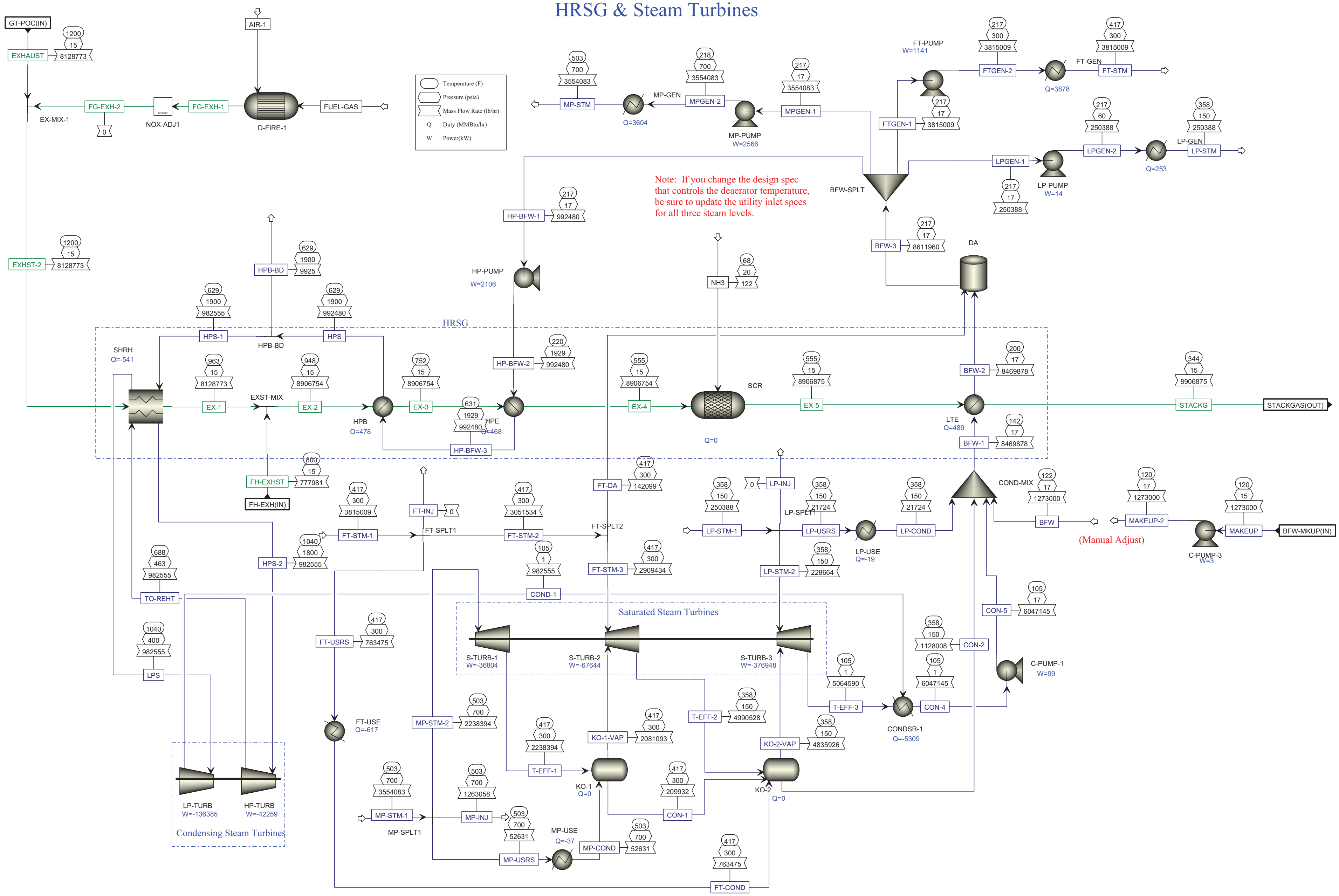


Gas Turbine

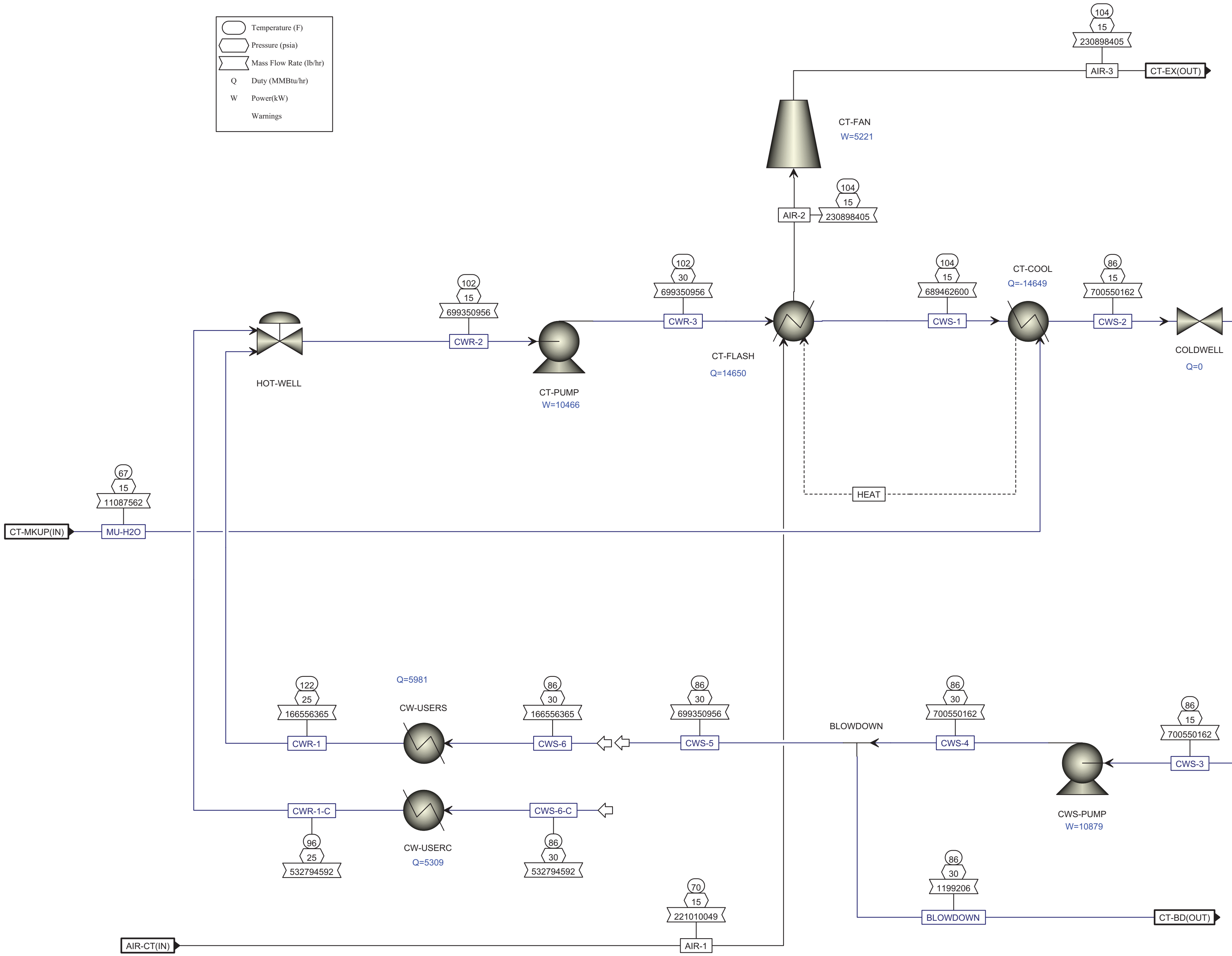


Temperature (F)
Pressure (psia)
Mass Flow Rate (lb/hr)
Q Duty (MMBtu/hr)
W Power(kW)
GT-COMP
W=360154
GT-TURB
W=-671330
70
15
7778823
AIR-GT(IN)
GT-AIR-1
FILTER
70
15
7778823
GT-AIR-2
709
184
7778823
GT-AIR-3
1200
15
8128773
EXHAUST
GT-POC(OUT)
709
184
839845
GT-AIR-4
GT-SPL-1
2321
175
7288928
GT-POC-1
99
17
1
GT-N2(IN)
GT-N2
N2-COMP
W=0
104
319
1
GT-N2-2
709
184
6938979
GT-AIR-5
N2-MIX
59
319
349950
GT-FUEL2
FUELCOMP
W=2597
118
435
349950
GT-FUEL3
GT-COMB
Q=0
2324
175
7288928
GT-NOX
GT-NOX
Q=-18
59
319
349949
TG-LPG-4(IN)
GT-FUEL

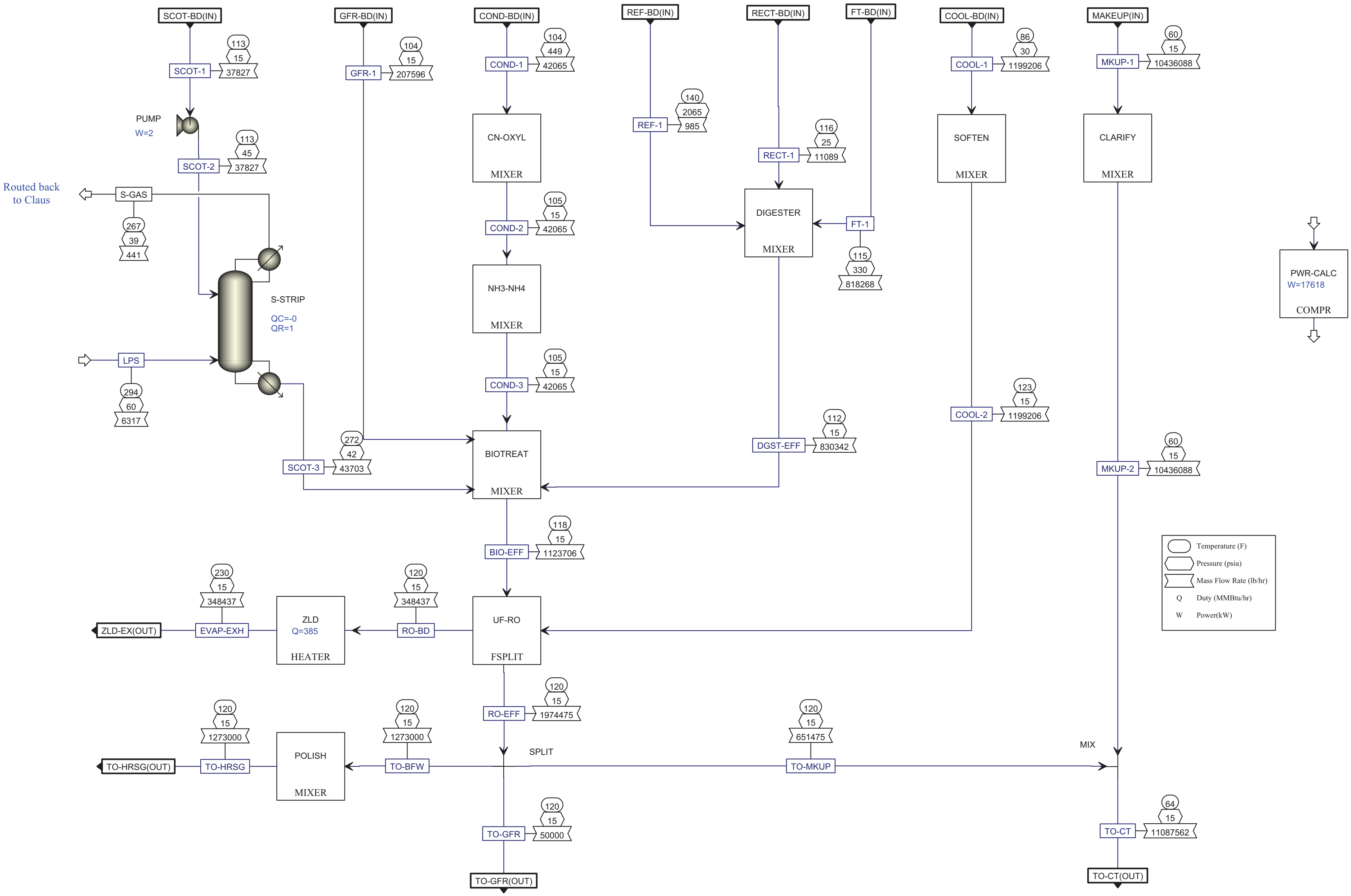
HRSG & Steam Turbines



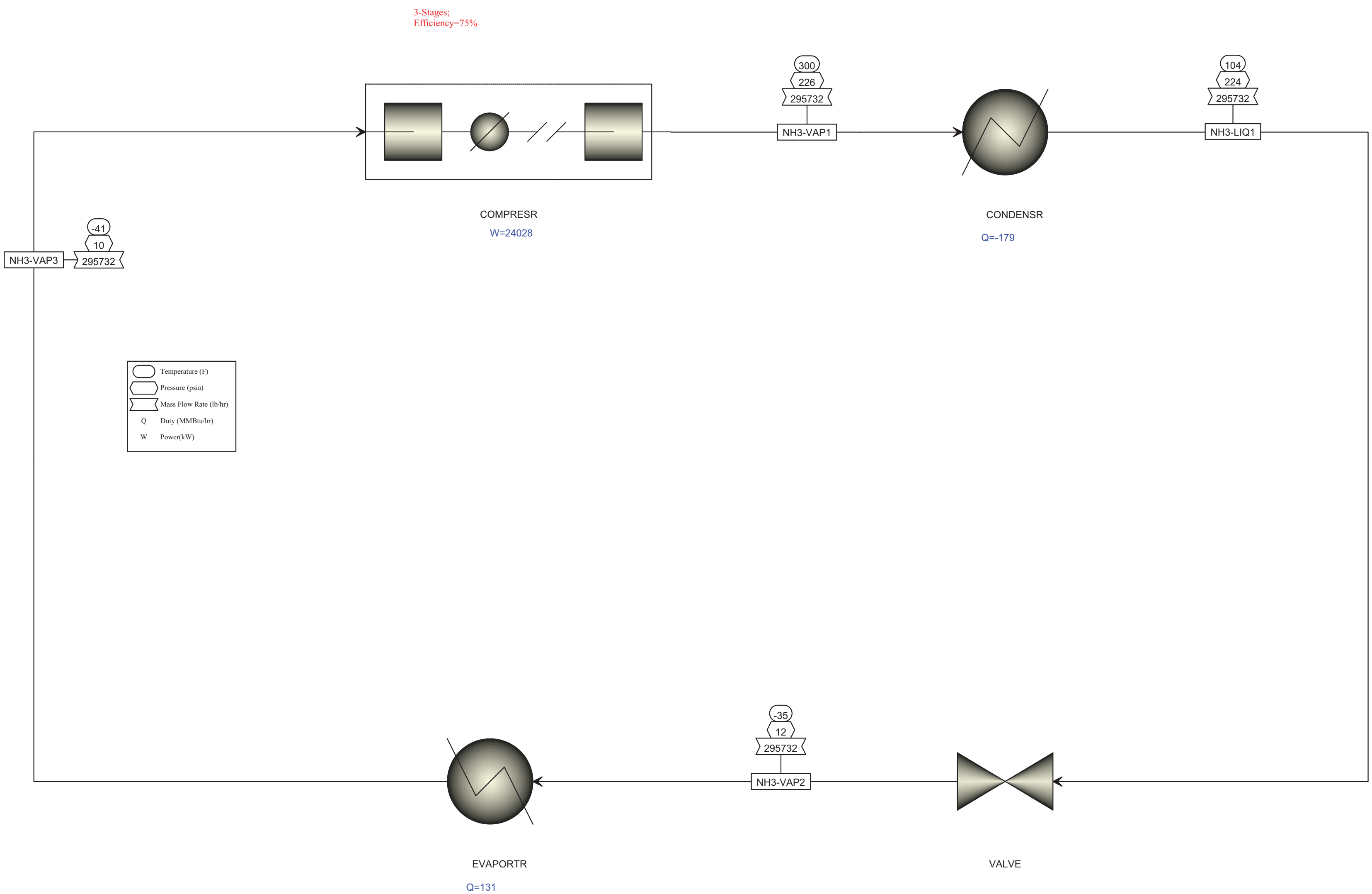
Cooling Tower

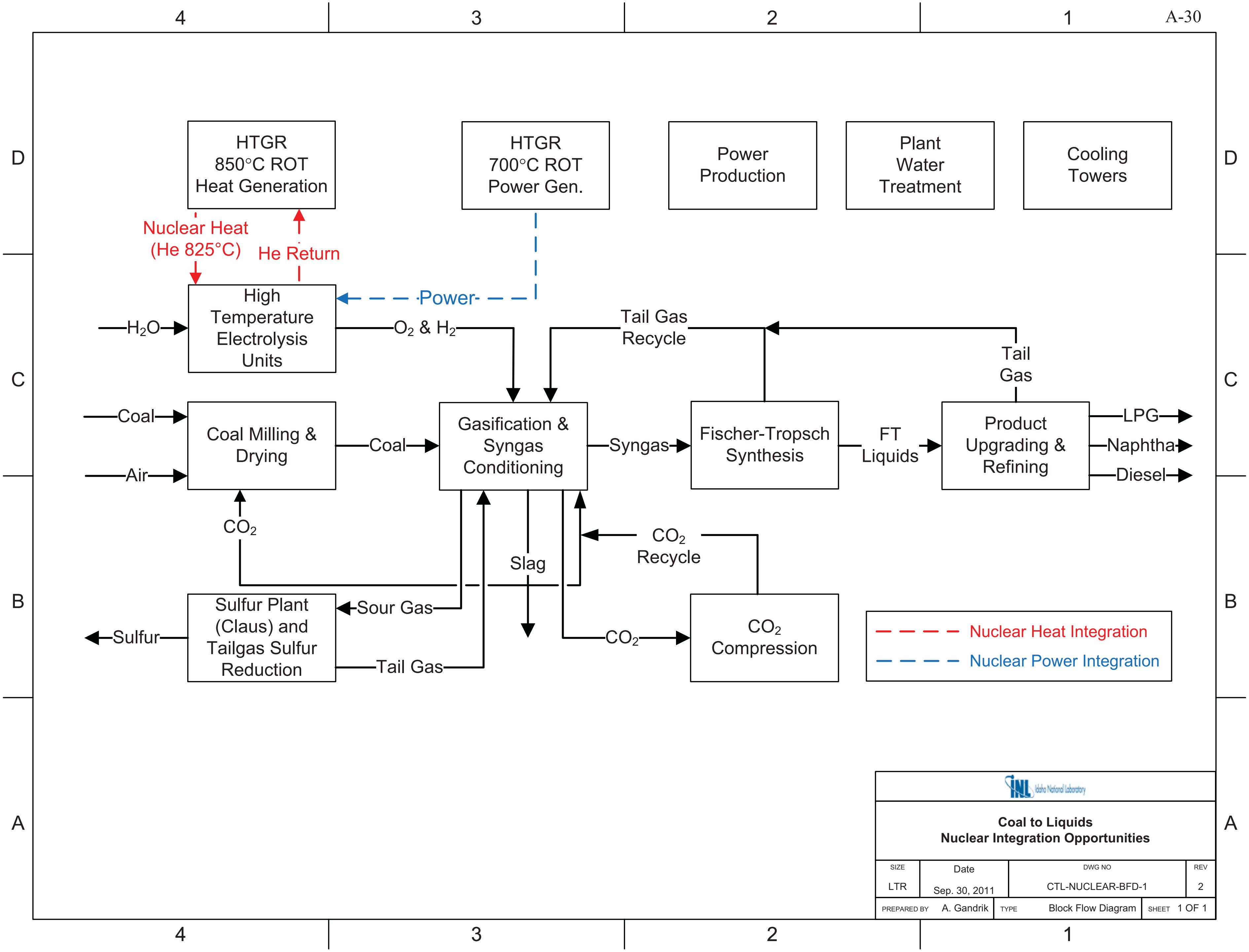


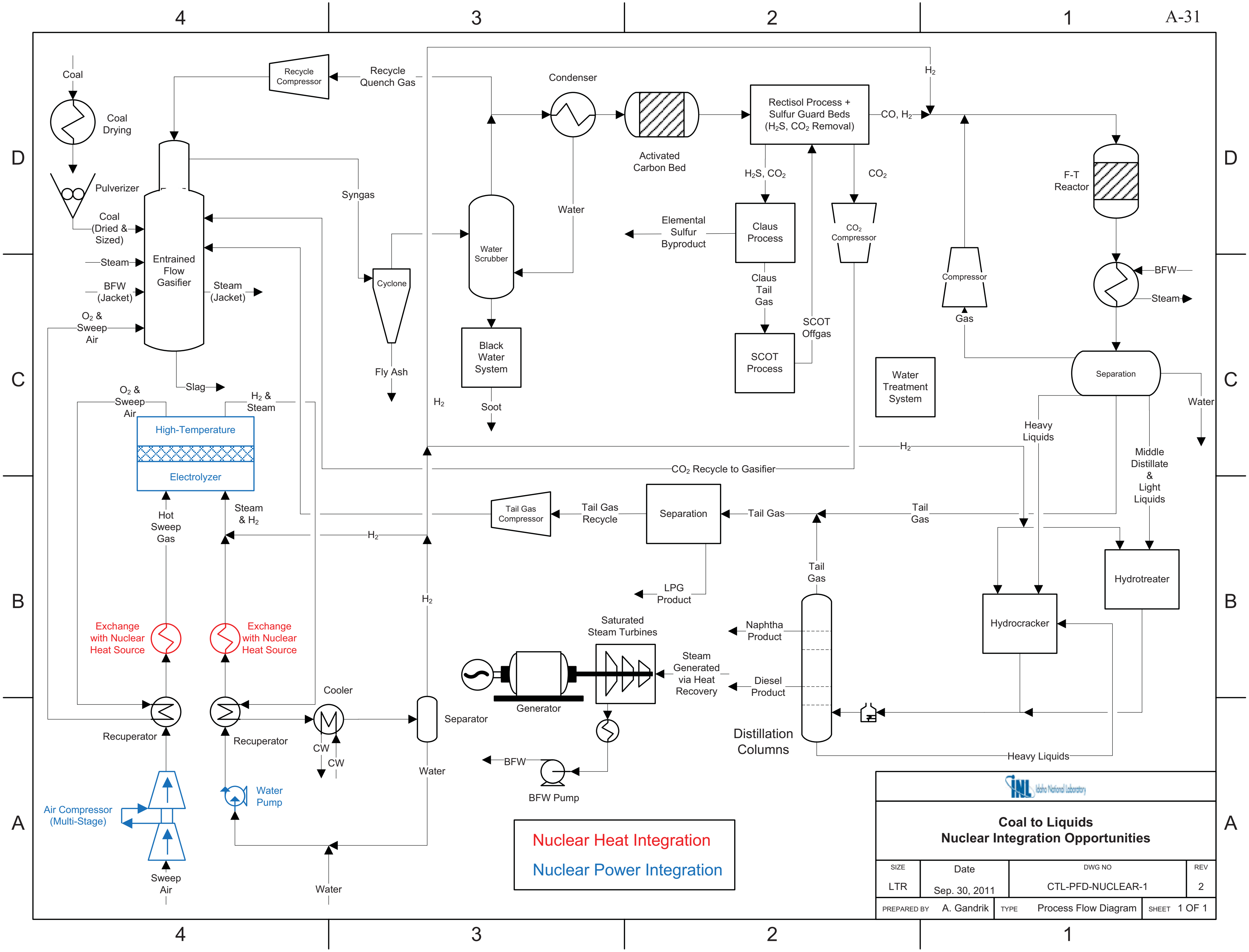
Simplified Water Treatment



Refrigeration Unit




HTGR 850°C ROT Heat Generation
HTGR 700°C ROT Power Gen.
Power Production
Plant Water Treatment
Cooling Towers
Nuclear Heat (He 825°C)
He Return
Power
High Temperature Electrolysis Units
H2O
O2 & H2
Tail Gas Recycle
Tail Gas
Coal
Air
Coal Milling & Drying
Coal
Gasification & Syngas Conditioning
Syngas
Fischer-Tropsch Synthesis
FT Liquids
Product Upgrading & Refining
LPG
Naphtha
Diesel
CO2
Slag
CO2 Recycle
Sulfur Plant (Claus) and Tailgas Sulfur Reduction
Sour Gas
Sulfur
Tail Gas
CO2
CO2 Compression
Nuclear Heat Integration
Nuclear Power Integration
Idaho National Laboratory
Coal to Liquids
Nuclear Integration Opportunities
SIZE
LTR
Date
Sep. 30, 2011
DWG NO
CTL-NUCLEAR-BFD-1
REV
2
PREPARED BY
A. Gandrik
TYPE
Block Flow Diagram
SHEET
1 OF 1


Coal
Coal Drying
Pulverizer
Coal (Dried & Sized)
Steam
BFW (Jacket)
O2 & Sweep Air
Entrained Flow Gasifier
Steam (Jacket)
Slag
Recycle Compressor
Recycle Quench Gas
Syngas
Cyclone
Fly Ash
Water Scrubber
Black Water System
Soot
Condenser
Water
Activated Carbon Bed
Rectisol Process + Sulfur Guard Beds (H2S, CO2 Removal)
CO, H2
H2
H2S, CO2
CO2
Claus Process
Elemental Sulfur Byproduct
Claus Tail Gas
SCOT Process
SCOT Offgas
CO2 Compressor
Compressor
Gas
F-T Reactor
BFW
Steam
Separation
Water
Water Treatment System
Heavy Liquids
Middle Distillate & Light Liquids
H2
CO2 Recycle to Gasifier
Tail Gas
Tail Gas Compressor
Tail Gas Recycle
Separation
Tail Gas
Tail Gas
LPG Product
Hydrotreater
Hydrocracker
O2 & Sweep Air
H2 & Steam
High-Temperature
Electrolyzer
Hot Sweep Gas
Steam & H2
H2
H2
Exchange with Nuclear Heat Source
Exchange with Nuclear Heat Source
Recuperator
Recuperator
Cooler
CW
CW
Separator
Water
Air Compressor (Multi-Stage)
Sweep Air
Water Pump
Water
Saturated Steam Turbines
Generator
Steam Generated via Heat Recovery
BFW
BFW Pump
Naphtha Product
Diesel Product
Distillation Columns
Heavy Liquids
Nuclear Heat Integration
Nuclear Power Integration
Idaho National Laboratory
Coal to Liquids
Nuclear Integration Opportunities
SIZE
LTR
Date
Sep. 30, 2011
DWG NO
CTL-PFD-NUCLEAR-1
REV
2
PREPARED BY
A. Gandrik
TYPE
Process Flow Diagram
SHEET
1 OF 1

```
Calculator Block ELECSUM

   ELECTROLYSIS SUMMARY:

     FEED SUMMARY:

       H2O FEED:
         MASS FLOW =                     1448890. LB/HR
           TEMPERATURE =                      70. DEG. F
           PRESSURE =                         14.7 PSI

     PRODUCT SUMMARY:

       H2 PRODUCT:
         MASS FLOW =                      163068. LB/HR
           H2 PURITY =                        99.92 MOL-%
         TEMPERATURE =                        79. DEG. F
         PRESSURE =                          710.7 PSI

       O2 PRODUCT:
         MASS FLOW =                     1285026. LB/HR
           O2 PURITY =                       100.00 MOL-%
         TEMPERATURE =                        79. DEG. F
         PRESSURE =                          710.7 PSI

     HEAT AND POWER SUMMARY:

       ELECTROLYSIS POWER REQUIREMENT =    2511.8 MW

       HEAT SUMMARY:

         REACTOR HEAT:
           DUTY REQUIRED =                  2330.2 MMBTU/HR
           HELIUM MASS FLOW =            2569270. LB/HR
             INLET TEMPERATURE =            1517. DEG. F
             OUTLET TEMPERATURE =            786. DEG F.
             PRESSURE DROP =                 -20.3 PSI

         TOPPING HEAT:
           DUTY REQUIRED =                     0.0 MMBTU/HR
           SYNGAS MASS FLOW =            4199040. LB/HR
             INLET TEMPERATURE =            1616. DEG. F
             OUTLET TEMPERATURE =              0. DEG F.
             PRESSURE DROP =                 600.0 PSI

Calculator Block SUMMARY

   FEED & PRODUCT SUMMARY:

     FEEDS:

       RAW COAL FEED RATE =                 9354.1 TON/DY
         COAL HHV AS FED =                 10934. BTU/LB
         COAL MOISTURE AS FED =               13.70 %

       PROXIMATE ANALYSIS:
          MOISTURE                            13.70 %
          FIXED CARBON                        40.12 %
          VOLATILE MATTER                     49.28 %
          ASH                                 10.60 %

       ULTIMATE ANALYSIS:
          ASH                                 10.60 %
          CARBON                              70.27 %
          HYDROGEN                             4.84 %
```

NITROGEN	1.36 %
CHLORINE	0.11 %
SULFUR	3.72 %
OXYGEN	9.10 %
SULFANAL ANALYSIS:	
PYRITIC	1.94 %
SULFATE	0.08 %
ORGANIC	1.70 %

INTERMEDIATES:

COAL FEED RATE AFTER DRYING =	8587.9 TON/DY
COAL HHV AFTER DRYING =	11910. BTU/LB
COAL MOISTURE AFTER DRYNG =	6.00 %
RAW SYNGAS MASS FLOW =	2188240. LB/HR
RAW SYNGAS VOLUME FLOW =	932. MMSCFD
RAW SYNGAS HHV (WET) =	239.4 BTU/SCF
RAW SYNGAS HHV (DRY) =	277.0 BTU/SCF
RAW SYNGAS COMPOSITION:	
H2	25.5 MOL.%
CO	46.7 MOL.%
CO2	7.7 MOL.%
N2	5.5 MOL.%
H2O	13.6 MOL.%
CH4	20. PPMV
H2S	7064. PPMV
QUENCHED SYNGAS MASS FLOW =	1997748. LB/HR
QUENCHED SYNGAS VOLUME FLOW =	827. MMSCFD
QUENCHED SYNGAS HHV (WET) =	268.9 BTU/SCF
QUENCHED SYNGAS HHV (DRY) =	273.2 BTU/SCF
QUENCHED SYNGAS COMPOSITION:	
H2	28.8 MOL.%
CO	52.8 MOL.%
CO2	9.6 MOL.%
N2	6.4 MOL.%
H2O	1.6 MOL.%
CH4	22. PPMV
H2S	7993. PPMV
CLEANED SYNGAS MASS FLOW =	1720988. LB/HR
CLEANED SYNGAS VOLUME FLOW =	1425. MMSCFD
CLEANED SYNGAS HHV (WET) =	310.4 BTU/SCF
CLEANED SYNGAS HHV (DRY) =	310.5 BTU/SCF
CLEANED SYNGAS COMPOSITION:	
H2	65.5 MOL.%
CO	30.6 MOL.%
CO2	0.1 MOL.%
N2	3.7 MOL.%
H2O	0.0 MOL.%
CH4	14. PPMV
H2S	0. PPMV

PRODUCTS:

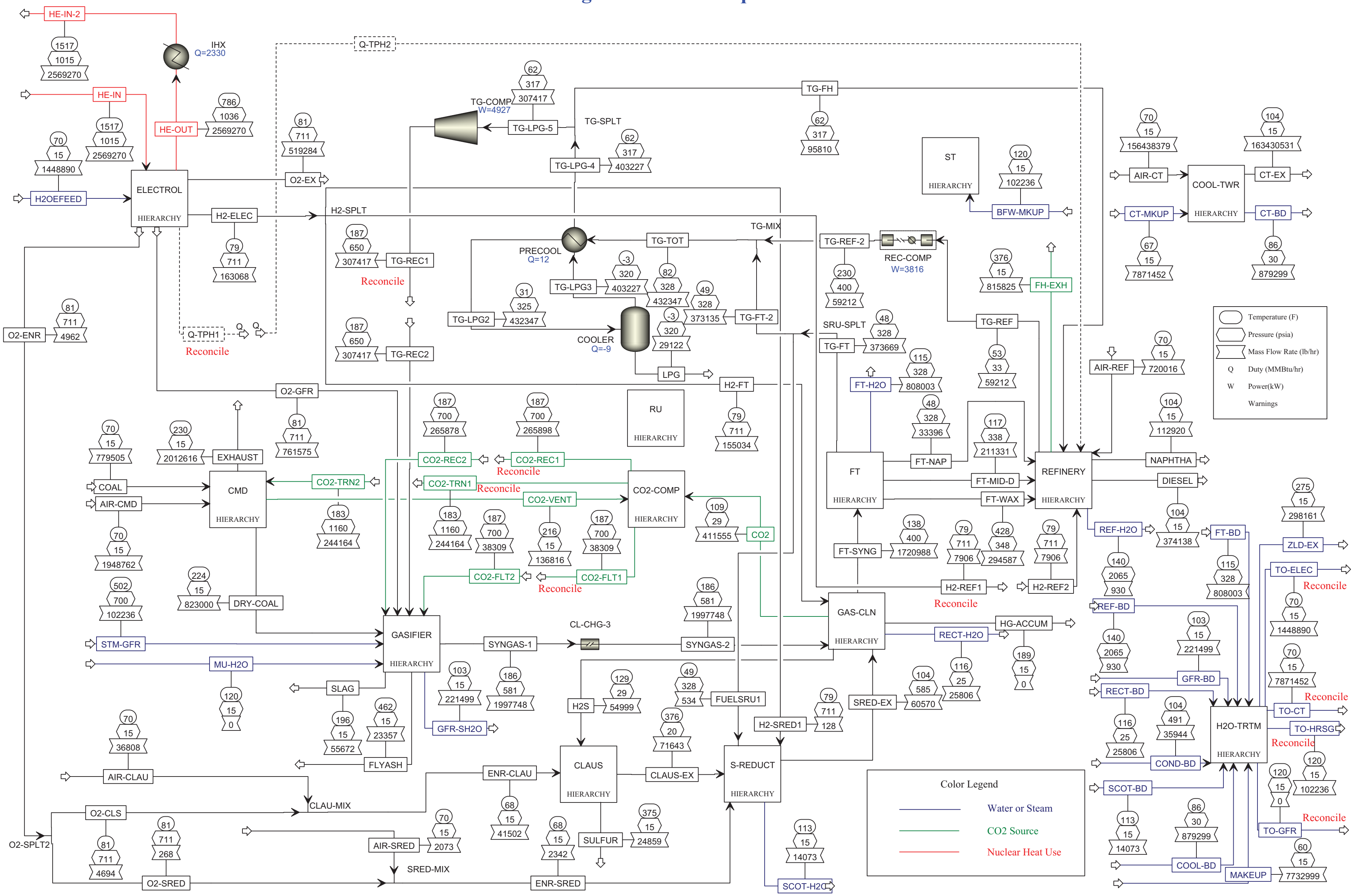
LIQUID PRODUCTS PRODUCED =	516180. LB/HR
LIQUID PRODUCTS PRODUCED =	6194.2 TON/DY
DIESEL =	374138. LB/HR
DIESEL =	4490. TON/DY
NAPHTHA =	112920. LB/HR
NAPHTHA =	1355. TON/DY
LPG =	29122. LB/HR
LPG =	349. TON/DY

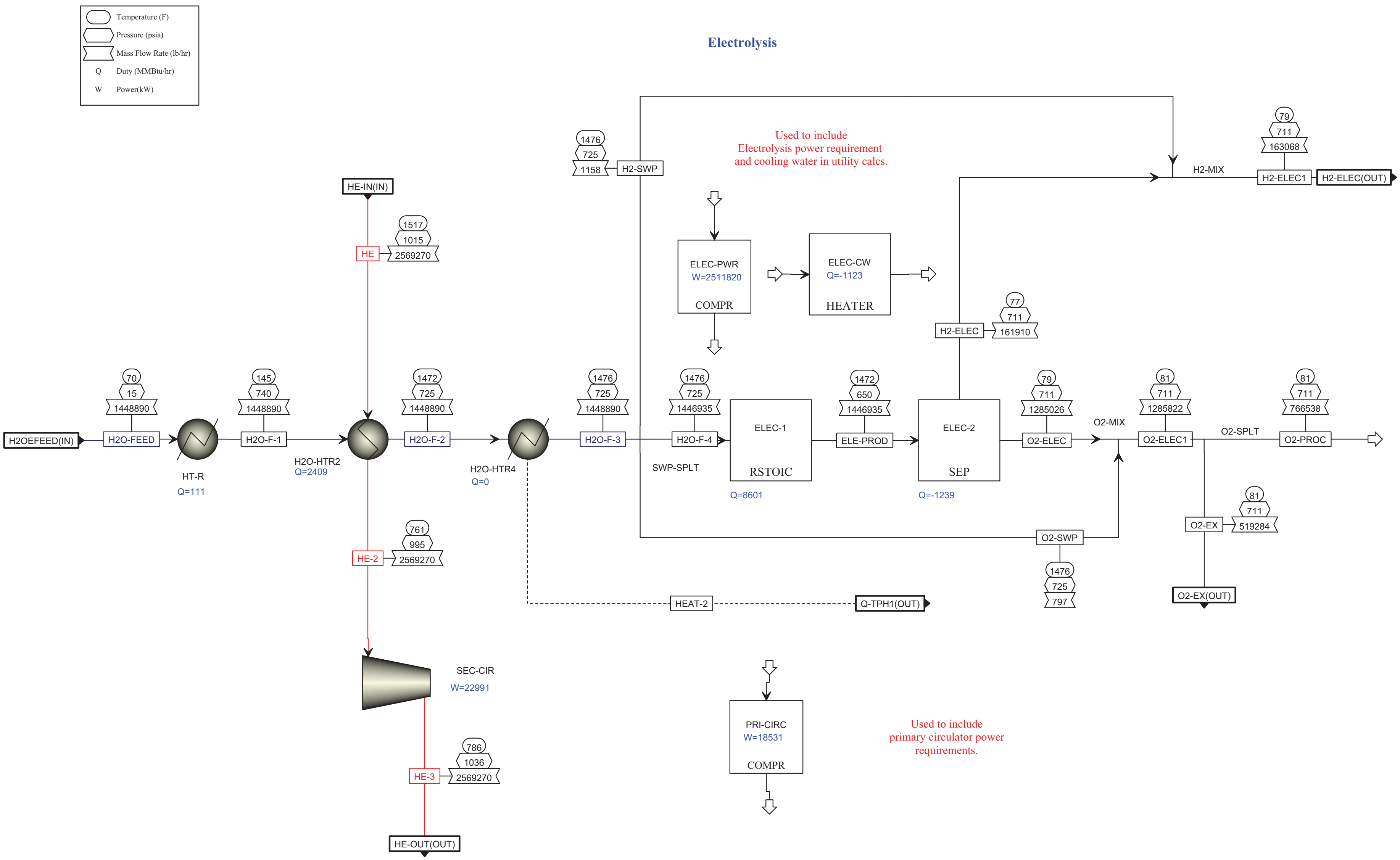
```
    LIQUID PRODUCTS PRODUCED =          50002. BBL/DY
      DIESEL =                          35194. BBL/DY
      NAPHTHA =                         11810. BBL/DY
      LPG =                              2998. BBL/DY
    LIQUIDS PRODUCED / COAL FED =           0.66 LB/LB
    LIQUIDS PRODUCED / COAL FED =           5.35 BBL/TON

  FUEL PROPERTIES:
                              DIESEL       NAPHTHA        LPG
    PROD. RATE, BBL/DAY    35194.        11810.         2998.
    LHV RATE, MMBTU/DAY   170016.        51098.        13285.
    MW                       188.2          80.8          59.8
    API GRAVITY               54.3          81.8
    DENSITY, LB/GAL            6.07          5.46          5.55
    CETANE NO.                94.3          28.3
    HHV CONTENT, BTU/LB    20369.        20351.        20542.
    LHV CONTENT, BTU/LB    18934.        18855.        19008.
    % CARBON                  84.7          82.2          81.4
    D86T CURVE, DEG. C:
        0%                  147.         -113.
        10%                 182.           21.
        20%                 200.           50.
        50%                 248.           81.
        90%                 327.          120.
        100%                355.          162.

POWER CALCULATIONS:

  POWER GENERATORS:
    SATURATED TURBINE POWER OUTPUT =     401.7 MW
  TOTAL POWER GENERATED =                401.7 MW

  POWER CONSUMERS:
    COAL PROCESSING POWER CONSUMPTION =    9.5 MW
    ELECTROLYSIS POWER CONSUMPTION =    2511.8 MW
    PRIMARY CIRC. POWER CONSUMPTION =     18.5 MW
    SECONDARY CIRC. POWER CONSUMPTION =   23.0 MW
    GASIFIER POWER CONSUMPTION =          18.1 MW
    GAS CLEANING POWER CONSUMPTION =      59.7 MW
    SCOT PROCESS POWER CONSUMPTION =       3.8 MW
    CLAUS POWER CONSUMPTION =              0.5 MW
    CO2 LIQUEF. POWER CONSUMPTION =       19.6 MW
    FISHER TROPSCH POWER CONSUMPTION =    28.0 MW
    REFINERY POWER CONSUMPTION =          15.1 MW
    POWER BLOCK POWER CONSUMPTION =        2.6 MW
    REFRIGERATION POWER CONSUMPTION =     26.2 MW
    COOLING TOWER POWER CONSUMPTION =     18.5 MW
    WATER TREATMENT POWER CONSUMPTION =   13.0 MW
  TOTAL POWER CONSUMED =                2767.9 MW

  NET PLANT POWER (+ GEN, - CONS)=     -2366.3 MW

WATER BALANCE:

  EVAPORATIVE LOSSES:
    CMD WATER NOT RECOVERED =             151.7 GPM
    COOLING TOWER EVAPORATION =         17086.5 GPM
    ZLD SYSTEM EVAPORATION =              595.8 GPM
  TOTAL EVAPORATIVE LOSSES =            17834.0 GPM

  WATER CONSUMED:
    ELECTROLYSIS FEED =                  2895.5 GPM
    GASIFIER ISLAND MAKEUP =                0.0 GPM
    BOILER FEED WATER MAKEUP =            204.3 GPM
    COOLING TOWER MAKEUP =              15730.3 GPM
```

```
  TOTAL WATER CONSUMED =                18830.1 GPM

  WATER GENERATED:
    GASIFIER ISLAND BLOWDOWN =            442.6 GPM
    SYNGAS CONDENSER BLOWDOWN =            71.8 GPM
    RECTISOL BLOWDOWN =                    51.6 GPM
    SULFUR REDUCTION BLOWDOWN =            28.1 GPM
    FT PROCESS BLOWDOWN =                1614.7 GPM
    REFINERY PROCESS BLOWDOWN =             1.9 GPM
    COOLING TOWER BLOWDOWN =             1757.2 GPM
  TOTAL WATER GENERATED =                3967.9 GPM

  PLANT WATER SUMMARY:
    NET MAKEUP WATER REQUIRED =         15453.6 GPM
    WATER CONSUMED / COAL FED =            9.92 LB/LB
    WATER CONSUMED / LIQUID PRODUCT =      10.6 BBL/BBL

BYPRODUCTS SUMMARY:

  SLAG =                                  668. TON/DY
  FLYASH =                                280. TON/DY
  SULFUR =                                298. TON/DY

CARBON BALANCE SUMMARY:

  % CARBON TO LIQUID FUEL =                91.7 %
  % CARBON TO SLAG & FLYASH =               0.4 %
  % CARBON TO SEQ OR EOR =                  0.0 %
  % CARBON TO HRSG TAILGAS =                7.1 %
  % UNACCOUNTED CARBON =                    0.8 %

  CO2 EMITTED =                          1473. TON/DY
  CO2 EMITTED =                            26. MMSCFD
    FROM REFINERY =                      1473. TON/DY
      LHV TO REFINERY =                 22012. MMBTU/DY
  CO2 EMMITED / LIQ PROD  =                0.24 LB/LB
  CO2 EMMITED / COAL FED =                 0.16 LB/LB

STARTUP FLARE SUMMARY:

  CO2 FROM FLARE =                        152. TON/DY
  LHV TO FLARE =                         1083. MMBTU/DY

EFFICIENCY CALCULATIONS:

    HEAT IN (HHV BASED):
        COAL HEAT CONTENT =              8523.3 MMBTU/HR

    HEAT OUT (HHV BASED):
        NET POWER =                     -8074.0 MMBTU/HR
        LIQUID HEAT CONTENT =           10517.0 MMBTU/HR

    PLANT EFFICIENCY (HHV BASED):
        EFFICIENCY =                       28.7 %

HTGR SUMMARY:

  850C SUMMARY - HEAT ONLY:
    850C ROT NET HEAT =                   664. MWT
    GROSS HEAT SUPPLIED =                 683. MWT
    PRIMARY CIRC. PWR =                    19. MWE

  700C SUMMARY - ELECTRICITY ONLY:
    700C ROT NET HEAT =                  5440. MWT
    NET PWR SUPPLIED =                   2366. MWT
```

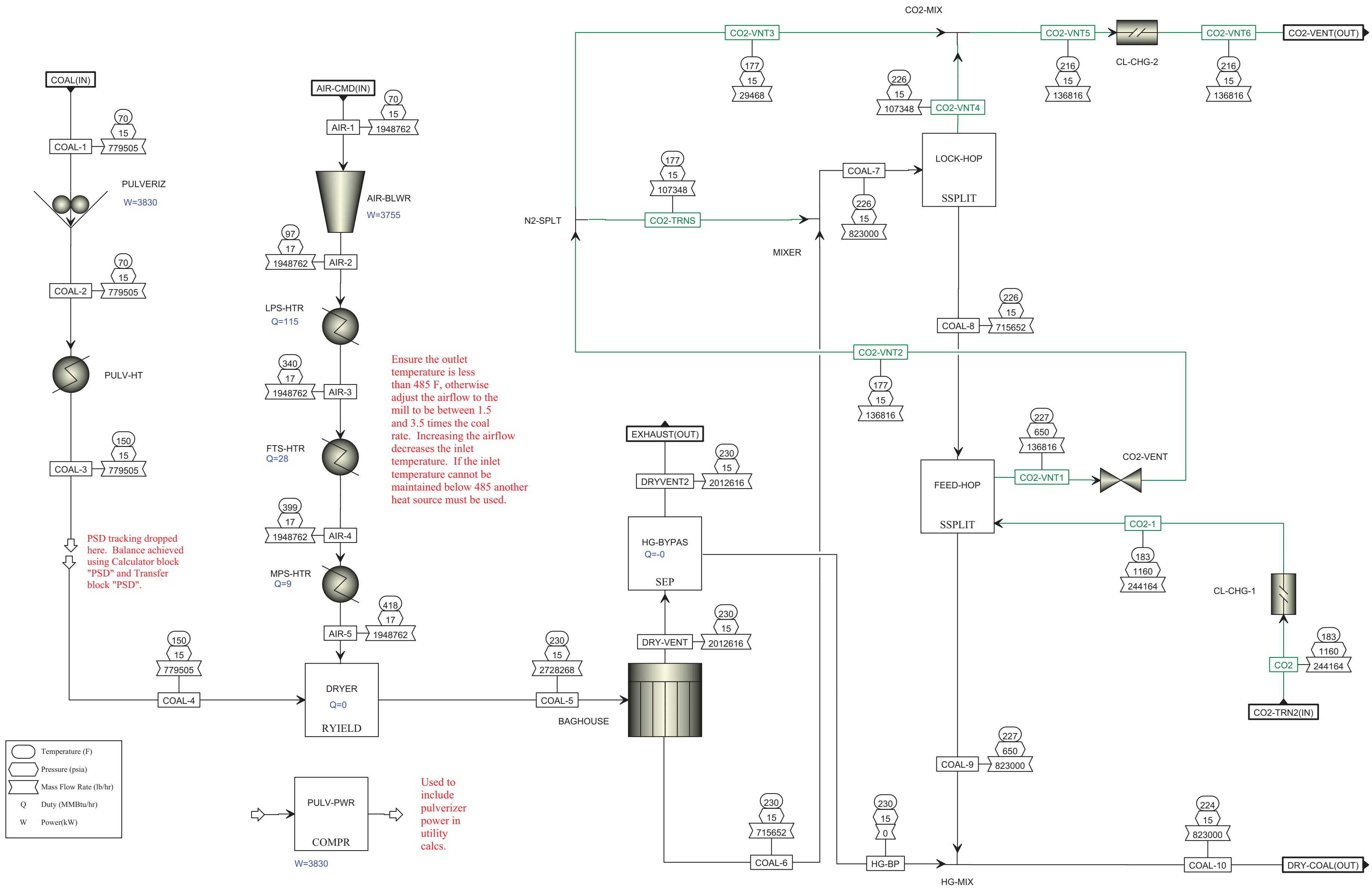
```
PWR TO 850C PRI CIRC. =                  19. MWE
PWR TO PROCESS =                       2348. MWE
```

Nuclear-Integrated Coal to Liquid Fuels

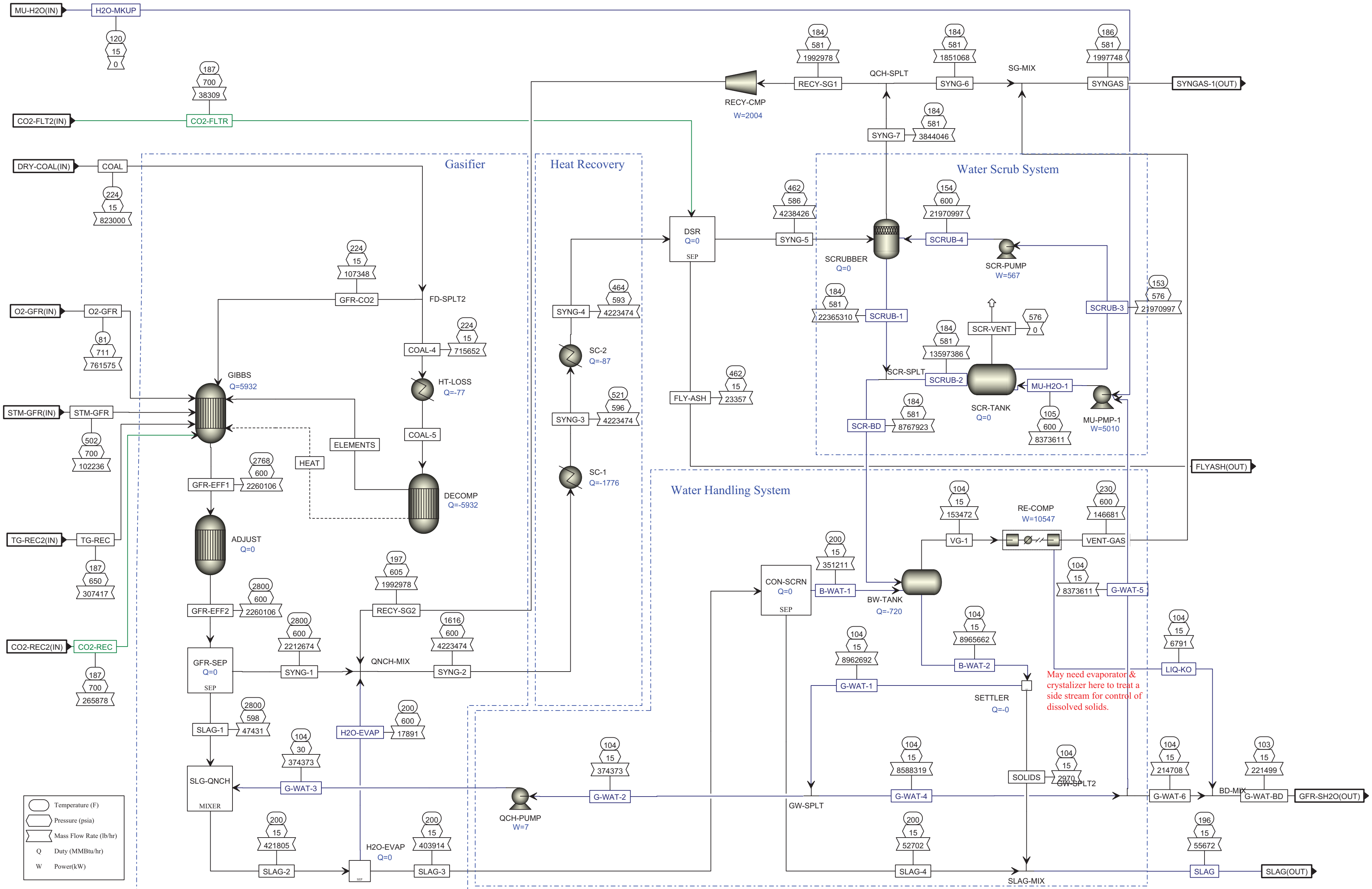



Temperature (F)
Pressure (psia)
Mass Flow Rate (lb/hr)
Q Duty (MMBtu/hr)
W Power(kW)
Electrolysis
Used to include
Electrolysis power requirement
and cooling water in utility calcs.
H2OEFEED(IN)
H2O-FEED
70
15
1448890
HT-R
Q=111
H2O-F-1
145
740
1448890
HE-IN(IN)
HE
1517
1015
2569270
H2O-HTR2
Q=2409
H2O-F-2
1472
725
1448890
H2O-HTR4
Q=0
H2O-F-3
1476
725
1448890
SWP-SPLT
H2O-F-4
1476
725
1446935
H2-SWP
1476
725
1158
ELEC-PWR
W=2511820
COMPR
ELEC-CW
Q=-1123
HEATER
ELEC-1
RSTOIC
Q=8601
ELE-PROD
1472
650
1446935
ELEC-2
SEP
Q=-1239
H2-ELEC
77
711
161910
H2-MIX
H2-ELEC1
79
711
163068
H2-ELEC(OUT)
O2-ELEC
79
711
1285026
O2-MIX
O2-ELEC1
81
711
1285822
O2-SPLT
O2-PROC
81
711
766538
O2-EX
81
711
519284
O2-EX(OUT)
O2-SWP
1476
725
797
HE-2
761
995
2569270
HEAT-2
Q-TPH1(OUT)
SEC-CIR
W=22991
HE-3
786
1036
2569270
HE-OUT(OUT)
PRI-CIRC
W=18531
COMPR
Used to include
primary circulator power
requirements.

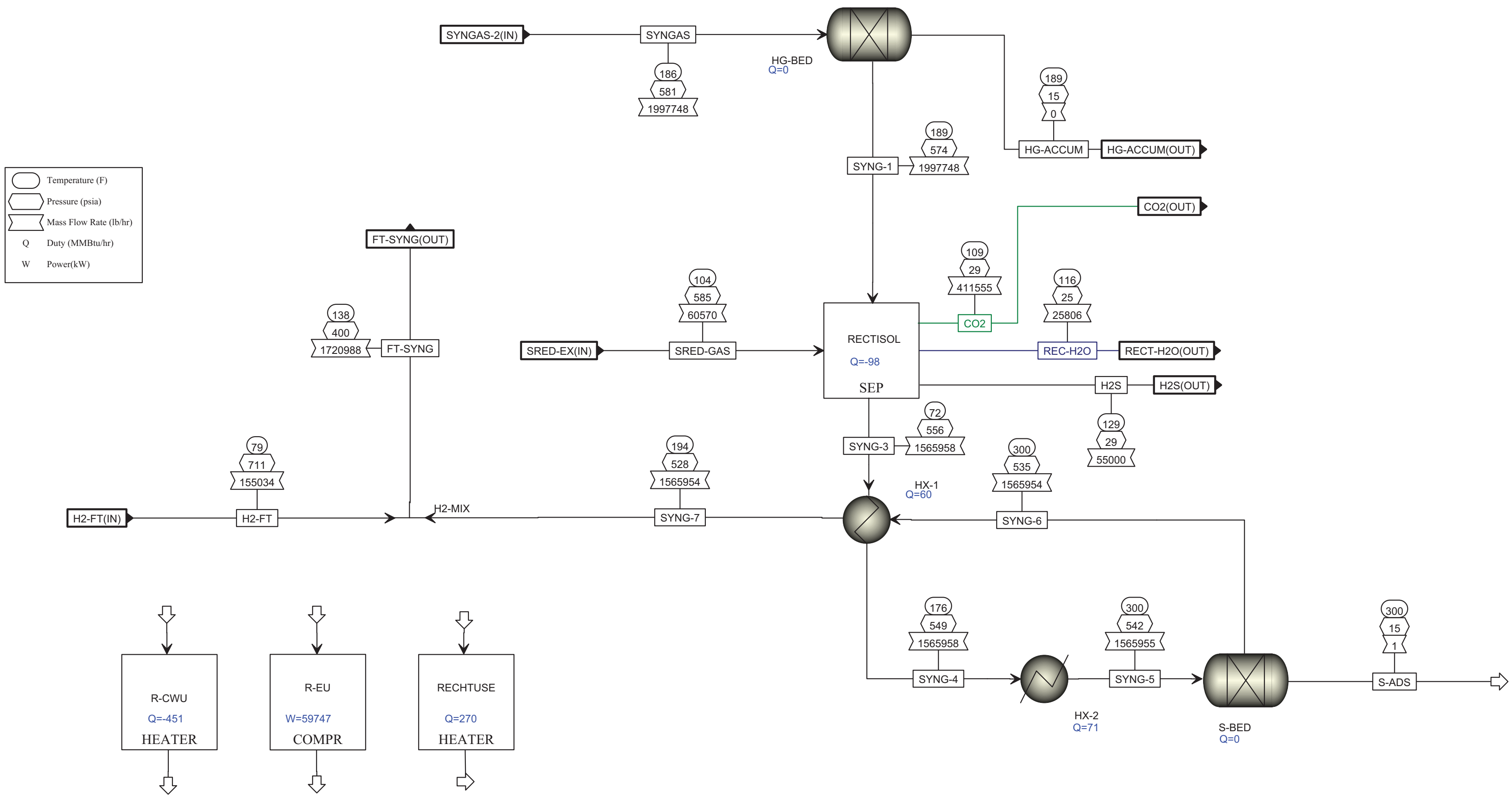
Coal Milling & Drying



Shell Gasifier w/ Heat Recovery

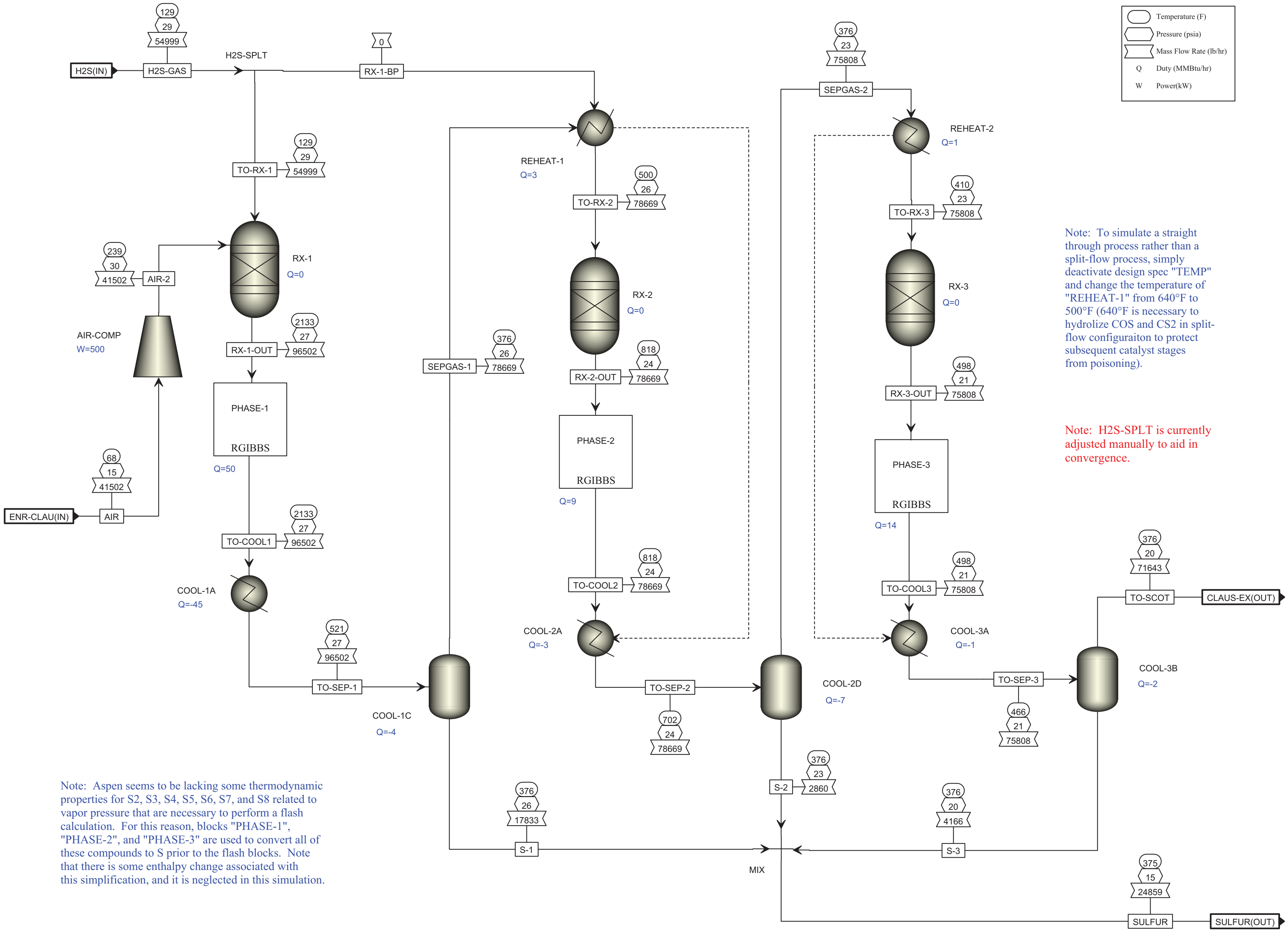


Syngas Cleaning & Conditioning



Used to include Rectisol coolng water, power, and steam requirements.

Claus Process

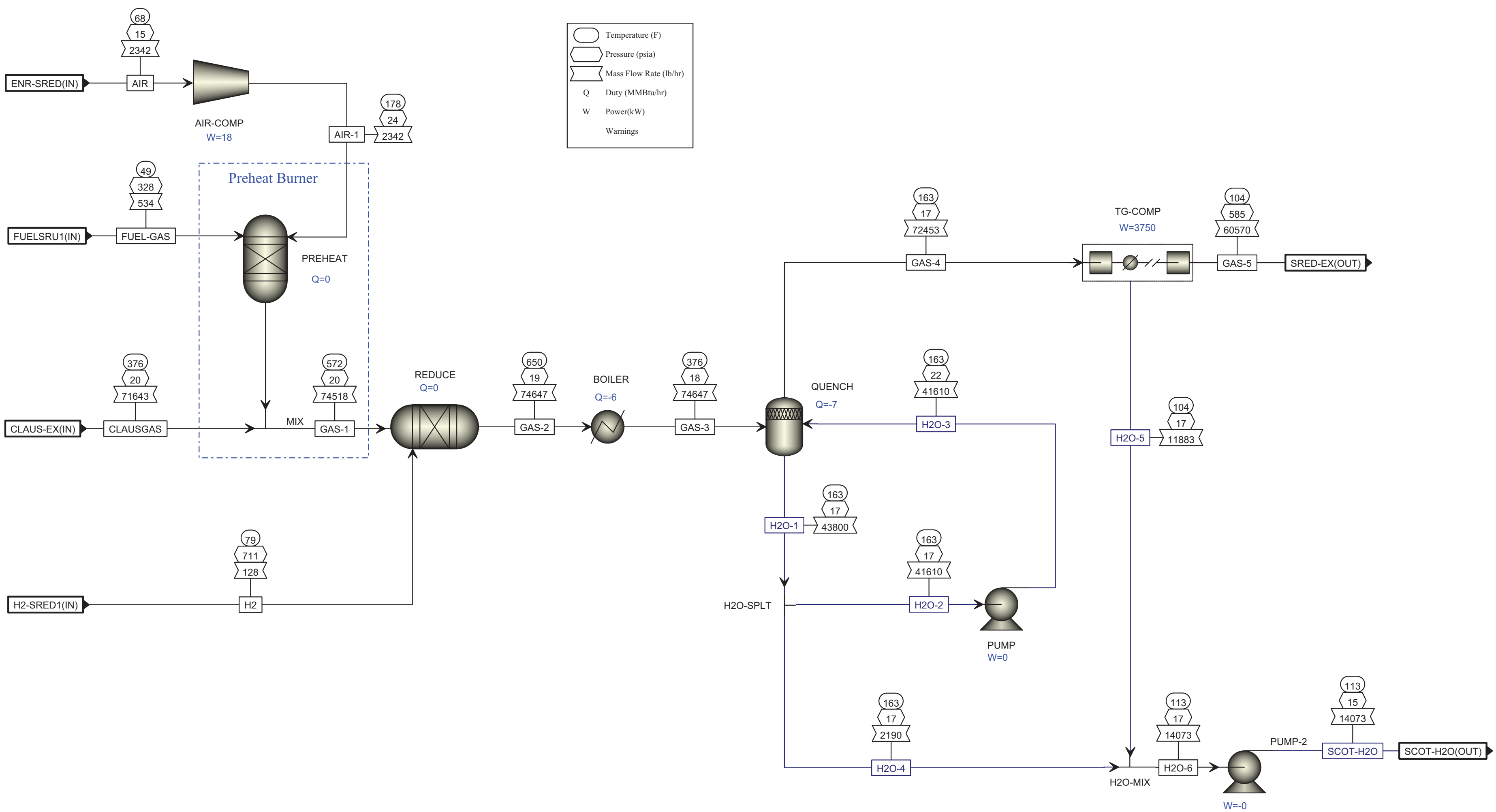


Note: To simulate a straight through process rather than a split-flow process, simply deactivate design spec "TEMP" and change the temperature of "REHEAT-1" from 640°F to 500°F (640°F is necessary to hydrolize COS and CS2 in split-flow configuraiton to protect subsequent catalyst stages from poisoning).

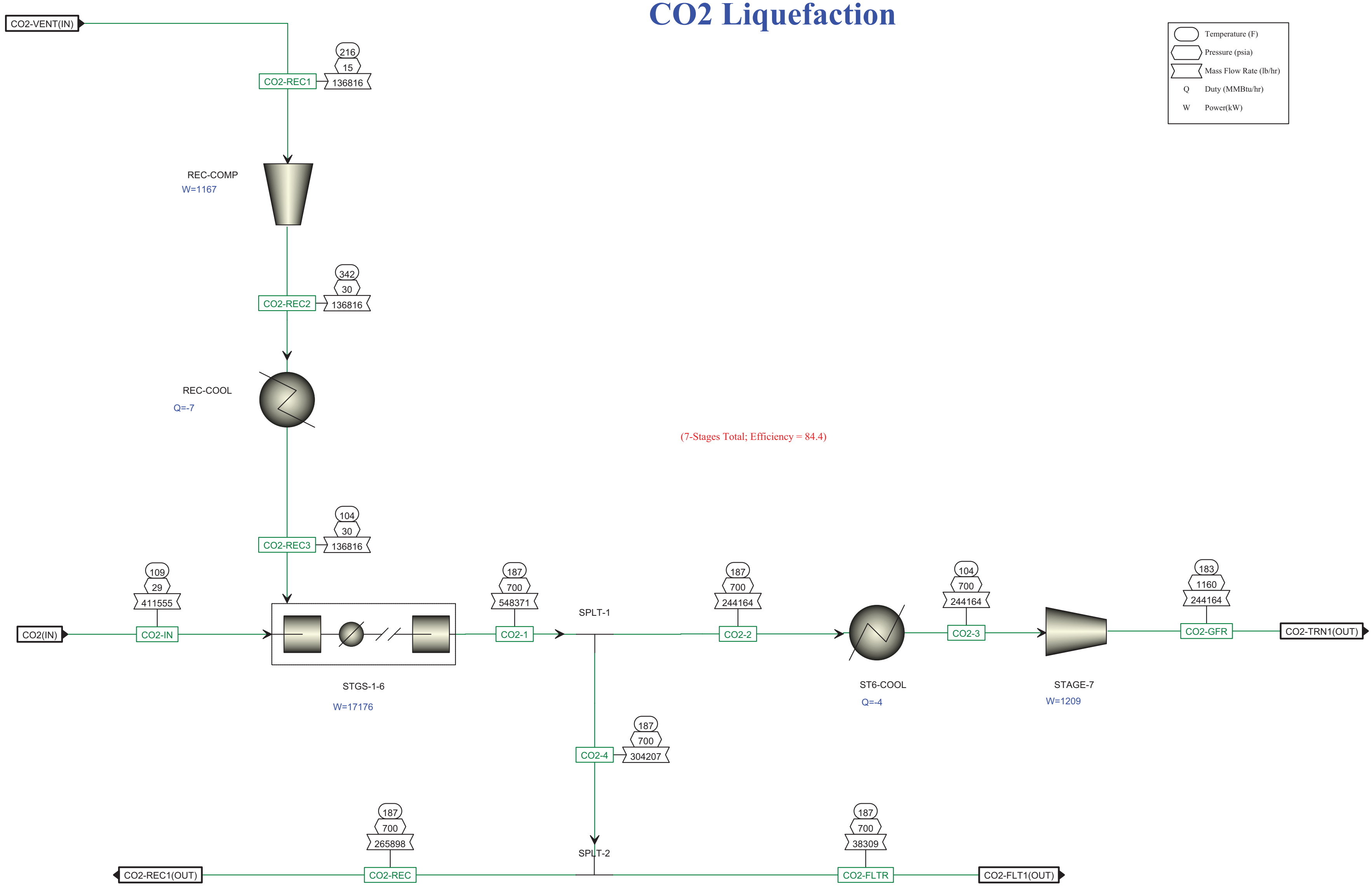
Note: H2S-SPLT is currently adjusted manually to aid in convergence.

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.

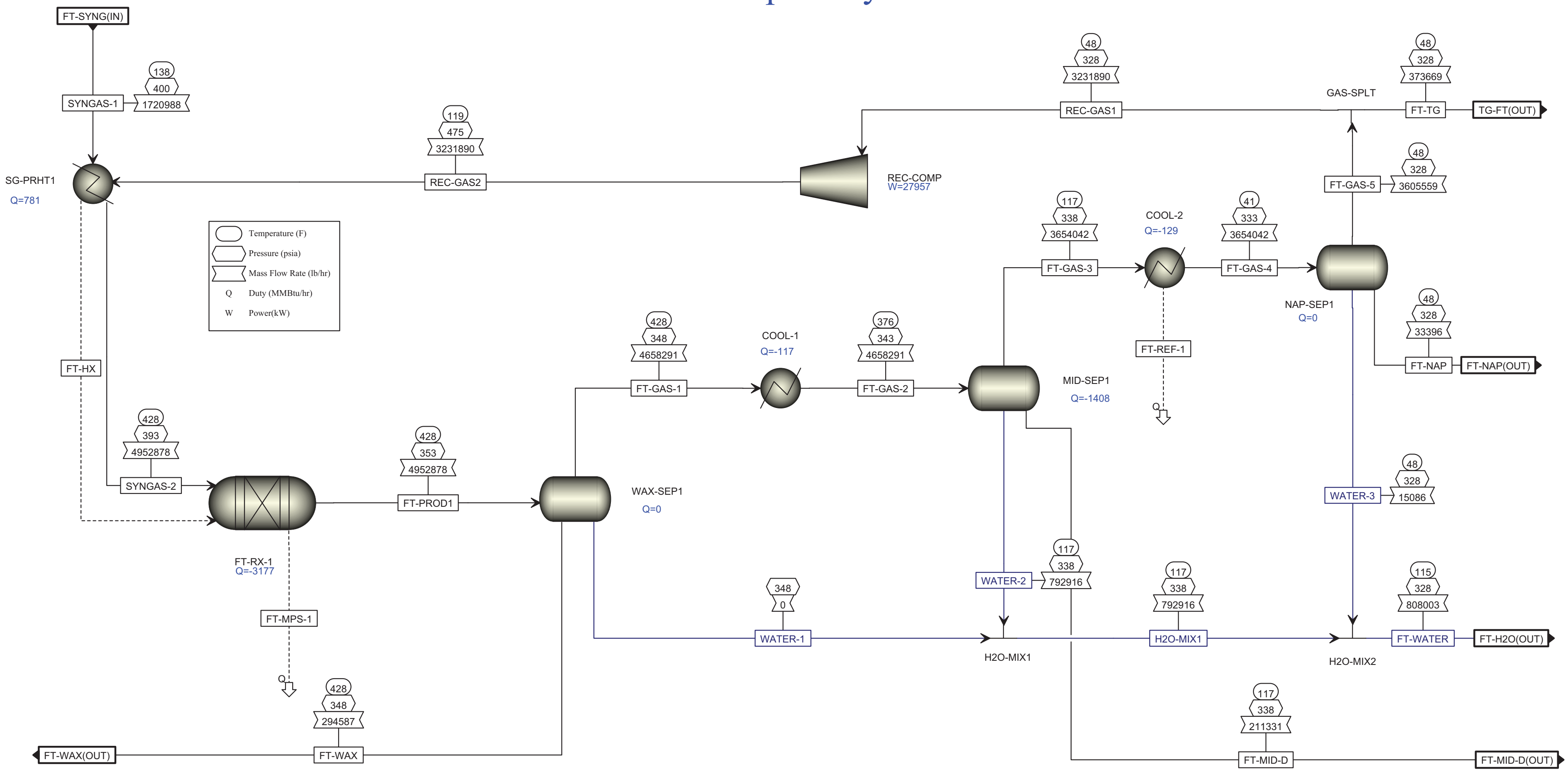
Catalytic Sulfur Reduction (SCOT or Beavon Process w/o H2S Absorber)



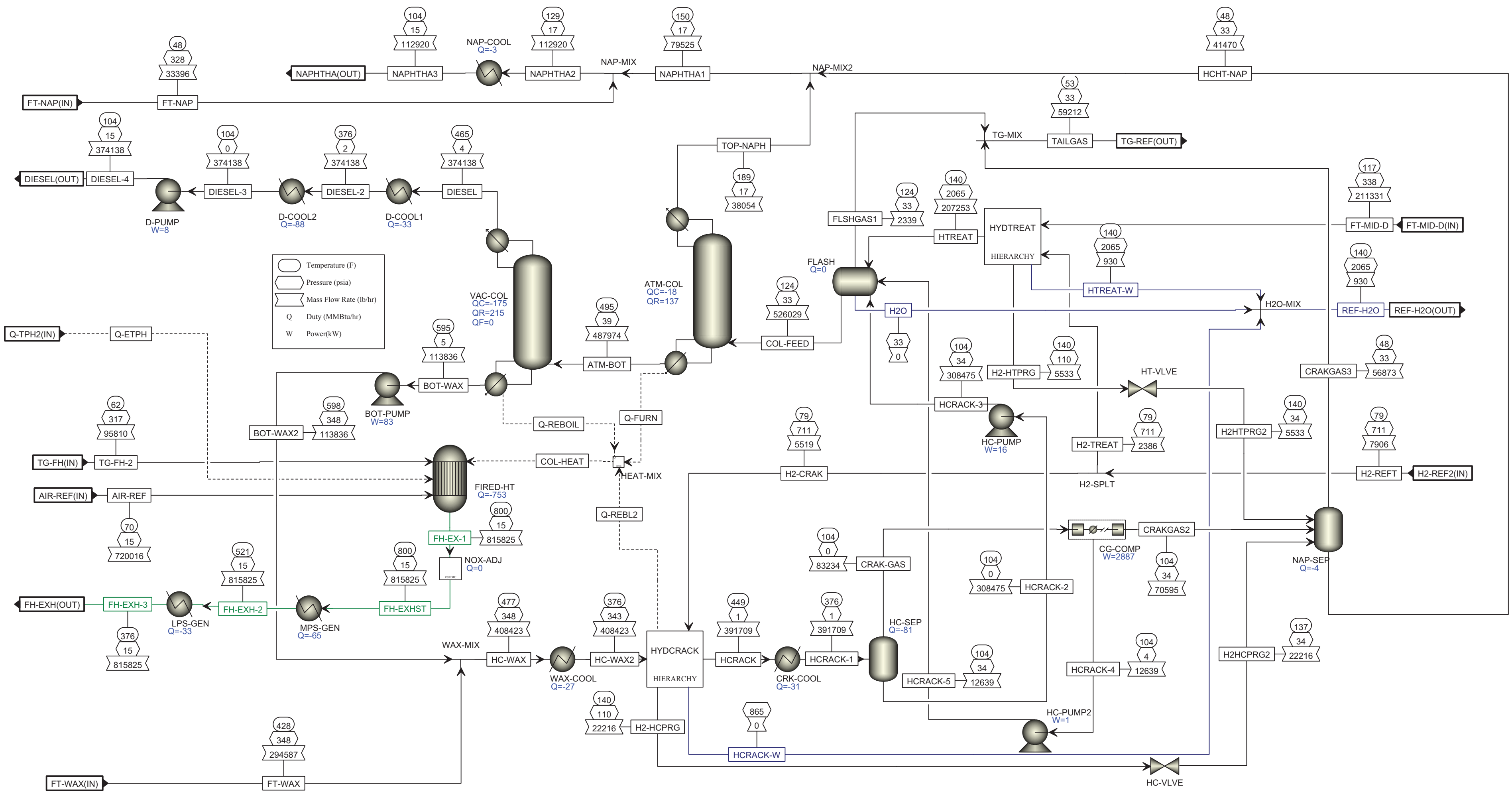
CO2 Liquefaction



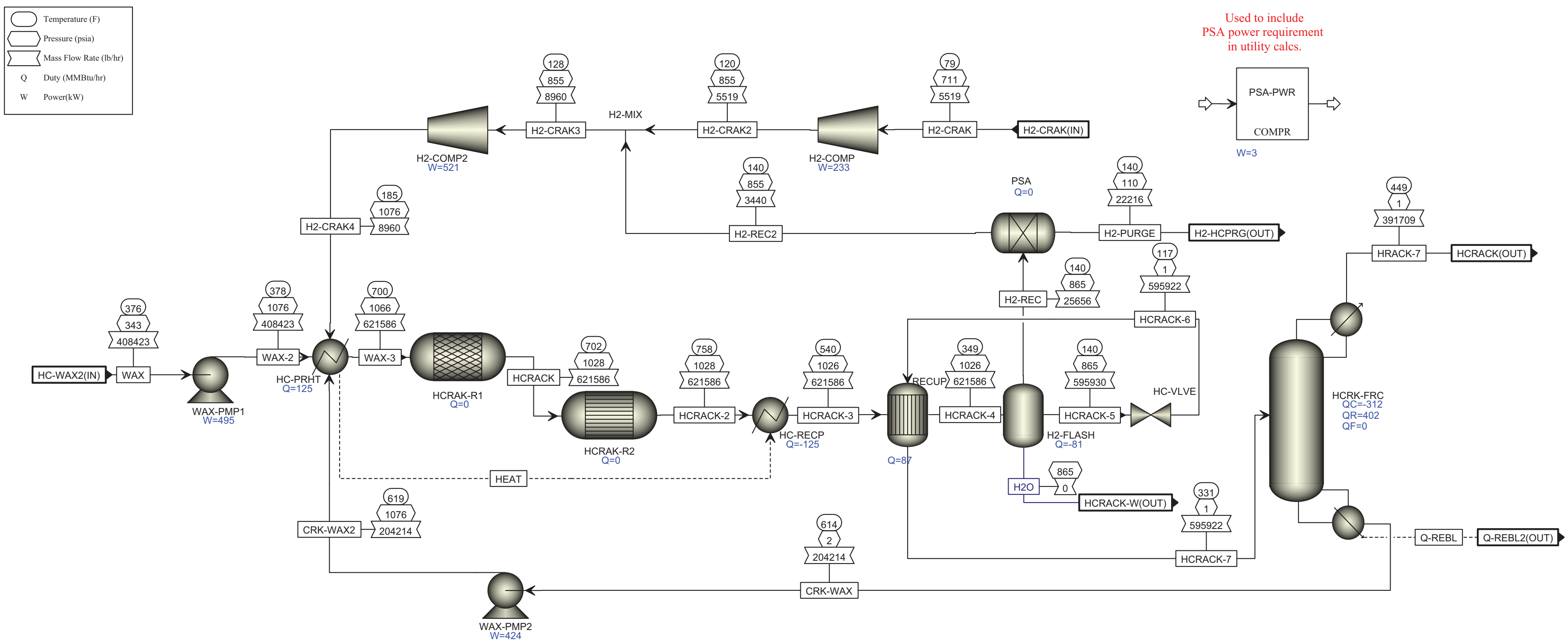
Fischer Tropsch Synthesis



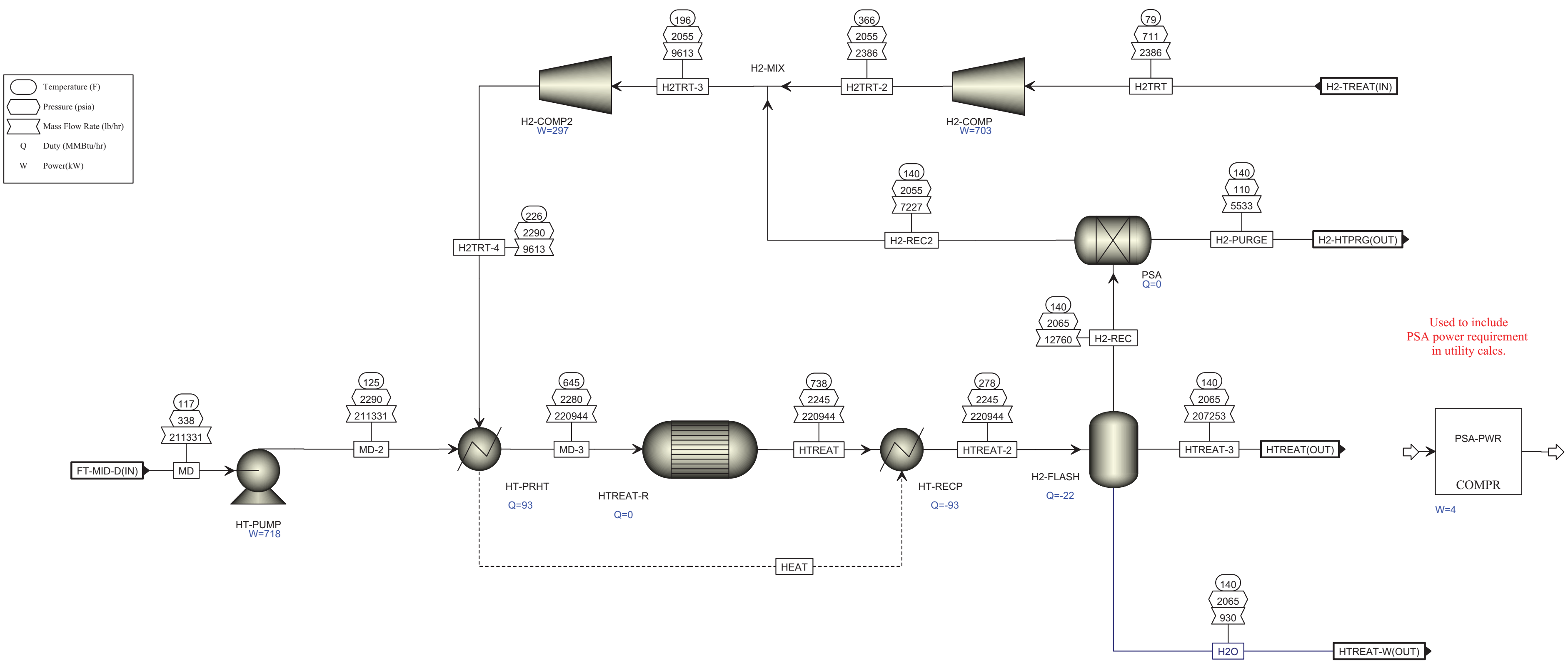
Product Upgrading and Refining



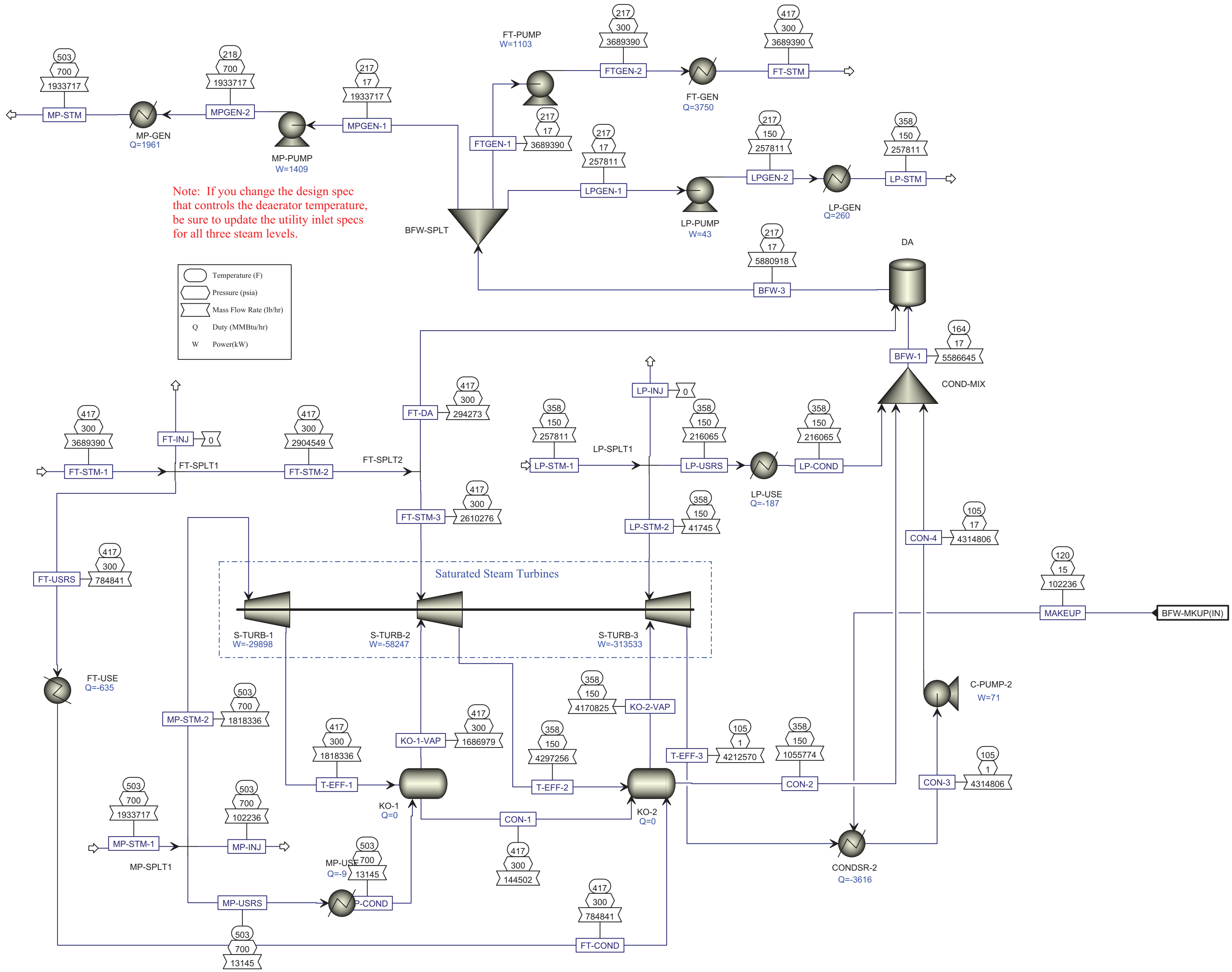
Hydrocracking



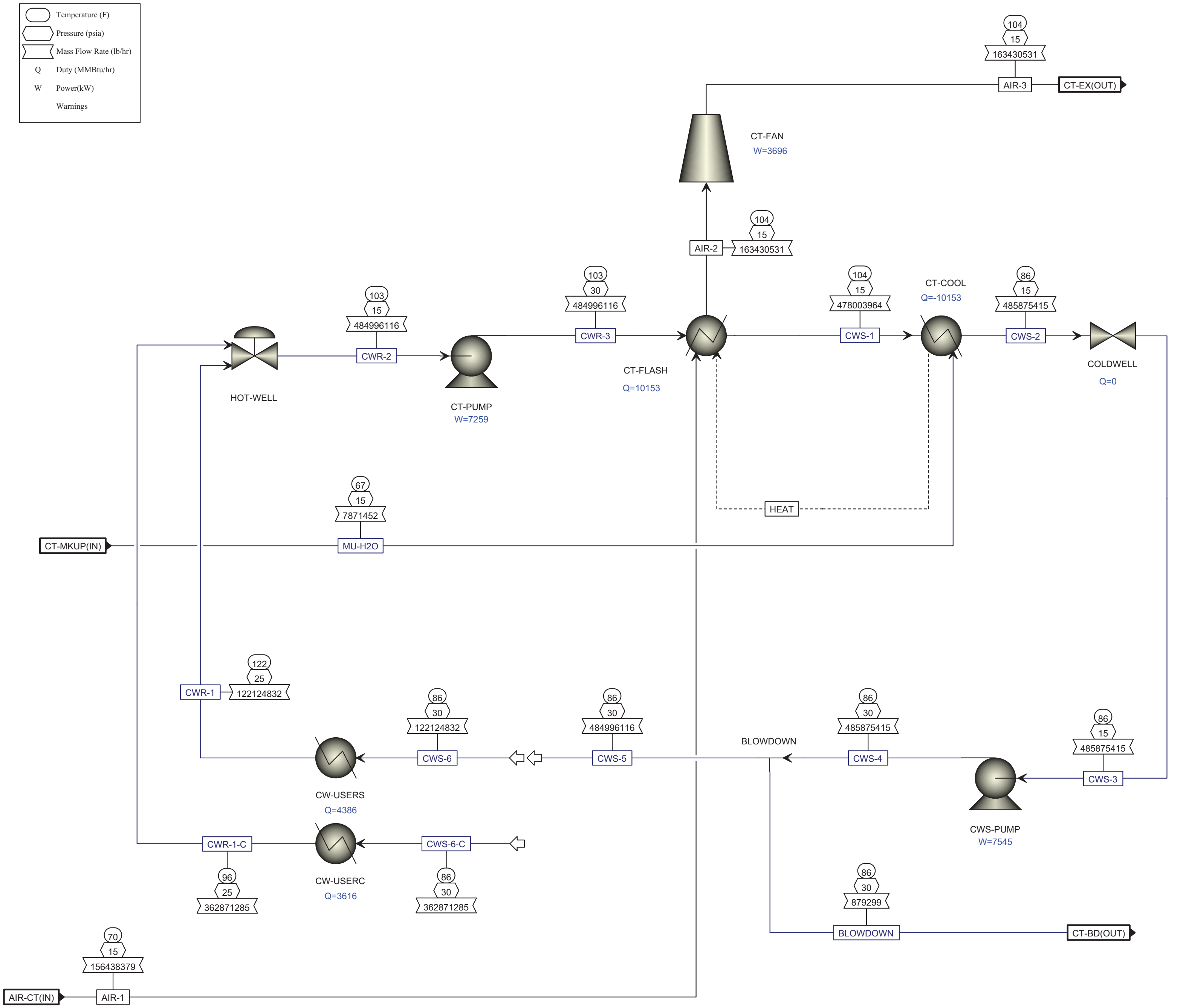
Hydrotreating



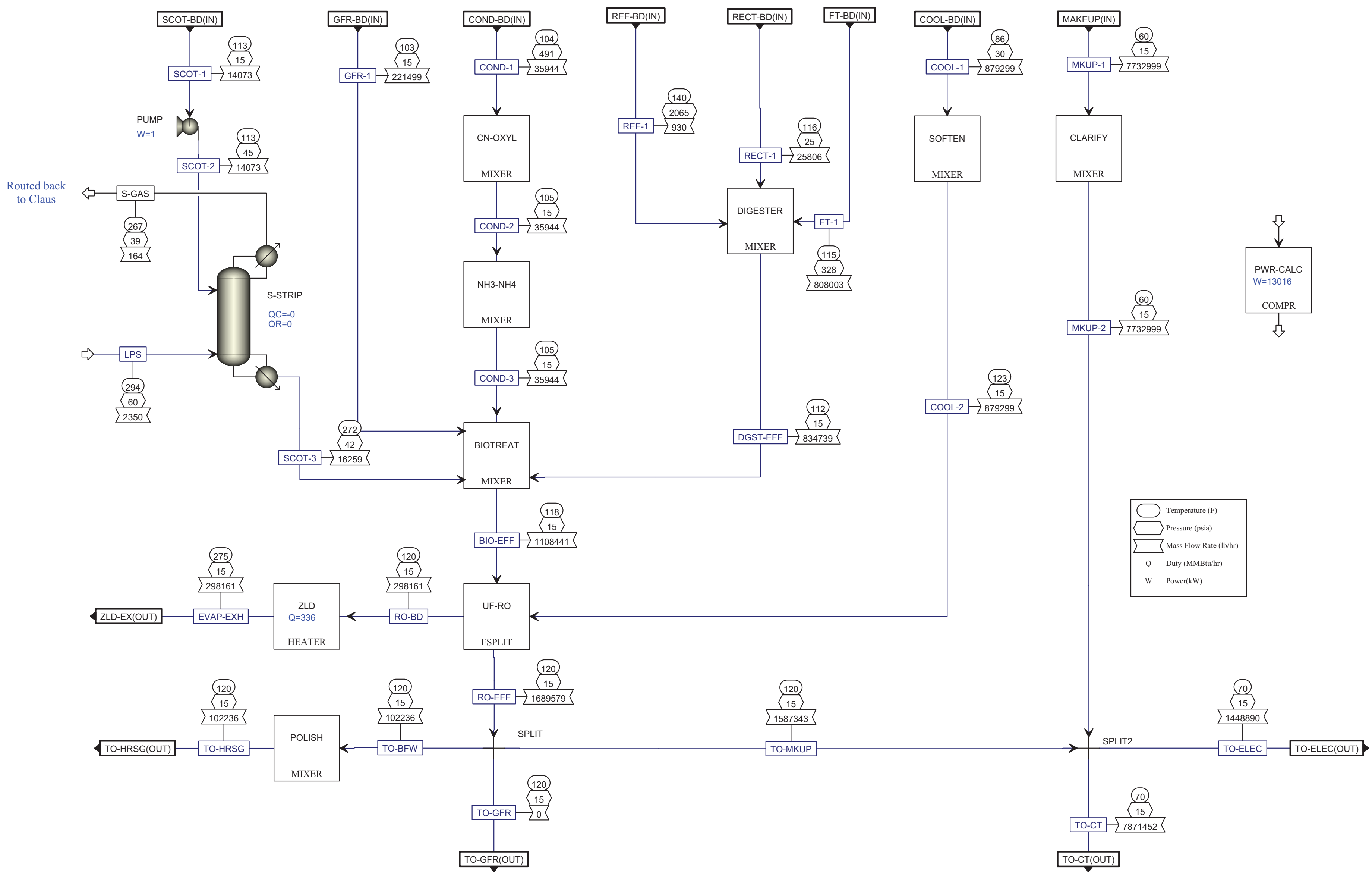
Steam Turbines



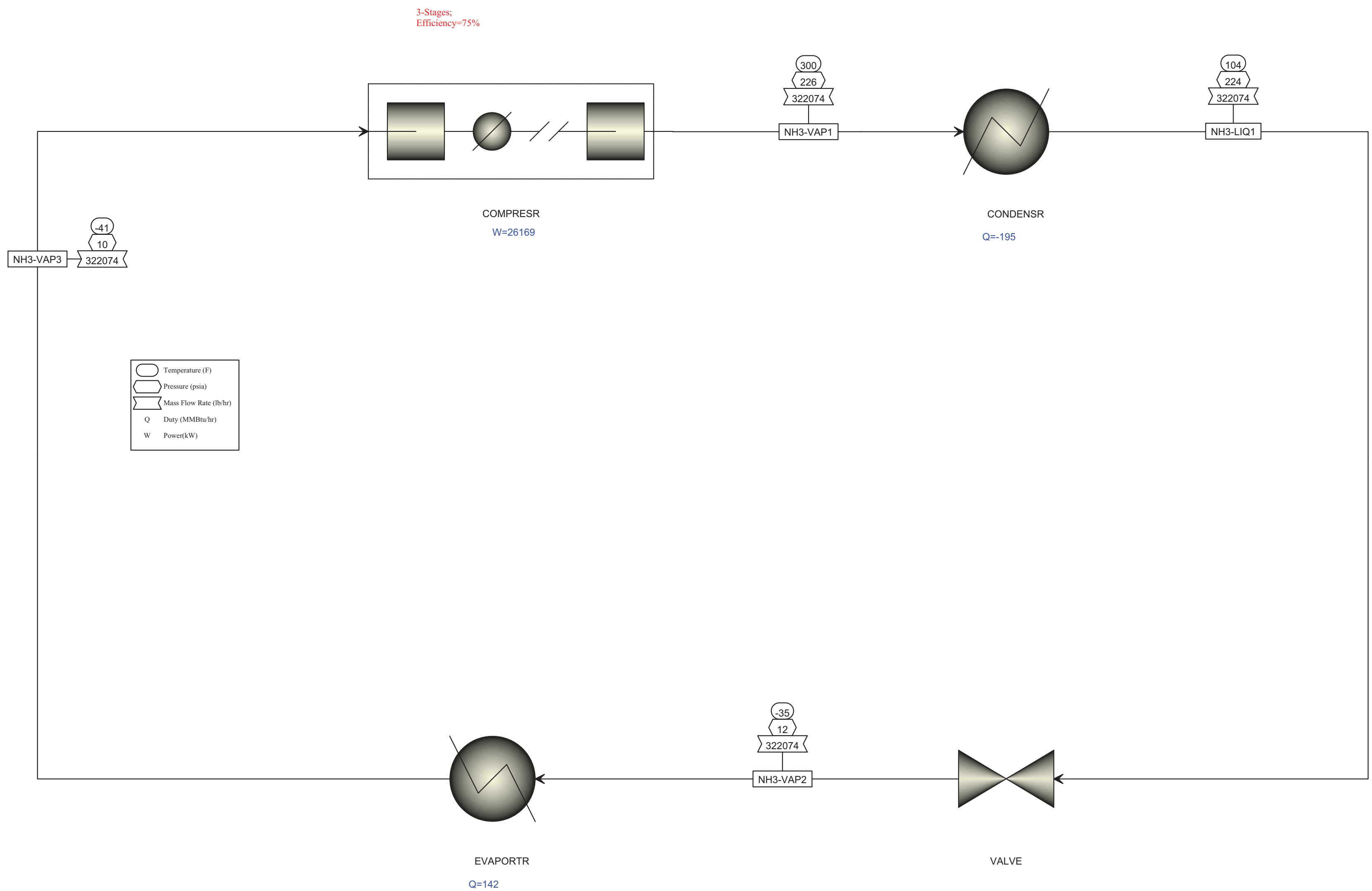
Cooling Tower



Simplified Water Treatment



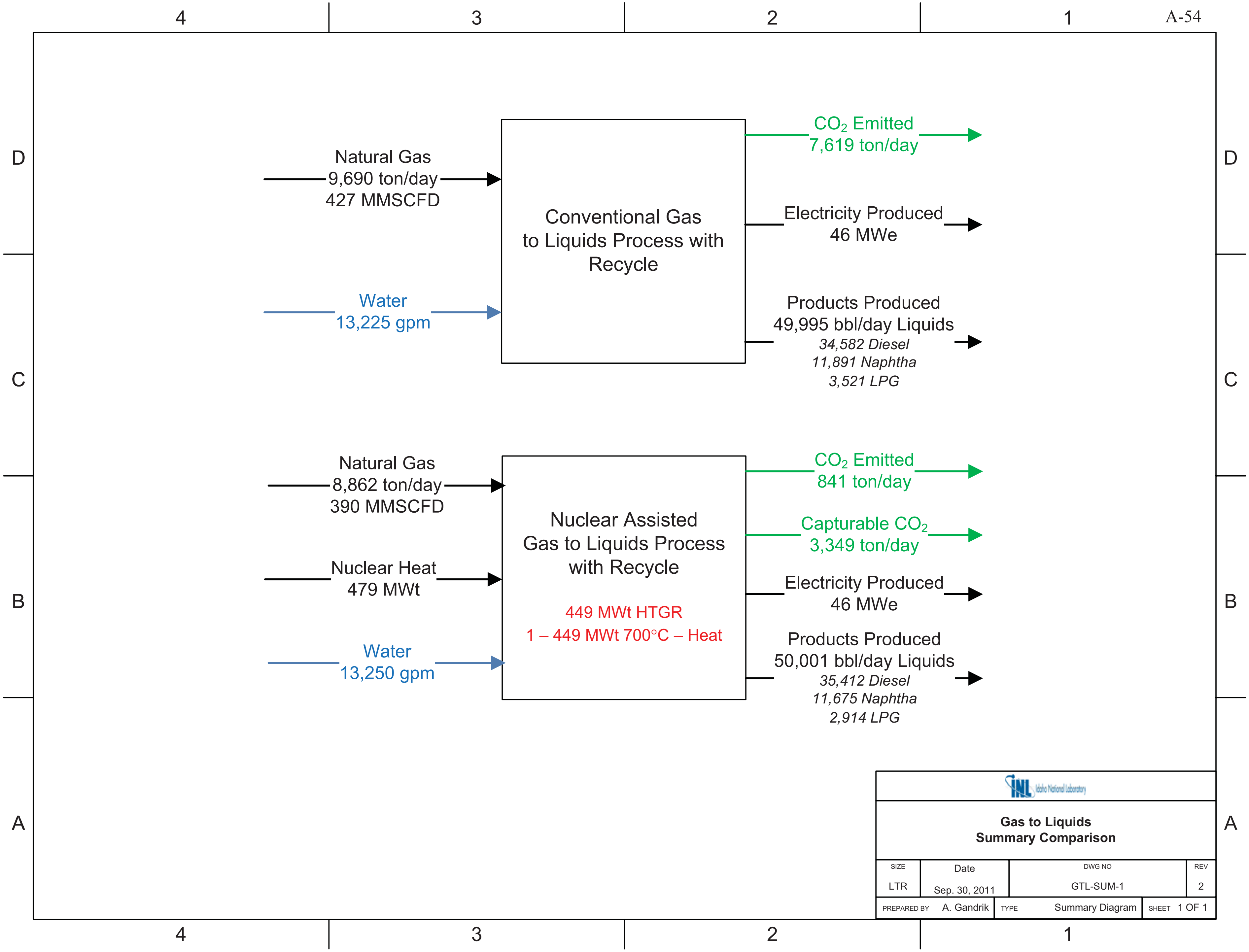
Refrigeration Unit

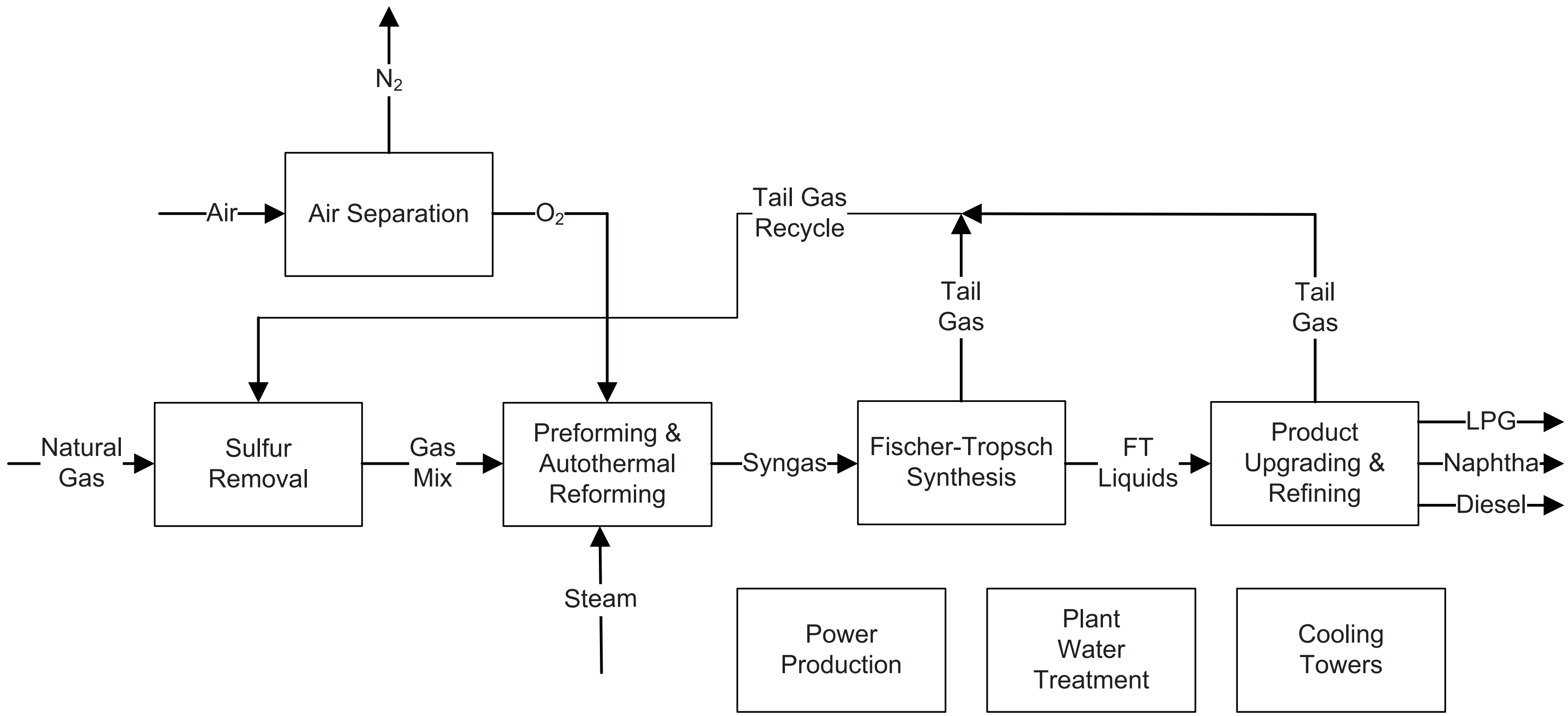


	Conventional GTL	Nuclear Integration GTL
Inputs		
Natural Gas Feed Rate (MMSCFD)[1]	427	390
% Carbon to Liquid Product	71.9%	79.3%
# HTGRs (600 MW_t)	N/A	0.75
Outputs		
Total Liquid Products (bbl/day)t	*49,994*	*49,998*
Diesel	34,581	35,410
Naphtha	11,892	11,674
LPG	3,521	2,914
Utility Summary		
Total Power (MW)	*66.6*	*69.7*
Power Consumed	-330.1	-402.3
Secondary Helium Circulator	N/A	-48.4
ASU	-132.7	-131.3
Natural Gas Reforming	-68.0	-68.9
CO_2 Compression/Liquefaction	N/A	-11.7
Fischer Tropsch & Refining Processes	-53.8	-60.3
Refrigeration	-41.5	-47.1
Cooling Tower	-18.8	-20.8
Water Treatment	-15.4	-13.9
Power Generated	396.7	471.9
Saturated Turbines	396.7	471.9
Water Requirements[2]		
Water Consumed (gpm)	13,790	14,552
Water Consumed/lb Feed (lb/lb)	8.55	9.86
Water Consumed/bbl Product (bbl/bbl)	9.5	10.0
CO_2 Summary		
Total CO_2 Produced (ton/day)	*7,164*	*4,190*
Emitted	7,164	841
Capturable	N/A	3,349
Nuclear Integration Summary		
Electricity (MW)	*N/A*	*-13.9*
HTGR House Loads	N/A	-13.9
Balance of Fossil Plant	N/A	N/A
HTGR Heat Use (MMBTU/hr)	*N/A*	*1,633*
Reformer	N/A	1,057
Refinery	N/A	741
From Secondary Circulator	N/A	-165

[1]Standard temperature of 60 degrees F.

[2]Does not include water usage for HTGR.


Natural Gas
9,690 ton/day
427 MMSCFD
Water
13,225 gpm
Conventional Gas to Liquids Process with Recycle
CO_2 Emitted
7,619 ton/day
Electricity Produced
46 MWe
Products Produced
49,995 bbl/day Liquids
34,582 Diesel
11,891 Naphtha
3,521 LPG
Natural Gas
8,862 ton/day
390 MMSCFD
Nuclear Heat
479 MWt
Water
13,250 gpm
Nuclear Assisted Gas to Liquids Process with Recycle
449 MWt HTGR
1 – 449 MWt 700°C – Heat
CO_2 Emitted
841 ton/day
Capturable CO_2
3,349 ton/day
Electricity Produced
46 MWe
Products Produced
50,001 bbl/day Liquids
35,412 Diesel
11,675 Naphtha
2,914 LPG
Gas to Liquids
Summary Comparison
SIZE
LTR
Date
Sep. 30, 2011
DWG NO
GTL-SUM-1
REV
2
PREPARED BY
A. Gandrik
TYPE
Summary Diagram
SHEET
1 OF 1

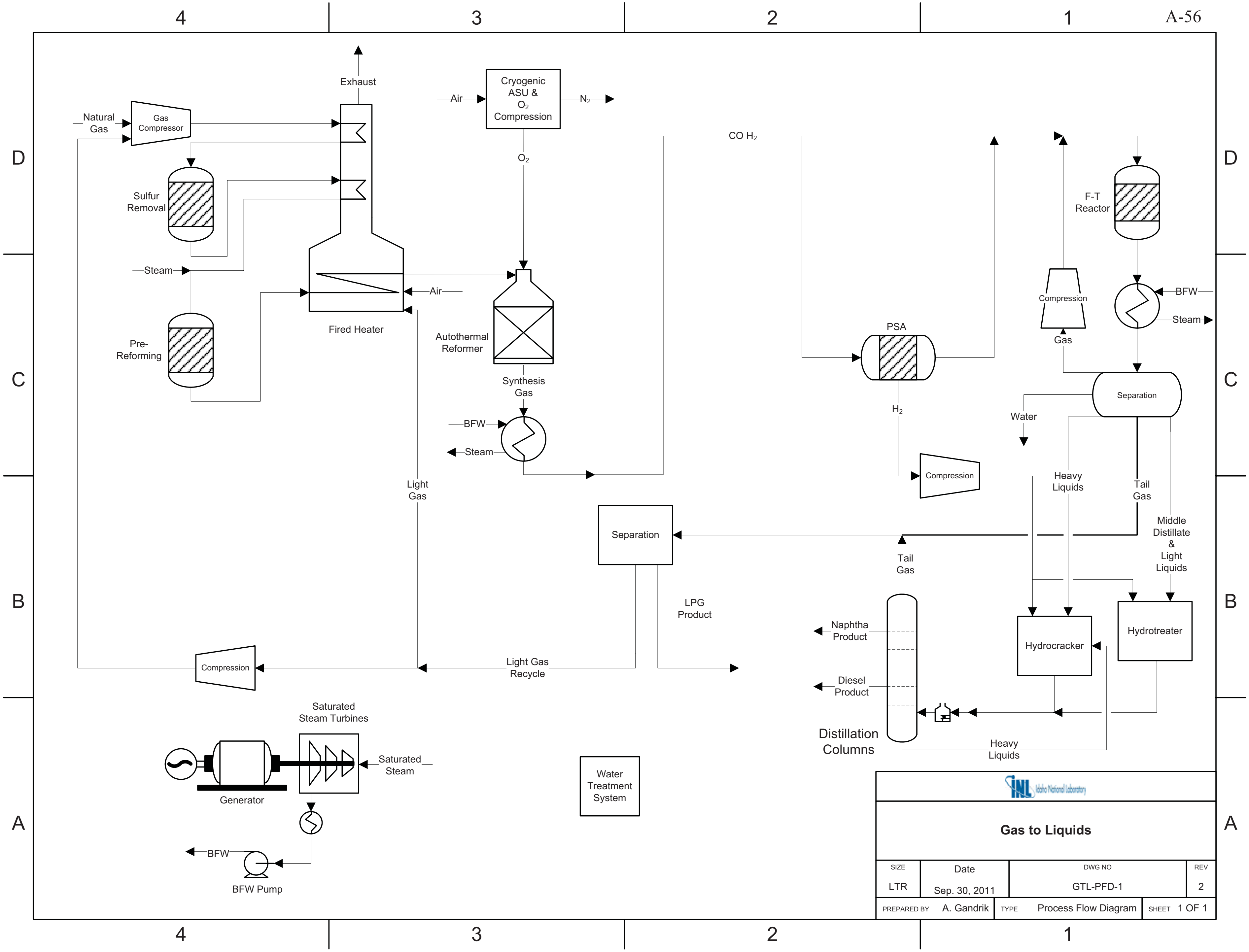


Idaho National Laboratory

Gas to Liquids

SIZE	Date	DWG NO	REV
LTR	Sep. 30, 2011	GTL-BFD-1	2

PREPARED BY	A. Gandrik	TYPE	Block Flow Diagram	SHEET	1 OF 1

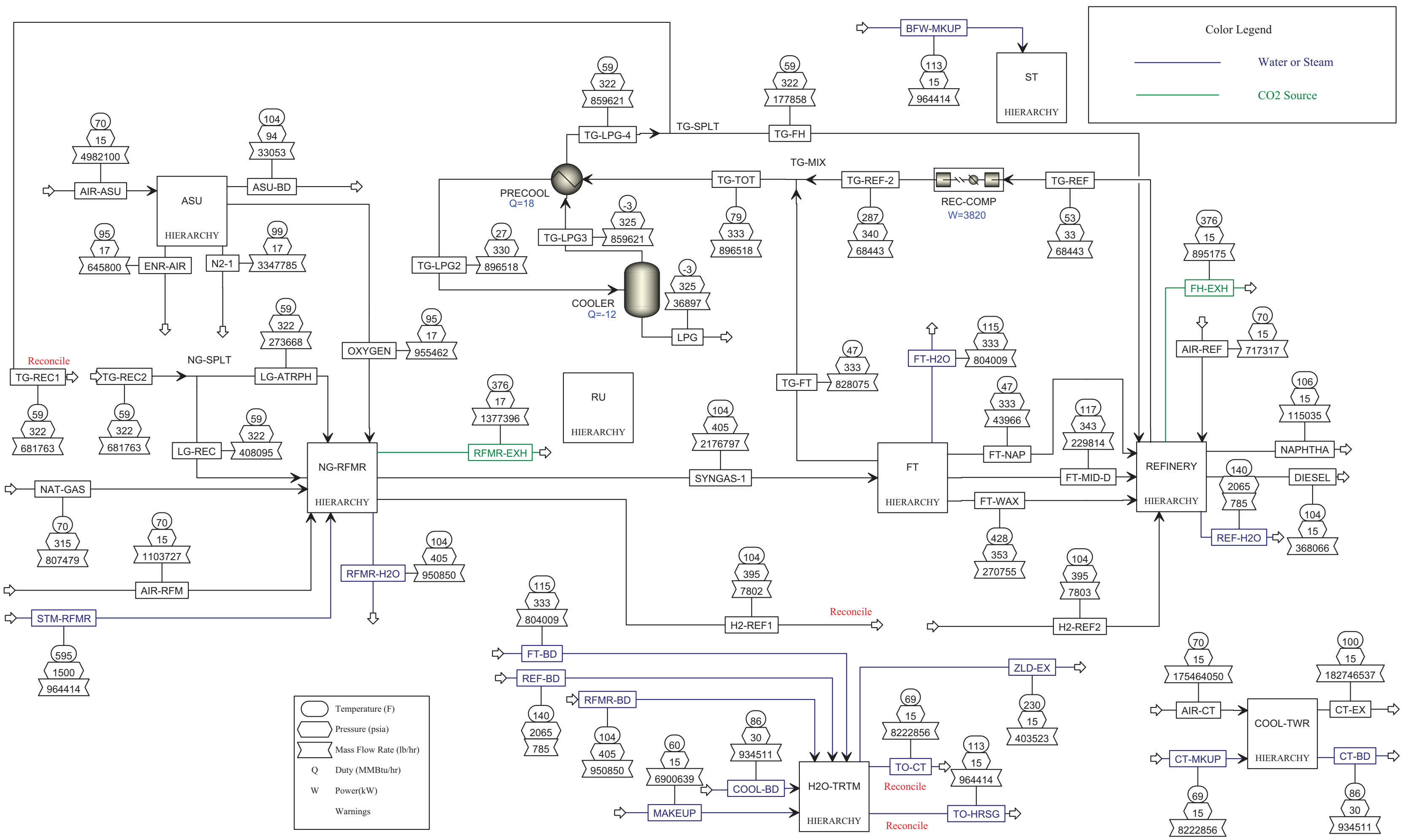

Natural Gas
Gas Compressor
Exhaust
Sulfur Removal
Steam
Pre-Reforming
Fired Heater
Air
Cryogenic ASU & O_2 Compression
N_2
O_2
Autothermal Reformer
Synthesis Gas
BFW
Steam
Light Gas
CO H_2
PSA
H_2
Compression
F-T Reactor
Gas
Separation
Water
Heavy Liquids
Tail Gas
Middle Distillate & Light Liquids
Hydrocracker
Hydrotreater
Naphtha Product
Diesel Product
Distillation Columns
LPG Product
Light Gas Recycle
Saturated Steam Turbines
Saturated Steam
Generator
BFW Pump
Water Treatment System
INL Idaho National Laboratory
Gas to Liquids
SIZE LTR
Date Sep. 30, 2011
DWG NO GTL-PFD-1
REV 2
PREPARED BY A. Gandrik
TYPE Process Flow Diagram
SHEET 1 OF 1

```
Calculator Block SUMMARY

   FEED SUMMARY:

     NATURAL GAS PROPERTIES:

       MASS FLOW =                             9690. TON/DY
       VOLUME FLOW =                            427. MMSCFD @ 60°F
       HHV =                                  23063. BTU/LB
       HHV =                                   1047. BTU/SCF @ 60°F
       ENERGY FLOW =                         446948. 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

     PRODUCTS:

       LIQUID PRODUCTS PRODUCED =            519998. LB/HR
       LIQUID PRODUCTS PRODUCED =              6240.0 TON/DY
         DIESEL =                            368066. LB/HR
         DIESEL =                              4417. TON/DY
         NAPHTHA =                           115035. LB/HR
         NAPHTHA =                             1380. TON/DY
         LPG =                                36897. LB/HR
         LPG =                                  443. TON/DY
       LIQUID PRODUCTS PRODUCED =             49994. BBL/DY
         DIESEL =                             34581. BBL/DY
         NAPHTHA =                            11892. BBL/DY
         LPG =                                 3521. BBL/DY
       LIQUIDS PRODUCED / NATURAL GAS FED =    0.64 LB/LB
       LIQUIDS PRODUCED / NATURAL GAS FED = 117.14 BBL/MMSCF

     FUEL PROPERTIES:
                                   DIESEL       NAPHTHA        LPG
       PROD. RATE, BBL/DAY       34581.        11892.         3521.
       LHV RATE, MMBTU/DAY      167237.        48894.        13289.
       MW                          189.9          78.6          56.2
       API GRAVITY                  54.1          94.9
       DENSITY, LB/GAL               6.08          5.53          5.99
       CETANE NO.                   95.9          35.7
       HHV CONTENT, BTU/LB       20366.        19110.        16216.
       LHV CONTENT, BTU/LB       18932.        17710.        15007.
       % CARBON                     84.7          79.0          70.1
       D86T CURVE, DEG. C:
           0%                     148.         -104.
           10%                    184.           16.
           20%                    202.           46.
           50%                    251.           83.
           90%                    327.          125.
           100%                   355.          177.

   POWER CALCULATIONS:

     POWER GENERATORS:
```

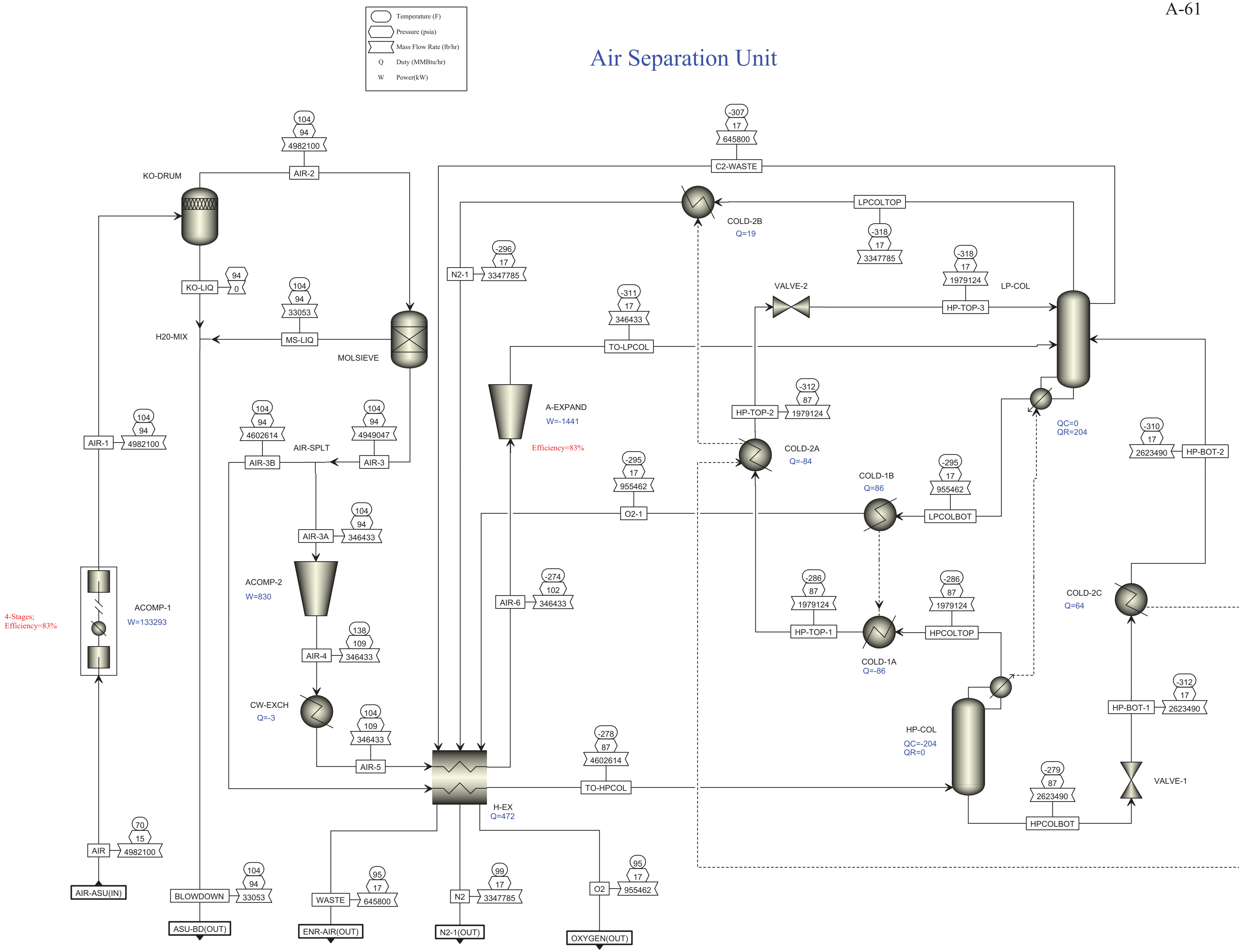
```
    SATURATED TURBINE POWER OUTPUT =      396.7 MW
  TOTAL POWER GENERATED =                 396.7 MW

  POWER CONSUMERS:
    ASU POWER CONSUMPTION =               132.7 MW
    NG REFORMER POWER CONSUMPTION =        68.0 MW
    FISHER TROPSCH POWER CONSUMPTION =     37.1 MW
    REFINERY POWER CONSUMPTION =           11.7 MW
    POWER BLOCK POWER CONSUMPTION =         5.0 MW
    REFRIGERATION POWER CONSUMPTION =      41.5 MW
    COOLING TOWER POWER CONSUMPTION =      18.8 MW
    WATER TREATMENT POWER CONSUMPTION =    15.4 MW
  TOTAL POWER CONSUMED =                  330.1 MW

  NET PLANT POWER (+ GEN, - CONS)=         66.6 MW

WATER BALANCE:

  EVAPORATIVE LOSSES:
    COOLING TOWER EVAPORATION =         16721.9 GPM
    ZLD SYSTEM EVAPORATION =              806.4 GPM
  TOTAL EVAPORATIVE LOSSES =            17528.3 GPM

  WATER CONSUMED:
    BOILER FEED WATER MAKEUP =           1927.3 GPM
    COOLING TOWER MAKEUP =              16432.6 GPM
  TOTAL WATER CONSUMED =                18359.9 GPM

  WATER GENERATED:
    NATURAL GAS REFORMING BLOWDOWN =     1900.2 GPM
    FT PROCESS BLOWDOWN =                1606.7 GPM
    REFINERY PROCESS BLOWDOWN =             1.6 GPM
    COOLING TOWER BLOWDOWN =             1867.5 GPM
  TOTAL WATER GENERATED =                5376.0 GPM

  PLANT WATER SUMMARY:
    NET MAKEUP WATER REQUIRED =         13790.2 GPM
    WATER CONSUMED / NATURAL GAS FED =     8.55 LB/LB
    WATER CONSUMED / LIQUID PRODUCT =      9.5 BBL/BBL

CARBON BALANCE SUMMARY:

  % CARBON TO LIQUID FUEL =                71.9 %
  % CARBON TO TAILGAS =                    27.3 %
  % UNACCOUNTED CARBON =                    0.7 %

  CO2 EMITTED =                          7164. TON/DY
  CO2 EMITTED =                           125. MMSCFD
    FROM REFINERY =                      2822. TON/DY
      LHV TO REFINERY =                 21740. MMBTU/DY
    FROM REFORMER =                      4342. TON/DY
      LHV TO REFORMER =                 33452. MMBTU/DY
  CO2 EMMITED / LIQ PROD  =                1.15 LB/LB
  CO2 EMMITED / NATURAL GAS FED =          0.01 LB/LB

STARTUP FLARE SUMMARY:

  CO2 FROM FLARE =                        134. TON/DY
  LHV TO FLARE =                         2073. MMBTU/DY

EFFICIENCY CALCULATIONS:

    HEAT IN (HHV BASED):
        NATURAL GAS HEAT CONTENT =      18622.9 MMBTU/HR
```

```
    HEAT OUT (HHV BASED):
        NET POWER =                          227.3 MMBTU/HR
        LIQUID HEAT CONTENT =              10292.7 MMBTU/HR

    PLANT EFFICIENCY (HHV BASED):
        EFFICIENCY =                          56.5 %


Calculator Block NG-RFMR  Hierarchy: NG-RFMR

  SULFUR REMOVAL CONDITIONS:

    INLET BED TEMPERATURE =                757. °F

  PREFORMER CONDITIONS:

    INLET TEMPERATURE =                    915. °F
    STEAM TO CARBON MOLAR RATIO =            1.00

  AUTOTHERMAL REFORMER CONDITIONS:

    INLET TEMPERATURE =                   1092. °F
    STEAM TO CARBON MOLAR RATIO =            0.94
    OXYGEN TO CARBON MOLAR RATIO =           0.57
    OUTLET TEMPERATURE =                  1870. °F
    H2/CO PRE PSA =                          2.219
    (H2 - CO2)/(CO + CO2) =                  1.520
    H2/CO POST PSA=                          2.138

    OUTLET COMPOSITION (PRE-CONDENSER):
      H2                                    47.3123 MOL.%
      CO                                    21.3198 MOL.%
      CO2                                    5.9098 MOL.%
      H2O                                   24.1853 MOL.%
      CH4                                    0.6451 MOL.%

    OUTLET COMPOSITION (POST-PSA):
      H2                                    61.3647 MOL.%
      CO                                    28.7014 MOL.%
      CO2                                    7.9550 MOL.%
      H2O                                    0.2753 MOL.%
      CH4                                    0.8685 MOL.%
      INERTS                                 8.7880 MOL.%
```

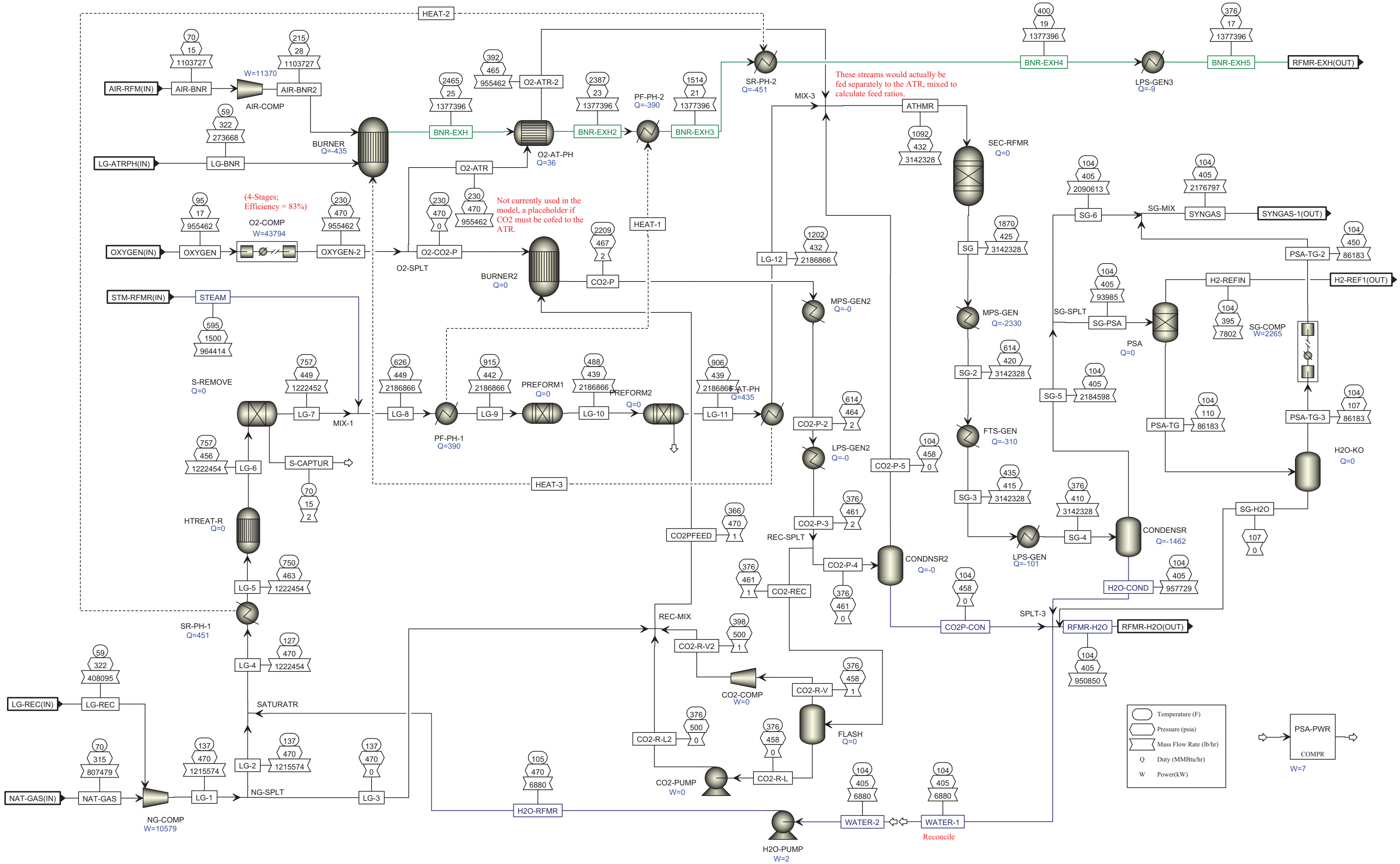
Conventional Natural Gas to Liquid Fuels



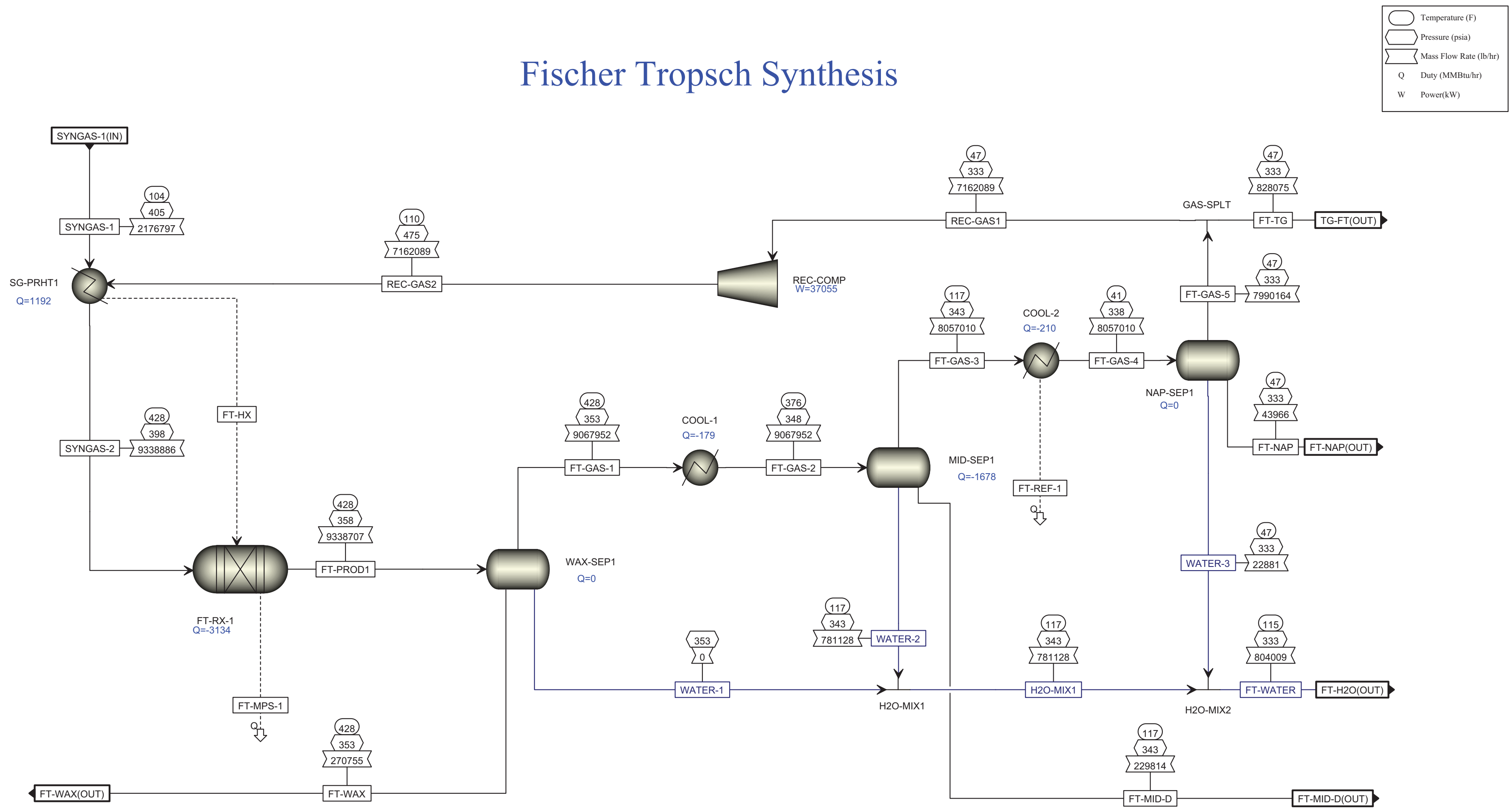
Air Separation Unit



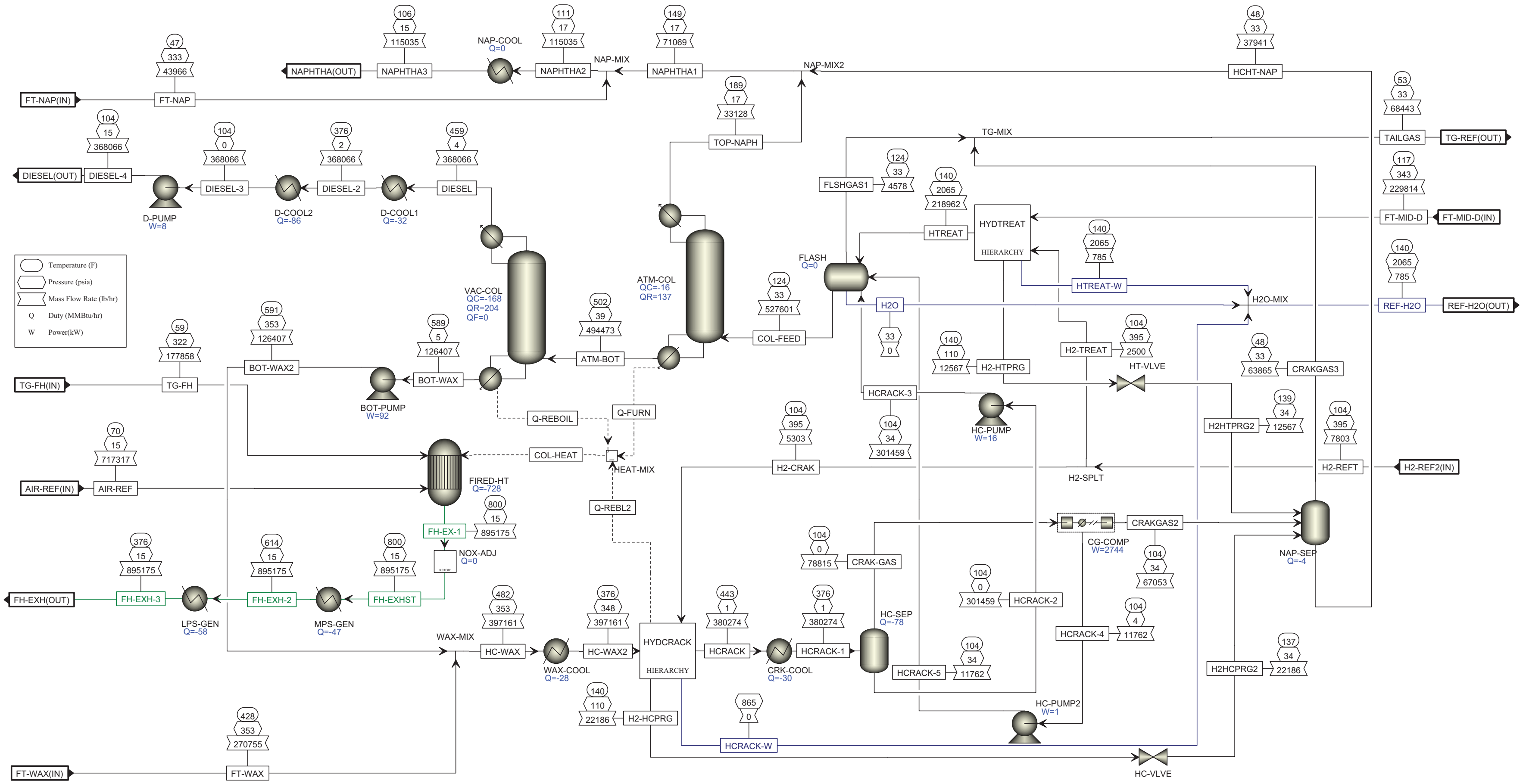
Natural Gas Autothermal Reforming



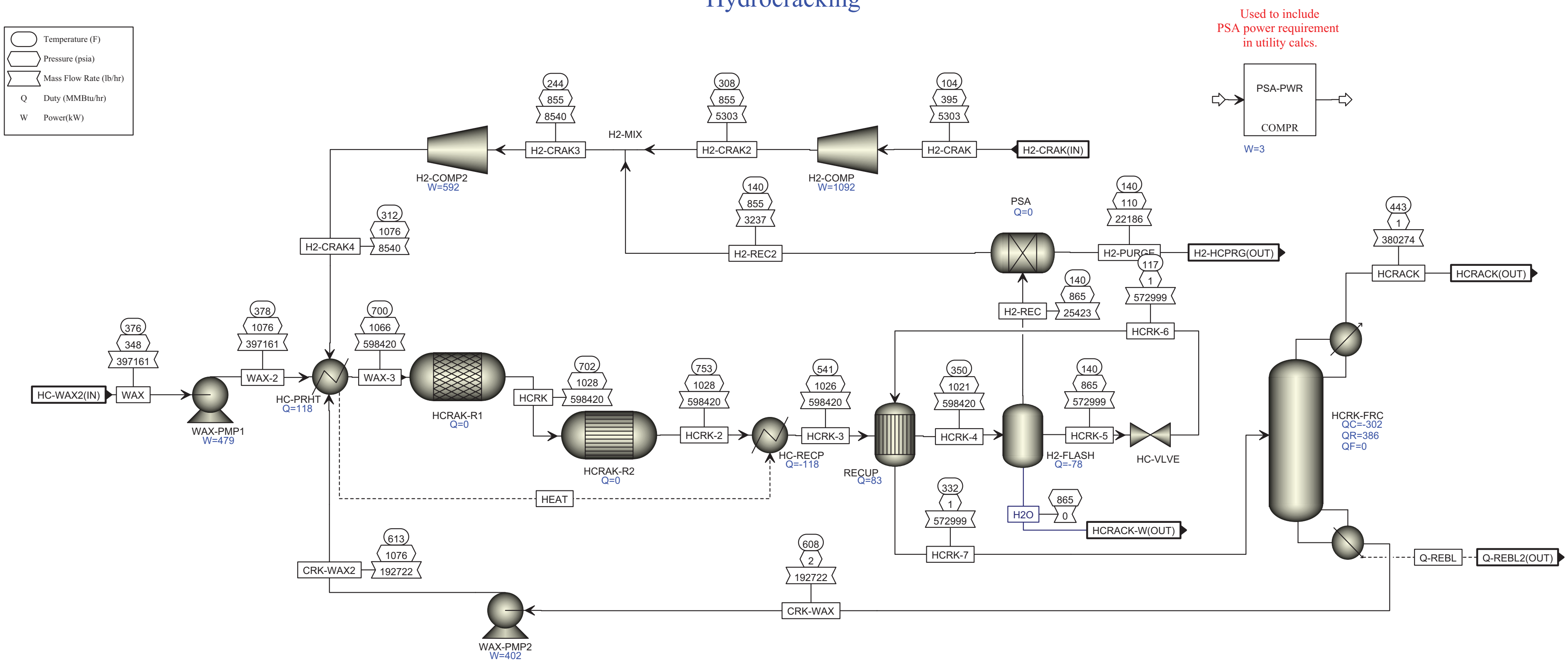
Fischer Tropsch Synthesis



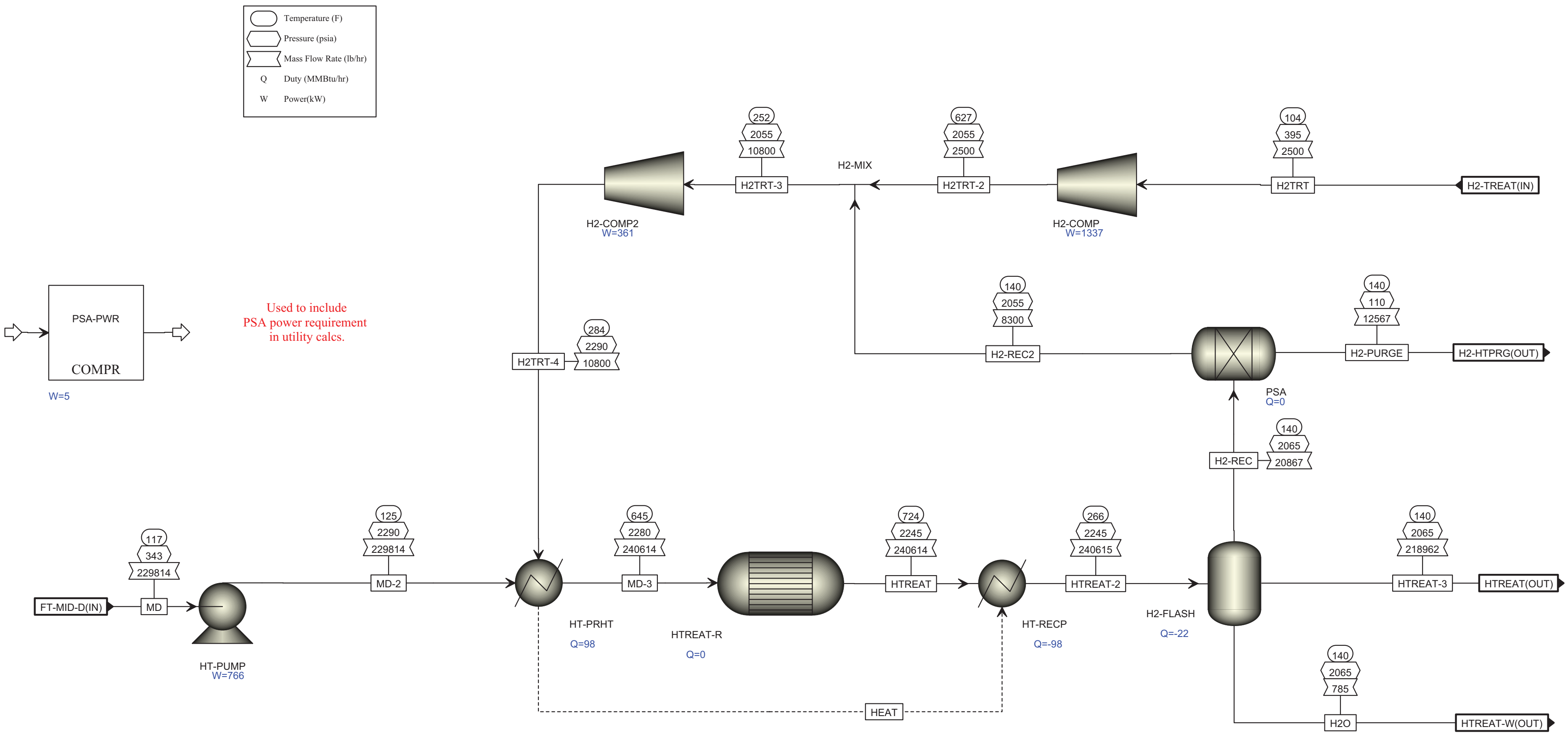
Product Upgrading and Refining



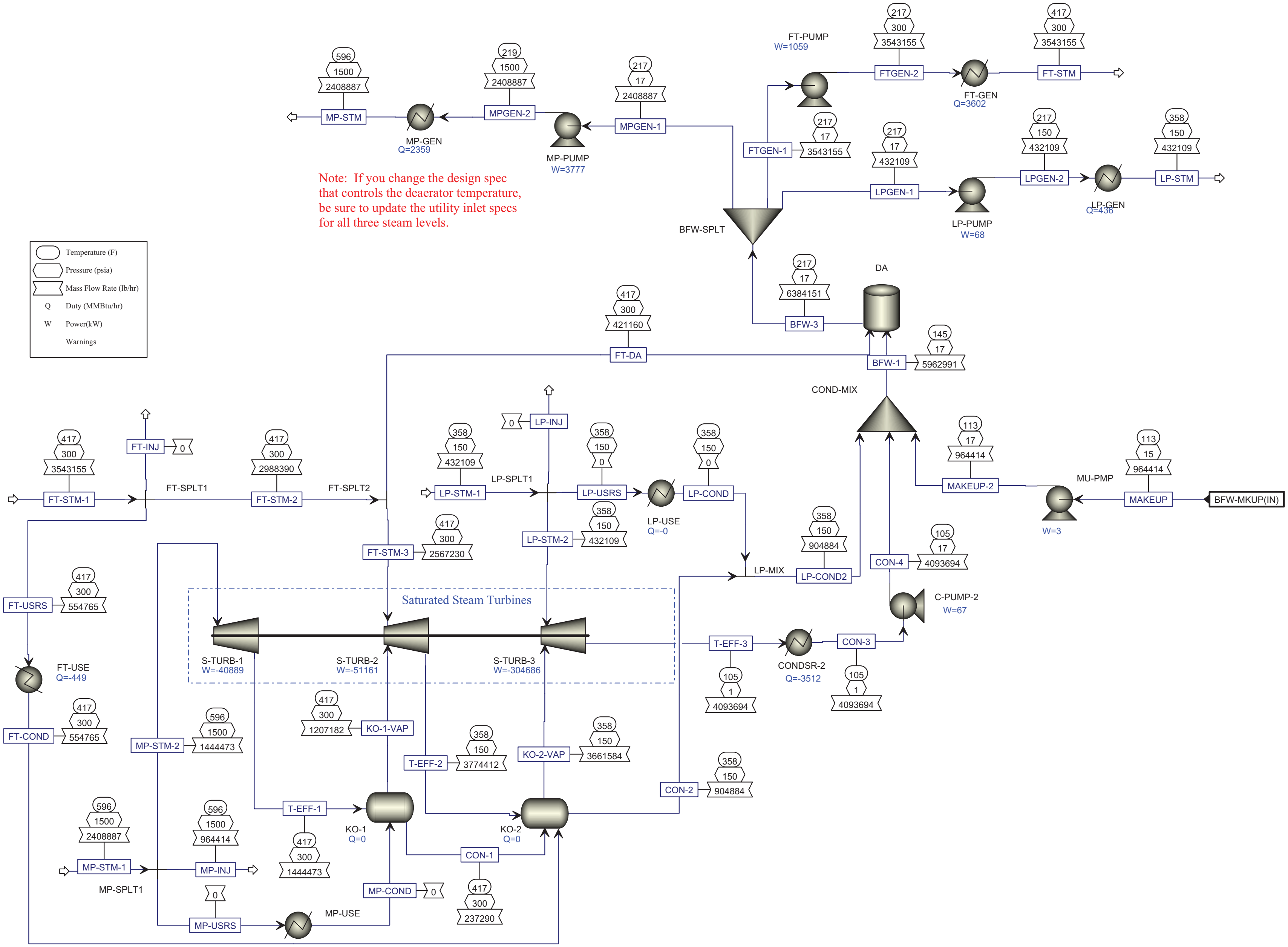
Hydrocracking



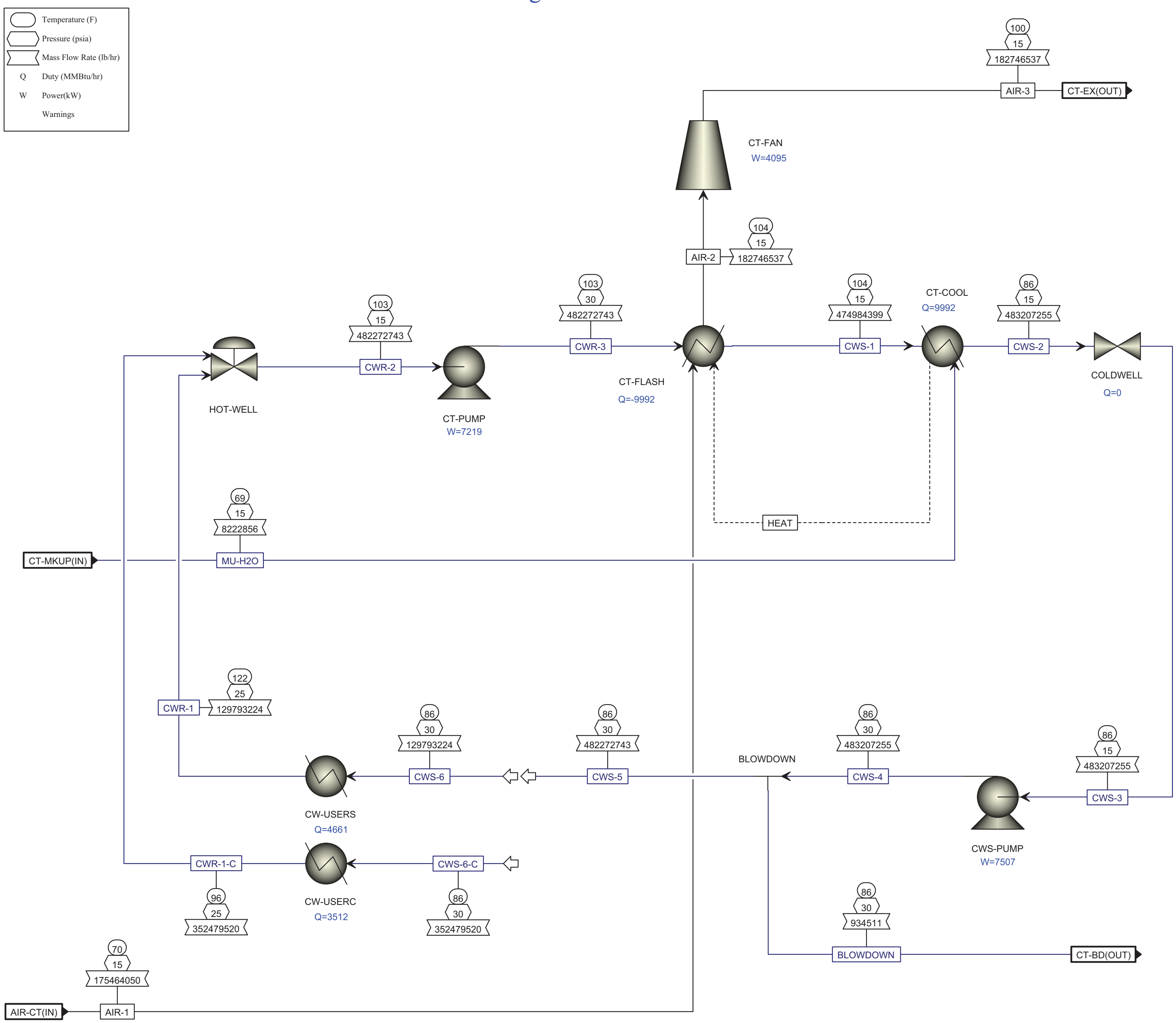
Hydrotreating



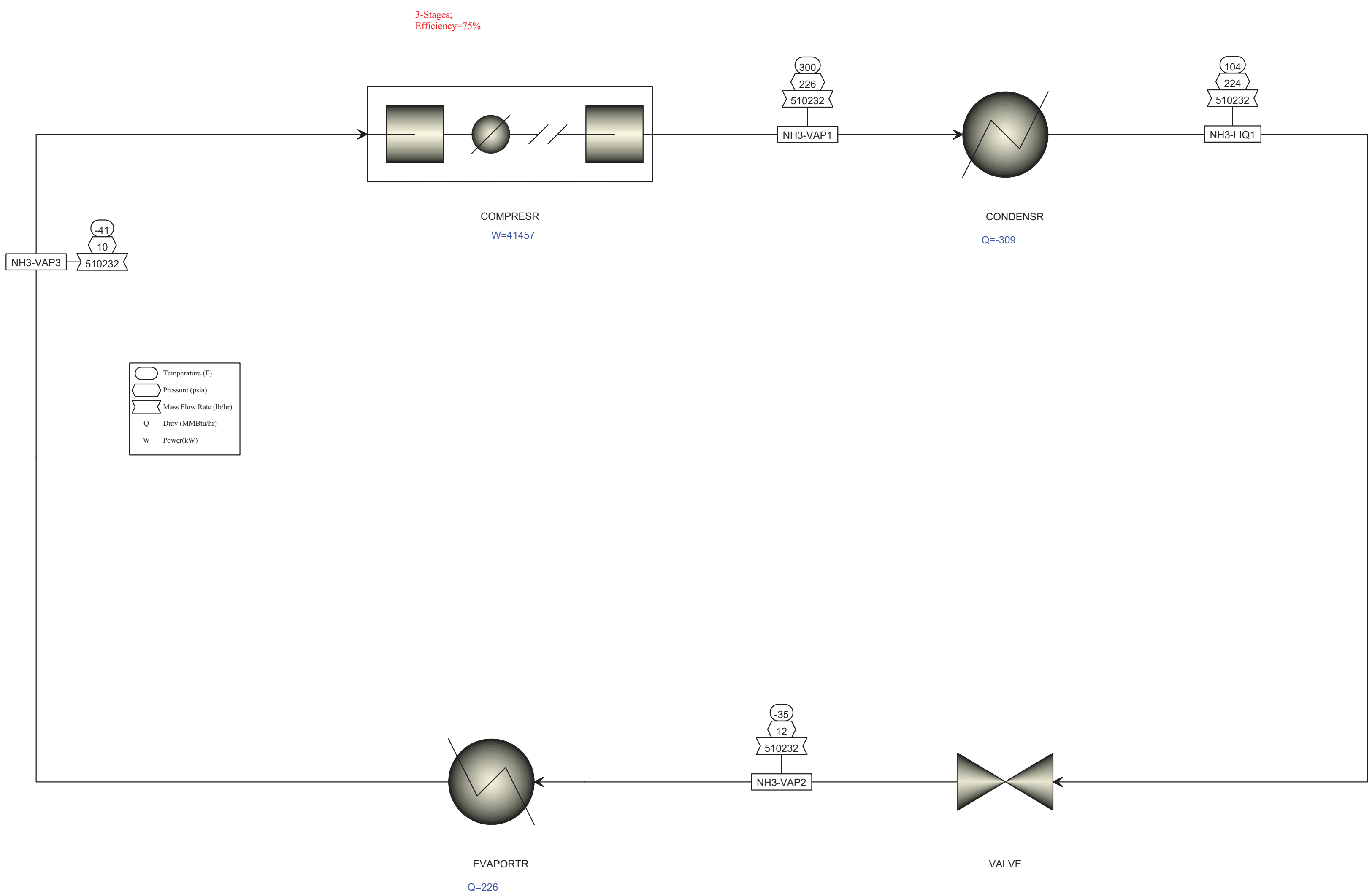
Steam Turbines



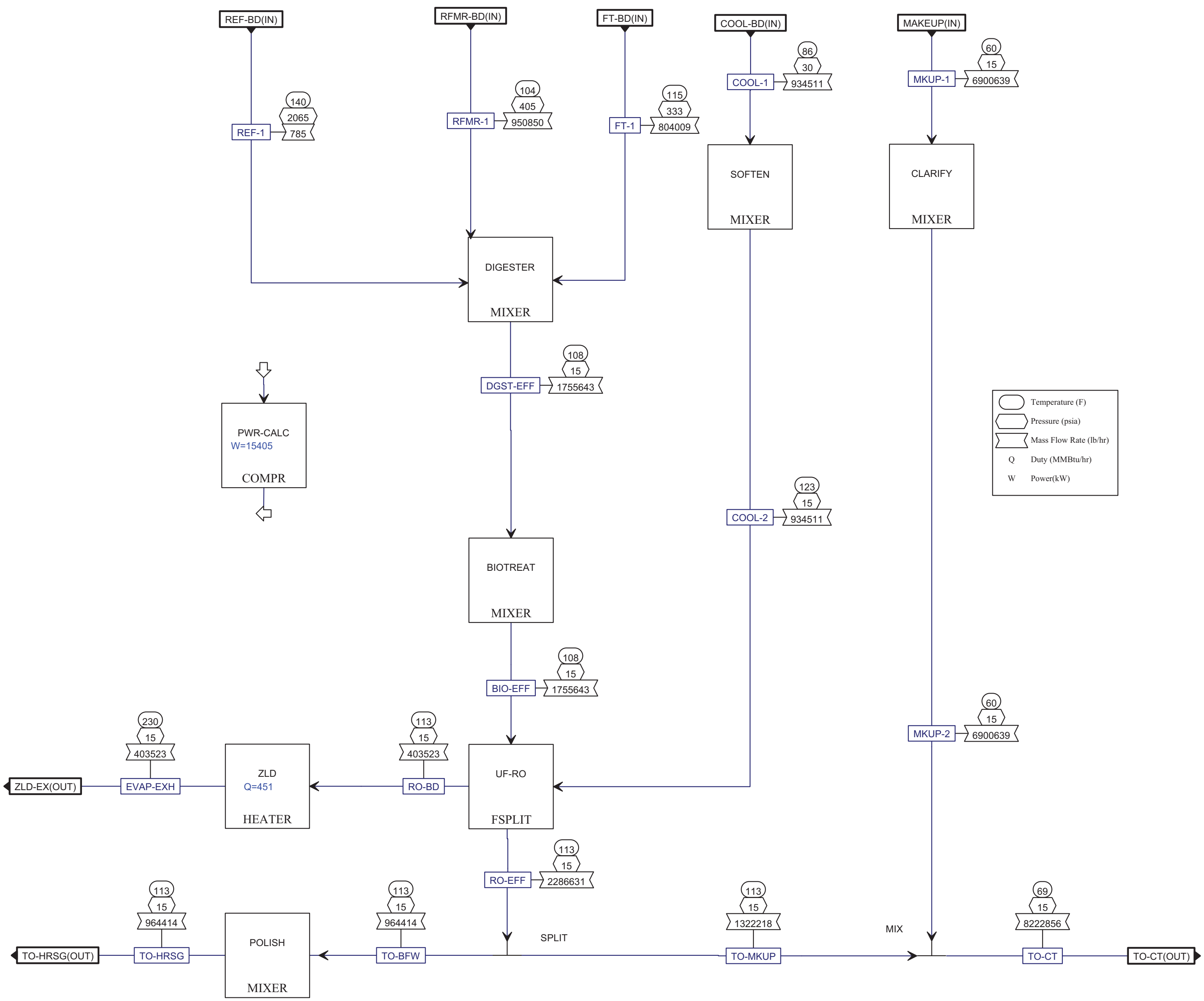
Cooling Tower

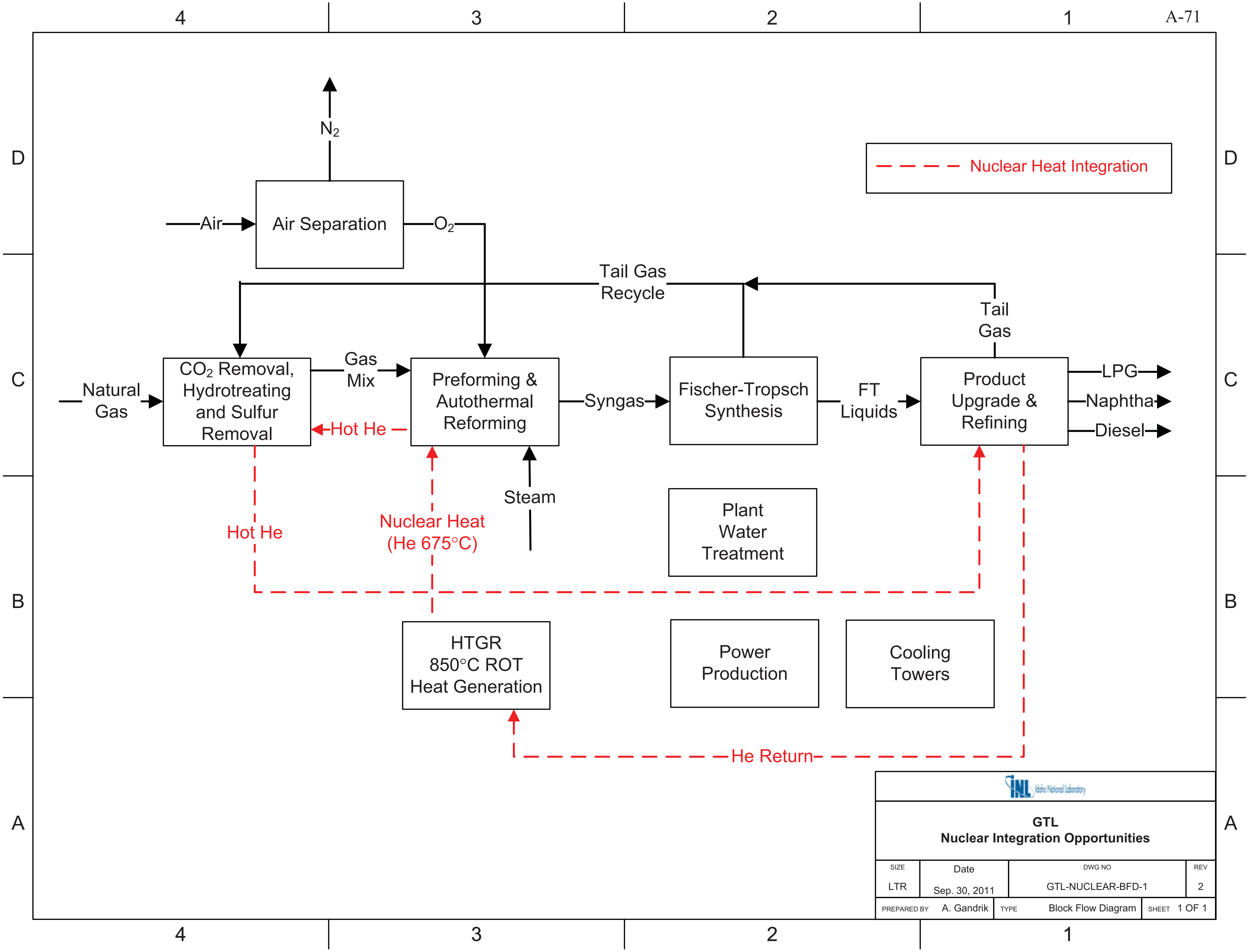


Refrigeration Unit

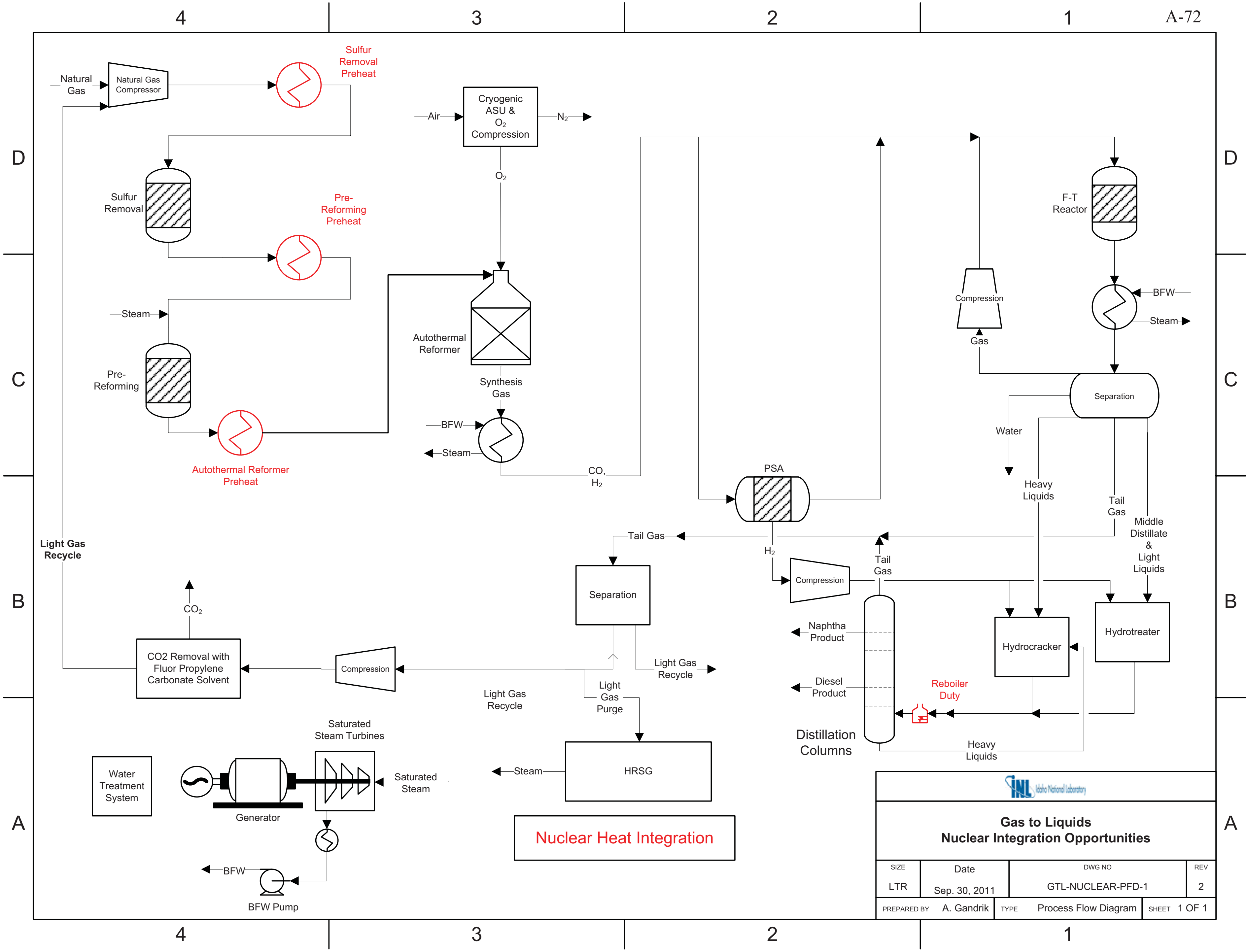


Simplified Water Treatment





Nuclear Heat Integration
N_2
Air
Air Separation
O_2
Tail Gas Recycle
Tail Gas
Natural Gas
CO_2 Removal, Hydrotreating and Sulfur Removal
Gas Mix
Preforming & Autothermal Reforming
Syngas
Fischer-Tropsch Synthesis
FT Liquids
Product Upgrade & Refining
LPG
Naphtha
Diesel
Hot He
Hot He
Nuclear Heat (He 675°C)
Steam
Plant Water Treatment
HTGR 850°C ROT Heat Generation
Power Production
Cooling Towers
He Return
INL Idaho National Laboratory
GTL
Nuclear Integration Opportunities
SIZE
LTR
Date
Sep. 30, 2011
DWG NO
GTL-NUCLEAR-BFD-1
REV
2
PREPARED BY
A. Gandrik
TYPE
Block Flow Diagram
SHEET
1 OF 1



Natural Gas
Natural Gas Compressor
Sulfur Removal Preheat
Cryogenic ASU & O_2 Compression
Air
N_2
O_2
Sulfur Removal
Pre-Reforming Preheat
Steam
Pre-Reforming
Autothermal Reformer
Autothermal Reformer Preheat
Synthesis Gas
BFW
Steam
CO, H_2
F-T Reactor
Compression
Gas
BFW
Steam
Separation
Water
Heavy Liquids
Tail Gas
Middle Distillate & Light Liquids
PSA
Tail Gas
H_2
Compression
Tail Gas
Light Gas Recycle
CO_2
Separation
CO2 Removal with Fluor Propylene Carbonate Solvent
Compression
Light Gas Recycle
Light Gas Recycle
Light Gas Purge
Naphtha Product
Diesel Product
Hydrocracker
Hydrotreater
Reboiler Duty
Distillation Columns
Heavy Liquids
Saturated Steam Turbines
Water Treatment System
Generator
Saturated Steam
Steam
HRSG
Nuclear Heat Integration
BFW
BFW Pump
Gas to Liquids
Nuclear Integration Opportunities
SIZE
LTR
Date
Sep. 30, 2011
DWG NO
GTL-NUCLEAR-PFD-1
REV
2
PREPARED BY
A. Gandrik
TYPE
Process Flow Diagram
SHEET
1 OF 1

CALCULATOR BLOCK NUC-SUM

REACTOR HEAT SUMMARY:

DUTY REQUIRED =	1633.6 MMBTU/HR
DUTY REQUIRED =	478.8 MWT
HELIUM MASS FLOW =	1925350. LB/HR
INLET TEMPERATURE =	1247. DEG. F
OUTLET TEMPERATURE =	563. DEG F.
PRESSURE DROP =	-20.3 PSI
TOTAL HEAT REQ. =	527.1 MWT
SEC. CIRC. HEAT GEN. =	48.4 MWT

Calculator Block SUMMARY

FEED SUMMARY:

NATURAL GAS PROPERTIES:

MASS FLOW =	8862. TON/DY
VOLUME FLOW =	390. MMSCFD @ 60˚F
HHV =	23063. BTU/LB
HHV =	1047. BTU/SCF @ 60˚F
ENERGY FLOW =	408781. 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

PRODUCTS:

LIQUID PRODUCTS PRODUCED =	518713. LB/HR
LIQUID PRODUCTS PRODUCED =	6224.6 TON/DY
DIESEL =	376925. LB/HR
DIESEL =	4523. TON/DY
NAPHTHA =	112135. LB/HR
NAPHTHA =	1346. TON/DY
LPG =	29654. LB/HR
LPG =	356. TON/DY
LIQUID PRODUCTS PRODUCED =	49998. BBL/DY
DIESEL =	35410. BBL/DY
NAPHTHA =	11674. BBL/DY
LPG =	2914. BBL/DY
LIQUIDS PRODUCED / NATURAL GAS FED =	0.70 LB/LB
LIQUIDS PRODUCED / NATURAL GAS FED =	128.09 BBL/MMSCF

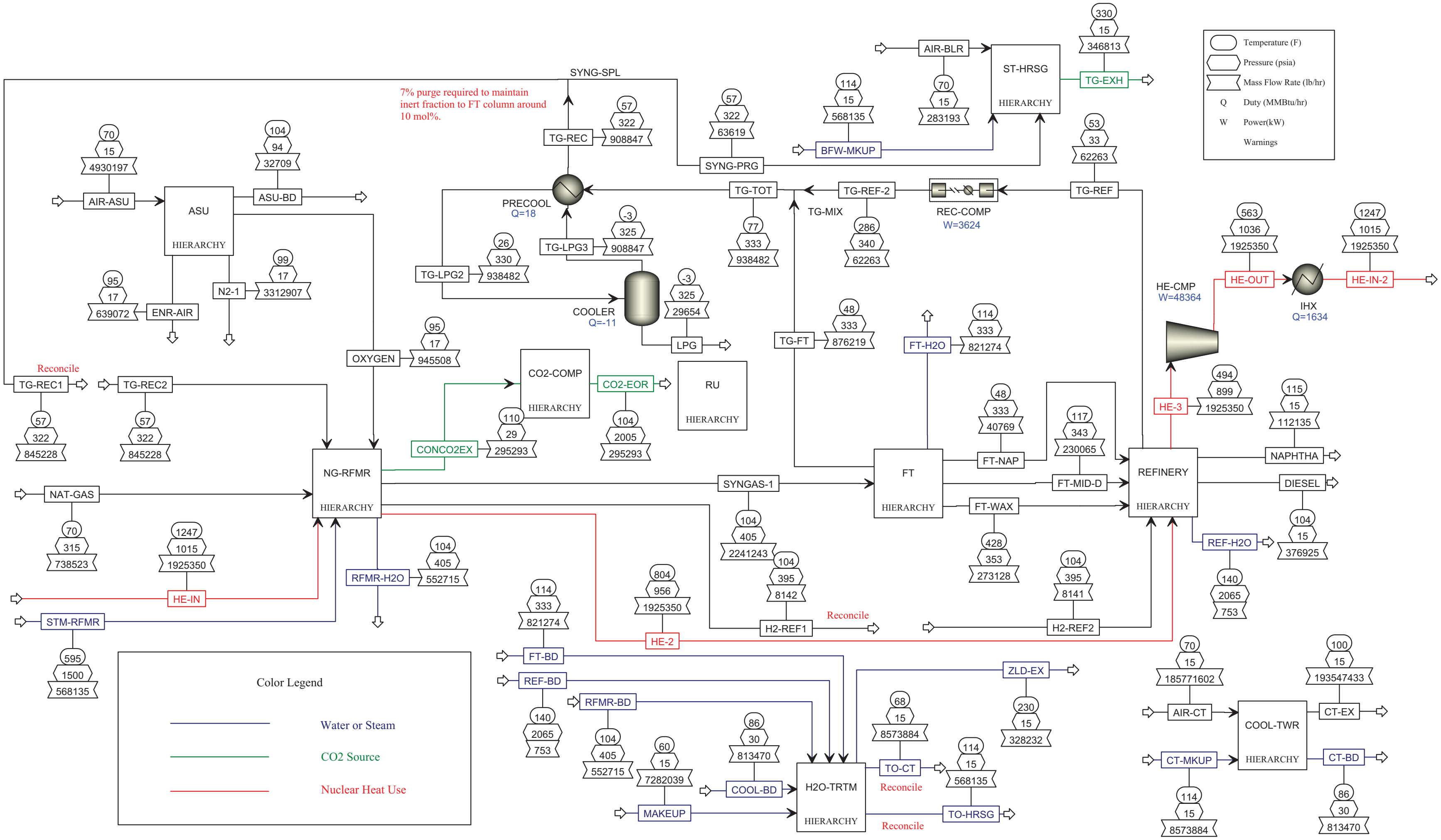
FUEL PROPERTIES:

	DIESEL	NAPHTHA	LPG
PROD. RATE, BBL/DAY	35410.	11674.	2914.
LHV RATE, MMBTU/DAY	171260.	49176.	12092.
MW	190.1	81.1	59.8
API GRAVITY	54.0	86.8	
DENSITY, LB/GAL	6.08	5.49	5.82
CETANE NO.	96.2	40.5	
HHV CONTENT, BTU/LB	20365.	19718.	18355.

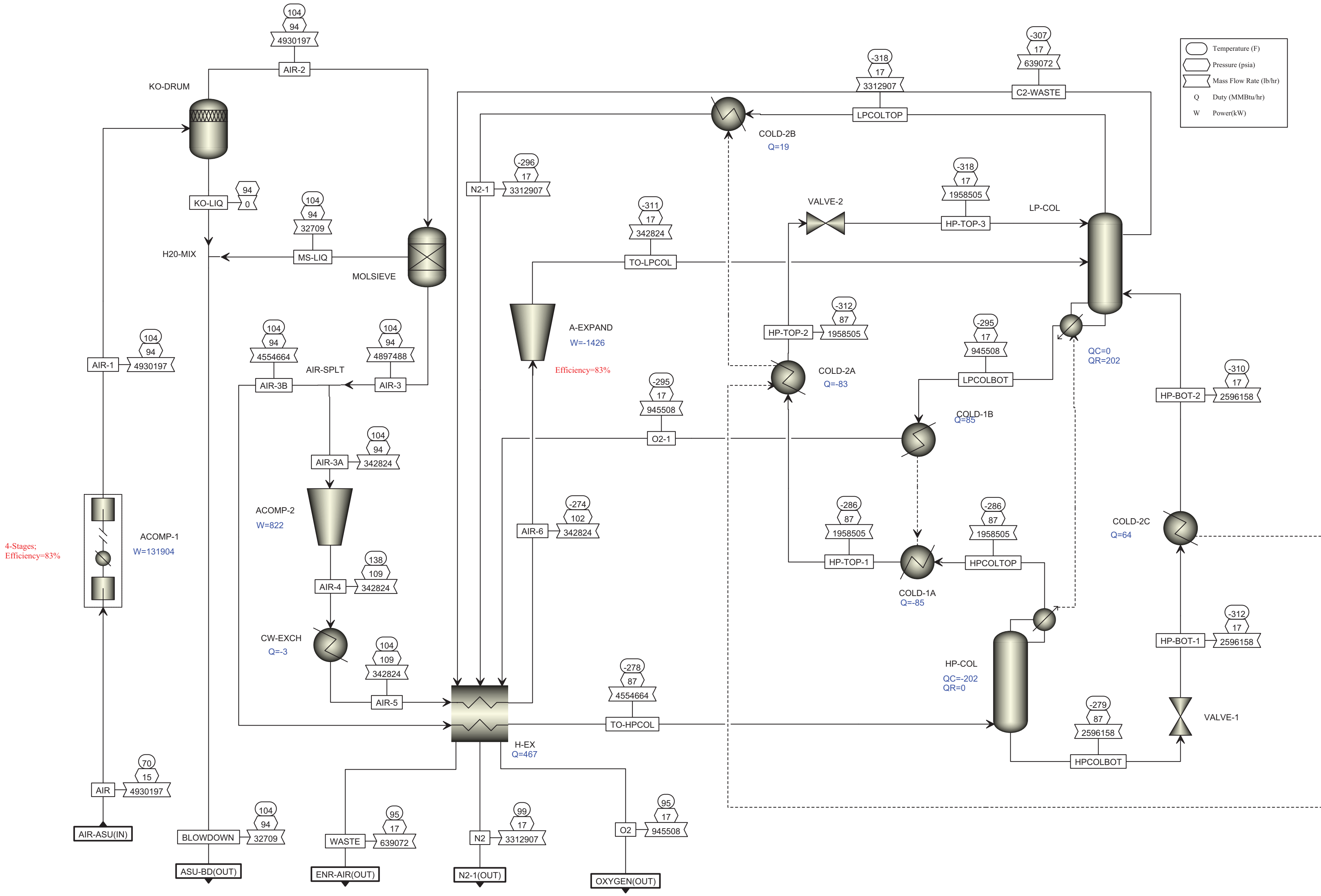
```
    LHV CONTENT, BTU/LB    18932.        18273.        16990.
    % CARBON                  84.7          80.6          75.7
    D86T CURVE, DEG. C:
        0%                  148.         -112.
        10%                 184.           24.
        20%                 203.           55.
        50%                 251.          100.
        90%                 327.          131.
        100%                355.          182.

POWER CALCULATIONS:

  POWER GENERATORS:
    SATURATED TURBINE POWER OUTPUT =     471.9 MW
  TOTAL POWER GENERATED =                471.9 MW

  POWER CONSUMERS:
    ASU POWER CONSUMPTION =              131.3 MW
    NG REFORMER POWER CONSUMPTION =       68.9 MW
    CO2 LIQUEFACTION POWER CONSUMPTION =  11.7 MW
    FISHER TROPSCH POWER CONSUMPTION =    44.0 MW
    REFINERY POWER CONSUMPTION =          11.6 MW
    POWER BLOCK POWER CONSUMPTION =        4.7 MW
    REFRIGERATION POWER CONSUMPTION =     47.1 MW
    COOLING TOWER POWER CONSUMPTION =     20.8 MW
    WATER TREATMENT POWER CONSUMPTION =   13.9 MW
    SEC. CIRCULATOR POWER CONSUMPTION =   48.4 MW
  TOTAL POWER CONSUMED =                 402.3 MW

  NET PLANT POWER (+ GEN, - CONS)=        69.7 MW

WATER BALANCE:

  EVAPORATIVE LOSSES:
    COOLING TOWER EVAPORATION =        17835.3 GPM
    ZLD SYSTEM EVAPORATION =             655.9 GPM
  TOTAL EVAPORATIVE LOSSES =           18491.2 GPM

  WATER CONSUMED:
    BOILER FEED WATER MAKEUP =          1135.4 GPM
    COOLING TOWER MAKEUP =             17134.1 GPM
  TOTAL WATER CONSUMED =               18269.4 GPM

  WATER GENERATED:
    NATURAL GAS REFORMING BLOWDOWN =    1104.5 GPM
    FT PROCESS BLOWDOWN =               1641.2 GPM
    REFINERY PROCESS BLOWDOWN =            1.5 GPM
    COOLING TOWER BLOWDOWN =            1625.6 GPM
  TOTAL WATER GENERATED =               4372.9 GPM

  PLANT WATER SUMMARY:
    NET MAKEUP WATER REQUIRED =        14552.4 GPM
    WATER CONSUMED / NATURAL GAS FED =     9.86 LB/LB
    WATER CONSUMED / LIQUID PRODUCT =     10.0 BBL/BBL

CARBON BALANCE SUMMARY:

  % CARBON TO LIQUID FUEL =               79.3 %
  % CARBON TO TAILGAS =                    3.5 %
  % CARBON TO CO2 REM. =                  16.4 %
  % UNACCOUNTED CARBON =                   0.8 %

  CO2 EMITTED =                         4190. TON/DY
  CO2 EMITTED =                           73. MMSCFD
    FROM FIRED HEATER =                  841. TON/DY
```

```
      LHV TO REFINERY =                    8514. MMBTU/DY
    FROM REFORMER =                        3349. TON/DY
      LHV TO REFORMER =                       0. MMBTU/DY
  CO2 EMMITED / LIQ PROD  =                   0.67 LB/LB
  CO2 EMMITED / NATURAL GAS FED =             0.01 LB/LB

STARTUP FLARE SUMMARY:

  CO2 FROM FLARE =                          123. TON/DY
  LHV TO FLARE =                           1896. MMBTU/DY

EFFICIENCY CALCULATIONS:

    HEAT IN (HHV BASED):
        NATURAL GAS HEAT CONTENT =     17032.5 MMBTU/HR

    HEAT OUT (HHV BASED):
        NET POWER =                      237.7 MMBTU/HR
        LIQUID HEAT CONTENT =          10431.6 MMBTU/HR

    PLANT EFFICIENCY (HHV BASED):
        EFFICIENCY =                      62.6 %

Calculator Block NG-RFMR  Hierarchy: NG-RFMR

SULFUR REMOVAL CONDITIONS:

  INLET BED TEMPERATURE =                  760. °F

PREFORMER CONDITIONS:

  INLET TEMPERATURE =                      915. °F
  STEAM TO CARBON MOLAR RATIO =              0.58

AUTOTHERMAL REFORMER CONDITIONS:

  INLET TEMPERATURE =                     1058. °F
  STEAM TO CARBON MOLAR RATIO =              0.50
  OXYGEN TO CARBON MOLAR RATIO =             0.54
  OUTLET TEMPERATURE =                    1870. °F
  H2/CO PRE PSA =                            2.218
  (H2 - CO2)/(CO + CO2) =                    1.773
  H2/CO POST PSA=                            2.138

  OUTLET COMPOSITION (PRE-CONDENSER):
    H2                                      51.7300 MOL.%
    CO                                      23.3247 MOL.%
    CO2                                      3.7404 MOL.%
    H2O                                     15.2905 MOL.%
    CH4                                      1.4578 MOL.%

  OUTLET COMPOSITION (POST-PSA):
    H2                                      60.0448 MOL.%
    CO                                      28.0842 MOL.%
    CO2                                      4.5034 MOL.%
    H2O                                      0.2740 MOL.%
    CH4                                      1.7552 MOL.%
    INERTS                                   9.8329 MOL.%
```

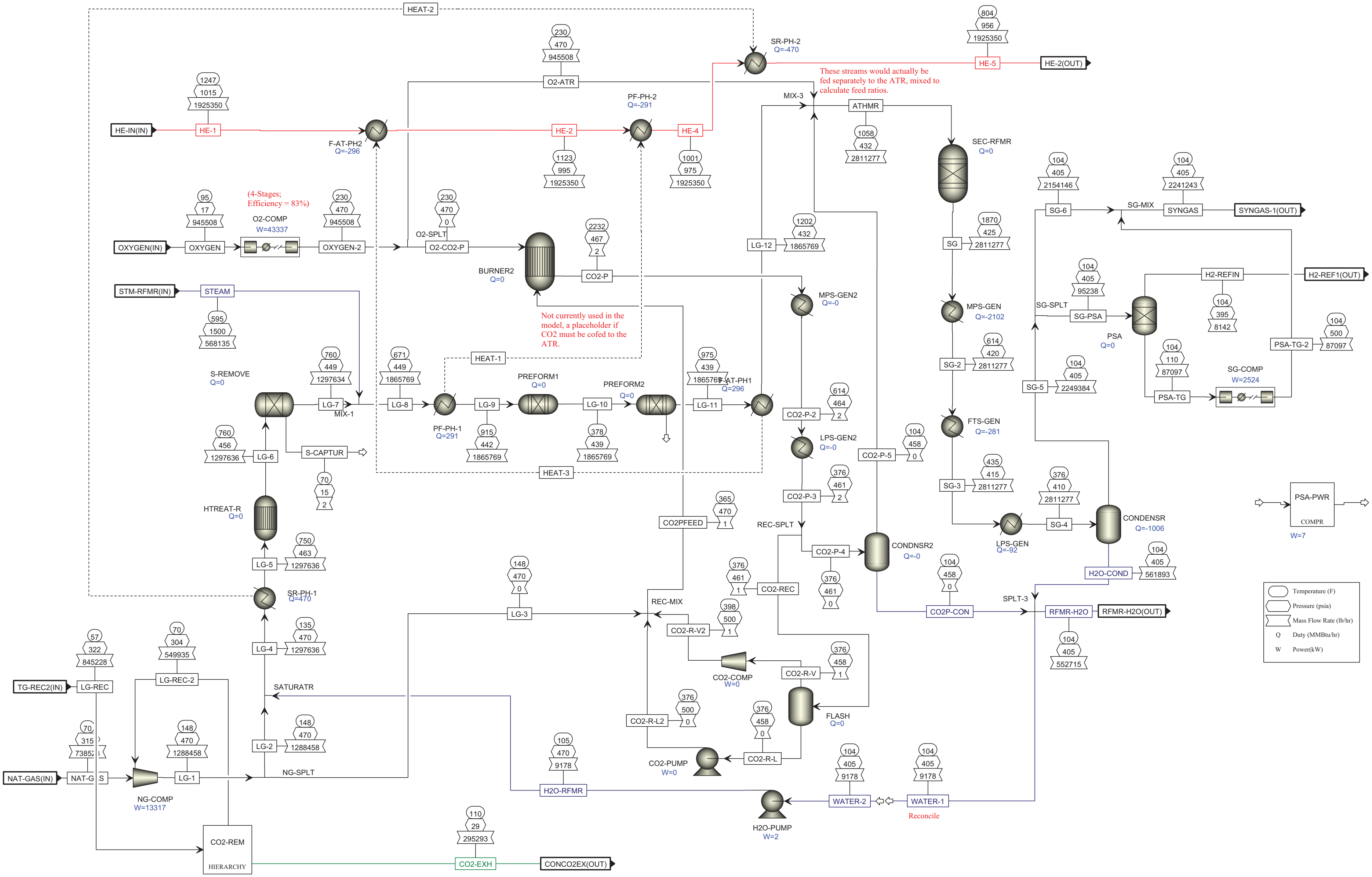
Nulcear-Integrated Natural Gas to Liquid Fuels



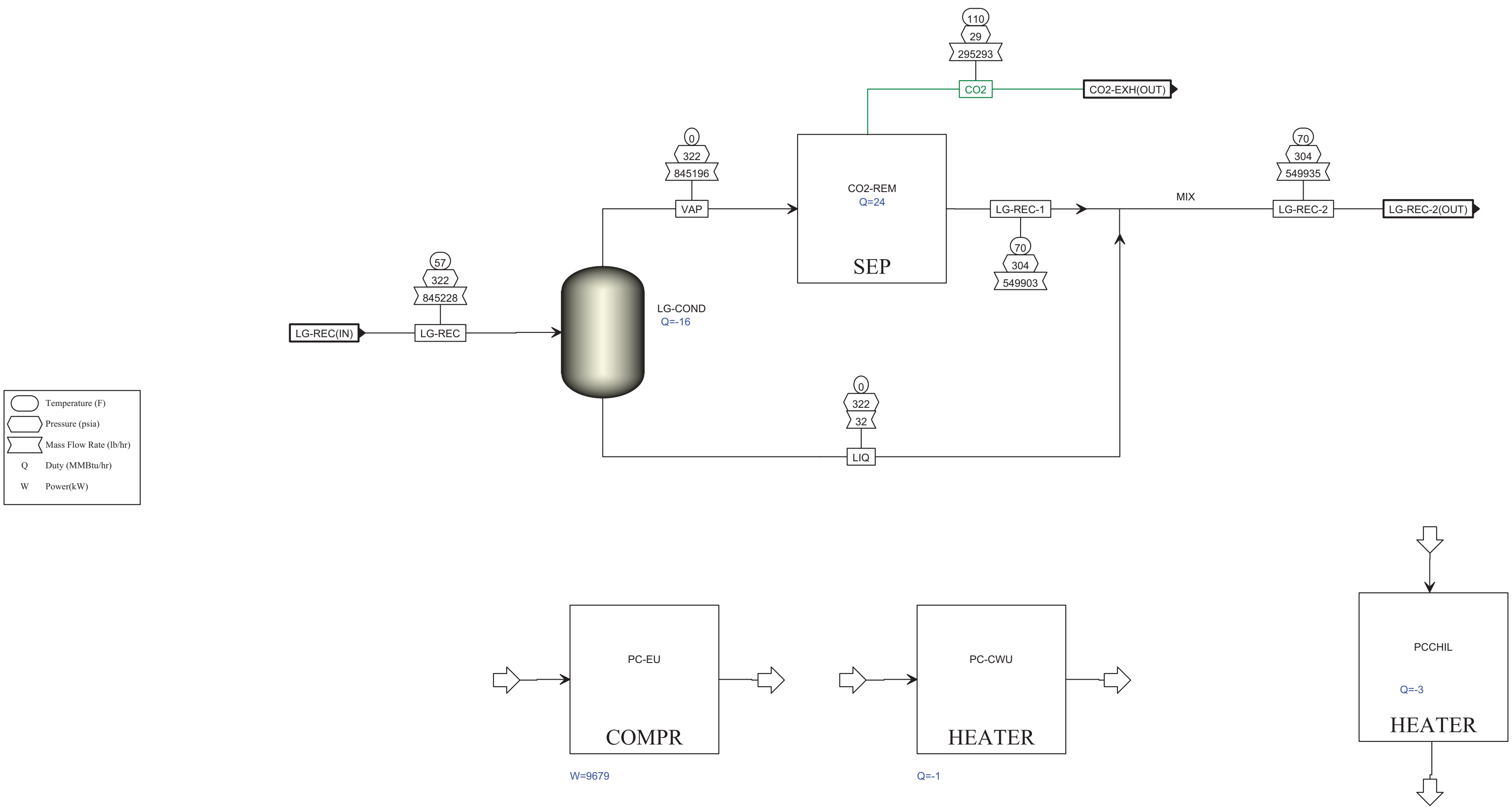
Air Separation Unit



Natural Gas Autothermal Reformer



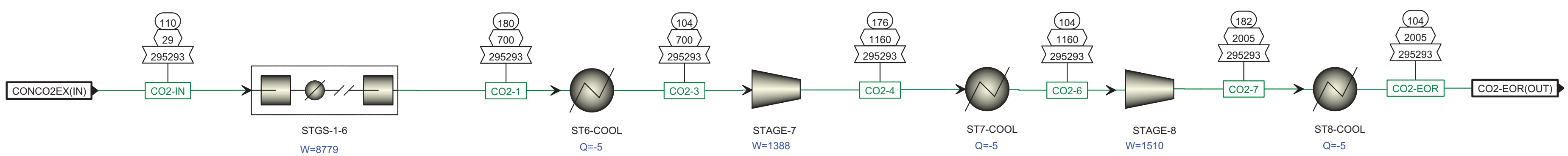
CO2 Removal with Propylene Carbonate (Fluor Solvent)



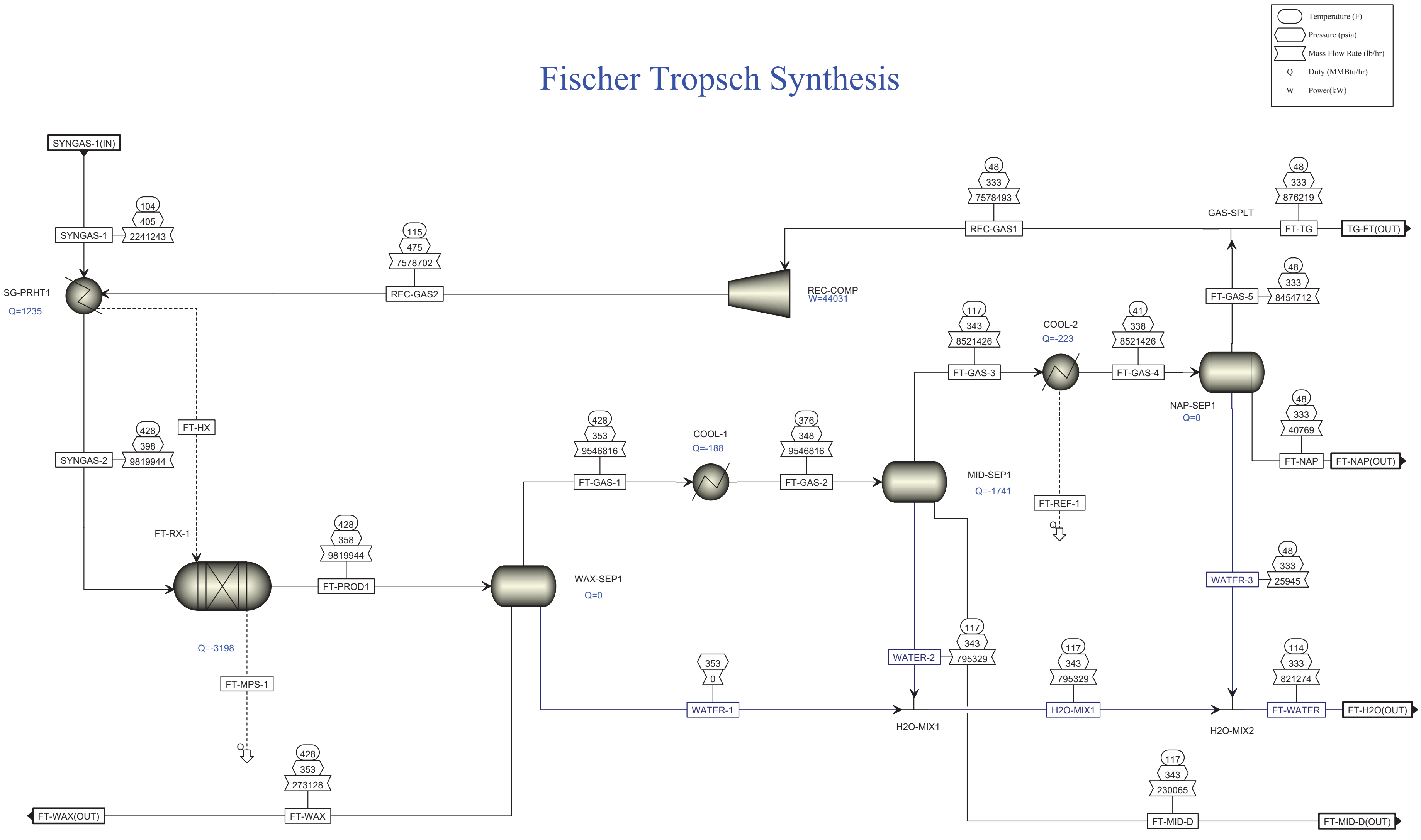
CO2 Compression/ Liquefaction

(oval)	Temperature (F)
(hexagon)	Pressure (psia)
(flag)	Mass Flow Rate (lb/hr)
Q	Duty (MMBtu/hr)
W	Power(kW)

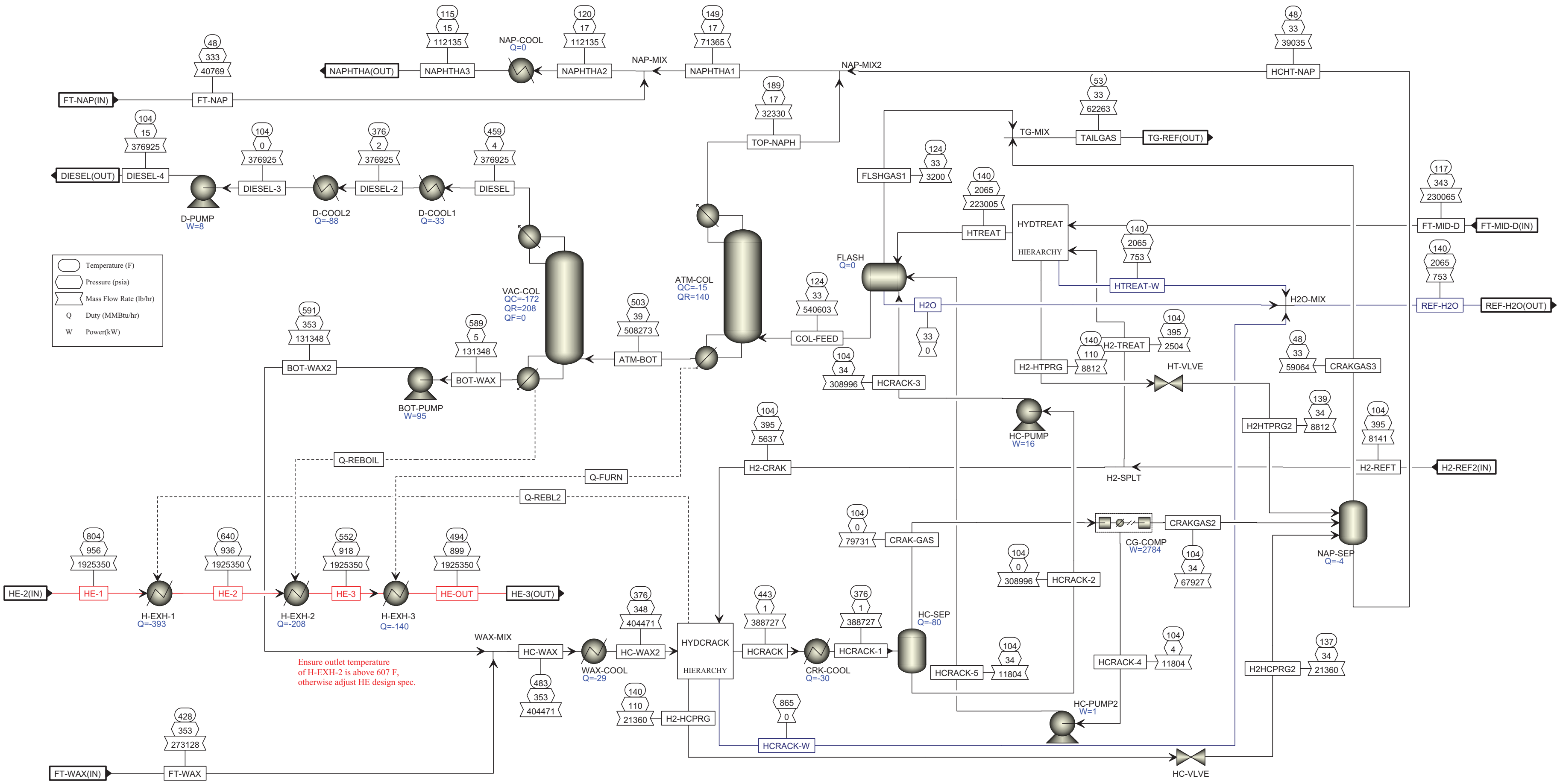
(8-Stages Total; Efficiency = 84.4)



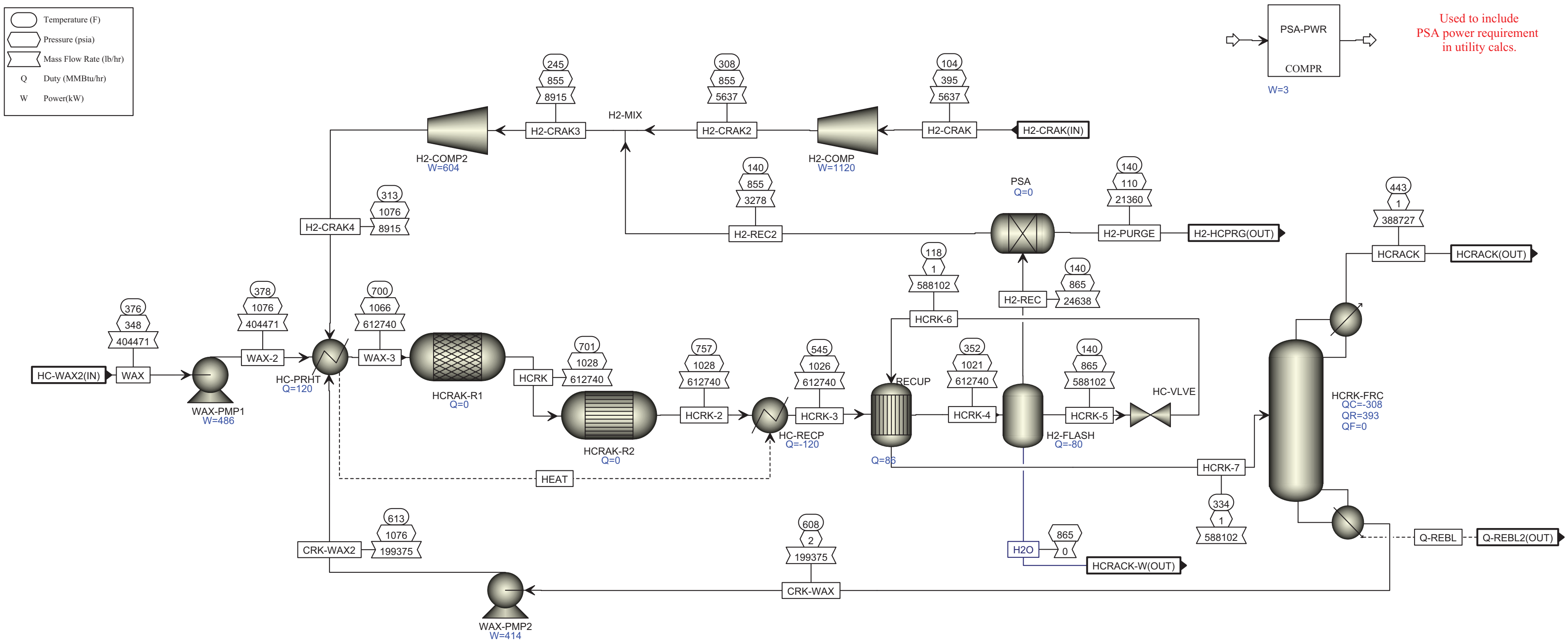
Fischer Tropsch Synthesis



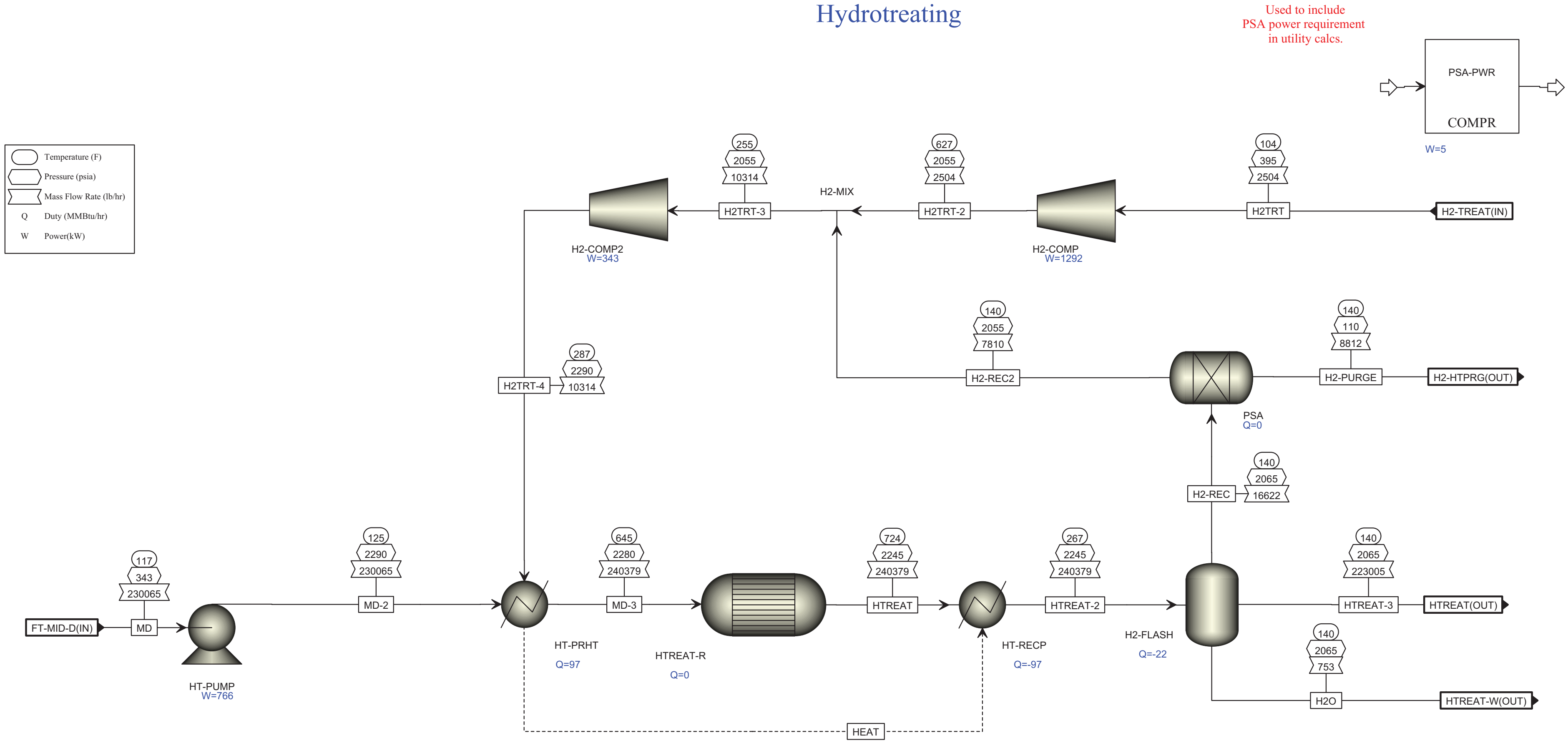
Product Upgrading and Refining



Hydrocracking

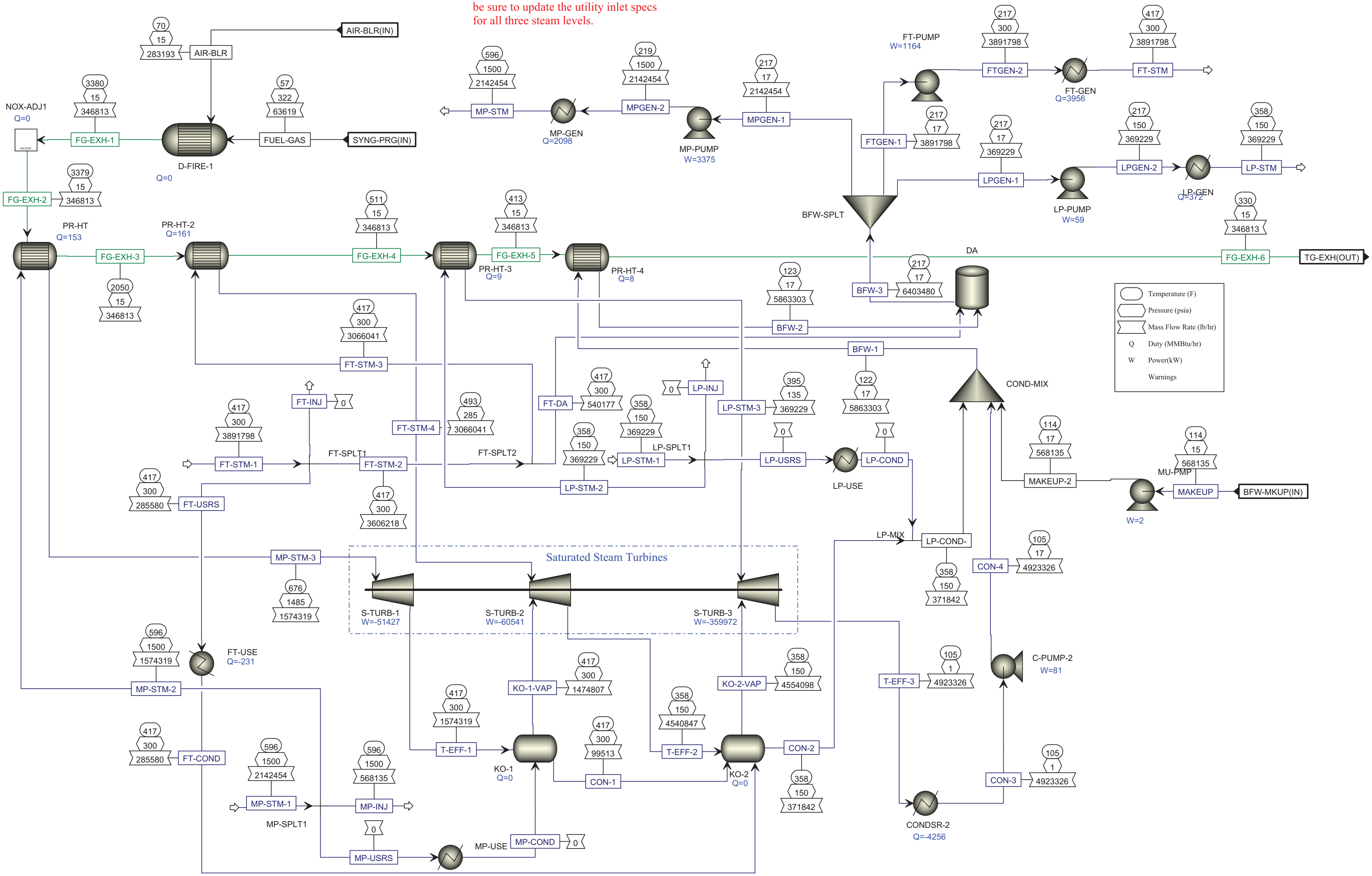


Hydrotreating

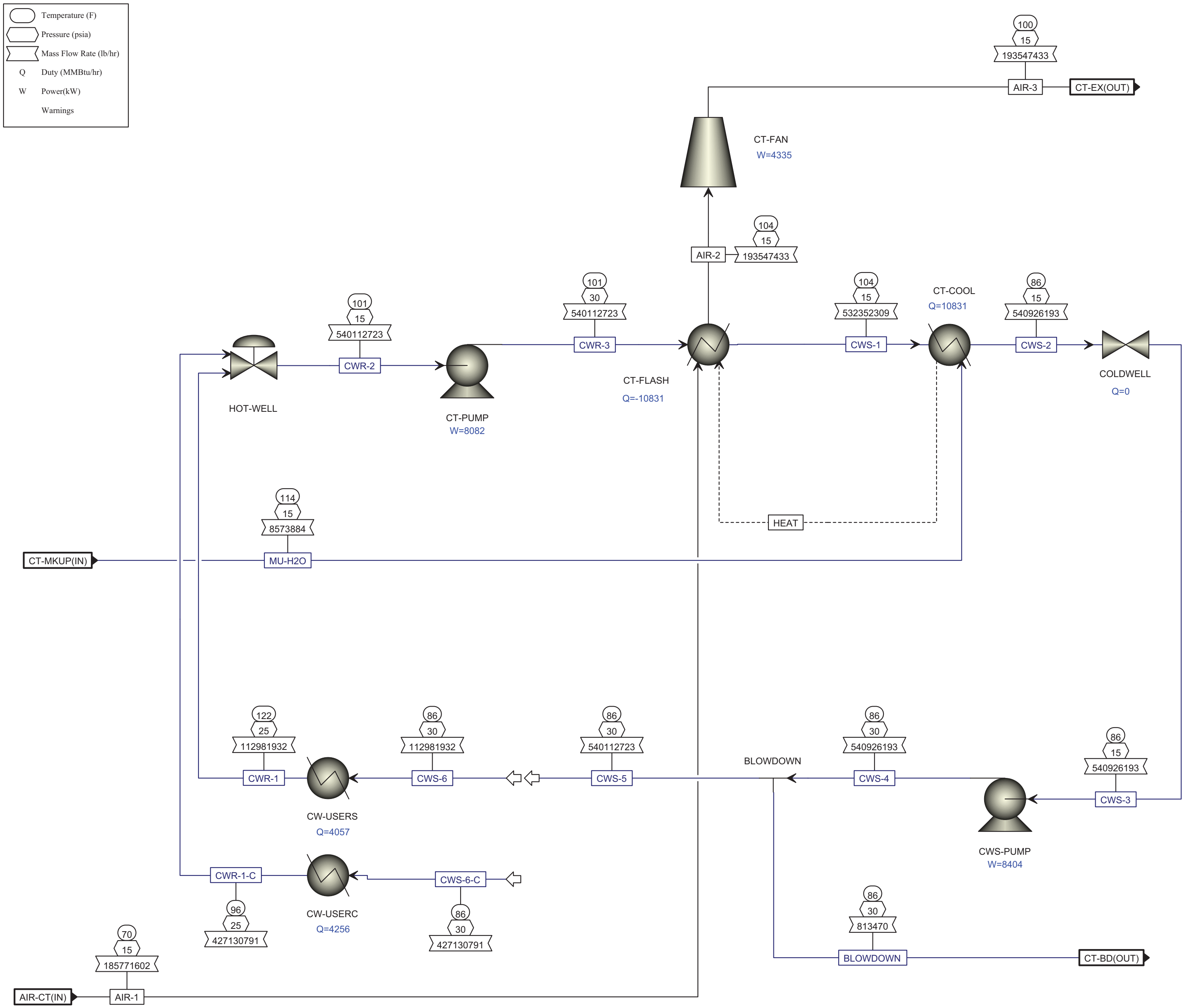


Steam Turbines

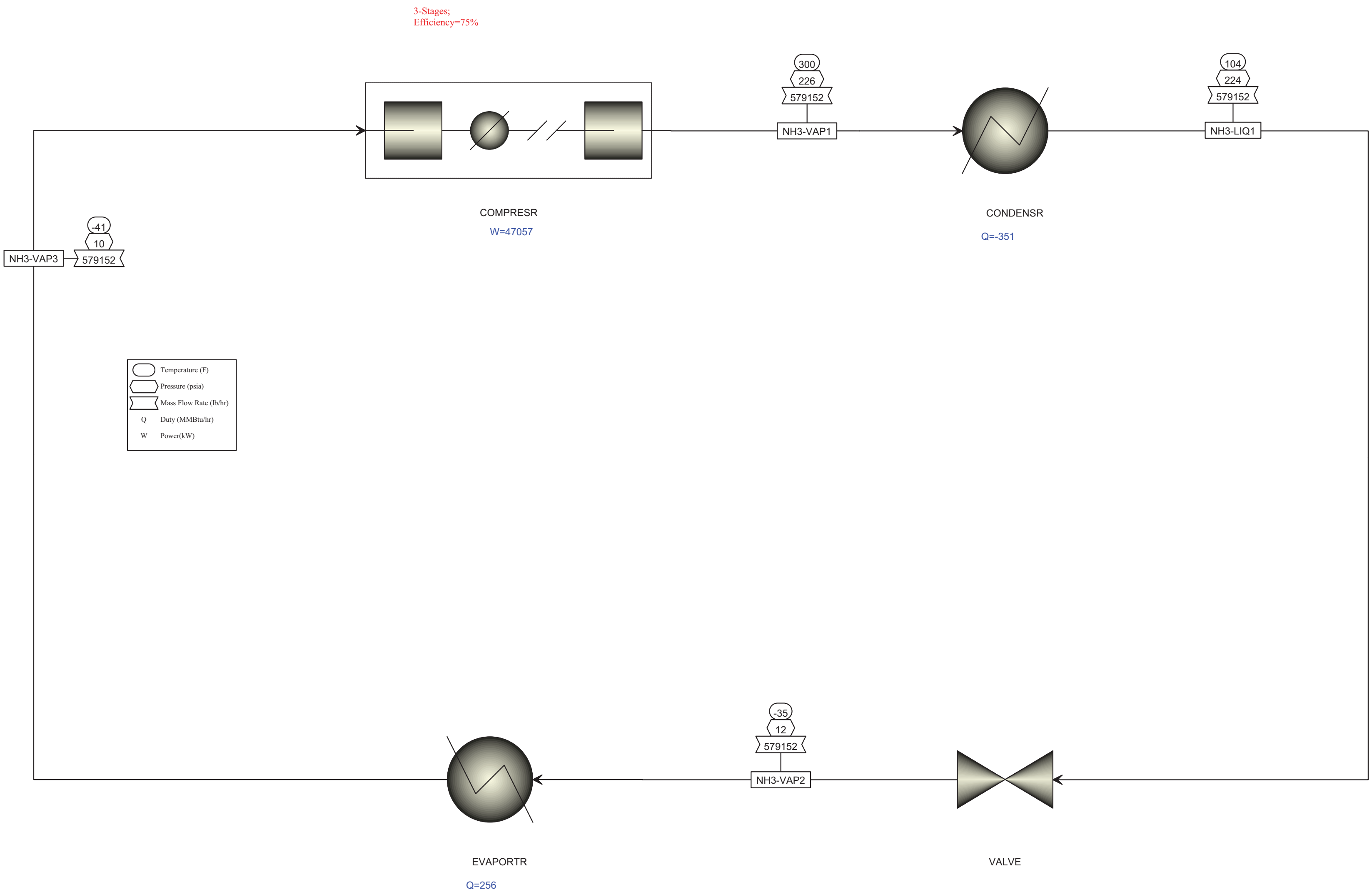
Note: If you change the design spec
that controls the deaerator temperature,
be sure to update the utility inlet specs
for all three steam levels.



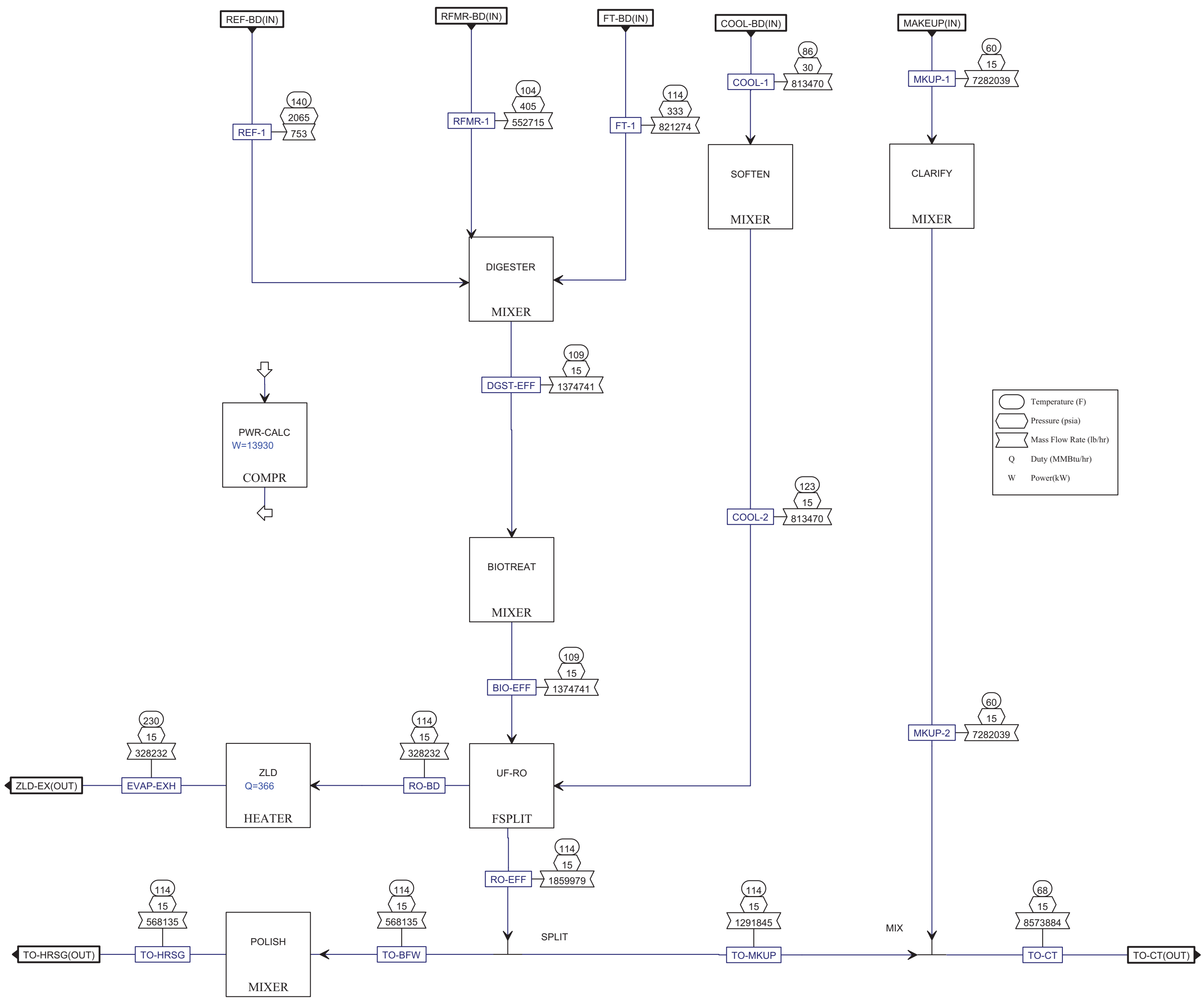
Cooling Tower



Refrigeration Unit



Simplified Water Treatment



Appendix B
CTL Baseline Stream Results.xlsx

[Electronic]

Appendix C
CTL HTGR Stream Results.xlsx

Electronic

Appendix D
GTL Baseline Stream Results.xlsx

[Electronic]

Appendix E
GTL HTGR Stream Results.xlsx

[Electronic]